FUELS AND THEIR COMBUSTION

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BY

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TO

Warren Kendall Lewis

AN INSPIRING AND SYMPATHETIC
TEACHER AND FRIEND

PREFACE

Efficiency in combustion and fuel utilization is daily assuming added importance as the most accessible supplies of our best fuels approach depletion, and as fuel costs advance. Recent years have seen such widespread adoption of new methods of fuel utilization that the science of combustion has not kept abreast of practice. This book—the outgrowth of a lecture course at the Massachusetts Institute of Technology—covers the field broadly and in so far as possible points out the underlying principles on which further progress must be based. It is intended to meet the needs of men using fuels in industry, and of students in technical schools and colleges. It is believed that this volume will serve in two ways: first, by giving a comprehensive treatment of the subject, and second, by means of numerous references guiding into further and more intensive study.

In order to give the reader the proper viewpoint, the first subjects treated are the origin, composition and production of the natural fuels—coal, petroleum and natural gas. In the chapters devoted to combustion equipment, representative types are described as illustrative of modern design. Since the interpretation of combustion data depends on calculations based on the common combustion reactions, this phase of the subject is presented in considerable detail. Secondary or manufactured fuels—producer, water, oil and coal gas—are given special prominence in view of their rapidly growing importance. Since the flow of air and flue gases and heat transfer are inseparably connected with the capacity and efficiency of combustion equipment, the book concludes with three chapters dealing with these subjects. The decision as to the amount of space to be allocated to each subject is a matter of individual judgment, and the authors, in making this decision, have been fortunate in securing the advice of many engaged in the industry.

Because the combustion of fuels is essentially a chemical change, an adequate treatment of the subject requires an understanding of the chemical principles involved. The authors have endeavored to make the volume self-contained in this respect.

Wherever possible, actual plant data have been used in the illustrative examples and the results of calculations are given only to the number of significant figures commensurate with the probable accuracy of the basic data. In a few instances the more rigorous scientific qualifications have been omitted for clearness of presentation.

Throughout the book some intentional repetition has been introduced for the sake of emphasis, or in order to avoid the necessity of turning back to previous chapters. A representative bibliography is given at the end of each chapter.

Data and material have been drawn from many sources Full acknowledgment has been given and any omission is unintentional. In addition to the acknowledgments given throughout the book, special thanks are due to Mr. H. O. Forrest of the Chemical Engineering Department of the Massachusetts Institute of Technology for the vast amount of work done in connection with Chapters X, XVII, Appendices I, II and III. Dr. E. W. Thiele, of the Standard Oil Company of Indiana, has rendered invaluable assistance in connection with Chapter XIX. Thanks are also due the following: Professors J. T. Ward and William P. Ryan of the Massachusetts Institute of Technology for reviewing and proof reading, Mr. R. L. Hershey for assistance in indexing, and Mr. Philip Byrne, of the American Gas & Electric Corporation, Mr. D. W. Wilson of the Iroquois Gas Company of Buffalo, New York, Mr. O. O. Malleis of the Koppers Company, Mr. F. H. Johnson of the Petroleum Heat & Power Company, Mr. M. E. Benesh of the Peoples Gas, Light and Coke Company of Chicago, Mr. Stewart Coleman of The Humble Oil and Refining Company of Houston, Texas, and Mr. L. J. Willien of Charles H. Tenney & Company for reviewing various chapters.

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CAMBRIDGE, MASSACHUSETTS,
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FUELS AND THEIR COMBUSTION

CHAPTER I

THE FUEL SITUATION

Efficiency in the use and production of heat and power is one of the major engineering problems of the present day. The world's energy demands are constantly increasing and to meet these expanding requirements we must look to coal, petroleum, and water power. Without the energy which is developed by the combustion of our fuels and by the utilization of our water power, civilization in its present state would not be possible.

In the early stages of social development the distasteful work of existence[1]* was done by a laboring class composed of slaves. In more recent industrial nations, however, this need has been met, in certain countries, by repression of the laboring class on the part of those in authority, making it practically impossible for them to rise above this class, or, as in the United States, by importing low-grade labor from other nations and by installing labor-saving devices. The labor situation, both in the United States and throughout the world, has altered materially since the outbreak of the World War and industry must now rely on an increased labor efficiency to offset the steadily mounting labor costs. Thus, in the United States, due somewhat to our recent immigration policy, we are gradually being forced to an increase in power consumption and in power efficiency through the economic pressure developed by the high price of labor. This increased effectiveness of human effort can be brought about in only one way—through the more efficient utilization of our natural resources in the production of energy. This energy to run our machines must be developed primarily from the combustion of our fuels.

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* Throughout this book reference figures in the text refer to the bibliographies at the end of each chapter.

Importance of Energy from Mineral Fuels and Water Power.—The major sources of energy are coal, petroleum, natural gas, and water power. It is extremely difficult to estimate the world's total power load, but we have available certain reasonably accurate figures for the United States showing the amount of energy derived from fuels and water power. Table I shows[2] the quantity of energy (from 1819 to 1924) produced annually in the United States from coal, petroleum, natural gas, and water power. The almost negligible quantity of energy supplied by the combustion of wood and by work animals has been omitted.

TABLE I

ANNUAL SUPPLY OF ENERGY FROM MINERAL FUELS AND WATER POWER IN THE UNITED STATES

(Figures represent trillions of B.t.u. Water power is represented by B.t.u. of coal necessary to produce the same amount of power.)

Year	Anthracite	Bituminous coal	Total coal	Domestic oil	Natural gas	Imported oil	Total oil and gas	Total mineral fuels	Water power (*a*)	Grand total, including water power
1819	(*b*)	......	(*b*)	. . .	....	. . .	.....	(*b*)	....	
1829	4	3	7	.....	. . . . .	. . .	. . .	6	No data	
1839	27	41	68	.....	....	.....	.....	68	No data	
1849	109	64	173	.....	....	.....	.....	173	No data	
1859	262	158	419	(*b*)	....	.....	.....	419	No data	
1869	464	415	879	25	....	.....	25	904	No data	
1879	822	993	1,815	119	(*b*)	.....	119	1,934	No data	
1889	1,239	2,507	3,746	211	(*c*) 268	.....	479	4,225	91	4,316
1899	1,643	5,065	6,708	342	(*c*) 240	.....	582	7,291	135	7,426
1909	2,205	9,949	12,155	1,099	517	0.4	1,616	13,771	411	14,182
1913	2,490	12,535	15,025	1,491	626	102	2,219	17,243	588	17,831
1918	2,688	15,180	17,868	2,136	775	226	3,137	21,005	837	21,842
1919	2,396	12,206	14,602	2,270	802	317	3,389	17,991	892	18,883
1920	2,437	14,899	17,336	2,658	858	637	4,153	21,489	971	22,460
1921	2,461	10,897	13,358	2,833	712	752	4,297	17,655	908	18,563
1922	1,487	11,063	12,551	3,345	820	764	4,929	17,480	1,024	18,504
1923	2,539	14,781	17,320	4,394	1,084	492	5,970	23,288	1,136	24,424
1924(*d*)	2,450	12,580	15,030	4,290	1,161	492	5,943	20,973	1,148	22,121

(*a*) No figures for water power are available prior to 1889. The fuel equivalent for water power is calculated from the reported horsepower of installed water wheels, assuming a capacity factor of 20 per cent for manufactures and mines, and of 40 per cent for public utilities, and assuming that the theoretical thermal equivalent of 1 hp.-hr. (2,547 B.t.u.) is 7 per cent of the B.t.u. that would have been consumed in generating 1 hp.-hr. from fuels, in practice. For 1919 to 1924, however, actual reports of the hp.-hours produced by water in electric utility plants have been used.

(*b*) Less than 0.5.

(*c*) Based on the amount of coal displaced by gas as estimated by the gas companies at the time.

(*d*) Preliminary figures.

The values in the table are all on the same basis, namely, the heating value of the fuels in trillions of B.t.u. Water power is represented by the heating value of the equivalent fuel.

The unit of energy used in the table, one trillion B.t.u., is too large to be intelligible, but the magnitudes of the various items in Table I show the relative proportion of the total power produced by the different fuels and by water power. Thus the energy from anthracite and bituminous coal is over 70 per cent of the total. The relative rates of growth[2] of the different sources of energy (with conditions in 1918 as the basis of comparison) are shown in Fig. 1. The figure shows that the consumption of bituminous coal increased steadily up to 1918, but since then has remained practically constant. Thus in 1918 the bituminous coal

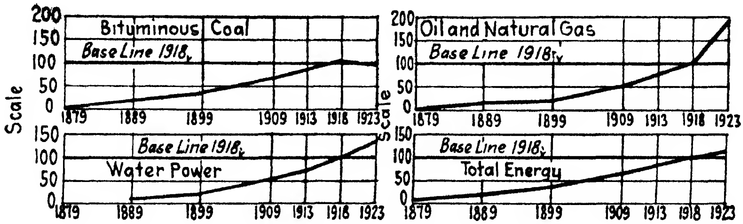

FIG. 1.—Relative growth of production of bituminous coal, oil, natural gas, and water power in the United States.

production of the United States was about 570,000,000 tons, while in 1923 it was slightly less, about 560,000,000 tons. In spite of this lack of recent growth, however, coal still remains the chief industrial fuel.

Contrast the rate of growth of coal consumption with the growth of energy derived from oil and natural gas. The production of petroleum in the United States more than doubled during the period from 1918 to 1923 and the consumption of natural gas also showed a substantial increase. The combined energy from domestic and imported oil and natural gas was 91 per cent greater in 1923 than in 1918. This increase reflects the tremendous growth of the automobile industry, which has expanded until over 300,000,000 hp.[3] is being generated in automobiles in the United States today. The total energy demand of the United States in 1923 was 12 per cent greater than in 1918. The annual fuel bill of the United States, from a conservative estimate, is probably about $5,000,000,000 at the point of fuel production, and about double this amount at the point of fuel consumption. In thinking of these huge sums it must be realized that the *average* efficiency in fuel utilization is probably not over 10 per cent. An increase in efficiency of 1 per cent (from 10 to 11 per cent) would mean a yearly saving of approximately $1,000,000,000.

Interest in fuel conservation, intensified by rising fuel prices, already has resulted in an increase in efficiency in a number of industries. In the production of coke for metallurgical work, beehive coking, which wastes about 35 per cent of the heat in the coal, has been replaced largely with by-product coking, which recovers all but 10 to 15 per cent of the heat. From 1913 to 1923 the amount of coke required, on the average, for the production of 1 ton of pig iron decreased over 10 per cent. Electric utilities burned 3.2 lb. of coal in 1919 to produce 1 kw.-hr., while in 1923 only 2.4 lb. of fuel were required. The railroads, prominent among the users of fuel, have materially reduced coal consumption per ton-mile. (From 0.146 lb. of coal per ton-mile in 1923 to 0.127 lb. in 1925.)

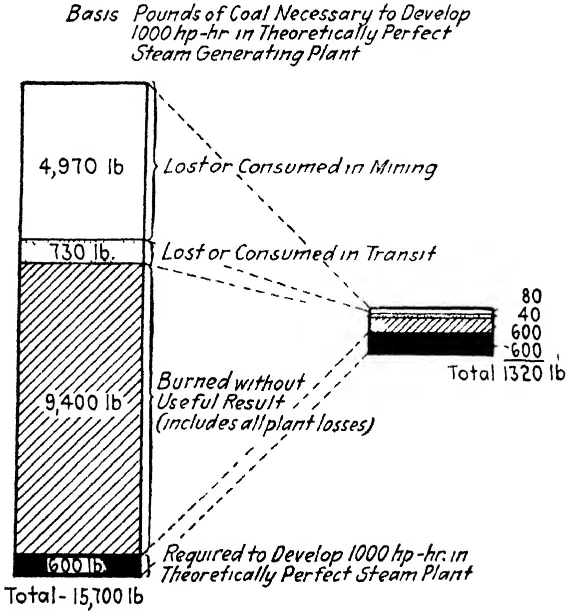

FIG. 2.—Losses in average practice of coal utilization compared with losses in best practice

Efficiency of Fuel Utilization.—These increased efficiencies, important as they are, hardly touch the vast and expanding field of fuel conservation and proper fuel utilization. Power, in universal use in factories, represents one of the four largest items of cost in almost every manufacturing enterprise and in some cases amounts to from 50 to 60 per cent of the total. Notwithstanding the encouragement of these cited economies and in spite of the magnitude of the world's fuel bill, the avoidable wastes in fuel use are appalling. Dr. George O. Smith[4] of the U. S. Geological Survey has used Fig. 2 to demonstrate forcibly the contrast between wasteful and economical use of our coal resources. In the big central stations, which have modern turbine equipment, and in which the effort is always toward the highest efficiency compatible with economy, it is found that for each 600 lb. of coal (theoretically equal to 1000 hp.-hr.) the plant burns another 600 lb. of coal from which it receives no return. In addition to the 600 lb. representing plant losses even under best conditions, 40 lb. additional are consumed in transit and 80 lb. are left in the coal mine or are consumed in mining.

These conditions may be contrasted impressively with average practice; the small steam plant with its poor load factor, high heat loss up the stack, inefficient equipment, and lack of proper supervision and control wastes actually in excess of 9000 lb. of coal for each 600 lb. that it really utilizes. Under average conditions about 30 per cent of the coal in the mine is left underground and this, coupled with various other inefficient practices at the mine, accounts for a loss of almost 5000 lb., a large part of which is preventable. Moreover, almost a third of the freight carried by our railroads is coal, and the losses and consumption of the coal in transit account for another 730 lb. Therefore, as shown by Fig. 2, in *average* practice 8 to 10 tons of coal are used to do the work that might be done with less than 1 ton of coal under the *best* modern conditions.

Fuel Resources.—The important wastes in fuel production and the glaring inefficiencies in fuel utilization which marked the earlier stages of industrial development were either ignored at the time or were but dimly recognized. Now, however, they take on a new and serious significance. Fuels have risen greatly in cost. The better class of our natural fuels, our best and most readily accessible coal, our petroleum and natural gas unquestionably approach depletion. It is of the utmost importance that efforts be made to use more efficiently the invaluable resources which remain in our custody as trustees for future generations.

Concerning our available coal, Smith[2] has said:

> The fact that more than half of the world's coal resources are believed to lie within the territory of the United States has led too many of us into unwarranted optimism. The total tonnage involves strings of figures hard for us to comprehend, but the tonnage remaining in the great producing fields of the East is so limited as to compel us to foresee their exhaustion within periods of the same magnitude as those which you executives figure as the expectancy of life for your industrial enterprises. For example, the Pittsburgh bed in Pennsylvania was estimated, 40 years ago, as good for 30 generations, but the rate of mining has so greatly increased that now we must measure the exhaustion of this largest bed in the Keystone State by the space of a single generation. In the Georges Creek field in Maryland this same bed, there called the "Big Vein," was believed 40 years ago to have a life of at least 150 years, but today the field is almost worked out. Even if you turn to a less nearly exhausted field, such as the Pocahontas, the earlier optimistic calculations of an expectancy of life of four or five generations are now reduced to three or even two.

The situation in respect to our supplies of petroleum and natural gas is even worse than in the case of coal. Three of the major petroleum-producing areas in the United States seem to offer no possibilities of increased production, and the estimated amount of petroleum left underground in the United States, if it could be produced as fast as wanted, would not be sufficient to meet the petroleum demands for more than 50 years. With increasing petroleum consumption, this country must look more and more to imported oil, except as advancing prices may make it an economic possibility for industry to obtain petroleum products by a method yet to be developed (see Chaps. VI and VII). Wasteful methods of use have exhausted a large proportion of the potential energy in the natural gas available at manufacturing centers, and for this high-grade fuel the point of maximum production has already been passed.

It is of the utmost importance that more economical methods of production and use of fuels be introduced to prevent our resources being used contrary to our best national advantage. "True fuel conservation lies not so much in hoarding, as in intelligent and efficient use."

The Development and Use of Power.—The most important source of energy is coal (as shown by Table I) and the development and use of power, until recently, have been intimately connected with the production of coal. Coal was known at least as early as the fourth century[5,6] B.C., and the beginnings of a regular coal trade appeared in the thirteenth century. It was not until about 1730, however, when the Abraham Darbys, father and son, revived the practice of smelting iron with coal instead of charcoal, that the modern use of coal really began.[7] This process had first been brought out by Dud Dudley in 1621, but had fallen into disuse only a few years after its inception. In the century following the Darbys' reintroduction of this process the annual iron production of Great Britain rose to about 750,000 tons.

The invention of James Neilson, about 1830, using preheated air in blast furnaces, lent further impetus to the iron industry and paved the way for the further developments in fuel economy in iron smelting. Just as the necessity of pumping the coal mines furnished one of the chief incentives for the originators of the steam engine, it was the development of steam power during the "Industrial Revolution" that started the coal industry toward its present position of extreme economic importance. The invention of Bessemer (1855) and the origination of the open-hearth furnace by the Siemens brothers in 1861 made possible the production of cheap steel. Cheap steel again increased coal production (since coke is a necessity for metallurgical work) and made possible cheap transportation. With the inventions of James Watt and others, with the increased power utilization following the Industrial Revolution, with the need for steel brought about by railways and other ramifications of industry in its new growth, came the insistent demand for an ever-increasing coal production and power development.

The rate of coal production of the United States and Great Britain over the period from 1890 to 1924 is shown in Fig. 3. In the United States the consumption of coal in the past 40 years has increased at the rate of about 6 per cent per annum. To bring the tremendous figures of coal production within the grasp of the mind, Steinmetz has pointed out that with the mass of coal produced in the United States in one year (nearly two-thirds of a billion tons) it would be possible to build a wall like the Chinese Wall all around the United States, following the Canadian and Mexican frontiers, the Atlantic, Gulf, and Pacific coasts. The chemical *energy* contained in the next year's coal production would be sufficient to lift this entire wall into space, 200 miles high.

Up to about 1870, more than half of the fuel consumed in the United States was anthracite coal. Gradually, however, the use of bituminous coal has increased to such an extent that its consumption has become four to five times the consumption of anthracite. Similarly, other fuels are now encroaching to a certain extent on the fields formerly served by bituminous coal. Figure 4 shows[2] the per cent of the total energy supply of the United States which has been contributed by each of the mineral fuels and by water power over the period from 1819 to 1924. The tremendous advances in the automobile industry since 1900 and the consequent demands for gasoline are reflected in the figure. In 1924, for example, the demand for gasoline, fuel oil, and other petroleum products had increased to such an extent that energy supplied from oil contributed about 23 per cent of

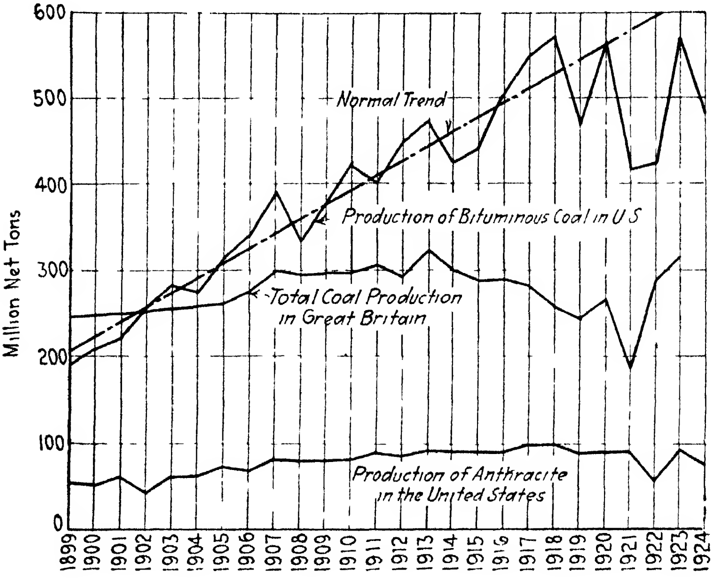

FIG. 3.—Coal production in the United States and Great Britain from 1899 to 1924.

the total requirements of the country. In 1918 oil supplied only about 15 per cent.

Natural gas, with its waning supply, contributes only about 5 per cent of the nation's fuel needs, and water power, although it has shown a substantial increase since 1900, still remains but a minor factor. Over 90 per cent of the power load of the United States is located east of the Rocky Mountains, whereas this area possesses only 30 per cent of the potential water power, only a small fraction of which has been developed. Fuel always will be the most important source of energy, since over half the energy required by man must be in the form of heat.

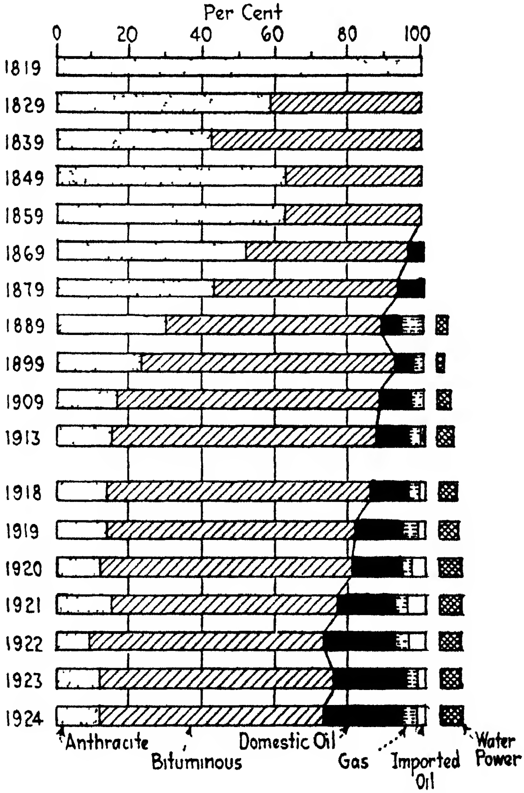

FIG. 4.—Per cent of total energy supplied by mineral fuels and water power. Figures for the United States.

INDUSTRIAL FUELS

In the production of heat and power three general types of fuels are used; solid, liquid, and gaseous. The more important, classified as to their being either natural or manufactured, are shown in Table II.

TABLE II

CLASSIFICATION OF FUELS

TYPE	(*a*) NATURAL, OR PRIMARY	(*b*) MANUFACTURED, OR SECONDARY
Solid..............	Anthracite coal Bituminous coal Lignite Peat Wood	Coke Charcoal Briquets
Liquid...............	Petroleum	Tar Petroleum distillates Petroleum residuums Alcohols Colloidal fuel
Gaseous.............	Natural gas	Illuminating gas Water gas Oil gas Producer gas Blast-furnace gas Acetylene

Solid Fuels.—*Coal.*—Of the solid fuels, coal is by far the most important, being the cheapest and most abundant. This class of fuels includes the anthracite and bituminous coals and the lignites. It is also the raw material for the majority of the *Secondary* Fuels. At present bituminous coal is the premier fuel for industrial purposes, anthracite being important in the United States primarily as a domestic fuel. In the future, with the exhaustion of our bituminous supplies, lignite and peat will become exceedingly important.

Coal contains inert matter in the form of ash and water, varying in amount from 8 to 40 per cent. The heating value ranges from about 16,000 for the better coals to about 11,500 B.t.u. per pound of actual organic substance for the poorer lignites.

Coke.—Coke, the residue left on the carbonization of coal, finds its major outlet in the metallurgical industries. It is also utilized extensively in the production of water gas and as a domestic fuel. It has very little volatile matter, and contains a higher percentage of ash than the coal from which it is made.

Wood.—Wood, on account of its high price and the large amount of moisture and combined water which must be evaporated, has ceased to be a commercially important fuel (except in the production of lime, in the case of waste from saw mills, etc.). The evaporative power is 3.5 to 4.2 lb. of water per pound of unseasoned wood and 4.2 to 5.1 lb. of water per pound of seasoned material. These figures are to be compared with an evaporative power of 7 to 10 lb. of water per pound of coal and 10 to 15 lb. of water per pound of oil.

Other Solid Fuels.—Charcoal (the residue from the distillation of wood) and briquets (fine coal bonded with a plastic material) are at present not particularly important in the United States, from a commercial viewpoint. The briquetting industry however, has grown rapidly in the past few years, and will become increasingly important with the necessity for more complete utilization of our coal supply.

Liquid Fuels.—*Petroleum.*—Crude petroleum is a mixture of hydrocarbons, the heating value ranging from 18,000 to 20,500 B.t.u. per pound, depending on the gravity. The composition of the petroleum varies considerably according to the district from which it comes. Most petroleum is subjected to partial distillation before being used as a fuel in order to save the more valuable light gasoline and the heavy lubricating stocks.

Petroleum Distillates.—The products obtained in the distillation of petroleum which are used for fuel oil contain the middle portions of the crude substance and are practically free from inert material.

Alcohol.—Alcohol, with a heating value of 12,390 B.t.u. per pound when pure, is not important commercially as a fuel at present. It is possible that it may be a future source of fuel supply, particularly as a substitute for gasoline.

Gaseous Fuels.—On account of varying composition, the heating values of the different gaseous fuels vary quite widely.

Natural Gas.—Natural gas, now rapidly becoming depleted, is composed chiefly of hydrocarbons, particularly methane, and some ethane, though some gases contain hydrogen, very small amounts of helium, and a varying per cent of nitrogen. At present, before burning, most natural gas is stripped of all readily condensable hydrocarbons, which are used in the blending of commercial motor fuels. The heating value is about 700 to 1400 B.t.u. per cu. ft.

The other gaseous fuels vary in composition and in heating value from about 550 B.t.u. per cu. ft. for illuminating gas through 125 to 185 B.t.u. per cu. ft. for producer gas to as low as 90 B.t.u. per cu. ft. for blast-furnace gas.

Gas formed by the destructive distillation or carbonization of coal is called either *illuminating gas*, *coal gas*, *manufactured gas*, or *city gas*. Such gas has a heating value of between 550 and 600 B.t.u. per cu ft. and is composed principally of methane, unsaturated hydrocarbons, and hydrogen. *Blue water gas*, consisting primarily of hydrogen and carbon monoxide, is made by the interaction of steam with incandescent carbon, and usually has a heating value of about 300 B.t.u. per cu. ft. Blue water gas enriched with the gaseous products of the thermal decomposition of oils is called *carburetted water gas*, has a heating value of from 400 to 600 B.t.u. per cu. ft., and is often mixed with coal gas. *Oil gas* is made by the thermal decomposition or "cracking" of hydrocarbon oils into lighter, fixed hydrocarbon gases; it ordinarily has a heating value of from 450 to 650 B.t.u. per cu. ft. *Producer gas*, made by the partial combustion of solid fuels with a mixture of air and steam, consists primarily of carbon monoxide, hydrogen, and nitrogen; due to the nitrogen content, the heating value usually is about 140 to 185 B.t.u. per cu. ft.

A more detailed description of the above fuels is given in the subsequent chapters.

BIBLIOGRAPHY

1. Gilbert and Pogue: "America's Power Resources," Century Company (1921).
2. Tryon, McKenney, and Finn: *Coal Age*, **27**, 91 (1925).
3. W. F. Durand in Program for Oil and Gas Power Week (published Feb. 14, 1925).
4. Paper read before the American Iron and Steel Institute, reproduced in *Combustion*, **4**, 22 (1921).
5. E. S. Moore: "Coal," John Wiley & Sons (1922).
6. F. R. Wadleigh: "Coal Manual," *Nat. Coal Mining News* (1921).
7. W. A. Bone: "Coal and Its Scientific Uses," Longmans, Green & Co. (1922).

CHAPTER II

THE ORIGIN AND COMPOSITION OF COAL

It is generally recognized that coal has been formed from vegetal matter and that the transformation took place in two stages: first, the formation of peat, and, second, the devolatilization of the peat. In the first stage, during the main coal-forming periods, the rate of plant growth was probably much more rapid than at the present time. This was due to a temperate, even climate, abundant and regular rainfall, high humidity which prevented excessive evaporation of water, and, possibly, to a high content of carbon dioxide in the air to stimulate plant growth. The formation of peat took place in two steps: (*a*) the accumulation of the plant material, and (*b*) a biochemical decay, which was brought about by the action of bacteria (micro-organisms). The biochemical decay was brought to an end either by the development of toxic substances which killed the bacteria or by the burial of the peat beneath sediments. The second stage in coal formation, which brought about the devolatilization and condensation of the peat, is termed dynamo-chemical, since a chemical action (the splitting off of water, carbon dioxide, etc.) was caused by a pressure exerted as the result of a movement of the earth's crust. A more extensive treatment of these theories follows:

Validity of Vegetal Origin of Coal.—Although formerly coal was supposed to have been formed by the inclusion of igneous seams of bituminous substances between layers of other material, an examination of the evidence at once shows that this view is untenable. As pointed out by Moore:[1]

> So definite is the evidence, however, that all coal has resulted from the alteration of vegetal matter in some form, that a theory of origin based on any other premise may be dismissed without consideration. Anyone questioning this conclusion has but to observe the transition from peat to lignite and from lignite to bituminous coal, with a gradual decrease in the distinctness of the plant remains in passing from the lower to the higher grades of coal, to be convinced regarding this matter.

He goes on to point out that coals which to the naked eye show no woody structure, when examined more closely under the microscope show unmistakably the remains of spores and wood tissue. Even anthracite seams have shown full trees gone to coal, still retaining, however, the markings of the original wood and bark.

STAGES OF COAL FORMATION FROM VEGETAL GROWTH

Accumulation of Vegetal Matter.--There are two main theories concerning the method by which the vegetal matter for the formation of coal was accumulated. The first considers that the deposits accumulated through the vegetal matter of the swamps falling where it grew and is called the *in situ* theory; while the second, the *transport* theory, considers that the deposits were formed from vegetal matter carried down streams and deposited in huge deltas at the river mouth. The evidence in favor of each may be summed up somewhat as follows:*

The following evidence favors the *in situ* theory as to the accumulation of the woody matter: (1) There are at the present time a number of swamps in which vegetal matter is accumulating on a scale sufficient to produce coal seams of considerable extent. (2) The coal should not be as free as it is from ash if the vegetal matter had been transported downstream along with other sediments. Periods of high water and floods, most favorable for transportation, also give rise to large amounts of sediment. (3) Trunks of trees are found in coals with roots piercing the underlying clays. (4) The large coal deposits were formed in land that was near the sea level, and a slight sinking of the surface would have allowed small amounts of sediment to be washed in. (5) Soils on which trees grew during the coal-forming periods are often found beneath the coal seams. (6) It would have been impossible for the vegetal matter to collect in the open sea, and in deltas where large amounts of mud are usually deposited. (7) The layers of cannel coal in bituminous coal indicate patches of open water in swamps where spores could collect in quantities.

Arguments in favor of the *transport* theory of accumulation are as follows: (1) Timber in huge quantities drifts down streams in virgin-forest regions. (2) Peat and lignite have been found in some modern deltas. (3) Marine fossils often are found in

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* For a more complete exposition of the theories regarding the formation of coal beds, the reader is referred to Moore,[1] p. 124 *et seq.*, and Stevenson.[2]

the strata above and below the coal. (4) Fish remains have been found in coal, especially in the English cannels. (5) It is difficult to determine whether a tree stump is located where it grew or was transported by water and buried by sand which washed in around it. (6) Trees have been found with tops headed downward. (7) It would appear difficult for large trees to take root in the enormous depths of vegetal matter that were necessary to produce some of our large coal seams.

Most of our larger coal deposits (such as the Pittsburgh seam) were undoubtedly formed by the *in situ* method. There are other coal beds, however (mainly occurring as local deposits, such as the Commentry coals in France), the vegetal matter for which may have been transported by water.

Formation of Peat.*—It is generally recognized that coal of all ages and varieties has been formed from peat, and, as pointed out by Davis,[3] a study of the peat deposits of the present day shows the processes through which the vegetal matter must pass to become coal.

A large amount of peat is forming at the present time, though these deposits, in spite of their magnitude, are small in comparison with the peat beds that gave rise to our large coal seams. The conditions most favorable to peat formation are regular and abundant rainfall and constantly high humidity of the air. The abundant rainfall supplies the water required to preserve the plant remains, and the high humidity tends to lessen evaporation. The climate should be temperate, since the saturation of cold air is much easier than hot air. The type of land formation is also exceedingly important, since there must be means of allowing water to collect. Usually the land had a very gentle slope, with consequent poor drainage, or possessed numerous depressions in which water could collect to offset periods of slight rainfall and high evaporation. The climatic requirements are usually best met in the cold or temperate regions, though peat deposits are also encountered to a lesser extent in the tropics.

In the peat bogs growth begins usually in a pond or lake, the process developing with the growth and decay of plants near the edge, the type of plant varying with the distance from the shore; that is, one type of plant may be able to grow partly or wholly submerged, while another may be able to grow only in

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* An excellent discussion of the mechanism and chemistry of peat formation is given by Thiessen.[4]

the shallower portions of the pond. In this way a series of zones of plant life are formed, varying in size and type with the depth of water. With the growth and partial decay of the plants, a deposit of peat forms along the shore and on the bottom of the pond, this material gradually increasing in amount until in time the pond may be almost filled with débris. With the decreasing depth of water there is a gradual change in the type of vegetal life, the pond weeds, bulrushes, and floating algae being displaced by moss, which is, in turn, followed by small trees. Periods of drought or flood may interrupt this progression, either hastening it or causing the type of plant matter to revert to its original form, thus giving rise to bands or zones of different appearance and structure in the resultant coal.

In addition to the peat bogs, there are a number of water-soaked plains, never completely under water, where peat is forming at the present time. In these deposits, where the water just reaches the surface, the type of vegetation is composed chiefly of the turf-forming plants. The "tule" marshes of the Pacific Coast of the United States, the Arctic tundra, the canebrakes, and cat-tail flag marshes are examples of this type.

The main examples of present-day peat accumulation are the Dismal Swamp of North Carolina and Virginia, the Okefinokee Swamp in Georgia, and large swamps in Java and Sumatra. The average water level in these swamps is sufficiently low so that the woody plant types (shrubs and trees) make up the major part of the plant life, and for this reason the peat is composed chiefly of the remains of this kind of vegetation. In the Dismal Swamp, for example, White[5] states that it is a little difficult to find places where a steel rod can be rammed down to a depth of 6 ft., the logs and stumps in the peat being so thick. It has been estimated that the available peat in this swamp is about 672,000,000 tons. There is also distinct evidence that sinking of the land level is slowly taking place at the present time, thus allowing the water level to increase with the increasing amount of peaty deposit.

Biochemical Decay.—During the formation of peat, the woody matter after falling becomes more or less disintegrated and the structure of the material, both physical and chemical, is greatly altered. This process is biochemical, the major part of the disintegration being brought about by the action of micro-organisms and fungi. Obviously, the extent of this action is dependent on the conditions prevailing, such as the amount and circulation of water, the temperature, the oxygen supply necessary for bacterial growth, and the development of toxins which affect the life of the bacteria. The biochemical process on the one hand may cease with only partial decomposition of the woody texture, leaving the peat in such condition that lignitic or woody coals are later formed, or, on the other hand, it may progress to more complete obliteration of the plant material, producing the peat in the proper condition for the formation of the "amorphous" coals. The biochemical process yields as a product an unconsolidated, hydrated, possibly jellified peat, varying in appearance and composition according to the extent and severity of the bacterial action.

The Rate of Peat Formation.—From the preceding discussion it is evident that the rate of peat formation in present-day peat bogs is dependent on the rate of growth of the vegetable matter and on the amount and type of decomposition which it undergoes. In stagnant water the plant substances may be fairly well preserved from decay, causing a relatively rapid deposition of peat, while if the vegetal matter can be attacked readily, the rate of accumulation will be very slow and only the most resistant portions will be left. Also, in practically every case, the rate varies at different stages even in the same bed, due to periods of drought or flood. These same conditions were in force at the time of the decomposition of the peat which gave rise to our coal seams. During these periods, however, the climate was relatively warm, there was more abundant and more regular rainfall, as contrasted with modern conditions of alternate wet and dry seasons, so that peat formation at the present time is slower than during the main coal-forming periods. White,[5] basing his estimates on the early rate of peat decomposition on a study of the modern subtropical coastal-swamp peat deposits, estimates (1) that at the present time in the North Temperate Zone about 1 ft. of peat is forming per century; and (2) 2 ft. of peat were formed in a century in the swamps of the Carboniferous coal fields.

Devolatilization of Peat by Thrust Pressure.—The second stage in the formation of coal is brought about by a dynamo-chemical process which greatly alters the physical and chemical properties of the hydrated peat formed during the biochemical decay. The process is called "dynamo-chemical" because a *chemical* action is brought about by a *movement* of the earth's crust. This process brings about a densification of the peat due to the load of soil on top, and results in a loss of water, certain of the decayed liquid products forming a dehydrated binder or paste. A further step in this process, brought about by the application of intense pressure, often resulting from an upward thrust of the earth's crust, causes a further loss of water, accompanied by devolatilization, whereby carbon dioxide, carbon monoxide, and other gases are evolved.

The best theory offered so far to explain the process of anthracitization and devolatilization of coals is the *thrust-pressure hypothesis,* and by its aid the origin of anthracite in all coal fields studied can be explained logically. The essential points of the theory as outlined by White[5,6,7] are given below.

The dynamo-chemical process follows the biochemical or peat-disintegration process, overlapping slightly on the latter, the gradual burial of the peat stopping the bacterial action, and initiating the consolidation and dehydration of the peat. This condensation of the peat, the driving off of gases and water, started by the mere pressure of the load on it, is in time increased by horizontal thrust pressure; the coal is made more dense, it decreases in volume, and is cemented together, gradually being dehydrated, devolatilized, and compacted. The horizontal thrust pressure has converted lignites (brown coals) successively into subbituminous, bituminous, semi-bituminous, semi-anthracite, anthracite, and even into graphitic coal. The extent to which this progression has occurred will depend on the intensity and the duration of the pressure applied. For example, two coals of the same rank may be formed, one by an intense pressure applied for a relatively short period, the other by the prolonged application of a relatively small pressure.

The temperature during devolatilization is also of importance, but the indications are that during the greater part of the coal formation it was relatively low, probably not over 212°F. (100°C.), although many earlier theories hold that the temperatures reached were somewhat higher.

In conclusion, it may be said that devolatilization by thrust pressure brought about by movements of the earth's crust has been the principal agency by which most of the high-grade coal of the earth has been produced from disintegrated peat. In many cases the per cent volatile matter increases progressively along a given coal seam away from a point of high thrust pressure, the per cent volatile matter being an index of the distance traversed from the point of pressure application.

Time Required for the Formation of Coal.—The amount of coal that can be derived from a given amount of peat is, of course, dependent on the variety of the coal. The chief factors in the transformation are the amount of pressure brought to bear by the earth layers over the coal seams, the loss of moisture, and the losses (largely unknown) of carbon, hydrogen, oxygen, and nitrogen during the process. In peat, the water content is often from 80 to 90 per cent and by the time this peat has changed to anthracite coal the moisture has dropped to about 3 per cent. The oxygen content of the original wood substance is over 40 per cent, and in the peat is about 30 per cent, while after anthracitization the coal contains less than 3 per cent of oxygen. Parr[8] gives the following to represent the changes of cellulose* into lignite and bituminous coal:

$$\underset{\text{cellulose}}{5C_6H_{10}O_5} = \underset{\text{lignite}}{C_{20}H_{22}O_4} + \underset{\text{methane}}{3CH_4} + \underset{\text{water}}{8H_2O} + \underset{\substack{\text{carbon}\\ \text{dioxide}}}{6CO_2} + \underset{\substack{\text{carbon}\\ \text{monoxide}}}{CO} \qquad (1)$$

$$6C_6H_{10}O_5 = \underset{\substack{\text{bituminous}\\ \text{coal}}}{C_{22}H_{20}O} + 5CH_4 + 10H_2O + 8CO_2 + CO \qquad (2)$$

The above equations merely indicate what is believed to be the magnitude of the losses of coal-forming substance that takes place in the coalification process.

Ashley[9] concluded that a 3-ft. layer of well-compacted old peat would form a 1-ft. layer of coal like that of the Pittsburgh seam in Pennsylvania, or 20 ft. of plant matter will form 1 ft. of coal. Accordingly, if 1 ft. of peat is formed per century, and 3 ft. of peat are required to form a foot of coal, 300 years would be necessary to form the peat for each foot thickness of our coal seams. Thus, the formation of the peat corresponding to the seams of Pennsylvania anthracite, some of which are 35 ft. thick, would have required 10,500 years.

The Causes of Variations in Coals.—In general, the variations in coal may possibly be due to one or more of the following causes:

1. Differences in types of vegetal matter.
2. Differences in the extent of biochemical decay.

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* Certain authorities now believe that cellulose is not the parent substance of coal (see p. 34).

3. Differences in the extent of devolatilization due to thrust pressure.

1. *Variations Due to Differences in Types of Vegetal Matter.*—All coals, except certain special types, such as cannel coals, contain light and dark bands, bits of bark, and other plant debris, all of which seem to point toward the same kind of plant material. Analyses of the various types of trees, both of the present time and the tree remains of different species found in coal (maximum variation in carbon in any case less than 3 per cent), show that it would be impossible to explain the wide difference in coals on this basis. Furthermore, as pointed out by White,[5] some of the beds of the anthracite fields in Pennsylvania are practically continuous from the Lehigh River on the east to the coal field in Dauphin County to the west. The plant remains are alike throughout, but in the east the coals are anthracites with very high fixed-carbon contents, the amount of fixed carbon decreasing toward the west until in Dauphin County the coals are only semi-anthracites. It should be recognized, however, that certain portions of the plant, such as spores or resin, containing high carbon percentages, if allowed to collect separately in large quantities, may form such varieties as cannel coal (see p. 22). The amount of such formation is small, however, and there is no evidence that differences in our major coal seams were brought about by differences in the type of vegetal matter. There are a few cases, on the other hand, where resinous trees, or the resinous portions of the vegetal matter, have caused increases in volatile matter, even in the same seam.

The fact that semi-bituminous, bituminous, and subbituminous coals occur in the same coal field seems to minimize the importance of climatic differences as an explanation of the variations existing in our coal beds. For example, in the Appalachian coal field, the fixed carbon content drops 20 per cent in passing from the Georges Creek basin near Cumberland, Md., a distance of a little over 70 miles, westward to Clarksburg, W. Va. It is obvious that the difference in climate between these two points must have been negligible.

2. *Differences Due to Extent of Biochemical Decay.*—If conditions of exposure are severe, if the wood is not under water and is exposed to the air, dry rotting is rapid and the woody material is decomposed first, leaving the more resistant bark. The presence of layers of mother of coal (fusain), for example, must

be attributed to these periods of excessive exposure. Thus, differences in exposure and consequent differences in biochemical action may give rise to the banded constituents of coal, but since these bands are found in practically all coals it is impossible to explain wide differences in variety by postulating differing times and types of exposure before burial. In general, the differences due to variation in extent of biochemical decay are mainly differences in the appearance of the coal.

FIG. 5.—Regional progressive carbonization in the Northern Appalachian trough. Heavy black lines (isocarbs) "contour" the fixed carbon in ash- and moisture-free coal; thin dotted lines outline coal fields; rows of small crosses mark thrust fields.

3. *Differences Due to Variation in Thrust Pressure.*—The map in Fig. 5 (from White[7]) shows the variation in fixed carbon* in the Appalachian trough of the eastern part of the United States. In the northern section of the trough are found a number of identical coal beds extending continuously about 300 miles from the eastern to the western side. In Ohio, on the

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* The percentages of fixed carbon represent the value calculated to a moisture-, ash-, and sulfur-free basis—that is, the per cent of fixed carbon in the "pure coal."

western side, the thrust pressures developed were relatively small and, as shown by the map, the resulting coals contain only 55 to 60 per cent fixed carbon. Moving eastward, coals with 60 per cent fixed carbon are found near Pittsburgh; still further east, near Connellsville, more intense thrust pressure produced coking coals with about 65 per cent fixed carbon; at Windber, only 70 miles east of Connellsville, the fixed carbon has increased to 83 per cent. In the extreme eastern section, where the most severe thrust pressures were developed when the Appalachian Mountains rose, the coals were devolatilized to such an extent that anthracites (with 97 per cent fixed carbon) have resulted.

It should be remembered, however, that the progression from east to west is frequently interrupted by local disturbances where the pressure was not applied entirely to the coal but was in part absorbed otherwise. That is, a coal may be of higher volatile matter than the coals around it, since part of the pressure might have been extended in folding the overlying earth layer without being exerted on the coal.

The impression is quite widespread that differences in the geologic period and age of coals are sufficient to explain differences in variety and quality. This would be true to a certain extent if all other factors were equal, since in the older coals the transforming processes would have had longer to act. However, the evidence previously presented with respect to the marked differences found even in the same coal bed is sufficient to show that difference in age does not necessarily mean difference in variety, but that the greatest differences are brought about by devolatilization due to thrust pressure.

The Origin of Cannel and Boghead Coals.—The cannel and boghead coals (see Chap. III) are characterized by their high percentage of volatile constituents, the absence of woody material, and their regular structure as contrasted to the various bands occurring in most coal. These coals (particularly the boghead and various types of oil shales) more closely resemble petroleum (see Chap. VI) than ordinary coals, on account of the high volatile content. They were originally supposed to have been formed from the remains of gelatinous plant or animal life. More modern researches have definitely established the fact that, except in rare instances, the cannels and bogheads consist of the spores, spore exines, and seed cases of the coal-forming plants, these small bodies collecting in the open water of the coal swamps and giving rise to the layers of this material that occur with some bituminous coal seams.

THE COMPOSITION OF COAL

From the preceding discussion of the origin and methods of formation of coal, it becomes evident that, in spite of the differences that are bound to occur in coal seams, there are certain fundamental bases of comparison. All coal contains various combustible materials, left from the alteration of the vegetal matter, moisture, and varying amounts of mineral matter which on burning appear as ash. The usual means of determining the composition of a coal are: (1) by an ultimate analysis and (2) by a proximate analysis.

Ultimate Analysis.*—The ultimate analysis of coal consists in determining the percentages of the "ultimate" constituents; namely, carbon, hydrogen, oxygen, sulfur, nitrogen, and ash in the *dry* coal. The amounts of carbon, hydrogen, and nitrogen are determined by the ordinary methods of organic quantitative analysis. The sulfur may be determined by the Eschka method or by analysis of washings from the bomb calorimeter used in determining the heating value of the coal. The amount of oxygen is always determined by difference, the sum of the percentages of carbon, hydrogen nitrogen and sulfur, plus the percentage of ash, being subtracted from 100 to give this constituent. This method has the disadvantage of making the reported percentage of oxygen subject to the cumulative error of all the direct determinations. For example, if the sulfur is originally present in the form of iron pyrites (FeS_2) this tends to make the oxygen too low due to formation of ferric oxide (Fe_2O_3) on burning, which would make the ash greater than the original mineral content of the coal. To obviate this particular difficulty it has been recommended by the Coal Analysis Committee of the American Chemical Society[12] assuming the pyrites to be converted to ferric oxide in the ash, that the oxygen be determined by subtracting from 100 the sum of carbon, hydrogen, nitrogen, ash, and five-eighths the sulfur.

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* Owing to the large number of texts which give detailed descriptions of methods of analysis, both ultimate and proximate, only a brief outline of the two methods is given here. For a complete description of standard methods, the reader is referred to references 1, 10, and 11 in Bibliography at the end of the chapter.

Proximate Analysis.—The proximate analysis of coal, which includes the determination of moisture, volatile matter, fixed carbon, ash, and sulfur, is much easier and quicker than the ultimate analysis, and is the one more commonly made, since it furnishes most of the data necessary to determine the commercially important properties of the coal. The amount of moisture is determined by the loss in weight of a small sample on heating for 1 to 1½ hr. at 220°F. (104°C.) to 230°F. (110°C.). The volatile matter is determined by heating a 1-gm. sample of the coal as purchased in a closed crucible at 1740°F. (950°C.) for 7 min. The percentage of volatile matter is the percentage loss as thus determined minus the per cent moisture. Ash is determined on the dried sample from the moisture determination by completely burning off all combustible matter until the ignited material reaches a constant weight. The percentage of fixed carbon is equal to 100 minus the sum of moisture, volatile matter, and ash. Sulfur is usually included in a proximate analysis, being carried out by the methods mentioned above under Ultimate Analysis.

The range in which the constituents of coals usually fall is given in Table III.

TABLE III

VARIATION IN PROXIMATE AND ULTIMATE ANALYSES OF COALS

	Ultimate Analysis Per Cent		Proximate Analysis, Per Cent
Carbon	50–95	Fixed carbon	16–93
Hydrogen	2–5.5	Volatile matter	3–50
Oxygen	2–40	Moisture	3–30
Sulfur	0.5–7.0	Ash	2–30
Nitrogen	0.5–3.0		
Ash	2.0–30.0		

Moisture.—Considerable confusion has resulted from the la of uniformity in methods of reporting analyses. The vario methods of reporting, such as "as received," "air dried," "as and moisture-free," etc., refer to the condition of the coal wh analyzed and have in many cases caused needless recalculati and confusion. Moisture should be reported on the coal received. It is always desirable to calculate the *total* water fro the actual moisture present, plus the combined water which for part of the coal substance itself. This combined water is repr sented by the oxygen content of the coal plus the equivale amount of hydrogen necessary to unite with it to form water. The remaining hydrogen in the coal is termed "free" or "available hydrogen," and on burning can produce heat by union with the oxygen of the air.

Ash.—The ash in coal is composed of two classes: (1) free ash and (2) fixed ash. The ash-forming constituents of coal are: (*a*) impurities present as an intimate mixture in the coal substance, derived from the ash in the original vegetal matter or from sedimentation, etc., during coal formation; (*b*) bits of clay, shale, "slate," pyrites, etc., that occur in the form of veins and partings; and (*c*) fragments from the roof and floor, etc., that are mixed with the coal in the process of mining. The "free ash" is mostly that formed by processes (*b*) and (*c*), and, consisting of dirt, slate, pyrites, etc., can be washed out of the coal by suitable coal-washing machinery. The amount of such ash in the coal varies considerably in different parts of the mine. The "fixed ash," formed mostly by process (*a*), is disseminated throughout the coal fairly uniformly, and usually amounts to from 2 to 3 per cent. This ash, since it cannot be readily washed out of the coal, therefore represents the minimum possible ash content of coal.

It is well known that in most coals the finer material contains larger percentages of ash than the lump material. For example, in certain anthracite coals, the egg size contains about 4 per cent ash, the nut size 9 per cent, and the pea size 15 per cent. Recent work by Parr[13] has indicated that, while a certain proportion of this increase in ash may be the result of mining methods, it is possible that one of the banded constituents of coals known as mother of coal (see page 40), being very friable and of high ash content, raises the ash content of the smaller sizes. Parr gives the following as the ash contents of the banded constituents of an Illinois coal:

	Ash, Per Cent
Bright band	3 22
Dull band	14 77
Mother of coal	22.45

Coal ash is a mixture of oxides, silicates, and sulfates, the relative percentages of which vary widely in different coals, the chief components being silica (SiO_2), alumina (Al_2O_3), iron oxides (FeO or Fe_2O_3), calcium oxide (CaO), with smaller quantities of magnesia (MgO), titanium oxide (TiO_2), and alkali compounds. The SiO_2, Al_2O_3, and TiO_2 come from sand, clay, slate, etc., the FeO and Fe_2O_3 mainly from pyrites (FeS_2) disseminated throughout the coal vein, and the CaO and MgO as sulfates and carbonates. The typical limits of some ash analyses are given in Table IV.

TABLE IV

Typical Limits of Ash Analyses

Constituent	Per Cent of Total Ash
SiO_2	40–60
Al_2O_3	20–35
Fe_2O_3	5–25
MgO	0.5– 4
$Na_2O + K_2O$	1.0– 4
SO_3	0.2–15
CaO	1.0–15

The fusibility of the ash is of distinct bearing on the question of clinkering troubles, but no method of general application has yet been developed for calculating the fusibility of an ash from its composition. Like clay, however, the softening point is, in general, lowered by increasing amounts of magnesia, lime, and iron compounds (particularly sulfides).

From the direct determination of the ash-fusion temperature of a large number of samples, coals have been divided into three classes.[14]

Class 1.—Coals having ash which fuses at temperatures from 2600 to 3100°F., and, in general, includes the anthracite coals and the lower and older bituminous beds. Ash from such coal is refractory, giving no trouble from clinkering.

Class 2.—Coals having ash which fuses at temperatures from 2200 to 2600°F., and, in general, includes the bulk of Pennsylvania bituminous beds. The ash from these coals is o medium fusibility.

Class 3.—Coals having ash which fuses at temperatures from 1900 to 2200°F. This ash is easily fusible and causes excessiv clinkering trouble unless used properly.

The fusion temperatures of the ash from a large number o American coals are given in Tables V, VI, and VII.

TABLE V

FUSIBILITY OF COAL ASH FROM PENNSYLVANIA COALS*

Series, bed	Number of mines sampled	Total number of samples	Average softening temperature, degrees Fahrenheit	Ash in dry coal, average per cent	Sulfur in dry coal, average per cent
Bituminous region					
Monongahela series:					
Pittsburgh bed	10	34	2360	7.17	1.43
Conemaugh series:					
Little Pittsburgh bed	1	2	2390	8.13	1.70
Allegheny series:					
Upper Freeport (E or Lemon) bed	33	85	2350	9.35	2.13
Lower Freeport (D or Moshannon) bed	21	70	+2390	8.52	2.06
Upper Kittanning (C Prime) bed	7	18	+2350	8.67	2.16
Middle Kittanning (C) bed	5	7	+2380	11.06	2.98
Lower Kittanning (Miller or B) bed	39	162	+2550	7.86	2.03
Fulton bed	4	12	+2940	7.36	1.18
Brookville (A) bed	3	3	+2800	12.98	1.86
Pottsville series:					
Bloss bed	3	3	2630	11.96	2.25
Anthracite region					
Northern field:					
Pittston district	1	1	+3010	6.03	0.58
Plymouth district	2	7	+3010	12.52	0.84
Scranton district	6	7	+3010	12.39	0.79
Wilkes-Barre district	3	4	+3010	13.17	0.78
Eastern Middle field:					
Hazelton district	1	1	2960	14.50	0.61
Western Middle field:					
Shamokin district	2	2	+2960	16.59	0.90
West Mahony district	3	3	+3010		
Southern field:					
East Schuylkill district	2	2	2990	11.19	0.78
West Schuylkill district	1	1	2730	18.07	0.82

Note.—A plus sign (+) placed before a given value denotes that the true value is above that indicated.

* From SELVIG and FIELDNER, *Chem. Met. Eng.*, **20**, 629 (1919).

TABLE VI

Fusibility of Coal Ash from West Virginia Coals*

Series, bed	Number of mines sampled	Total number of samples	Average softening temperature, degrees Fahrenheit	Ash in dry coal, average per cent	Sulfur in dry coal, average per cent
Monongahela series:					
Sewickley (Mapleton) bed	1	5	2080	9.61	3.99
Redstone bed	2	9	2120	6.08	1.95
Pittsburgh bed	15	63	2170	7.49	2.45
Conemaugh series:					
Mahoning bed	4	4	2160	5.62	1.89
Allegheny series:					
Upper Freeport bed	1	1	2190	6.17	1.97
Lower Freeport bed	2	2	2090	9.84	3.14
Middle Kittanning bed	1	1	2110	10.93	4.06
Lower Kittanning bed	7	22	+2490	8.48	3.28
Pottsville series:					
Kanawha group:					
Coalburg (Buffalo Creek) bed	2	7	+2990	8.94	0.80
Winifrede bed	3	4	2950	7.83	0.78
Cedar Grove bed (Thacker bed)	10	24	+2540	5.96	1.21
No. 2 Gas (Campbell Creek, Island Creek, Upper War Eagle) bed	6	27	+2680	5.60	0.97
Eagle (No. 1 Gas, Middle War Eagle) bed	1	3	+2940	4.40	0.77
New River group:					
Sewell (Davy) bed	33	95	+2580	4.36	0.66
Welch (Tug River) bed	1	5	2840	7.41	0.62
Beckley (War Creek) bed	22	101	+2800	4.82	0.68
Fire Creek (Quinnimont) bed	3	17	2440	6.26	0.81
Pocahontas group:					
Pocahontas No. 6 bed	1	6	2380	2.88	0.70
Pocahontas No. 5 bed	2	3	+2610	5.45	0.60
Pocahontas No. 4 bed (Thin Vein Pocahontas)	7	23	+2570	6.27	0.63
Pocahontas No. 3 bed (Thick Vein Pocahontas)	66	246	2460	4.81	0.60

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* From Selvig, Ratliff and Fieldner, *Chem. Met. Eng.*, **20**, 275 (1919).

TABLE VII

Fusibility of Coal Ash from the Interior Province Coals*

Region, state, bed	Number of mines sampled	Total number of samples	Average softening temperature, degrees Fahrenheit	Ash in dry coal, average per cent	Sulfur in dry coal, average per cent
Eastern region:					
Illinois:					
No 1 bed	1	6	2110	11.74	4.86
No. 2 bed	6	15	2010	9.97	3.58
No. 5 bed	9	27	1990	10.84	3.28
No. 6 bed	36	160	2160	10.27	2.30
No. 7 bed	2	13	2050	10.62	2.69
Indiana:					
No. 3 bed	3	11	2090	10.61	4.34
No. 4 bed	7	30	2390	8.17	1.62
No. 5 bed	9	29	2130	10.23	3.54
No. 6 bed	1	5	2040	9.91	2.65
Minshall bed	1	1	2120	9.60	2.99
Western Kentucky:					
No. 6 bed	1	4	2130	8.81	2.97
No. 9 bed	20	78	2030	10.53	3.67
No. 10 bed	2	2	1990	11.99	4.18
No. 11 bed	7	39	2030	9.57	4.08
No. 12 bed	3	17	2150	10.20	2.30
Western region:					
Kansas:					
Bevier bed	2	6	1980	14.83	4.71
Cherokee bed	3	8	2110	9.42	3.18
Leavenworth bed	1	3	2020	18.26	5.46
Weir-Pittsburgh bed	3	7	2010	11.68	5.31
Missouri:					
Bevier bed	16	42	1960	13.47	4.90
Bowen bed	1	3	1940	13.18	4.61
Cainsville bed	1	3	1980	12.71	5.78
Cherokee bed	1	3	2150	7.51	1.97
Jordan bed	4	12	2010	12.74	4.42
Lexington bed	14	39	2000	13.48	4.04
Lower Rich Hill bed	3	8	1940	15.39	5.43
Lower Weir-Pittsburgh bed	2	6	1940	10.78	4.45
Mulberry bed	2	5	1990	14.58	3.18
Mulky bed	4	9	1940	11.28	5.25
Rich Hill bed	1	3	1970	15.47	6.12
Tebo bed	8	23	2040	11.64	4.66
Waverly bed	1	2	2020	17.43	8.29
Oklahoma:					
Dawson bed	3	3	1920	8.95	3.91
Henryetta bed	2	2	1980	8.03	1.59
Lehigh Coal bed	3	15	2150	11.46	4.17
Lower Hartshorne bed	9	27	2020	6.03	1.43
McAlester bed	6	28	2180	6.94	1.67
McCurtain bed	1	6	2110	6.92	0.84
Panama bed	2	4	2160	6.81	1.46
Stigler bed	5	5	2050	5.13	1.91
Upper Hartshorne bed	3	7	2170	6.15	1.51
Arkansas:					
Denning bed	1	3	2200	7.38	2.45
Hartshorne bed	5	9	2120	11.59	1.40
Paris bed	1	3	2140	10.12	3.28
Shinn Basin bed	1	1	2180	10.36	2.23
Southwestern region:					
Texas:					
Santo Tomas	2	2	2580	19.21	1.98

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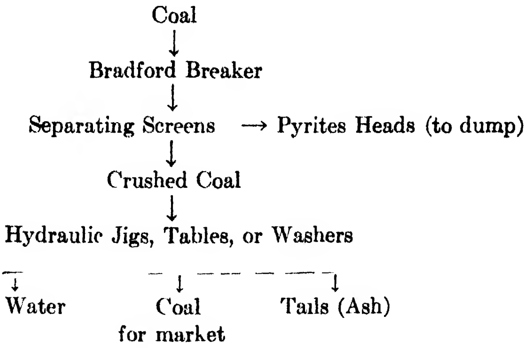
* From Selvig, Ratliff and Fieldner, *Chem. Met. Eng.*, **20**, 275 (1919).

The practice of coal washing has for its object the removal of as much *free* ash as possible. A number of processes are in use and depend for the most part on differences in specific gravity (and in some cases to other properties, such as the differences in coefficient of friction) of the coal and mineral matter. A diagrammatic representation of one method of operating is as follows:



Such a process successfully removes a considerable quantit of ash at a comparatively low cost. Bone[11] cites the example an installation handling 100 tons of coal per hr., which reduc the ash from 22.5 to about 7.0 per cent, with a loss of less tha 2 per cent of the coal substance and a water consumption of 1 gal. per ton of coal washed. The labor cost per ton is smal

Sulfur.—The presence of sulfur in coal is always harmfu since it affects the quality of coke for metallurgical uses, increas the corrosion of boilers, and usually affects the fusibility of th ash. The sulfur content of coals, which may vary from 0. to 7 per cent, is mainly derived from extraneous material cro ping into the coal seam. It occurs in three forms: (1) pyriti sulfur, (2) organic sulfur, and (3) sulfate sulfur.

1. Pyritic sulfur, present in coal mainly as iron pyrites "fools gold" (FeS_2), is usually distributed in the form of banc running parallel with the bedding or in small veins across th seam, and sometimes in quite large lumps (pyrites heads Thiessen[15] has shown that, in addition to the large particles pyrites visible to the naked eye, there is almost always a certai amount present in the coal which only shows up under the micr scope. The pyrites is mainly formed by the intrusion of roc but Thiessen points out that in almost all plants there is sulfu in the proteins present and this on putrefaction yields H_2S, whic could precipitate the heavy metal sulfides. The amount pyritic sulfur varies widely and may be lowered by prope

2. Organic sulfur is a component of the organic compounds present in coal. It is probably derived from the protein sulfur mentioned above and may be the source of the pyritic sulfur.

3. Sulfate sulfur is commonly present in coal as gypsum ($CaSO_4.2H_2O$), though iron, copper, and magnesium sulfates are sometimes found. These sulfates were mainly formed by the oxidation of the sulfides present. Pyrites, on weathering, undergoes oxidation according to the reaction:

$$2FeS_2 + 7O_2 + 2H_2O = 2FeSO_4 \text{ aq.} + 2H_2SO_4 \text{ aq.}$$

The amount of sulfur present as sulfate is, therefore, to a certain extent, an indication of the amount of weathering to which the coal has been subjected. In some cases of weathered coal containing originally a high percentage of pyritic sulfur, the iron sulfate formed by the above reaction deposits as a white powder over much of the surface of the coal lumps.

Nitrogen.—The nitrogen content of coals is of great economic importance, because of the value of the nitrogen obtained as ammonia in the manufacture of by-product coke. The nitrogen, present as a constituent of the organic compounds which go to make up the coal substance, is mainly derived from the original vegetal matter (the N_2 content of trees varying from 1 to 3 per cent), although a small proportion may originally have come from animal matter or from the air.

The Rational Analysis of Coals.—In addition to the ultimate and proximate analyses of coals, which give the percentages of the various elements present and in a general way indicate the commercial value of the fuel, it is of great importance to know how the elements are combined and how the specific compounds affect the properties of the coal substance itself. This, "the rational analysis" of coals, has been the subject of a great deal of work, and, while it is still impossible to predict the influence of any constituent on the behavior of a coal, distinct progress is being made.

From our knowledge as to the formation of coal, it is seen that coal may contain degradation products of three types:

1. *Cellulosic material.* This is derived from the woody part of the plant or tree.

2. *Resinic material,* such as gums, waxes, resins, etc. These may have been derived from the sap of the original vegetal matter.

3. *Nitrogenous material,* compounds of carbon, hydrogen, oxygen, nitrogen, and sulfur.

From the discussion of the method of coal formation, it becomes evident that the above three classes of compounds must all have undergone varying degrees of change or degradation dependent on the conditions of vegetal accumulation and decay, and the subsequent pressure and temperature to which they were subjected. There are four main methods for determining in what form the degradation products of these materials are present in coal:

1. Action of chemicals.
2. Selective solution.
3. Effect of heat.
4. Microscopical analysis.

1. Action of Chemicals.*—Chemical attack by means of reagents which alter the coal substance has been used by a number of investigators. The reagents are usually oxidizing agents, such as nitric acid, sulfuric acid, bromine, ozone, caustic alkalies, alkaline potassium permanganate, or chromic acid. The object in using reagents is to transform the coal into another degradation product, the relationship of which to the original coal substance can be determined. For example, when peat, brown coal, or lignite is digested in a solution of caustic alkali, part of the material apparently dissolves, yielding a deep-brown liquor. The dissolved material separates out on acidification of the liquor, and is known as "ulmic acid" or "humic acid." The amount of this material in a coal gives a rough indication of its geological rank, peats yielding 50 per cent or more, while the bituminous and anthracite coals are practically unaffected.

In general, while some interesting data have been obtained,† the

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* For a complete and exhaustive account and discussion of the major investigations which have been undertaken following the different methods of "rational analysis of coals," see STOPES and WHEELER,[16] BONE,[11] and ILLINGWORTH.[17] An excellent and more recent review has been prepared by FIELDNER and DAVIS.[18]

† Although not undertaken to develop an analytical method, the researches of Bergius in Germany,[21] should be mentioned here. In the Bergius process coal dust suspended in oil, or tar, pitch or asphalt are hydrogenated and reduced to relatively light, low boiling products by heating with hydrogen at pressures of 100–200 atmospheres and at about 750°F. (400°C.). A carrier or accelerator such as Fe_2O_3 (5 per cent) is added to the suspended coal before hydrogenation. This process may convert as high as 50 per cent of the coal to oil and 70 per cent to oil and gas. The oils are essentially non-phenolic. The Bergius method may become of distinct commercial importance, particularly in countries having coal but little oil.

results of this method of investigation have been scanty, due to the complex nature of coal and the difficulty in tracing back to the original coal substance the relationship of the compounds formed.

2. Selective Solution.—When it can be used, the action of solvents is preferable to other known methods, since it brings about the minimum chemical alteration of the coal substance. The usual agents employed, benzene, chloroform, petroleum ether, acetone, pyridine, aniline, or phenol, are open to the objection, however, that the proportion of the coal substance dissolved is usually very small, and it is difficult to find solvents which will selectively dissolve one type of compounds without having a depolymerizing or breaking-down effect on other portions.

Some consider that a practically complete separation of the cellulosic and resinic materials in the coal substance has been brought about by combined pyridene and chloroform extraction. The results of solvent action, and their interpretation, as quite generally accepted by the English workers in this field, have been outlined by Stopes and Wheeler[19] somewhat as follows: Two main fractions of the coal substance, one apparently derived from the cellulosic constituents of the vegetal matter and the other of the resinic portions, may be separated by means of solvents. These groups are called ($\alpha + \beta$) compounds (which later can be further separated into α and β compounds which differ but little) and γ compounds, respectively. These three types of compounds are specifically differentiated by the successive action of pyridene and chloroform as follows:

1. The alpha (α) compounds are insoluble in pyridene; on destructive distillation they yield gas and very small quantities of tar, but do not fuse; they are obtained as a brown powder.
2. The beta (β) compounds are soluble in pyridene but are left as the insoluble residue from a subsequent chloroform extraction of the portion soluble in pyridene; they are distinguishable from the alpha (α) compounds only by their solubility in pyridene.
3. The gamma (γ) compounds consist of the chloroform-soluble portion of the pyridene extraction; these compounds are obtained in the form of a resin or lac, and melt at about 212°F. (100°C.); on destructive distillation they yield gas and tars (composed of olefines, paraffins, and naphthenes), and the residue is well fused and coked.

Wheeler believes that the alpha and beta compounds represent the degradation products of cellulose, while the gamma compounds are derived chiefly from the resinous material present in the coal-forming vegetal matter. Thiessen,[4] and Fischer and Schrader,[20] however, have adduced strong evidence tending to show that most of the cellulose disappears during peat formation, being decomposed into methane, carbon dioxide, and water, and that the major portion of the coal substance is derived from the degradation products of lignin.

Another important method of solvent extraction has been employed by Fischer[21] and his co-workers in Germany. The method, which consists in treatment of the coal with benzol at high pressures and temperatures, yields an extract containing the compounds responsible for the coking characteristics of the coal (*i.e.*, similar to the γ compounds). The following abstract[21] gives the major conclusions from Fischer's work. The extraction of coal with benzol at 482°F. (250°C.) under pressure yields an extract which can be separated into two main components by means of petroleum ether (gasoline). The first component, an oil, is soluble in gasoline; the second component, a solid brown substance, is insoluble.

Recently it has been discovered that the oily body is the principal source of the coking quality of the coal and apparently consists only of hydrocarbons. In contrast to this, the solid body which is decomposed by heat with gas evolution has been shown to be the principal source of the swelling of coal. It consists of compounds containing hydrogen and resins. The oily body has been named oily bitumen; the solid body, solid bitumen; the sum of the two, total bitumen.

The older the coal is the more oily bitumen it contains compared to the quantity of solid bitumen. The absolute quantity of total bitumen, however, decreases considerably with the older coals. The decomposition point of the solid bitumen rises with the age of the coal; it is of more significance for the inception of swelling than the quantity of solid bitumen. If the decomposition point lies at the temperature at which the coal is plastic, then the swelling is greatest. With regard to the oily bitumen, on the other hand, its quantity is of special significance.

With repeated successive extraction of strongly swelling coal, first, the swelling property disappears, but the coal still remains coking. Finally, the ability to coke is also lost.

If the total extract is added to the extracted coal substance, the original coal is reconstructed, with relation to its behavior on coking. If the oily bitumen only is added, then non-swelling coking coal results. If the solid bitumen only is added, poorly coking but strongly swelling coal is yielded.

The experiments were made with coals from the Ruhr, the Saar, and Upper Silesia. It would be interesting to see whether these generalizations apply also to American coals, that is, whether they are of general significance.

In general, it may be said that, while a great deal of work on solvent action has been done, and more is still being carried on, the results are thus far mainly of academic interest, and their applications to practice can only come after prolonged further research on this subject.

3. The Effect of Heat.—The effect of dry distillation of coal at both high and low temperatures is of great importance, not only from the practical viewpoint, in its application to the production of gas and coke, but also on account of the bearing this method of attack has had upon knowledge of the composition of coal. The general methods and results of the carbonization processes are taken up in detail in Chap. XX. For the purposes of the present chapter only those factors which influence the understanding of the chemical composition of the coal substance will be considered. The investigations discussed in the present chapter include work not only on coal as such, but also on the constituents of the coal substance separated by solvents and other means, the distillations being carried out at (1) low temperature and (2) high temperature.

Among the most important investigations concerning the thermal decomposition of coal at *low temperatures* are those of Burgess and Wheeler.[22] These investigators distilled bituminous coal under vacuum at temperatures of 212, 392, 572, 752, 932°F. (100, 200, 300, 400, 500°C.), determining the yields of gaseous and condensible products at each temperature. Their results may be summarized as follows: (1) Occluded gases, mainly paraffin hydrocarbons, are continuously evolved up to 300 to 400°F. (150 to 200°C.). (2) At 390°F. (200°C.) there is a copious evolution of water of constitution, which continues up to 840°F. (450°C.). (3) Between 520 and 570°F. (270 and 300°C.), H_2S is freely evolved, due to decomposition of the organic sulfur compounds. (4) At 570°F. (300°C.) higher olefines (above

C_2H_4) make their appearance, and at 660°F. (350°C.) the amount of these drops off. (5) A critical decomposition point occurs at about 660°F. (350°C.), which is marked by a rapid increase in gas evolution and the appearance of a viscous oil. (6) The gaseous hydrocarbons methane (CH_4), ethane (C_2H_6), propane (C_3H_8), butane (C_4H_{10}), and even pentane (C_5H_{12}), as well as the olefines (C_nH_{2n}), are present in the fraction distilled off up to 840°F. (450°C). (7) No benzene or homologues of benzene could be detected below 840°F. (450°C.).

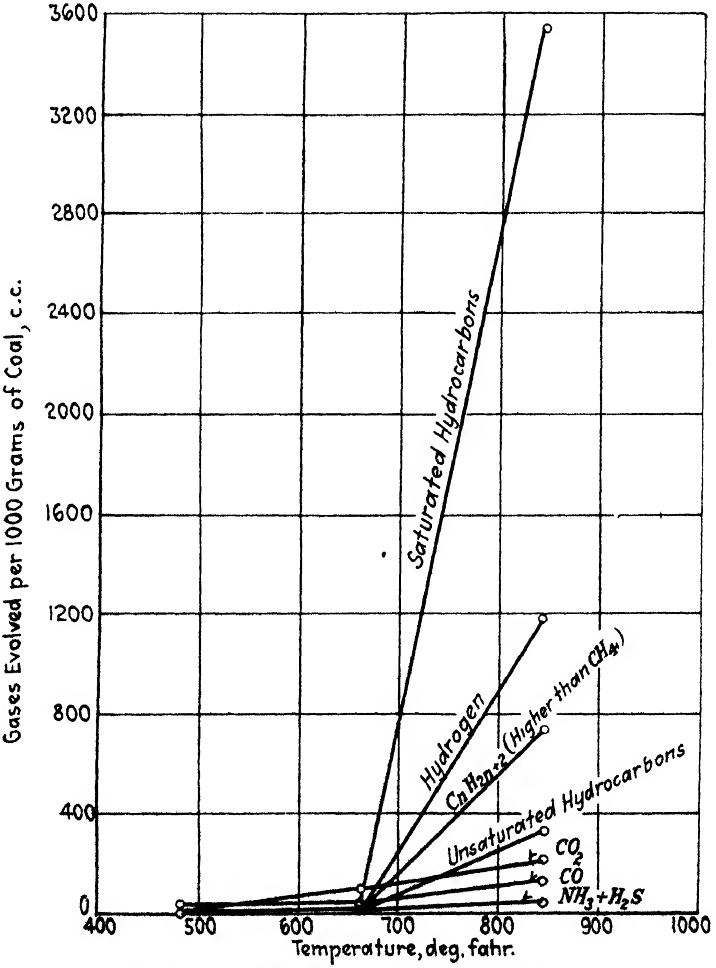

FIG. 6.—Quantities of various gases evolved from New River (W. Va.) coal distilled in vacuo.

Similar results were obtained by Taylor and Porter[23] at low temperatures, the yields of various gases obtained from New River (West Virginia) coal being shown in Fig. 6.

Burgess and Wheeler also investigated the amount and composition of the gaseous and condensible products obtained during the distillation of four English coals at *high temperatures*, namely, from 840 to 1830°F. (450 to 1000°C.). Figure 7 shows the amounts of gaseous products obtained from one of the coals at various temperatures and is representative of the results obtained from all the coals.

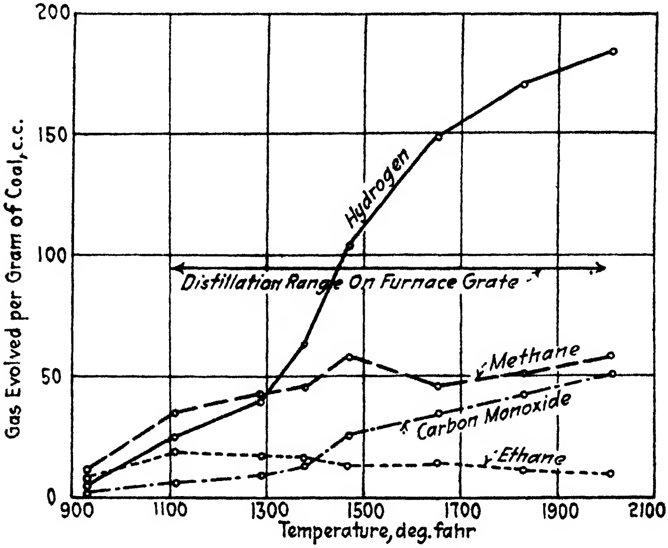

Fig. 7.—Volumes of gases evolved from coal at different temperatures.

The main points to be brought out in connection with this latter work are as follows: (1) Ethane (C_2H_6), propane (C_3H_8), butane (C_4H_{10}), and higher members of the paraffin series form a large percentage of the gases at temperatures below 840°F. (450°C.). (2) With all coals from anthracite to bituminous there is a well-defined temperature between 1300 and 1470°F. (700 and 800°C.) at which hydrogen begins to be evolved rapidly, probably due to decomposition of hydrocarbons. (3) This evolution drops off at about 1700°F. (925°C.) with bituminous coals and at 2000°F. (1100°C.) with anthracite. (4) The evolution of the paraffin hydrocarbons ceases entirely at temperatures above

1300°F. (700°C.). (5) The formation of carbon monoxide (CO) is continuous throughout the period of distillation, the rate increasing with increasing temperature (probably due to hydrocarbons and carbon giving CO and H_2).

From these results Burgess and Wheeler drew the following conclusions: (1) that coal contains two types of compounds of different degrees of ease of decomposition; the *least stable* of these *yields mainly paraffin hydrocarbons* and not hydrogen; the *more stable yields mainly hydrogen* or possibly hydrogen and oxides of carbon; (2) that the chemical differences of various coals depend on the proportions of these two compounds present; (3) that the presence of CO in the gases is probably due to the reaction of carbon and the water given off from the hydroxy compounds; (4) that the main "paraffin-yielding" compounds represent the resinous substances originally present in the coal, and that the main "hydrogen-yielding" compounds are the degradation products of the celluloses.

The principal results of Porter and Taylor[23] (from their experiments on the distillation of various American coals in vacuo) are (1) More than two-thirds of the organic coal substance is decomposed below 930°F. (500°C.). Different coals vary in ease of decomposition, the older coals of the Appalachian field being less rapidly and easily broken down than the younger coals of the middle western and western fields. (2) The first decomposition, as the temperature is raised, marks the breaking down of certain oxygen-containing substances related to cellulose, the products being chiefly H_2O, CO_2, and CO. (3) In general, their results at low temperatures agree with those of Burgess and Wheeler, the main difference occurring in the high-temperature distillation. This difference, which centers around the explanation of the "critical point" of 1380°F. (750°C.), is attributed by Porter and Taylor to a secondary decomposition of the primary volatile products into hydrogen. The liquid hydrocarbons in the tar undergo decomposition more readily than the gases, and at 1350°F. (730°C.) yield hydrogen, methane, ethylene, hydrocarbons, and carbon.

More recent investigations by Wheeler and his co-workers and by S. R. Illingworth, on the action of heat on the extracted portions of the coal substance, have led to the conclusions previously given under the heading, Selective Solution (p. 33). Illingworth investigated the relation of the α, β, and γ groups to the coking property of coals, determining the proportion of each group in the residues from heating typical coals under varying conditions. He concluded that the β compounds (soluble in pyridene, insoluble in subsequent chloroform extraction) were less stable than the γ, or resinic compounds (the chloroform soluble portion of the pyridene extraction), and postulated, as others have done, that the γ compounds form the cementing material of the coke, since it becomes fluid at temperatures below its decomposition point, fills the interstices, and cements the individual particles within.

The conclusions from the action of heat on coals may be summarized as follows:

1. Copious evolution of water and oxides of carbon begins at 390°F. (200°C.) and continues at least to 840°F. (450°C.).
2. Hydrogen sulfide begins to be evolved at 520 to 570°F. (270 to 300°C.).
3. Olefines (higher than ethylene) are given off at 570°F. (300°C.).
4. The first critical temperature, at 660°F. (350°C.), is characterized by (*a*) marked increase in gas evolution, (*b*) appearance of a viscous oil, and (*c*) the appearance of ammonia.
5. Up to 840°F. (450°C.) the gases evolved are mainly paraffins (C_nH_{2n+2}) and olefines (C_nH_{2n}), with some CO.
6. Paraffins cease coming off at 1290°F. (700°C.).
7. Hydrogen begins to come off at 1290°F. (700°C.), the second critical temperature, and continues to 1650 to 2010°F. (900 to 1100°C.), depending on the type of coal.
8. At low temperatures no aromatic hydrocarbons of the benzene series are found.

4. Microscopic Studies.—Mention must also be given the microscopic studies that have been made of the banded constituents of coals (or, more properly, the constituents of banded coals). The pioneer work in the United States was done by Thiessen,[25] who developed a method of cutting sections of coal thin enough for complete examination under the microscope. The bright part of the coal, which represents coalified trunks, branches, and twigs, which were buried in the peat débris, he terms anthraxylon, meaning wood coal. Thiessen[4] has also introduced the term "attritus" to describe the highly macerated plant débris consisting of bits of leaves, bark, spores, etc., mixed with resinous, carbonaceous, and mineral matter. This material is dull in appearance and the dull bands in coal consist of relatively large amounts of attritus with only small fragments of the woody material (anthraxylon). The bright bands in coal are composed principally of anthraxylon (in many cases also containing small amounts of attritus). One other banded constituent is recognized in the United States, namely, the mother of coal or mineral charcoal, so called on account of its resemblance to manufactured charcoal. The mother of coal is dull in appearance, but is generally softer than the other dull bands and is sooty and crumbling. It is fibrous in structure as contrasted to the granular appearance of the dull bands. Parr and Knapheide[25] consider the mother of coal of distinct importance in studies of carbonization problems with coals of the mid-continental field, since it acts as a non-coking constituent and a vehicle for oxygen (see p. 44 *et seq.*).

The English workers in this field of investigation recognize four banded constituents which are given by Stopes[26] somewhat as shown in Table VIII.

TABLE VIII

CHARACTERISTICS OF CONSTITUENTS OF BANDED COALS

(*After Stopes*)

1. Fusain—the equivalent of "mother of coal," "mineral charcoal," etc.
2. Durain—the equivalent of "dull" hard coal of various authors, the "Mattkohle" of the Germans
3. Clarain
4. Vitrain

} Clarain and vitrain are together the equivalent of "bright" or "glance" coal of various authors. Sometimes the "bright coal" seems to be vitrain only. Clarain is said to contain more recognizable plant tissues than any other part of the coal. Vitrain has a conchoidal fracture, is brilliant in appearance, is structureless, and is of uniform texture.

Thiessen, however, has been unable to confirm the structureless nature of vitrain. He has found that with sufficiently thin sections and with proper illumination, typical plant cell structure can always be observed in bright coal seams. It would thus seem that clarain and vitrain are, in reality, the same material and that the difference believed to exist by the English workers may be one of degree or a question of technique in preparing thin sections and observing them under the microscope.[18] Further evidence tending to confirm this view is mentioned by Fieldner and Davis.[18]

These views as to the constituents of banded coals are summarized in Table IX.

TABLE IX

Summary* of Views as to the Banded Constituents of Coals

General appearance	Views of English investigators	Views of American investigators
Bright bands. . . .	1. Clarain—contains much recognizable plant structure 2. Vitrain—vitreous appearance; structureless	Anthraxylon, or anthraxylon containing very small amounts of attritus
Dull bands. . .	Durain	Mainly attritus, usually containing some small particles of anthraxylon
Dull, thin, charcoal-like bands.	Fusain	Mother of coal: mineral charcoal

THE EFFECT OF OXYGEN IN COAL

"In all the chemical changes marking both stages in the conversion of organic débris into coal, the preeminently important result to man is elimination of oxygen. Economically this is the most essential incident of the process."* This elimination of oxygen in the progression from wood to anthracite is the most noticeable chemical change in coal during the process. Table X and Fig. 8 show the amount of this decrease in the case of a number of typical coals, calculated to an ash- and moisture-free basis:

TABLE X

Oxygen Content of Wood, Peat and Typical Coals

(Ash- and moisture-free basis.)

Fuel	Per Cent Oxygen
Wood	43.15
Peat	33.53
Lignite	19.60
Coal:	
Subbituminous	17.01
Bituminous	5.18
Semi-bituminous	2.64
Semi-anthracite	2.17
Anthracite	2.13

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* For a detailed exposition of the effect of oxygen and the tabulation of a large number of coals with regard to O_2 content, H:O ratio, etc., see reference 27.

It is important to see the effect of the oxygen content on the two main uses of coal, namely (1) to produce heat, and (2) to serve as a raw material for coke.

Effect of Oxygen on Heating Value.—To determine the effect of oxygen on the heating value of coal, White made a study and tabulation of the complete analyses and calorimeter tests (heating values) of over 300 coals, representing practically all the coal fields in the United States. The coals came from 27 states, as well as from Alaska and Mexico. Practically all ranks of coal from lignite to anthracite were included. In the coals studied the ash ranged from 2.38 to 29.38 per cent and the oxygen from 1.82 to 38.57 per cent, and from the tabulation it was found that 1 per cent of oxygen had the same effect as 1 per cent of ash, *i.e.*, oxygen and ash appear to have the same anti-calorific effect.

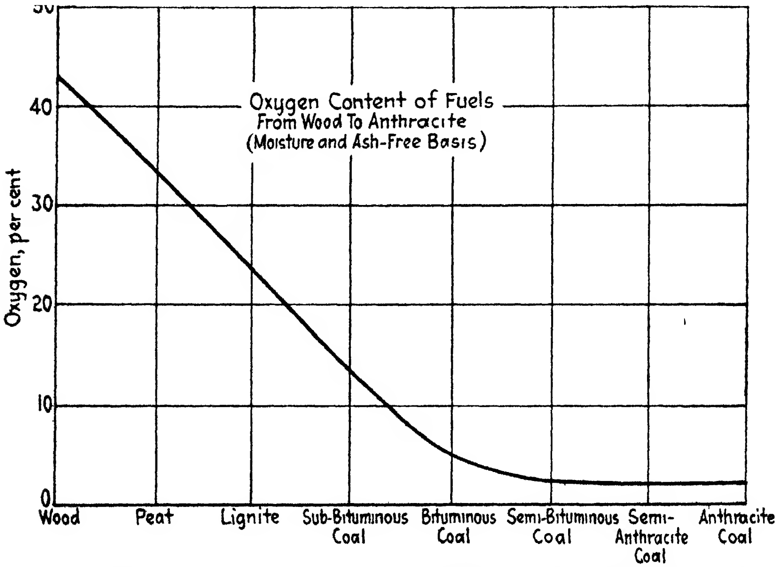

FIG. 8.—Oxygen content of fuels from wood to anthracite (moisture- and ash-free basis).

The per cent carbon changes practically inversely with oxygen due to the high combining power of oxygen compared with hydrogen, and for this reason the effect of hydrogen on the heating value changes but little in all coals. As carbon is the variable heat-forming element, and oxygen and ash are the non-heat-forming elements, White plotted the heating value of the coals (B.t.u. per lb.) against the ratio $\frac{\text{Carbon}}{\text{Oxygen} + \text{Ash}}$, obtaining the curve shown in Fig. 9.

The ratio of carbon to oxygen plus ash is a close approximation to the ratio of combustible to inert material in the coal. This curve may be expressed mathematically by the equation

$$\text{B.t.u. per pound of coal} = 16{,}780 - \frac{17{,}230}{r + 0.98}.$$

$$\text{where } r = \frac{\text{Carbon}}{\text{Oxygen} + \text{Ash}}.$$

This same formula applies equally well to dry coal, air-dried coal, or wet coal, the B.t.u., of course, being for the coal *as calculated*, *i.e.*, for 1 lb. of coal dry, air dry, or wet, as the case may be. This relationship holds, since a change in water content not only effects the $\frac{\text{Carbon}}{\text{Oxygen} + \text{Ash}}$ ratio, but correspondingly alters the heating value. Since oxygen is determined by the difference between carbon, hydrogen, nitrogen, sulfur, ash and 100, and therefore requires the carrying out of five determinations, the above formula serves as a means to calculate the per cent of oxygen by the relatively easy method of a determination of carbon, ash, and the heating value from a bomb calorimeter test: Thus,

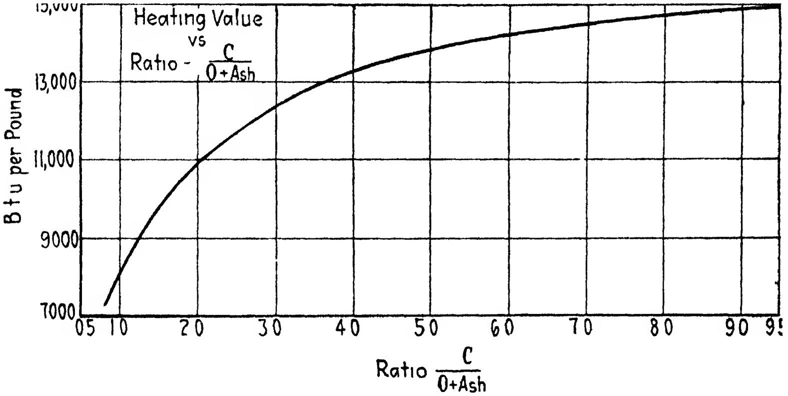

Fig. 9.—Relation between the heating value of coals and the ratio $\frac{\text{carbon}}{\text{oxygen} + \text{ash}}$

$$\frac{\text{Carbon}}{\text{Oxygen} + \text{Ash}} = \frac{17{,}230}{16{,}780 - \text{B.t.u.}} - 0.98,$$

or

$$\text{Per cent Oxygen} = \frac{16{,}780C - (H)C - 0.98A(H) - 810A}{0.98\ \text{B.t.u.} + 810},$$

where C = per cent carbon, A = per cent ash, and H = heating value in B.t.u. per lb.

The Effect of Oxygen on Coking of Coal.—The second major use of coal is to serve as a raw material for the production of coke. Two widely used rules for identifying coking coals are:

1. The volatile matter must be between 20 and 30 per cent. This often breaks down, however, and is quite unsatisfactory.

2. The *available** (net or free) hydrogen should be about 3.20 to 3.80 per cent. This "rule" is also untenable. White, however, in his study of the effect of oxygen in coals, found that when the fixed carbon in the pure coal† is less than 79 per cent and the ratio of $\frac{\text{Per cent Hydrogen} \times 100}{\text{Per cent Oxygen}}$ is 59 or more, with but one or two exceptions the coals make coke by the ordinary commercial process. While coal with a H:O ratio of 59 or over is *almost certain* to give a good coke, and while coke of good quality *may usually* be expected from coal with a ratio as low as 55, poor results *are usually* obtained with coals having a ratio below 55. The best cokes obtained in ordinary commercial processes are made from coals having a ratio of 60 or over. If the per cent of fixed carbon in the pure coal is more than 79 per cent, coking is possible only if the amount of carbon in the volatile matter is large.

The application of the H:O ratios and the use of the tables contained in White's work seems to offer valuable suggestions concerning the methods of mixing coals to obtain coke of good quality. However, the use of the H:O ratio does not correctly classify all varieties of coal as to their coking quality.

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* The term "net hydrogen" is discussed in detail in a later chapter. It may be defined here as the portion of the hydrogen in the coal above that equivalent to the oxygen content, *i.e.*, net hydrogen = total hydrogen − $\frac{1}{8}$ oxygen.

† The coal substance, according to Grout,[28] is the fixed carbon plus the volatile matter, and therefore the fixed carbon in the pure coal substance is:

$$\frac{(\text{Fixed carbon})(100)}{(\text{Fixed Carbon}) + (\text{Volatile Matter})}.$$

BIBLIOGRAPHY

1. E. S. Moore: "Coal," John Wiley & Sons (1922).
2. Stevenson: *Proc.* Am. Phil. Soc., **50,** 1 (1911).
3. C. A. Davis: "Origin and Formation of Peat," U. S. Bur. Mines, *Bull.* 38 (1913).
4. R. Thiessen: "Constitution of Coal," Am. Inst. Mining Met. Eng., Preprint (1925).
5. David White: "Origin of Coal," U. S. Bur. Mines, *Bull.* 38 (1913).
6. David White: "Environmental Conditions of Deposition of Coal," Am. Inst. Mining Met. Eng., Preprint (1925).
7. David White: "Progressive Regional Carbonization of Coals," Am. Inst. Mining Met. Eng., Preprint (1925).
8. Illinois Geol. Survey, *Bull.* 3 (1906).
9. G. N. Ashley: "Maximum Rate of Deposition of Coal," *Econ. Geol.*, **2,** 34 (1907).
10. S. W. Parr: "Fuel, Gas, Water and Lubricants," McGraw-Hill Book Co., Inc. (1922).
11. W. A. Bone: "Coal and Its Scientific Uses," Longmans, Green & Co. (1918).
12. J. Am. Chem. Soc., **21,** 1129 (1899).
13. S. W. Parr: "Distribution of Ash in the Coal Seam," Paper read before Milwaukee Meeting Am. Chem. Soc. (1923).
14. Selvig and Fieldner: "Fusibility of Ash from Coals of the United States," U. S. Bur. Mines, *Bull.* 209.
15. *Trans.* Am. Inst. Mining Met. Eng., **63,** 913 (1920).
16. Stopes and Wheeler: "Constitution of Coal," Dept. Sci. Ind. Research (Brit.) (1918).
17. S. R. Illingworth: "Researches on the Constitution of Coal," *Colliery Guardian* (1923).
18. Fieldner and Davis: "Modern Views of the Chemistry of Coals of Different Ranks as Conglomerates," Am. Inst. Mining Met. Eng., Preprint (1925).
19. *Fuel in Science and Practice,* **2,** 30 (1923); see also a series of articles by Stopes and Wheeler, **3** (1924).
20. Fischer and Schrader: *Brennstoff-Chem.,* **3,** 65–72 (1922).
21. Fischer *et al., Ges. Abhandl. Kenntnis Kohle,* **1–4** (1917 to 1922); see also paper by Fischer Ind. Eng. Chem. **17,** 707 (1925);
22. *J.* Am. Chem. Soc., **97,** 1917 (1910); **99,** 649 (1911); **105,** 131 (1914); and Clark and Wheeler, **103,** 1704 (1913).
23. U. S. Bur. Mines, *Tech. Paper* 140; also *Proc.* Am. Gas Inst., **9,** 234 (1914).
24. U. S. Bur. Mines, *Bull.* 117 (1920); etc.; see also reference 3.
25. Paper read before Milwaukee Meeting, Am. Chem. Soc., 1923.
26. *Fuel in Science and Practice,* **1,** 97 (1922).
27. David White: U. S. Bur. Mines, *Bull.* 29 (1911).
28. *Can. Geol.,* **2,** 240 (1907).

CHAPTER III

TYPES OF COAL AND THEIR CLASSIFICATION

The methods of classifying coal are discussed in the latter part of this chapter. Before considering these classifications, however, it is necessary to describe briefly the varieties of coal recognized scientifically and commercially. It should be borne in mind that these different varieties cannot be differentiated sharply, since the properties of one class gradually merge into the properties of another without rigid dividing lines. Two coals may differ considerably in composition and still be of the same class.

TYPES OF COAL*

Peat.—Since peat is the first stage in coal formation, its description has been included, though it is not a variety of coal. Peat is an accumulation of compacted and partly devolatilized vegetable matter containing relatively high percentages of oxygen and water. It varies from a woody material to a black, jelly-like substance and in a few cases contains so much mineral matter that it will not burn freely. It always contains a large amount of water, and to make a good fuel it usually has to be compressed and dried. The water content may run from about 60 to 90 per cent, though in most peats it is 80 to 90 per cent. The water content may be reduced to 50 to 60 per cent by pressure and to 30 to 35 per cent by air drying. The following table gives a moisture- and ash-free analysis of a typical peat:

TABLE XI

ANALYSIS OF A TYPICAL PEAT
(Ash- and moisture-free basis.)

	PER CENT
Volatile matter	70.06
Fixed carbon	29 94
Sulfur	1.71
Hydrogen	5.44
Carbon (total)	56.37
Nitrogen	2.95
Oxygen	33.53
Heating value	9610 B.t.u. per lb.

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* A more extended discussion of the various types of coal is given by MOORE[1] and CAMPBELL.[23] The coal characteristics deemed most important from the point of view of combustion are discussed in this chapter.

The sulfur content is usually below 1 per cent. Some peats contain such a high percentage of nitrogen that their value is enhanced considerably.

Lignite.—In America the words "lignite" and "brown coal" are used interchangeably, since both are of the amorphous woody type and their chemical properties are nearly alike. The term "subbituminous," as stated below, has been used by the U. S. Geological Survey to define the glossy "black lignite," while the term "lignite" has been reserved for the lignitic brown coals which are of distinct woody or clay-like appearance. These coals usually contain from 30 to 40 per cent of water as they come from the mine, and on losing this water through weathering they split up into slabs, disintegrating more readily and completely than subbituminous coal Owing to their high water content, lignites are usually burned near the mine The following table gives the ash- and moisture-free analysis of a typical Montana lignite

TABLE XII

ANALYSIS OF A TYPICAL LIGNITE

(Ash- and moisture-free basis)

	PER CENT
Volatile matter	52 91
Fixed carbon	47 09
Sulfur	1 53
Hydrogen	5 10
Carbon (total)	70 54
Nitrogen	1 53
Oxygen	21 50
Heating value	12,160 B t u per lb.

The moisture averages from about 10 to 40 per cent and the ash from 5 to 12 per cent. Lignite as mined usually has a heating value of from 5500 to 8000 B.t.u. On a moisture- and ash-free basis this corresponds to 10,000 to 13,000 B.t.u per lb.

Subbituminous.—This term has been adopted by the U. S. Geological Survey for the "black lignite" coals between bituminous coal and brown woody lignite. These coals are usually glossy black in color and can be distinguished from the lignites by their apparent absence of woody structure. According to Campbell,[2] they may be distinguished from the bituminous coals by the loss of moisture and the consequent "slacking" or crumbling parallel to the bedding that they undergo when weathered. Due to the loss of combined water, subbituminous coals check and crack irregularly on weathering as compared with the bituminous which splits along definite cleavage planes. On account of the pitch-like appearance, black lignite is known as Pechkohle in Germany. Subbituminous coals vary greatly in both appearance and chemical composition, some of them being banded like the bituminous coals and others having the appearance of the cannel types. The analysis[4] of a typical Wyoming subbituminous coal, calculated to an ash- and moisture-free basis, is given below:

TABLE XIII

Analysis of a Typical Wyoming Subbituminous Coal

(Ash- and moisture-free basis.)

	Per Cent
Volatile matter	40.51
Fixed carbon	59.49
Sulfur	0.45
Hydrogen	4.81
Carbon (total)	76.61
Nitrogen	1.10
Oxygen	17.03
Heating value	13,030 B.t.u. per lb.

Moisture usually averages from 7 to 25 per cent and the ash from 3 to 30 per cent.

Bituminous.—The main distinguishing feature of bituminous coals is their higher content of volatile matter as compared with the semi-bituminous coals. On this account they burn more freely with a long yellowish flame, giving off a heavy yellow smoke of characteristic bituminous odor. This coal is always laminated in structure, having numerous dull and bright bands occurring in varying proportions. Bituminous coal is the main fuel for industrial purposes, being used under boilers for steam raising, in the manufacture of coal gas and producer gas, and for direct heating of furnaces and kilns.

Bituminous coals are designated as caking or non-caking, depending on their action when heated. If on heating the coal softens or melts and solidifies into a more or less solid cake, leaving a hard, coherent mass on further heating, the coal is called a caking or coking coal. If, on the other hand, the bituminous coal does not soften seriously on heating and if the resultant residue is friable and non-coherent, the coal is called a non-caking or non-coking coal.

The composition of a typical bituminous coal (Connellsville, Pa.) is shown in the following table calculated to an ash- and moisture-free basis:

TABLE XIV

ANALYSIS OF A TYPICAL BITUMINOUS COAL (CONNELLSVILLE, PA.)

(Ash- and moisture-free basis.)

	PER CENT
Volatile matter	30.20
Fixed carbon	69.80
Sulfur	1.06
Hydrogen	5.94
Carbon (total)	87.10
Nitrogen	1.37
Oxygen	5.12
Heating value	15,560 B.t.u. per lb.

The moisture usually ranges from 2 to 10 per cent and ash from 2 to 12 per cent.

Cannel coal (a variety of non-coking, bituminous coal) has a particularly high percentage of volatile matter. It burns with a bright flame, and was originally called "candle coal," because splinters of the best grades will ignite easily when touched with a lighted match. The richness of cannel coal is due to the fact that it is composed for the greater part of spores, spore cases, and seed coats of plants, and the resinous or waxy portion of the vegetal matter. On account of its richness it was once used extensively for the distillation of oil. Owing to the absence of woody material and its regular structure, it breaks like glass with a shell-like fracture. The following is an analysis[4] of a Kentucky cannel coal which is characteristic of the class.

TABLE XV

ANALYSIS OF A KENTUCKY CANNEL COAL

(As received basis.)

	PER CENT
Moisture	2.36
Volatile matter	48.40
Fixed carbon	38.75
Ash	10.49
Sulfur	1.20
Hydrogen	6.47
Carbon (total)	71.98
Nitrogen	1.16
Oxygen	8.70
Heating value	13,770 B.t.u. per lb.

Semi-bituminous.—The term "semi-bituminous" is an unfortunate choice of words, since to many it implies that this variety of coal is of a lower grade than bituminous, while as a matter of fact it denotes a coal of *higher* rank—really, a *super-bituminous* coal. The term was devised to cover the "dry," low-volatile coals as opposed to the "fat" or high-volatile bituminous coals. Semi-bituminous coal contains a fairly high percentage of fixed carbon, which makes it practically a "smokeless coal" and in many sections of the country it is frequently sold as a domestic fuel. The best coals of this variety have a higher heating value than any other type of coal and are, in consequence, in great demand for steam-raising purposes. The analysis of a typical high-grade semi-bituminous coal from the Virginian Appalachian region, calculated to a moisture-and ash-free basis, is given below:

TABLE XVI

ANALYSIS OF A TYPICAL HIGH-GRADE SEMI-BITUMINOUS COAL
(Ash- and moisture-free basis.)

	PER CENT
Volatile matter	16.80
Fixed carbon	83.20
Sulfur	0.79
Hydrogen	4.75
Carbon (total)	90.31
Nitrogen	1.17
Oxygen	2.98
Heating value	15,880 B.t.u. per lb.

The water content usually varies from 2 to 6 per cent and the ash from 4 to 11 per cent.

Semi-anthracite.—Semi-anthracite, although a hard coal, is not so hard as anthracite. The name was adopted to cover the coals between the hard anthracites and the soft semi-bituminous coals. Semi-anthracite is not so high in fixed carbon as anthracite, and, owing to a higher percentage of volatile matter, it kindles more quickly and gives off a slightly yellow flame when ignited. It is not found in very large quantities and is usually sold as anthracite for domestic purposes. The following table gives an analysis, on a moisture- and ash-free basis, of a typical semi-anthracite coal from the Pennsylvania region (Sullivan County).

TABLE XVII

Analysis of a Typical Semi-anthracite Coal

(Ash- and moisture-free basis.)

	Per Cent
Volatile matter	9.92
Fixed carbon	90 08
Sulfur	0.77
Hydrogen	3.72
Carbon (total)	91.60
Nitrogen	1.27
Oxygen	2.64
Heating value	15,410 B.t.u. per lb.

The moisture content may vary from about 2 per cent to 8 per cent, and ash from 4.3 to 14.5 per cent.

Anthracite.—In America this coal is sometimes called "hard coal" and in Wales "stone coal." It burns with a short blue flame, gives off but little odor, and does not soften or cake on burning. These properties, together with its high apparent (bulk) density (55 lb. per cu. ft.), make anthracite an exceptionally good domestic fuel. The heating value of anthracite is not so high as some of the best semi-bituminous or bituminous coals. Except in the smaller sizes unsuited for domestic use, anthracite is not used for steam raising or general manufacturing unless a non-smoking or low-sulfur-content coal is required. The following table gives an analysis on a moisture- and ash-free basis of a typical anthracite coal from the Pennsylvania region (Luzerne County).

TABLE XVIII

Analysis of a Typical Anthracite (Pennsylvania Region)

(Ash- and moisture-free basis.)

	Per Cent
Volatile matter	6.18
Fixed carbon	93.82
Sulfur	0.62
Hydrogen	2.68
Carbon (total)	94.01
Nitrogen	0.99
Oxygen	1.74
Heating value	14,070 B.t.u. per lb.

The moisture content usually varies from 2.2 to 4 per cent and ash from 1.5 to 10 per cent (and higher in the small sizes used as an industrial fuel).

Comparison of the Different Coals.—The following table* and Fig. 10 show the relative composition of wood, peat, and coals, the coals being calculated to an ash- and moisture-free basis.

TABLES XIX

Composition of Wood, Peat, and Coals

(Ash- and moisture-free basis.)

	Proximate analysis, per cent		Ultimate analysis, per cent					
	Volatile matter	Fixed carbon	Sulfur	Hydrogen	Total Carbon	Nitrogen	Oxygen	Heating value (B.t.u. per lb.), moisture- and ash-free
Wood	......	......	.....	6.25	49.50	1.10	43.15	5,800
Peat	70.06	29.94	1.71	5.44	56.37	2.95	33.53	9,610
Lignite	60.67	39.33	1.89	4.74	72.79	0.98	19.60	12,172
Subbituminous	38.12	61.88	0.50	4.74	76.28	1.47	17.01	12,938
Bituminous	30.26	69.74	1.06	5.39	87.00	1.37	5.18	15,527
Semi-bituminous	16.12	83.88	2.52	4.37	89.07	1.40	2.64	15,683
Semi-anthracite	9.95	90.05	0.74	3.76	92.15	1.18	2.17	15,457
Anthracite	1.29	98.71	1.00	1.77	94.39	0.71	2.13	14,882

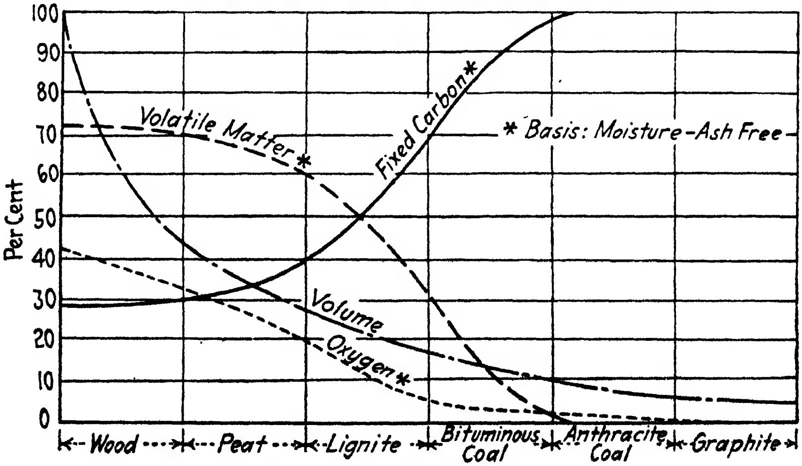


Fig. 10.—Graphic representation of the transformation of fuels from wood to anthracite. (*Adapted from "The Making, Shaping and Treating of Steel," Carnegie Steel Company* (1920).)

The gradual increase in fixed-carbon content and the simultaneous decrease in the amount of volatile matter are strikingly brought out by the figure.

* Adapted from Moore.[1]

As previously pointed out, one of the most interesting features in the progression from wood to anthracite is the gradual decrease in oxygen content. As shown in the above table and in Fig. 10, the oxygen content of wood is about 40 per cent, falling off to little over 2 per cent in the case of the anthracite. In order to show the wide difference in the analyses of coals "as received" and "calculated to an ash- and moisture-free basis," the following table is given, the analyses being for the same coals as in the preceding table.

TABLE XX

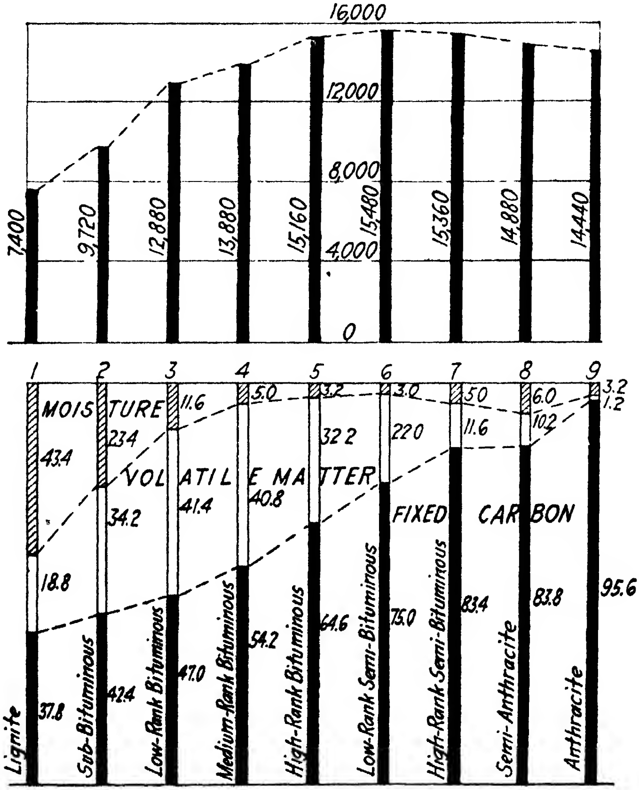
COMPARISON OF WOOD, PEAT, AND COALS
(Analyses "as received." *Data from E. S. Moore*[1])

Kind of fuel	Per cent									Heating value, B.t.u per lb.
	Moisture	Volatile matter	Fixed carbon	Ash	Sulfur	Hydrogen	Total Carbon	Nitrogen	Oxygen	
Wood	.....	.....	.....	.....	.....	6.25	49.50	1.10	43.15	5,800
Peat	56.70	26.14	11.17	5.99	0.64	8.33	21.03	1.10	62.91	3,586
Lignite	34.55	35.34	22.91	7.20	1.10	6.60	42.40	0.57	42.13	7,090
Subbituminous	24.28	27.63	44.84	3.25	0.36	6.14	55.28	1.07	33.90	9,376
Bituminous	3.24	27.13	62.52	7.11	0.95	5.24	78.00	1.28	7.47	13,919
Semi-bituminous	2.03	14.47	75.31	8.19	2.26	4.14	79.97	1.26	4.18	14,081
Semi-anthracite	3.38	8.47	76.65	11.50	0.63	3.58	78.43	1.00	4.86	13,156
Anthracite	2.80	1.16	88.21	7.83	0.89	1.89	84.36	0.63	4.40	13,298

A more striking method of comparison is that adopted by Campbell.[3] He represents graphically the proximate chemical composition of certain coals as shown in the accompanying diagram (Fig. 11), the figures being taken from actual analyses, and calculated over to an ash-free basis.

The coals represented are: (1) North Dakota lignite; (2) Wyoming subbituminous; (3) bituminous coal from Indiana; (4) bituminous coal from Ohio; (5) bituminous coal from Pittsburgh; (6) and (7) semi-bituminous coals from the Windber district, Pennsylvania; (8) semi-anthracite; (9) Pennsylvania anthracite.

The diagram shows an increase in fixed carbon in the ascending scale from lignite to anthracite, and that the moisture content, after decreasing rapidly from lignite to medium-grade bituminous, is relatively small in the case of the better coals. It should be noted that the volatile matter is a maximum in the intermediate ranges, and that the heating value of anthracite is *less* than some of the lower-rank coals. As pointed out by Campbell, there is a tendency for people to believe that, on account of its high price, anthracite should have the highest heating value. Of course, anthracite has the highest percentage of fixed carbon, but the heating value of coal is chiefly determined by the percentage of carbon plus hydrogen. Since the heating value of carbon is only 14,544 B.t.u. per pound and that of hydrogen is 62,000 B.t.u. per pound, it can readily be seen that some of the lower-rank



Upper Diagram: Comparative heat value of same coal represented in lower diagram, ash-free basis.
Lower Diagram: Variation in fixed carbon, volatile matter, and moisture of coals, from lignite to anthracite, as received, ash-free basis.

FIG. 11.—Comparison of the heating values and compositions of various coals.

coals with their relatively high percentages of net hydrogen* can have higher heating values than the anthracites.

The distinguishing chemical and physical characteristics of various ranks of coal have been summarized by Fieldner and Davis[5] as follows:

TABLE XXI

PHYSICAL AND CHEMICAL CHARACTERISTICS OF COALS

(After Fieldner and Davis.)

Chemical characteristics			Physical characteristics
Coal	Approximate per cent normal mine moisture	Ratio of fixed carbon to volatile matter	
Lignite	30–45	.	Distinctly brown; either markedly clay-like or woody in appearance. Falls into pieces on exposure to weather
Subbituminous	18–30	.....	Black; no distinct woody texture; disintegrates and loses moisture on exposure to weather, but less rapidly than lignite
Bituminous	3–15	3	But slightly affected by exposure to weather
Semi-bituminous	3–6	3–7	But slightly affected by exposure to weather
Semi-anthracite	3–6	6–10	But slightly affected by exposure to weather
Anthracite	2–3	10–60	But slightly affected by exposure to weather

THE CLASSIFICATION OF COALS

From a knowledge of the method of formation of coal, it is fairly obvious that any attempt at classification must be more or less arbitrary and that boundary lines cannot be rigid, since the

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* The term "net hydrogen" has been defined as the portion of the hydrogen in the coal above that equivalent to the oxygen content. The oxygen with the equivalent amount of hydrogen to form water is, of course, valueless as regards the production of heat, while the hydrogen in excess of this amount (net hydrogen) yields 62,000 B.t.u. per pound on being burned.

various types shade off into each other so gradually. There are, however, a number of names, such as anthracite, bituminous, and lignite, which have been used for a long time in the coal trade to designate certain varieties of coal. These terms are much too broad, however, and with the rapid expansion of the coal industry a number of attempts have been made to define the varieties more closely for practical and scientific purposes. Some of the more important of the proposed methods of classification* are given below.

Frazer's Classification.—Among the earliest attempts to classify coals on anything like a scientific basis was that of Frazer,[6] utilizing the ratio of fixed carbon to volatile matter as obtained from the proximate analysis. He attempted to differentiate between the various trade varieties and found that the fields overlapped considerably. He finally evolved a classification for the following ranks of coal, clearly recognizing that it was decidedly arbitrary:

TABLE XXII

FRAZER'S CLASSIFICATION OF COAL

	$\text{Ratio} = \frac{\text{Fixed Carbon}}{\text{Volatile Matter}}$
Anthracite	100–12
Semi-anthracite	12– 8
Semi-bituminous	8– 5
Bituminous	5– 0

It was soon pointed out that, while Frazer's classification served quite well for the immense coal deposits of the eastern United States, it would not serve to classify the great coal fields of the West. At the time this attempt at classification was made, the "younger" coals of the western part of the country were not of so much importance, but with our national expansion their use began to increase, thus making Frazer's classification of little general value.

Ralston's Classification.—A classification based on the percentages of carbon, hydrogen, and oxygen in ash-, moisture-, sulfur-, and nitrogen-free coal has been evolved by O. C. Ralston.[7] The percentages of the three elements (carbon, hydrogen, oxygen) are plotted on trilinear coordinates, the coal being assumed to be only carbon, hydrogen, and oxygen. Thousands of analyses were plotted by Ralston, and the general trend of the curve is

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* Several methods not discussed in this book are described by MOORE.[1]

shown in Fig. 12, the zones occupied by the different coals being shown rather than the actual analyses.[8] The small triangle represents the complete diagram, with each of the coordinates running to 100. Since all the fuels fall in the lower left-hand corner (shaded), this portion of the diagram is enlarged in the lower figure. To illustrate the method of reading the plot, the junction of the lignites and peats is represented by the point 30 per cent oxygen, 64 per cent carbon, and 6 per cent hydrogen. The diagram shows in a rather striking manner the rapid decrease in oxygen from wood and the lignites through to anthracite.

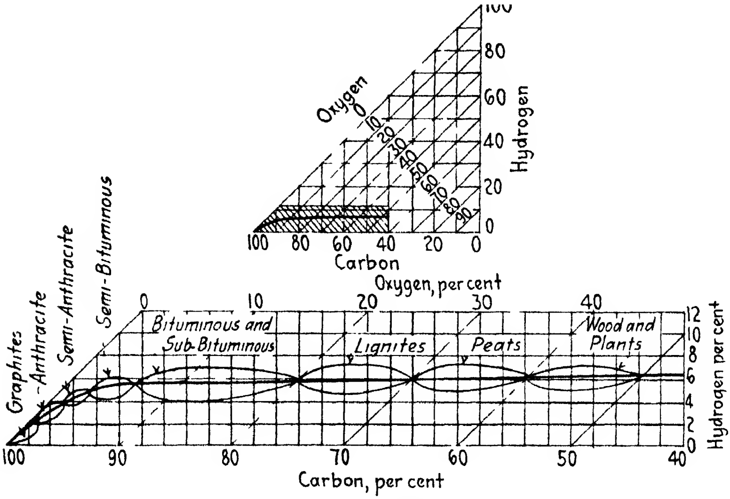

FIG 12 —Ralston's classification of coals.

Campbell's Classification.—After critically examining the analyses of a large number of coals for the purpose of deriving a workable system of classification, Campbell[9] came to the following conclusions: (1) Fuel ratio, $\frac{\text{Per cent Fixed Carbon}}{\text{Per cent Volatile Matter}}$, is not a satisfactory method of separation except for anthracite, semi-anthracite, and semi-bituminous coals. (2) The calorific value is an unsatisfactory method, especially in the higher or better ranks of coal. (3) The percentage of fixed carbon, while fair, fails in detail, and is thus unsatisfactory. (4) Although the percentage of hydrogen is in itself valueless as a means of classification, the carbon-hydrogen ratio is satisfactory except for cannel, some anthracite and semi-anthracite coals. Campbell[2] has also advocated what might be termed a descriptive method of classification of the coals below bituminous. The essential points of this classification are the appearance of the coal and its behavior on weathering.

On weathering, bituminous coals break up into prisms along their regular cleavage planes, while subbituminous coals, on weathering, check and break irregularly into smaller pieces and also separate into flat plates (the face of the plates having been originally parallel to the floor of the fuel seam). Separation of subbituminous coal from lignite is made on the basis of appearance, the former being black and shiny while the latter is brown and woody.

White's Classification.—White,[10] in his work on the effect of oxygen in coal, plotted the $\frac{\text{Carbon}}{\text{Oxygen} + \text{Ash}}$ ratios against calorific values (see Fig. 9, Chap. II) and found that the points fell on a smooth curve, except for (1) weathered coals, (2) those having more than 79 per cent fixed carbon, and (3) the boghead-cannel group. He considered that the H:O ratio is of marked value in distinguishing the coking coals when the per cent of fixed carbon is less than 79 per cent. According to White, coals with a H:O ratio of 0.59 or over are almost certain to coke; coke of good quality may usually be expected from coals with a ratio between 0.55 and 0.59, while poor results are usually obtained when the ratio is lower than 0.55. This and other points in White's work have already been discussed in Chap. II, and, in spite of the importance of oxygen content, this method does not readily serve to classify the types known to the coal industry.

Classification of the Twelfth International Geological Congress. The Twelfth International Geological Congress,[11] in an attempt to determine the coal resources of the world, sent questionnaires to a large number of authorities. The questionnaire classified the coal along the lines shown in the following table:*

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* The classification given here is in abridged form. All the essential points are included, however.

TABLE XXIII

CLASSIFICATION OF THE TWELFTH INTERNATIONAL GEOLOGICAL CONGRESS

Class	Commercial name	Appearance on burning	Per cent volatile combustible	Fuel ratio, or $\frac{\text{Fixed Carbon}}{\text{Volatile Matter}}$	Heating value, B.t.u. per lb.	Mean composition, per cent			Remarks
						C	H_2	$O_2 + N_2$	
A_1	**Anthracite**	Short blue flame	3– 5	12 and over	14,500–15,000	93–95	2.0–4.0	3.0– 5.0	
A_2	**Semi-anthracite**	Slightly luminous short flame, little smoke	7–12	7 –12	15,000–15,500	90–93	4.0–4.5	3.5– 5.0	Does not cake
B_1	**Anthracite and high-carbon bituminous**	Short luminous flame	12–15	4 – 7	15,200–16,000	80–90	4.5–5.0	5.5–12.0	Does not coke readily
B_2	**Bituminous**	Luminous flame	12–26	1.2– 7	14,000–16,000	75–90	4.5–5.5	6.0–15.0	Generally cokes
B_3	**Low-carbon bituminous**	Burns freely with long flame	Up to 35	....	12,000–14,000	70–80	4.5–6.0	18.0–29.0	Withstands weathering, but fractures readily and occasionally has moisture content up to 6 per cent; porous, tender coke; split volatile,* = 2.5 – 3.3
C	Cannel	Long smoky flame	30–40	....	12,000–16,000	. . .	. . . . . .	. .	Very porous coke; fracture generally resinous
D_1†	Lignite or sub-bituminous	....	.....	....	10,000–13,000	60–75	6.0–6.5	20.0–30.0	Moisture in fresh-mined commercial output up to 20 per cent. Fracture generally conchoidal; drying cracks irregular, curved lines; color generally black, occasionally brown. Split volatile* 1.8 to 2.5
D_2	Lignite	....	.....	. ...	7,000–11,000	45–65	6.0–6.8	30.0–45.0	Moisture in commercial output over 20 per cent; fracture generally earthy and dull; drying cracks generally separate along bedding planes and often show woody structure; color brown, sometimes black

$$\text{* Split volatile} = \frac{\text{Fixed Carbon} + \frac{1}{2}\text{ Volatile}}{\text{Hygroscopic Moisture} + \frac{1}{2}\text{ Volatile}}.$$

† D_1 and D_2 generally contain over 6 per cent moisture; disintegrate on drying; streak, brown or yellow; cleavage, indistinct.

The classification as given in the above table serves well in defining and classifying the coals into groups corresponding to the usual trade names.

Regnault-Grüner Classification.—In their method of classification, Regnault and Grüner make use of the ultimate analysis and two determinations from the proximate analysis (per cent volatile matter at 900°C. and per cent fixed carbon). Their classification, which also specifies the chief uses and type of coke, has been revised by Bone[12] as follows:

TABLE XXIV

Classification of Coal
(*After Regnault-Gruner.*)
(All figures refer to ashless, moisture-free coal.)

Genus	Class	Chief uses	Per cent composition			Per cent volatile at 1652°F. (900°C.)	Per cent fixed carbon	Character of carbonaceous residue
			C	H	O+N+S			
A Lignites	Non-caking		60–75	±5.0	20–35	Above 45	Below 55	Non-coherent
B Bituminous	B1 Non-caking, long flame	Reverberatory furnaces	75–80	4.5–5.5	15–20	40–45	55–60	Non-coherent
	B2 Caking, long flame	Gas making	80–85	±5.6	10–15	32–40	60–68	Very porous coke (domestic)
	B3 Hard coking	Coke manufacturing	84–89	5.0–5.6	5.5–11.0	26–32	68–72	Dense coke (by-product)
	B4 Hard coking, short flame	Coke manufacturing and steam raising	88–90	4.5–5.5	5.5–6.5	18–26	74–82	Very dense coke
BC Semi-bituminous	Non-caking, short flame	Steam raising	90–92	4.0–4.5	4.0–5.5	15–20	80–85	Weakly caking or non-coherent
C Anthracitic and anthracites	C1 Anthracitic non-caking	Steam raising	92–94	3.0–4.0	3.0–4.5	8–15	85–92	Pulverulent
	C2 Anthracites non-caking	Domestic and central-heating				Below 8	Above 92	

The following factors should be noted with regard to the progression from lignite to anthracites:

1. The carbon content increases.
2 Hydrogen is about constant until anthracite is reached.
3. Oxygen continually decreases.
4. Volatile matter continually decreases.
5. Fixed carbon increases.

A serious objection to all the preceding methods of classification, except that of Frazer, is that they require at least one of the constituents that can be determined only from the utlimate analysis. The advantages of a workable classification based on the more readily accessible and quicker proximate coal analysis are obvious. A classification of this type was devised by Dowling,[13] who used what he termed the split volatile ratio, *i.e.*,

$$\frac{\text{Fixed Carbon} + \frac{1}{2}\text{ Volatile Combustible}}{\text{Moisture} + \frac{1}{2}\text{ Volatile Combustible}}.$$

This was adopted in part by the classification of the Twelfth International Geological Congress as previously shown. Two other methods which do not require an ultimate analysis are given below.

Ashley's Classification.—G. H. Ashley[14] has proposed what he terms "a use classification of coal," primarily intended for people in the coal business. Ashley uses two ratios, namely, the fuel ratio $\left(i.e., \frac{\text{Fixed Carbon}}{\text{Volatile Matter}}\right)$, and a "fixed carbon–moisture ratio." That is, two ratios are used, and only a proximate analysis on the "as received" basis is required. In following through this method, Ashley has taken the names of the localities from which the coal was mined. The following table given by Ashley shows the application of this method to various coals.

TABLE XXV

Ashley's Classification of Coal

Coal	Fuel ratio, $\frac{\text{Fixed Carbon}}{\text{Volatile Matter}}$	$\frac{\text{Fixed Carbon}}{\text{Moisture}}$ ratio
Anthracite	10+	10+(30+)
Bernice coal	7 -10	10+(27+)
Brushy Mountain, Virginia, coal	5 -7	10+(26+)
Pocahontas coal	3.5 -5	10+(24.5)
Sewell, New River, coal	2.5 -3.5	10+(23)
Connellsville coal	1.85-2.5	10+(21.5)
Pittsburgh coal	1.4 -1.85	10+(19.5)
Beaver River, Pennsylvania, coal	1.4	10+(17)
Gallatin County, Illinois, coal	1.4	6.0 -10.0
St. Clair County, Illinois, coal	1.4	4.0 6.0
Sangamon County, Illinois, coal	1.4	2.5 - 4.0
Grundy County, Illinois, coal	1.4	2.0 - 2.5
Sheridan, Wyoming, coal	1.4	1.7 - 2.0
Carney, Wyoming, coal	1.4	1.4 - 1.7
Gillette, Wyoming, coal	1.4	1.0 - 1.4
Wood County, Texas, lignite	1.4	0.85-1.00
Houston County, Texas, lignite	1.4	0.65-0.85
Williston, North Dakota, lignite	1.4	0.5 -0.65

In the above table it will be seen that the $\frac{\text{Fixed Carbon}}{\text{Moisture}}$ ratio for the lignites falls between 0.5 and 2 and the $\frac{(\text{Fixed Carbon})}{(\text{Moisture})}$ ratio for the principal subbituminous coals falls between 2 and 10 −. The coals above subbituminous are separated on the basis of the fuel ratio, $\frac{(\text{Fixed Carbon})}{(\text{Volatile Matter})}$.

Ashley has introduced several rather interesting features, such as the use of locality names for designating types of coal, making the type name end in *ite*. Thus, Pocahontas coal is called Pocahontite, and in addition to coal from this district of West Virginia it would also mean coal of the same grade from any locality. In view of the hold that the present names (bituminous, anthracite, etc.) have on the coal trade, it is doubtful whether any attempt to change the method of nomenclature radically would be generally adopted.

Parr's Classification.—After having previously suggested a classification[15] which necessitated the use of an ultimate analysis, S. W. Parr has more recently published a method of classification[16] based on the heat value of what he calls "unit coal substance." The determination of the amount of pure coal substance is obtained by correction from the ordinary proximate analysis and consists in allowing for ash, moisture, and sulfur as well as the inert portion of the volatile matter. The heating value of unit coal substance is given by the following equation:*
Heating Value =

$$\frac{\text{Indicated B.t.u. per lb.} - 5000 \times \text{fraction Sulfur}}{1.00 - (1.08 \times \text{fraction Ash} + \frac{22}{40} \times \text{fraction Sulfur})}.$$

On the basis of heat value of unit coal substance, Parr drew up the following table:

TABLE XXVI

Heating Value of Unit Coal Substance
(*After Parr.*)

	Heating Value of Unit Coal Substance, B.t.u. per Lb.
Cellulose and wood	6,500– 7,800
Peat	7,800–11,500
Lignite, brown	11,500–13,000
Lignite, black, or subbituminous coal	13,000–14,000
Bituminous coal (mid-continental field)	14,000–15,000
Bituminous coal (eastern field)	15,000–16,000
Semi-anthracite and semi-bituminous	15,500–16,000
Anthracite	15,000–15,500

Since this table is not completely satisfactory, due particularly to the lack of regular progression in the higher grades of coal, Parr then plotted the heating value of unit coal against the "per cent volatile matter in unit coal"

$$\text{Unit Fixed Carbon} = \frac{\text{Fixed Carbon as determined}}{1.00 - (1.08 \text{ Ash} + \frac{22}{40} \text{ Sulfur})}$$

and Per Cent Volatile Matter in Unit Coal = 1.00 − Unit Fixed Carbon) for a large number of coals and obtained the diagram shown in Fig. 13. The dividing lines were chosen arbitrarily, but, as Parr points out, there is a marked thinning

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* Derivation of the formula is shown in reference 15, Bibliography.

out of the group representatives at the divisions, though this effect is much less marked in the case of the lignites than with the higher-rank coals.

This is the only classification evolved so far that satisfactorily differentiates between the eastern and western United States bituminous coals.

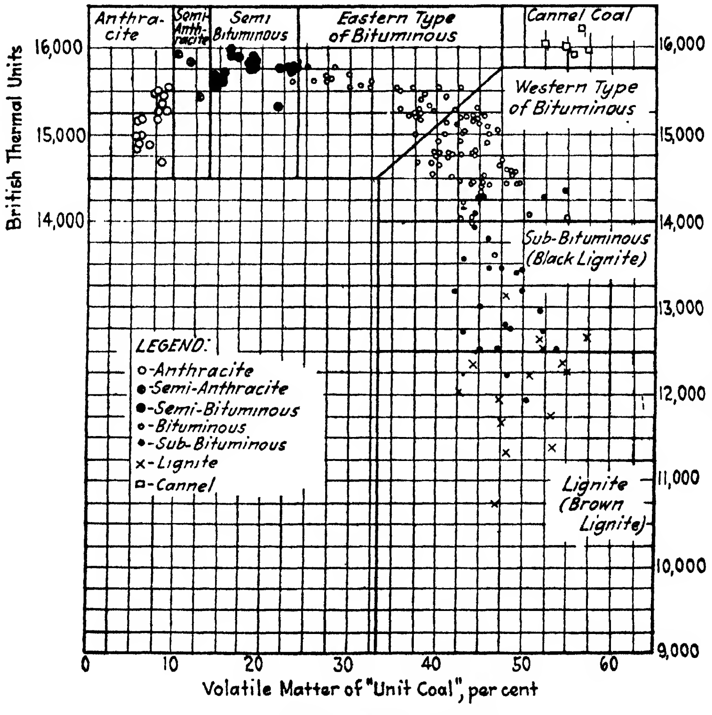

FIG. 13.—Parr's classification of coals.

Summary of Methods of Classification.—The following table shows the basis of the methods of coal classification mentioned, together with a summary of the advantages and disadvantages of each:

TABLE XXVII

SUMMARY OF METHODS OF COAL CLASSIFICATION

Classification	Basis of classification	Advantages and disadvantages
Frazer's	1. Ratio $\frac{\text{Fixed Carbon}}{\text{Volatile Matter}}$	1. Only proximate analysis needed 2. Unsuited for western coals
Ralston's	1 Per cent composition (carbon, hydrogen, and oxygen) on a moisture-, sulfur-, nitrogen-, and ash-free basis	1. Analysis difficult 2. Lines of demarcation not sharp
Campbell's	1. $\frac{\text{Carbon}}{\text{Hydrogen}}$ ratio 2. Appearance of coal 3. Action of coals on weathering	1 Gives an easy method for differentiating the bituminous, subbituminous, and lignite coals from each other 2. Ultimate analysis required
White's	1. Per cent oxygen on moisture- and ash-free basis 2. $\frac{\text{Oxygen}}{\text{Hydrogen}}$ ratio	1 Per cent oxygen hard to determine 2. Gives some idea as to coking property of the coal
Twelfth International Geological Congress	1. Proximate and ultimate analysis	1. Classifies coals into the well-recognized commercial grades
Regnault-Gruner	1. Ultimate and proximate analysis 2. Coke characteristics 3. Chief uses	1. Requires long ultimate analysis 2. Very complete
Ashley's	1. Fuel ratio and $\frac{\text{Fixed Carbon}}{\text{Moisture}}$ 2. Advocates new names for coals based on localities	1. Requires only proximate analysis on the "as received" basis 2. Names proposed of doubtful value
Parr's	1. Per cent volatile matter and heating value of unit coal substance	1. Requires only proximate analysis 2. Differentiates the eastern from western bituminous coals

COAL ANALYSES

The following tables give a series of proximate analyses of typical anthracites and steam-raising coals from the eastern part of the United States:

TABLE XXVIII

PENNSYLVANIA ANTHRACITE[17]

(As sampled by the U. S. Bureau of Mines.)

	Per cent moisture	As received B.t.u. per lb.	Dry, B.t.u. per lb.	Per cent ash	Number of analysis
Broken	3.87	12,782	13,920	11.11	13
Egg	4.60	12,658	13,268	11.30	4
Stove	3.81	12,585	13,084	12.05	19
Nut	4.80	12,365	12,988	12.48	6
Pea	3.85	12,150	12,637	15.25	119
Buckwheat No. 1	4.66	11,449	12,009	18.54	49
Buckwheat No. 2	7.56	11,365	12,295	16.77	18

(Range of Analyses, as Shipped.[12])

	Ash, per cent	Fixed carbon, per cent	Volatile matter, per cent	Moisture, per cent	Sulfur, per cent
Broken	7–10	81–84	4–6	1–2	0.7
Egg	8–11	80–83	4–6	1–2	0.7
Stove	9–12	79–81	4–6	1–2	0.7
Nut	10–13	77–80	4–6	3–4	0.7
Pea	13–15	76–78	4–6	3–4	0.7
Buckwheat No. 1	16–20	71–75	4–6	3–4	0.7
Buckwheat No. 2	18–22	69–73	4–6	3–4	0.7
Buckwheat No. 3	20–25	66–71	4–6	4–6	0.7

TABLE XXVIII.—(*Continued*).

Steam Coals[17]

(Pocahontas and New River Coals.)

	Per cent					B.t.u. per lb.	Ash-fusing temperature, degrees Fahrenheit
	Moisture	Volatile	Fixed carbon	Ash	Sulfur		
1. Pocahontas run of mine ..	Dry	19 07	76.00	4.93	0.61	14,960	2550
2. Pocahontas run of mine....	Dry	18.56	74.80	6.14	0.68	14,751	
3. New River run of mine.....	Dry	20.20	75.96	3.83	0.75	15,074	2600
4. New River run of mine... .	Dry	20.32	74.00	5.58	0.82	14,789	

Note.—All analyses by U. S. Bureau of Mines.
1. Average of 111 samples from 27 mines.
2. Average of 4905 samples from actual shipments.
3. Average of 181 samples from 55 mines.
4. Average of 3206 samples from actual shipments.

(Pennsylvania Steam Coals.)

	Per cent				B.t.u. per lb.
	Volatile	Fixed carbon	Ash	Sulfur	
Best run of mine....	23.25	70 37	6.38	1.42	14,643
Best run of mine ...	15.81	77.62	6.57	1.07	14,787
Second run of mine...	15.98	75.20	8.82	1.35	14,315

Locomotive Steam Coals[17]

(Samples from Shipments Received by Railroads.)

	Per cent					B.t.u. per lb. dry basis	Ash-fusing temperature, degrees Fahrenheit
	Moisture	Volatile	Fixed carbon	Ash	Sulfur		
Pennsylvania, low-volatile.....	0.58	21.28	69.88	8.26	1.28	14,777	2,900
West Virginia, low-volatile.....	1.92	17.95	75.98	6.07	0.75	14,739	3,450
West Virginia, Fairmont.......	Dry	36.84	54.24	8.92	2.06	13,874	2,270
West Virginia, Kanawha.......	1.19	36.55	56.03	6.22	0.66	14,200	2,600
West Virginia, Kanawha.......	1.51	33.55	57.35	7.59	1.19	13,957	2,590
Virginia, southwest Banner seam.......................	Dry	36.23	56.53	7.24	0.57	14,303	2,490
Pennsylvania, Pittsburgh......	1.18	34.88	56.24	7.70	2.16	14,267	2,550
Pennsylvania, Pittsburgh......	1.20	31.59	58.02	9.19	1.44	14,140	2,550

TABLE XXIX

ANALYSIS OF REPRESENTATIVE COALS OF THE UNITED STATES
(All analyses on "as received" basis.)

Name*	Type†	Proximate analysis, per cent				Ultimate analysis, per cent					Heating value, B.t.u. per lb. as received
		Moisture	Volatile matter	Fixed carbon	Ash	Sulfur	Hydrogen	Carbon	Nitrogen	Oxygen	
Alabama:											
Bibb, Belle Ellen	B	3.2	31.0	59.6	6.2	1.2	5.3	78.3	1.4	7.6	14,141
Jefferson, Bessemer	B	2.4	24.4	68.4	4.8	0.7	4.7	81.7	1.5	6.5	14,623
Jefferson, Palos	B	2.10	27.5	59.4	11.0	0.7	5.0	75.4	1.6	6.2	13,419
Shelby, Aldrich	B	2.3	38.6	51.9	7.2	0.8	5.2	75.0	1.0	10.0	13,648
Alaska:											
Moose Creek	B	4.7	35.5	50.9	8.9	0.3	5.3	67.6	1.9	15.9	12,146
Arkansas:											
Hartford, Central No. 10	B	2.9	19.3	67.3	10.5	1.1	4.1	77.4	1.6	5.3	13,271
Huntington, Central No. 6	SeB	3.2	18.1	69.7	9.0	1.9	4.4	78.7	1.6	4.4	13,703
Colorado:											
Gunnison, Somerset	B	4.3	39.7	46.7	9.3	0.4	5.5	70.6	1.5	12.7	12,627
Routt, Mt. Harris	B	11.0	37.7	45.7	5.6	0.4	5.8	65.3	1.6	21.3	11,455
Weld, Erie	Sub B	24.6	29.8	42.0	3.6	0.3	6.3	54.8	1.2	33.8	9,517
Illinois:											
Christian, Pana	B	13.0	37.0	39.3	10.7	3.7	5.6	59.8	1.1	19.1	10,856
Franklin, Christopher	B	9.8	34.2	47.4	8.6	0.8	5.3	66.2	1.5	17.5	11,725
Franklin, Orient	B	7.3	36.7	47.9	8.1	1.0	5.4	69.0	1.6	15.0	12,164
Franklin, Sesser	B	10.1	33.4	48.2	8.4	1.1	5.5	66.8	1.3	17.0	11,817
Perry, Duquoin	B	10.8	33.1	47.4	8.8	0.8	5.4	64.5	1.4	19.2	11,524
Williamson, Herrin	B	9.4	33.0	49.0	8.6	0.9	5.2	67.1	1.5	16.7	11,858
Indiana:											
Greene, Jasonville	B	13.5	36.3	42.9	7.3	1.1	5.8	64.5	1.5	19.8	11,543
Knox, South Bruceville	B	9.5	38.3	41.4	10.9	3.2	5.5	62.3	1.0	17.1	11,542
Sullivan, Dugger	B	12.6	36.0	44.5	6.9	1.5	5.9	65.7	1.5	18.6	11,792
Sullivan, Vandalia	B	14.9	34.3	44.1	6.7	1.3	5.9	63.8	1.4	20.9	11,416
Iowa:											
Lucas, Chariton	....	15.4	30.5	41.5	12.6	3.2	5.7	55.8	1.1	21.5	10,243
Polk, Altoona	....	13.9	37.0	35.2	14.0	6.2	5.5	54.7	0.8	18.8	10,244
Kansas:											
Cherokee, Stone City	B	5.1	34.5	52.2	8.3	3.3	5.2	71.8	1.2	10.2	13,082
Crawford, Edison	B	3.9	34.2	50.1	11.8	4.6	4.9	68.8	1.2	8.7	12,496

Kentucky:											
Christian, Mannington	B	9.2	33.7	46.4	10.7	3.1	...	....	...	....	11,682
Webster, Clay	....	5.4	34.9	50.4	9.3	1.1	5.1	70.4	1.6	12.6	12.501
Maryland:											
Allegany, Frostburg	SeB	2.2	17.2	69.1	10.9	1.1	4.3	76.9	1.9	4.9	13,432
Allegany, Hoffman	SeB	2.4	19.0	71.2	7.4	1.0	4.4	79.8	1.9	5.5	14,087
Allegany, Allegany	B	2.3	22.0	67.9	7.8	4.4	4.7	79.3	1.8	4.9	14,031
Allegany, Lord	SeB	2.2	20.5	70.2	7.1	1.0	4.7	81.2	1.9	4.2	14,231
Allegany, Ocean	SeB	1.2	17.9	73.2	7.7	1.0	4.5	81.0	1.9	4.0	14,186
Montana:											
Carbon, Washoe	Sub B	10.5	34.7	43.7	11.2	1.1	5.6	59.8	1.3	21.0	10,545
Musselshell, Roundup	Sub B	13.6	32.9	45.5	8.1	0.7	5.6	62.0	1.0	22.7	10,687
New Mexico:											
San Juan, Farmington	B	6.9	38.1	43.0	11.9	1.3	...	..	...	...	11,628
North Dakota:											
Ward, Burlington	L	36.9	24.9	27.7	10.4	0.2	6.4	37.4	0.6	45.0	6,010
Williams, Roy	L	42.3	25.0	28.7	3.9	0.4	7.3	37.8	0.7	50.0	6,331
Williams, Wheelock	L	42.1	25.0	24.4	8.5	1.3	7.1	35.2	0.5	47.5	5,994
Ohio:											
Columbian, New Salisbury	B	3.5	36.7	48.9	10.9	4.3	5.2	69.9	1.4	8.3	12,731
Jefferson, Irondale	B	2.2	39.6	47.5	10.7	5.0	5.3	71.0	1.3	6.7	13,025
Jefferson, Yellow Creek	B	3.4	36.3	49.6	10.7	5.1	5.2	69.7	1.4	8.0	12,722
Oklahoma:											
Coal, Lehigh	B	6.6	38.6	42.9	11.9	4.3	5.0	62.8	1.5	14.5	11,264
Haskell, McCurtin	B	3.0	22.8	68.2	6.0	1.0	4.8	80.9	1.8	5.6	14,148
Latimer, Degnan	B	3.7	36.8	53.8	5.6	0.9	...	....	..	....	13,630
Pittsburg, Alderson	B	2.7	38.2	54.2	4.9	0.7	5.4	77.4	1.9	9.7	13,867
Pittsburg, Ridgway	B	3.8	38.0	50.6	7.6	1.7	5.4	73.8	1.8	9.6	13,284
Oregon:											
Coos, Beaverhill	Sub B	16.1	31.1	39.6	13.2	0.8	5.5	51.1	1.2	28.2	9,031
Pennsylvania:											
Allegheny, Creighton	B	2.9	33.8	53.1	10.2	1.1	5.3	73.8	1.5	8.2	13,217
Armstrong, Montgomeryville	C	1.8	32.8	37.3	28.2	3.7	4.5	56.9	1.1	5.6	10,462
Armstrong, W. Kittanning	B	3.4	35.5	51.3	9.7	3.1	5.3	71.4	1.3	9.1	13,037
Beaver, Smith's Ferry	B	4.4	37.2	53.5	4.9	1.4	5.4	75.4	2.0	11.0	13.653
Bedford, Hopewill	B	1.6	16.3	70.0	12.1	1.0	4.1	77.4	1.4	3.4	13,813
Blair, Gen. White	B	2.8	30.0	60.6	6.7	1.9	5.1	78.8	1.4	6.2	14,063
Butler, Goff Station	B	3.2	37.4	45.6	13.8	3.5	4.9	67.5	1.0	9.3	12,294
Cambria, Bakerton	SeB	2.2	22.2	70.3	5.3	1.1	. .	... .	...	....	14,460
Cambria, Beaverdale	SeB	1.7	18.0	71.2	9.2	2.0	..	... .	...	....	13,877
Cambria, Nanty Glo	SeB	2.5	20.0	71.5	6.1	1.9	...	... .	...	....	14,380
Cambria, Somman	SeB	3.3	18.3	72.0	6.4	1.8	4.7	80.7	1.4	5.0	14,171
Cambria, St. Michael	SeB	3.4	15.4	75.5	6.1	1.0	4.4	81.7	1.3	5.4	14,243
Cambria, Windber	SeB	2.6	17.0	73.7	6.9	1.2	4.3	83.9	1.3	2.4	14,621
Clearfield, Munson	B	3.0	22.8	64.6	9.5	3.1	4.7	76.7	1.3	4.7	13,626

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* In each case the first name indicates the county, the second name the mine.
† A = Anthracite. SeA = Semi-anthracite. SeB = Semi-bituminous. B = Bituminous. Sub B = Subbituminous. C = Cannel. L = Lignite.

TABLE XXIX.—(*Continued*).

Name*	Type†	Proximate analysis, per cent*				Ultimate analysis, per cent					Heating value, B.t.u. per lb. as received
		Moisture	Volatile matter	Fixed carbon	Ash	Sulfur	Hydrogen	Carbon	Nitrogen	Oxygen	
Jefferson, Punxsutawney	B	2.6	30.4	59.1	7.9	2.0	5.1	76.6	1.2	7.2	13,856
Somerset, Cairnbrook	SeB	2.8	17.8	67.6	11.9	1.9	...	....	...	....	13,230
Somerset, Seanor	SeB	2.4	17.3	71.4	9.0	2.5	4.3	78.5	1.2	4.5	13,738
Rhode Island:											
Providence, Cranston	A	4.5	3.0	78.7	13.8	0.9	0.5	82.4	0.1	1.8	11,624
Texas:											
Webb, Dolores	C	4.4	46.0	30.5	19.0	2.1	5.8	59.3	1.2	12.7	11,065
Virginia:											
Montgomery, Blacksburg	SeB	1.9	14.0	68.9	15.2	0.5	3.6	75.3	0.9	4.8	12,739
Montgomery, Merrimac	SeA	3.6	9.5	67.5	19.4	0.5	...	....	.	....	11,851
Pulaski, Guntan Park	SeA	3.8	9.4	62.2	24.0	0.8	...	....	.	...	10,955
Pulaski, Parrott	SeB	1.6	13.3	61.5	23.6	0.7	3.3	67.4	0.8	4.3	11,403
Russell, Darte	B	2.4	34.7	55.8	7.2	0.6	5.3	78.2	1.5	7.3	13,928
Tazewell, Pocahontas	SeB	3.0	20.3	72.2	4.5	0.5	4.7	84.0	1.2	5.2	14,605
Wise, Josephine	B	2.6	33.8	53.6	10.0	0.9	5.1	73.7	1.6	8.8	13,270
Washington:											
Kittitas, Ellensburg	B	10.3	30.4	43.4	15.9	1.4	5.5	61.3	1.5	14.4	11,005
Lewis, Centralia	Sub B	22.8	32.9	29.5	7.8	0.6	6.6	43.9	0.8	40.2	7,547
Thurston, Tono	Sub B	21.7	34.8	33.3	10.3	1.2	6.3	49.9	0.9	32.4	8,696
West Virginia:											
Brook, Collier	B	4.4	37.4	50.1	8.1	2.6	5.3	72.1	1.4	10.5	12,940
Fayette, Layland	SeB	2.6	18.8	73.5	5.1	0.8	4.7	82.8	1.6	5.1	14,456
Grant, Bismarck	SeB	3.0	16.7	71.0	9.3	2.7	..	....	..	...	13,590
Hancock, Chester	B	4.2	34.8	54.4	6.7	1.8	5.3	74.2	1.3	10.8	13,302
McDowell, Arlington	SeB	2.3	18.9	74.9	4.0	0.4	4.6	84.9	1.2	5.0	14,756
McDowell, Coalwood	SeB	1.0	17.0	76.1	6.0	0.8	4.3	84.2	1.2	3.6	14,562
McDowell, Jenkin Jones	SeB	2.0	21.2	70.1	6.7	0.5	4.6	82.3	1.0	4.9	14,346
Mercer, Simmons	SeB	2.6	19.1	74.0	4.3	0.6	4.6	84.2	1.1	5.1	14,603
Mineral, Emoryville	SeB	2.4	16.1	65.9	15.6	2.5	...	....	...	...	12,600
Mingo, Thacher	B	2.4	34.4	56.9	6.3	1.0	5.2	77.9	1.5	8.0	13,898
Ohio, Elm Grove	B	3.0	41.4	45.6	8.1	3.6	5.3	72.8	1.4	8.8	13,201
Raleigh, Oswald	SeB	2.7	21.4	72.0	3.9	0.9	4.8	83.4	1.6	5.4	14,625
Wyoming, Alpoca	SeB	2.2	18.7	74.8	4.4	1.3	4.6	83.9	1.4	4.5	14.663
Wyoming:											
Lincoln, Elkol	Sub B	20.8	35.4	40.6	3.2	0.7	6.3	57.8	0.9	31.1	10,084
Lincoln, Green River	B	3.6	38.4	52.5	5.5	1.0	5.3	74.5	1.3	12.5	13,313

* In each case the first name indicates the county, the second name the mine.

† A = Anthracite. SeA = Semi-anthracite. SeB = Semi-bituminous. B = Bituminous. Sub B = Subbituminous. C = Cannel. L = Lignite.

BIBLIOGRAPHY

1. E. S. Moore: "Coal," John Wiley & Sons (1922).
2. M. R. Campbell: "A Practical Classification of Low-grade Coals," *Econ. Geol.*, **3**, 134 (1908).
3. M. R. Campbell: "Coal Fields of the United States," U. S. Geol. Survey, *Prof. Paper* 100-A (1917).
4. "Analyses of Coals in the United States," U. S. Bur. Mines, *Bull.* 22 (1912).
5. Fieldner and Davis: "Modern Views of the Chemistry of Coals of Different Ranks as Conglomerates," *Am. Inst. Mining Met. Eng.*, Preprint (1925).
6. Pa. 2nd Geol. Survey, *Rept.* M. M., 879.
7. O. C. Ralston: "Graphic Studies of Ultimate Analyses of Coal," U. S. Bur. Mines, *Tech. Paper* 93 (1915).
8. Kreisinger, Augustine and Ovitz: "Combustion of Coal and Design of Furnaces," U. S. Bur. Mines, *Bull.* 135 (1917).
9. "Report on the Operation of the Coal-testing Plant," U. S. Geol. Survey, *Prof. Paper* 48, part 1, 156 (1906).
10. D. White: "The Effect of Oxygen in Coal," U. S. Bur. Mines, *Bull.* 29 (1911).
11. "Coal Resources of the World," vol. 1 (1913). Morang & Co.
12. W. A. Bone: "Coal and Its Scientific Uses," Longmans, Green & Co. (1918).
13. D. B. Dowling: "Classifications of Coals by the Split Volatile Ratio," *Can. Mining J.*, **29**, 143 (1908).
14. G. H. Ashley: "A Use Classification of Coal," *Trans.* Am. Inst. Mining Met. Eng., **63**, 782 (1920).
15. S. W. Parr: "The Classification of Coals," *J.* Am. Chem. Soc., **28**, 1425 (1906); *J. Ind. Eng. Chem.* **1**, 636 (1909).
16. S. W. Parr: "The Classification of Coal," *J.* Ind. Eng. Chem., **14**, 921 (1922).
17. *Nat. Coal Mining News*, Dec. 19 (1923).
18. U. S. Bur. Mines, *Bull.* **22** (1912); **85** (1914); **123** (1918); **193** (1922), and *Prof. Paper* 100-A.

CHAPTER IV

SPONTANEOUS COMBUSTION AND STORAGE OF COAL

With exposure to the air all coals, except anthracite, even at ordinary temperatures absorb oxygen and slowly combine with it, evolving heat. Spontaneous combustion of coal is caused by this absorption of oxygen. Even when precautions are taken to dissipate any heat that may be liberated, the oxidation still progresses. The discussion in Chaps. II and III has shown that an increase in oxygen content of coal decreases the heating value and impairs the coking properties. This is true whether the oxygen is present as a natural component left from the wood, or whether it comes from the air on weathering. Since *any* oxidation that takes place previous to burning represents a loss, it can be seen that the question is one of considerable economic importance, affecting the handling of the coal at the mine, during transportation, and during storage.

THE SPONTANEOUS COMBUSTION OF COAL

The Influence of Sulfur.—Under certain conditions coal is subject to spontaneous combustion—that is, will ignite and burn without the application of heat from any external source. Up to about the year 1880 fires of spontaneous origin were ascribed to the heat liberated by the oxidation of pyrites (FeS_2) in the coal, rather than to oxidation of the coal substance itself. According to these early views, enough heat would be liberated by the reaction

$$2FeS_2 + 2H_2O + 7O_2 = 2FeSO_4 + 2H_2SO_4 + 1{,}155{,}000 \text{ B.t.u. per lb.-mol.}$$

to raise the temperature of the coal pile to a point where vigorous reaction between the coal and oxygen would begin, and the whole mass then ignite. This view of the process is wrong, however,[1] since the whole pyrites content of the coal (2 per cent) on oxidizing would only raise the temperature of the pile to about 260°F. (125°C.), even if all the sulfur were oxidized immediately and none of the heat generated were removed by radiation or conduction. This temperature is below the ignition point of coal.

The fact that spontaneous combustion cannot be entirely attributed to pyrites (and that in many cases the pyritic content is *not* an index of the ease with which a coal will spontaneously ignite) has been demonstrated by a number of investigators,[2,3,4,5,6] The opinion of authorities now is that pyrites is merely a supplementary factor in spontaneous combustion. It has been definitely established that the real cause lies in the direct oxidation of the coal substance by the adsorption of atmospheric oxygen.

The oxidation of pyrites, however, gives off heat, thus to some extent raising the temperature of the coal and by this means may increase the speed at which the oxygen and the coal unite. Furthermore, on oxidation the pyrites increases in volume, thereby breaking the coal into small particles. This increases the amount of coal surface exposed, and therefore increases the rate of oxidation of the coal itself, since oxidation is a reaction that takes place on the surface of the coal.

The oxidation of pyrites requires the presence of water, as shown by the reaction given above. Thus, when moisture is present, pyritic oxidation becomes possible and trouble from this cause may ensue. In view of the influence of pyrites on spontaneous combustion, it seems advisable to guard against excessive sulfur and moisture.

Direct Action of Coal and Oxygen.—All coals, except anthracites, even at ordinary temperatures, combine with oxygen, and spontaneous combustion may result from the heat generated by this oxidation. The absorption of oxygen by the coal begins as soon as it is broken out of the mine, and is most rapid with freshly mined coal. A summary[5,7,8,9] of the more important established points follows:

1. The oxidation of coal is continuous over a wide range of time and conditions.

2. The rate of oxidation increases greatly with increasing temperature (see Fig. 14) and with *decreasing* size of coal particles.

3. For a given coal there is a critical temperature below which spontaneous combustion cannot start. The continuance (and initial attainment) of this temperature depends on external conditions. If these conditions are withdrawn, combustion ceases. Above this critical temperature, oxidation *is independent of* external factors and combustion is self-propelling or spontaneous.

4. Weathered coals[7] (under a given set of conditions) have a higher critical temperature and are less liable to ignite spontaneously than green coal.

5. Coals containing a high percentage of textural moisture (*i.e.*, moisture which is retained in the pores of the coal and does not cause the coal to feel wet) lose this moisture on exposure to the air, thus offering the large surface of the pores for oxygen absorption. Coals with high textural moisture,[7] and coals of high oxygen* content[10] have relatively low critical temperatures and are particularly liable to ignite spontaneously.

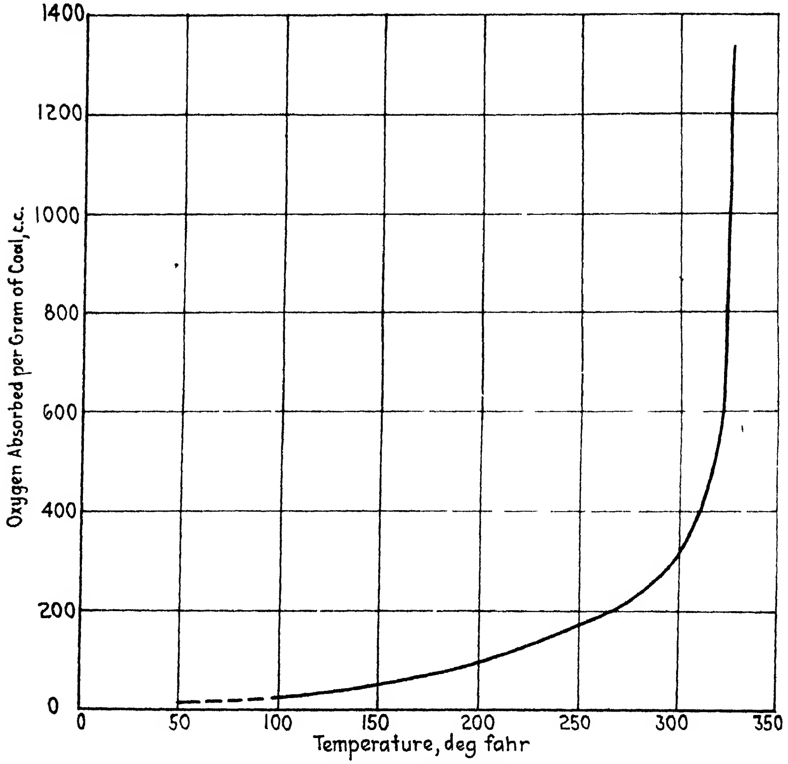

FIG. 14.—Relative rate of oxidation of dry coal at various temperatures. (*From S. W. Parr, in "Industrial Coal," a report of the American Engineering Council, Ronald Press Co.* (1924).)

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* It is interesting to note that coals with high textural moisture contents are usually high in oxygen.

6. Heating the coal to the temperature of autogenous oxidation is brought about by the net effect of several temperature-affecting variables. These variables, any one of which may be dangerous if the heat liberated is not dissipated, are:

(*a*) External sources of heat, such as steam pipes, the sun's rays, etc.

(*b*) Fineness of division of the coal, which speeds up the oxidation by offering a greater amount of active coal surface to the oxygen of the air.

(*c*) The heat generated by the addition of oxygen to unsaturated compounds in the coal.

(*d*) The heat generated by the oxidation of pyrites (FeS_2).

(*e*) Moisture. This is mentioned[5] separately because its importance is apt to be underestimated. It is essential to the oxidation of pyrites and, without exception, in an extensive series of tests by Parr and Kressman, *the wetting of the coal increased the rate at which oxygen was absorbed.*

7. The oxidation and spontaneous combustion of coal takes place in five more or less distinct stages. At first the coal begins to absorb oxygen slowly until a temperature of about 120°F. is reached. The second stage begins at 120°F. and more rapid absorption of oxygen continues until the coal has reached 175 to 280°F. (80 to 140°C.). The temperature depends upon the quality of the coal, on its fineness of division and other factors. The third stage begins at this latter temperature and differs from the second stage in that CO_2 and water vapor are given off as the absorption of oxygen accelerates. This rapid absorption of oxygen and liberation of CO_2 and water vapor continue to a temperature of about 450°F. (230°C.) at which temperature the fourth stage begins. In this stage, the oxidation is spontaneous (*i.e.* does not require heat from an external source) and the liberation of water vapor and CO_2 is much more rapid. The fifth stage, namely the actual combustion or kindling of coal, takes place at about 660°F. (350°C.) The coal then continues to burn vigorously as long as air is supplied.

Tables XXX and XXXI show the kindling temperatures for various types of coals as given by Fayol[3] and Lewes.[4] The temperatures reported by Lewes are considerably higher than those reported by Fayol (probably on account of the fineness of the coal samples used by Fayol).

TABLE XXX

IGNITION TEMPERATURES OF POWDERED FUELS IN HOT AIR

(*After Fayol.*)

Fuel	Ignition temperature	
	Degrees Fahrenheit	Degrees Centigrade
Lignite	302	150
Gas coal	392	200
Hard coking coal	482	250
Anthracite	572 or above	300 or above

TABLE XXXI

KINDLING TEMPERATURES OF SOLID FUELS

(*After Lewes.*)

Fuel	Approximate ignition temperature	
	Degrees Fahrenheit	Degrees Centigrade
Cannels	698	370
Lignites	842	450
Steam coals	878	470

Summary of Information on Spontaneous Combustion.—In reviewing the information collected in the various investigations of spontaneous combustion, it seems to be fairly definitely established:

1. That the oxidation of the coal itself is the real cause of spontaneous combustion and that this oxidation is dependent on the fineness of division of the coal; *i.e.*, the fines and more dusty coals take fire more readily than lump or nut coal.

2. That this oxidation is more rapid and more serious for freshly mined coals than for coals which have been weathered.

3. That the temperature of the coal pile is of great importance in determining the rate of this oxidation.

4. That although the oxidation of pyrites (FeS_2) is not a major factor in all cases, it often exerts a very definite accelerating

action, both from heat liberated and from the increased surface area due to splitting of the coal.

5. Moisture is an aid to pyritic oxidation and it has an additional accelerating effect in causing spontaneous combustion of the coal itself.

EFFECTS OF WEATHERING OTHER THAN ON SPONTANEOUS COMBUSTION

The effects of weathering on the properties of coal (other than effects on spontaneous combustion) can be most readily discussed under the following headings: (1) appearance, (2) size, (3) loss in heating value, and (4) firing and coking qualities of weathered coal.

1. Appearance of Weathered Coal.—Some coals on exposure become covered with a frost-like surface film of white iron sulfate. According to Parr this is not a sign of deterioration, however, since only a portion of the sulfur has been oxidized to the white sulfate, and this is dissolved and washed away by rain water. It can be seen that the ash content will be lowered slightly by this process, but it is only a surface effect, and the major portion of the coal is unaffected.

2. Change in Size on Weathering.—On being stored, coal is generally broken up to a certain extent, and, as already shown, fineness of division has a distinct bearing on the question of spontaneous combustion. This breaking is brought about in two ways: (*a*) by a physical and chemical change due to weathering, and (*b*) by mechanical means through handling. Bituminous coals, on prolonged weathering, break up into small pieces, splitting along the cleavage planes. Subbituminous coals and the lignites (see Chap. III) check irregularly, breaking up into thin plates. This effect extends much further with lignites and subbituminous coals (over an equal weathering period) than with the higher-rank coals. The effects of reduction in size, whether from weathering or mechanical shock during handling, have already been discussed under Spontaneous Combustion, and are further mentioned under Firing Qualities below.

3. Loss in Heating Value.—In view of the possible economic importance of losses in heating value on weathering, a number of tests have been run[11,12,13] to determine the magnitude of this effect in practice. The results of these tests may be summarized as follows:

(*a*) There is no loss in heat value in the case of anthracites.

(*b*) All the coals except anthracite, when freshly mined, absorb oxygen and suffer a decrease in heating value per pound, the rate of deterioration, however, decreasing rapidly with time (see[12] Fig. 15). The bituminous coals may lose from 1 to 5 per cent in heating value on weathering. The coals of low oxygen content, such as those from West Virginia and Pennsylvania, lose from 1 to 3 per cent in five years, the greater part of this loss occurring shortly after mining. With increasing oxygen content of the coal, the decrease in heating value becomes greater, and with the Illinois bituminous coals the losses may amount to from 4 to 6 per cent over a five-year period of exposure. In almost any case, if the temperature is not allowed to rise above

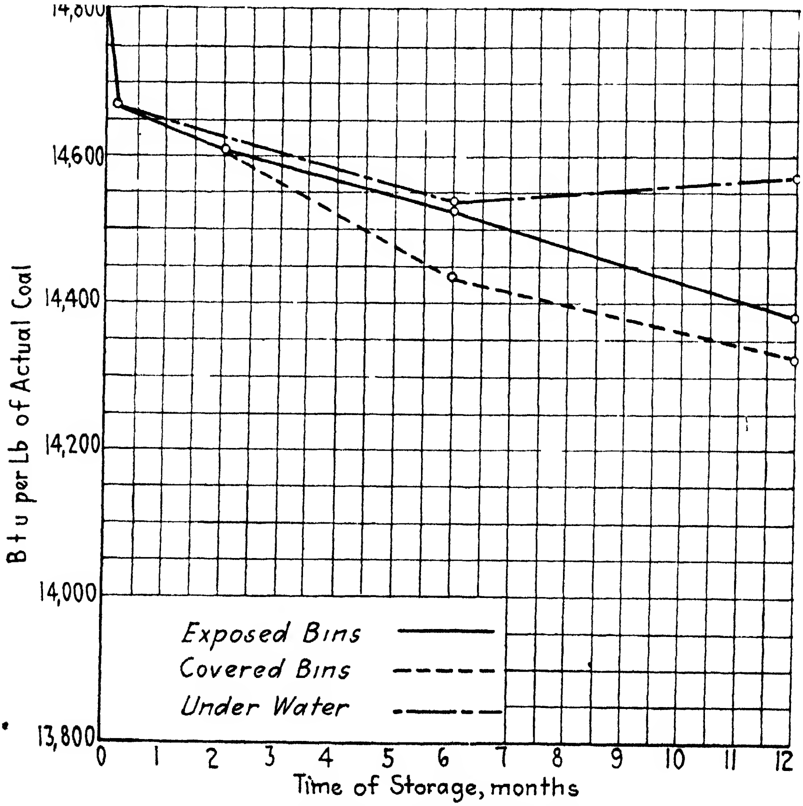

Fig. 15.—Decrease in heating value of Illinois coal on weathering.

140°F. (60°C.), bituminous coals suffer relatively small depreciation in heating value.

(*c*) The heating value of subbituminous coals and lignites is usually lowered on weathering more than that of bituminous coals.

(*d*) A distinction must be drawn between the "loss of heating value per pound of coal" and the "loss per pile." That is, the oxygen absorption which goes on during ordinary weathering (below 120°F. (50°C.)) causes an *increase* in the weight of the pile, only small amounts of the coal substance itself being lost during the process. The total amount of potential heat in the coal pile is, therefore, relatively unaffected, although the heating value of a pound of the weathered coal may be distinctly less than that of the freshly mined coal (*i.e.*, the absorbed oxygen acts as a diluent).

If, however, the coal is allowed to reach 150°F. (65°C.), oxygen absorption, of course, becomes more rapid, and although the weight of the coal pile continues to increase, there is a considerable evolution of water and CO_2. If the coal is allowed to remain at this temperature for even a few days, there is a marked decrease in heating value both per pound and per pile, the firing qualities are affected, and there is a serious impairment of the coking value of the coal.

(*e*) The storage of bituminous coal under water is not to be recommended for the sake of preventing loss of heating value. The advantage of submergence lies in the absolute prevention of spontaneous combustion and preserving the firing qualities.

4. Firing and Coking Qualities of Weathered Coal.—The opinion is prevalent among a great many practical men that weathered coal is inferior in its firing quality. This is due to a considerable extent to the slacking and breaking that takes place, the resulting fine coal requiring a greater draft than is necessary with an unweathered coal. Although it is possible,[14] by firing in a thinner fuel bed and by maintaining the necessary higher draft, to burn weathered coal with the same efficiencies as fresh coal, the weathered coal fire requires greater attention. As pointed out by Beagle,[15] the high-oxygen, middle west coals do not ignite so readily after weathering, and under ordinary firing conditions* it is not uncommon to find a 10 per cent decrease in actual plant efficiency due to the inferior firing qualities of these coals after weathering. With the higher-grade

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* In this case on chain-grate stokers.

coals from the eastern part of the United States this condition is far less serious.

The *coking qualities* of many coals are seriously affected by weathering and only a small amount of exposure may cause the resultant coke to have a distinctly weakened structure. It seems probable that the bonding material of the coke (present only in small amounts) is oxidized and thus but little weathering is required to give a poor coke. From six to eight months' weathering may be very serious as regards coking quality, and it is practically impossible to obtain a good coke from coal which has been stored in the open for over two years.

Summary of Weathering Effects Other Than Spontaneous Combustion.—In conclusion, it may be said that the deterioration, from weathering, in size, heating value, and coking qualities *increases* with the original oxygen content of the coal. There is no loss in the heating value of anthracite and very little with the low-oxygen bituminous coals. The losses increase, however, with the subbituminous coals and lignites. Losses in heating value and coking characteristics become serious if the coal temperature is allowed to reach 150°F. (65°C.) over any considerable period. It is more difficult to reach the same efficiency with weathered western coals than with unweathered or freshly mined coal. It is unnecessary to store high-grade bituminous coals under water to prevent loss in heating value from weathering; the chief advantage of this type of storage is the absolute prevention of spontaneous combustion.

PRECAUTIONS IN STORING COAL

It is clearly recognized that the storage of coal by both producer and consumer is to the best economic advantage of all concerned. The major reasons for storing coal have been outlined by H. H. Stoek[13] somewhat as follows: (1) to assure the consumer of an adequate supply of fuel at all times, thus protecting him against strikes and uncertain railroad deliveries; (2) to take advantage of low freight rates, or of low prices for coal at certain seasons of the year; (3) to permit the railroads* to utilize

* Industry has lost far more through transportation delays resulting from the overlapping of the coal movement with the late fall and early winter shipment of other commodities than it has through fluctuation of coal prices themselves.[9]

their cars and equipment to the best advantage; and (4) to maintain a uniform rate of production* at the mines.

For devising any system of storage there are four main rules, based on the causes for spontaneous combustion as previously outlined. These are:

1. Keep down the temperature and dissipate any heat that may be formed.
2. Prevent oxygen from getting at the coal, or, if this is impossible, allow such free access of air that it will carry away any heat generated.
3. Keep the amount of surface exposed as small as possible.
4. Avoid the possible catalytic effect of moisture.

A number of specific precautions to be observed in storing coal, based on the four general rules given above, are as follows:[9,13,16,17,19]

1. Pile so that lumps and the fine coal are distributed as evenly as possible. Do not allow lumps to roll to the bottom, forming air passages. It is a good plan to build the coal pile in 2- to 3-ft. layers, leveling with scrapers after each layer has been deposited in order to prevent segregation of coarse and fine coal.

2. If possible, store only screened nut coal. In any event, keep out dust as much as possible and reduce handling to a minimum.

3. Provided air is either excluded or allowed circulation free enough to dissipate any heat generated, coal may be piled, theoretically, to any desired height without danger. In practice, however, to minimize danger due to careless piling, it is advisable to pile the coal to moderate heights only (12 to 15 ft.). By observing this precaution it becomes possible to move quickly and to consume any section of the pile that shows a tendency to heat.

4. Any temperature rise is dangerous. Therefore:

(*a*) Do not pile coal in very hot weather,

(*b*) Do not store near any external sources of heat, such as steam pipes, etc.

(*c*) Inspect the pile regularly. A temperature of 150°F. (65°C.) represents the danger point for many coals.

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* Seasonal demand is responsible for 47 per cent of the idle time of the coal industry, and is responsible to a large extent for the overdevelopment of mine capacity; coal production capacity in the United States is now twice as great as consumption capacity.[9]

5. Since green coal oxidizes rapidly, allow at least six weeks' seasoning before final storing.

6. Avoid air leakage to the pile through interstices around foreign objects (such as posts, brickwork, etc.), or through porous bottoms, such as coarse cinders. The practice of ventilating with pipes may cause local overheating and often does more harm than good.

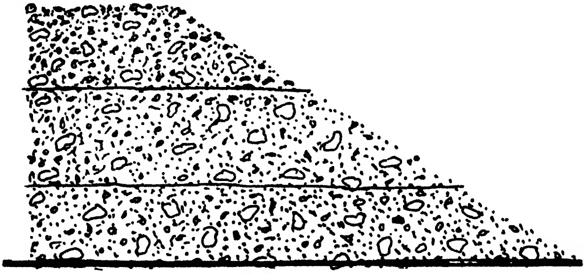

Fig. 16.—Correct method of stacking coal. Lumps and fine coal well mixed and leveled off in 2-ft. layers.

Fig. 17.—Correct method of dumping coal. Lumps and fine coal evenly mixed.

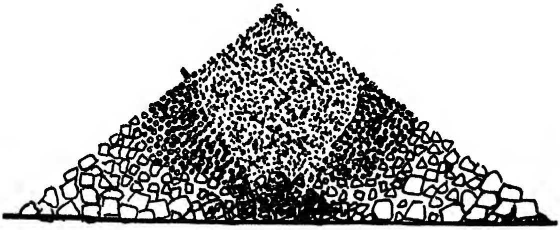

Fig. 18.—Incorrect method of stacking coal. Lumps and fine coal allowed to segregate. Figures 16, 17 and 18 are (*from J. B. C. Kershaw, "Fuel Economy and Smoke Prevention," courtesy of Constable & Company, Ltd., London.*)

7. Avoid alternate wetting and drying of the coal, as this may speed up pyritic oxidation.

METHODS OF STORING COAL

By far the greater proportion of coal is stored on land, with access to the air. In this type of storage it is evident that there

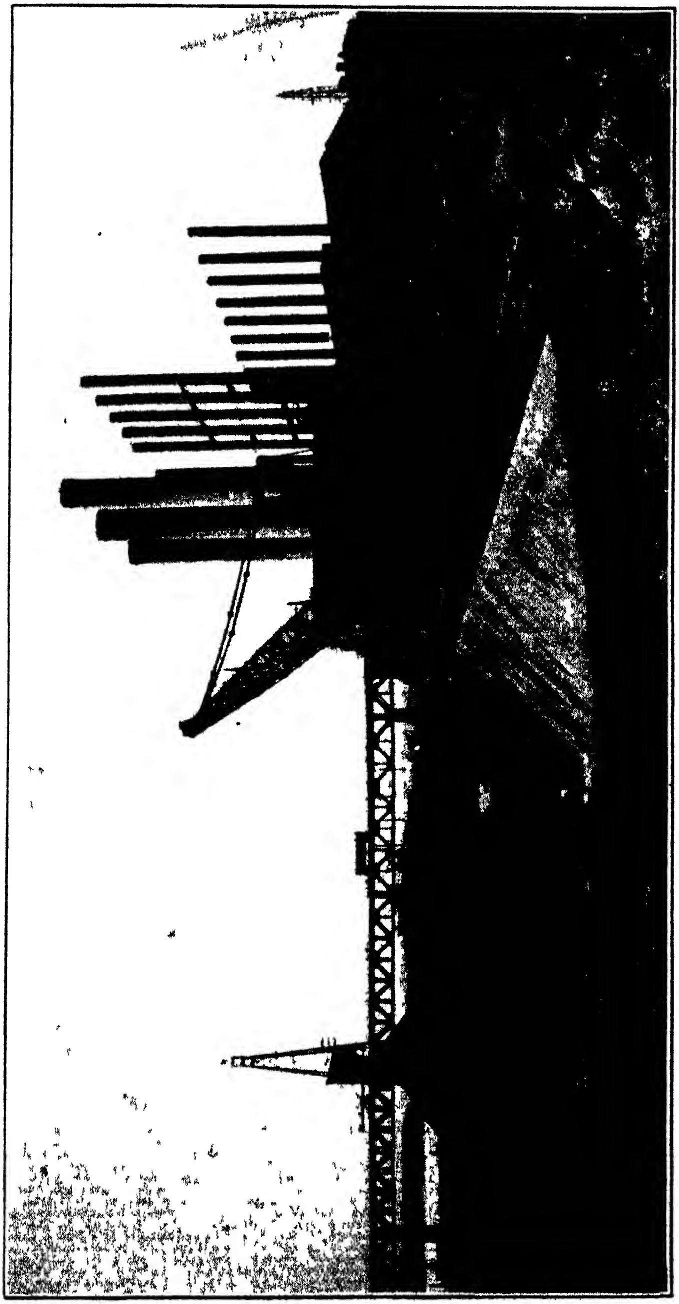

FIG. 19.—Concrete basin for storage of coal under water. (*Courtesy of the Duquesne Light Company.*)

should be no segregation of coarse and fine material, as this may cause unequal ventilation and local overheating. Diagrams of correct and incorrect methods of storing[18] are shown in Figs. 16, 17, and 18. In open storage of this kind it is highly important either to level off each layer with a scraper, or, if the coal is put on the pile from a clam shell, to lower the clam shell to the surface of the pile before discharging the contents

In recent years a number of installations have featured either underwater storage or storage on land with more or less complete exclusion of air. On removal from an underwater storage pit, coal does not retain enough moisture to affect seriously its firing or other qualities. The submerged-storage system of the Duquesne Light Company of Pittsburgh is shown in Fig. 19. This basin has a capacity of 100,000 tons when the coal is completely submerged and 150,000 tons when it is piled up.

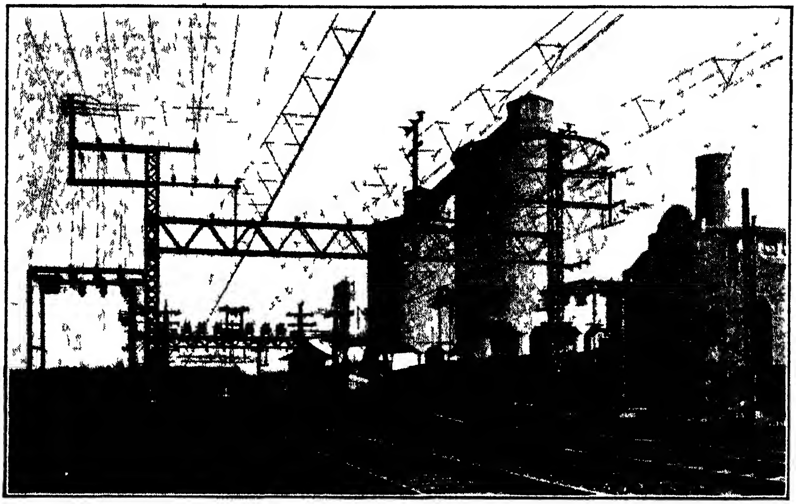

FIG. 20.—Coal-storage silos of the Cos Cob power station. (*Courtesy of the New York, New Haven & Hartford R. R*)

An interesting type of storage that seems to be gaining in favor is a tall concrete bin or silo, such as those of the Cos Cob Power Station of the New York, New Haven and Hartford R. R. The Cos Cob installation is shown in Fig. 20. These silos are about 40 ft. in diameter and about 50 ft. high, the coal being raised to the top by conveyors The bins are roofed and air is fairly well excluded, so that danger from spontaneous combustion is at a minimum. The silos are equipped, however, with pipes for the introduction of water or steam to smother any fires that might occur.

BIBLIOGRAPHY

1. W. A. BONE: "Coal and Its Scientific Uses," Longmans, Green & Co. (1918).
2. RICHTER: "*Dinglers polytechn. J.*, **190,** 368; **193,** 34; **194,** 264; **195,** 315, 449.
3. FAYOL: "*Bull.* soc. l'industr. min. (1879); see also THRELFALL: *J.* Soc. Chem. Ind., **28,** 763 (1909).
4. LEWES: *J. Gas Lighting,* **55,** 645.
5. PARR and KRESSMAN: Univ. Illinois Eng. Expt. Sta., *Bull.* 46 (1910).
6. *Rept.,* First New South Wales Commission.
7. PARR and COONS: *Ind. Eng. Chem.,* **17,** 118 (1925).
8. PARR: "Spontaneous Combustion of Coal in Storage," *Ind. Eng. Chem.,* **17,** 120 (1925).
9. PARR: "Chemical Factors of Deterioration of and Heating of Coal in Storage," in "Industrial Coal Purchase, Delivery and Storage," *Rept.* of Am. Engineering Council, Ronald Press Company (1924).
10. STOPES and WHEELER: *Fuel in Science and Practice,* **2,** 29, 83, 122 (1923).
11. PORTER and OVITZ: "Deterioration in the Heating Value of Coal during Storage," U. S. Bur. Mines, *Bull.* 136 (1917).
12. S. W. PARR: "Effects of Storage upon the Properties of Coal," Univ. Illinois Eng. Expt. Sta., *Bull.* 97 (1917).
13. H. H. STOEK: "The Storage of Bituminous Coal," Univ. Illinois Eng. Expt. Sta., *Circ.* 6 (1918).
14. A. P. KRATZ: "A Study of Boiler Losses," Univ. Illinois Eng. Expt. Sta., *Bull.* 79 (1915).
15. BEAGLE: "The Deterioration of Mid-west Coals in Storage," *Ind. Eng. Chem.,* **17,** 123 (1925).
16. U. S. Bur. Mines, *Tech. Paper* 16 (1912).
17. "The Coal Catalog," 1925 ed., pp. 1188, *et seq.,* Keystone Publishing Co.
18. KERSHAW: "Coal Storage and the Avoidance of Spontaneous Combustion Dangers," *Fuel in Science and Practice,* **3,** 115 (1924).
19. "Storage of Coal and Spontaneous Combustion," Round Table Discussion, *Ind. Eng. Chem.,* **17,** 115–130 (1925).

CHAPTER V.

COAL RESOURCES AND COAL PRODUCTION

Of the world's total available supply of solid fuel, which is estimated at about 8,150,000 million short tons, only about 60 per cent is composed of the high-grade coals. The supply of anthracite is small (less than 7 per cent of the total) and the seams easiest to mine will soon be exhausted. The remainder of the solid fuels, the bituminous coals, and the low-grade subbituminous coals and lignites, make up more than 90 per cent of the total. At the present time the more valuable bituminous coals are mined far more than the subbituminous and lignitic coals, but with the exhaustion of the best of our bituminous seams the low-grade fuels in the future will become exceedingly important.

COAL RESOURCES

Resources by Continents.—The coal resources* of the world by continents, according to the most complete and reliable estimates[1] available, are shown in Fig. 21.

The main features to be noted in the figure are the large coal reserves in America (69 per cent of the world's available supply) and the anthracite in Asia (practically all in China). Although it is probable that some of the coal in China has been classified as anthracite, when in reality it is semi-anthracite or semi-bituminous, the anthracite resources of that country nevertheless surpass the reserves of this type of coal in all other countries combined.

Resources of America.—The distribution of coal† in America[1] is shown in Fig. 22. The figure shows the large amount of valu-

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* Unless otherwise noted the figures in diagrams of coal resources are estimates of all seams 1 ft. and over in thickness, less than 4000 ft. below the surface; seams 2 ft. and over in thickness, between 4000 and 6000 ft. deep, are also included.

† Owing to the meager data obtainable on the amounts of the different varieties of coal which go to make up South America's 35,285 million tons, only the total estimate is given, which shows up in the grand total, but for obvious reasons does not appear in the total estimated resources of anthracite, bituminous, etc.

able bituminous and subbituminous coal in the United States (52 per cent of the world's supply and 75 per cent of the resources in America) and the fairly large reserves of these two coals in

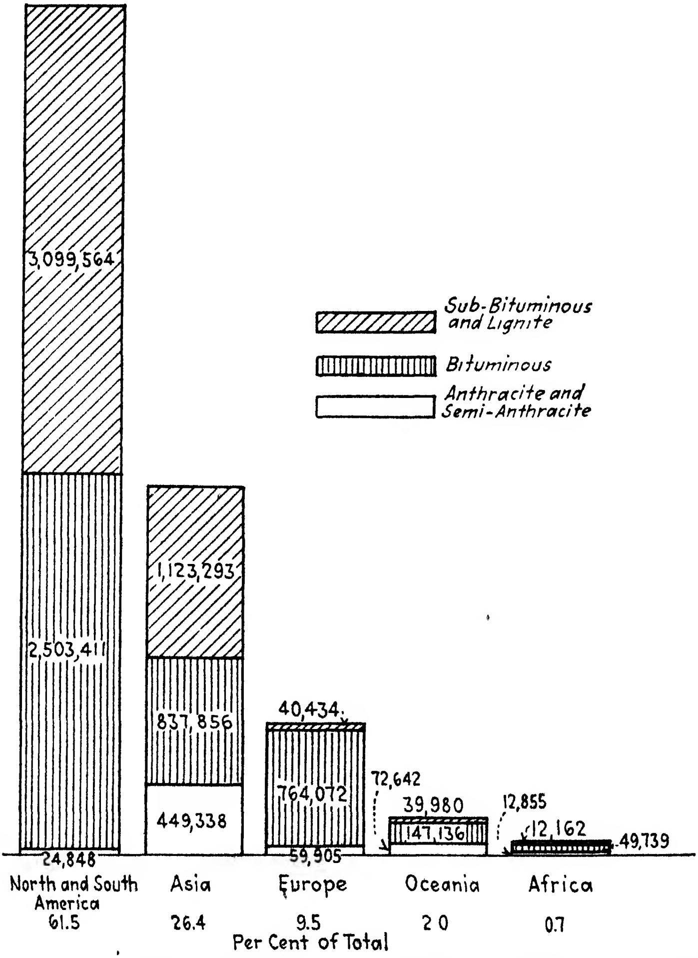

FIG. 21.—Coal resources of the world, by continents. Figures are in millions of net tons.

Canada. Mexico is thought to have considerable coal reserves, but the actual amounts are largely a matter of conjecture, since very little field work has been done. It is remarkable that South America has such small known supplies of coal, the estimated total, including all the coals from lignite to anthracites, being only 0.6 per cent of the world's supply.

Resources of the United States.—For convenience in describing coal areas, the U. S. Geological Survey[2] recognizes four classes, which, in ascending order of magnitude, are known as district, field, region, and province. A coal district (usually named from the leading town) is generally small and the term is applied only to developed coal areas. A coal field is next larger than a "district," but still signifies a fairly compact area. Examples are the New River field and Pocahontas field of West Virginia, and the Windber field in Pennsylvania. Coal fields are grouped

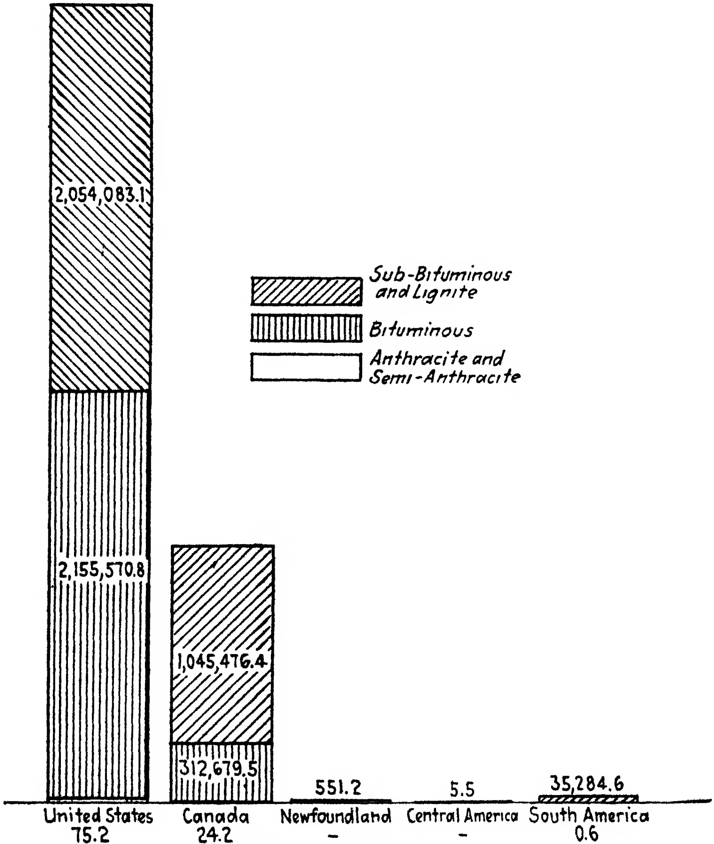

Fig. 22.—Coal resources of America. Figures are in millions of net tons.

together into the larger "coal regions," as, for example, in the anthracite region in Pennsylvania. The largest division in classifying coal areas is called a "coal province." Six of these coal provinces have been recognized in the United States, and their principal characteristics are given in the following paragraphs.

1. *The Eastern Province** contains probably nine-tenths of the high-rank coal of the country and includes (*a*) the anthracite regions of Pennsylvania and Rhode Island; (*b*) the Atlantic Coast region, which takes in the fields of Virginia and North Carolina; and (*c*) the Appalachian region, which extends from northern Pennsylvania into Alabama, and includes parts of Ohio, Maryland, Virginia, West Virginia, Kentucky, Tennessee, and Georgia.

The Rhode Island and Massachusetts anthracite region, although mined intermittently, thus far has been of little commercial importance, since the coals were subjected to such high thrust pressures (this high pressure also creating a somewhat higher than normal temperature) during coal formation (see Chap. II) that many of them were transformed into graphite. The coal is usually of high ash content, much of it in excess of 25 per cent. This ash is quite uniformly and intimately distributed through the coal, and therefore cannot be washed out by crushing, sizing, and hydraulic classification. Due to the excessive movements of the earth during formation, this coal is largely broken and appears on the market in small sizes only.

The Pennsylvania anthracite region in the eastern part of the state produces over 99 per cent of the anthracite mined in the United States and contains 91 per cent of America's anthracite supply. This region contains most of the finest available anthracite in the world.

The Appalachian region contains a large proportion of the high-rank coals of the United States, and the development of this coal supply has made possible the concentration of the

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* For both proximate and ultimate analyses of coals from mines not referred to in the table at the end of Chap. III, see reference 3 in Bibliography. In addition to giving proximate and partial ultimate analyses of practically all the seams mined in the United States, the "Coal Catalog"[4] gives (1) a list of dealers supplying the different coals, (2) a list of seams producing coal suitable for each industrial use, and (3) maps showing the various mining districts of the individual states.

steel industry and other manufacturing enterprises in the eastern part of the United States. The map,[5] Fig. 23, clearly shows the distribution of coal in this and the other coal provinces and indicates the contours of the Appalachian region.

In general, the coal is semi-bituminous or bituminous, considerable variation being found even in the same bed. The percentage of fixed carbon usually decreases from east to west, away from the point of greatest thrust pressure, and, as pointed out in Chap. II (Fig. 5), the semi-bituminous coals in the eastern part gradually change to bituminous toward the western margin of the region. In West Virginia are found some of the highest-grade coals of the world (New River and Pocahontas fields). These coals have a very high heating value, and the low content of volatile matter (about 19 to 21 per cent) makes them extremely satisfactory for hand-fired and Scotch marine types of boilers and, in general, wherever steam is to be raised with the production of relatively small quantities of smoke. These coals are used extensively as a domestic fuel.

North of these fields, in eastern Pennsylvania (Clearfield, Cambria, and Somerset counties), lie other low-volatile, high-grade steaming coals, while immediately to the west (Jefferson, Indiana, and Westmoreland counties) lie large beds of medium- (28 to 35 per cent) volatile coals. The ash from the low-volatile eastern Pennsylvania coals has a higher fusing point than the low-volatile West Virginia coals, an important point if the coal is to be used in underfeed types of stokers (see Chap. XIV). In between the West Virginia and eastern Pennsylvania fields of low-volatile coals lies the famous Georges Creek field (Allegany and Carrett counties, Maryland, and the northern part of Mineral, Grant, and Tucker counties, West Virginia). These coals have 14 to 20 per cent volatile matter and a low percentage of sulfur and ash. The ash in coals from this district has a very high fusion point (class A, fusion point above 2600°F.).

West Virginia (Kanawha district), Pennsylvania (Irwin gas coal district, Westmoreland County), and Ohio furnish good high-volatile coals suitable for the manufacture of city gas.

Very satisfactory gas-producer coals are found in Ohio and eastern Kentucky.

In western Pennsylvania, around Pittsburgh, are found the very highest-grade coking coals, suitable for producing hard, dense, blast-furnace coke from either beehive or by-product

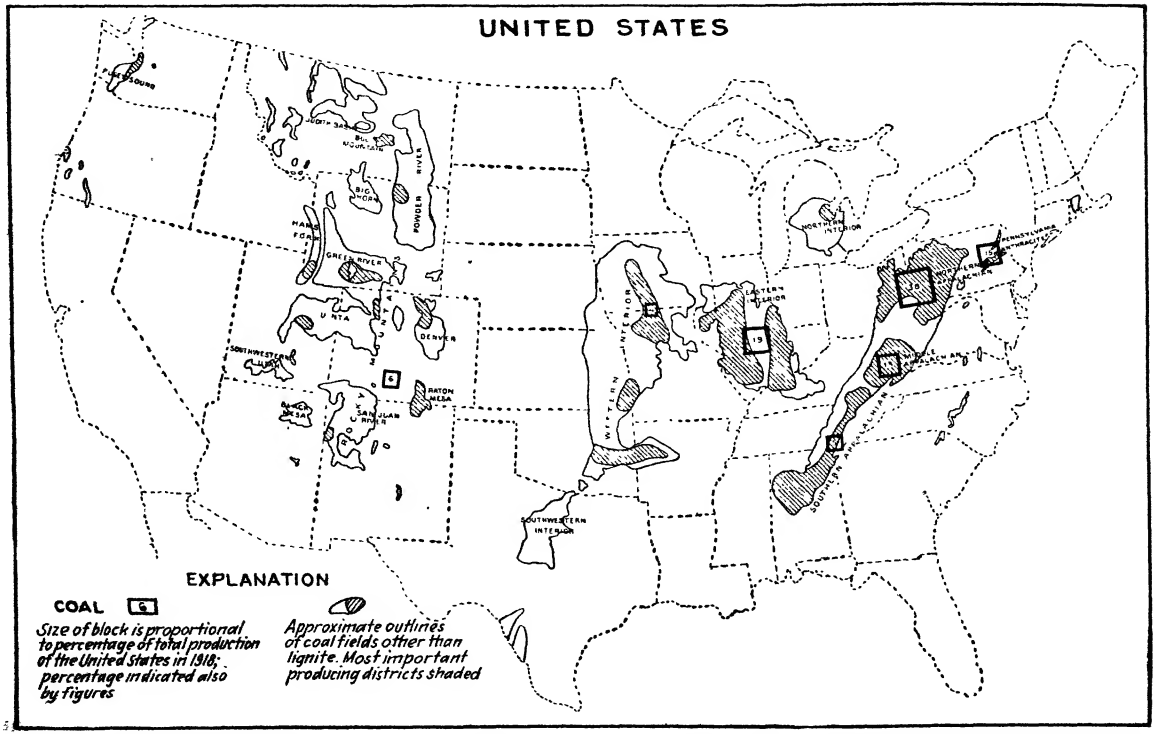

Fig. 23.—Outlines of coal fields in the United States and coal production in 1918. (*From United States Geological Survey, "World Atlas of Commercial Geology," plate* 8, (1921).)

coke ovens. The Connellsville district, famous for being the birthplace of the coke industry, lies to the southeast of Pittsburgh, Pa.

Proximate and ultimate analyses of representative coals from the principal fields of all the provinces are shown in Table XXIX at the end of Chap. III.

2. *The Interior Province* includes: (*a*) the Northern region, containing the Michigan coal fields; (*b*) the Eastern region, containing the Illinois, Indiana, and western Kentucky fields; (*c*) the Western region, containing the coal fields in Iowa, Missouri, Nebraska, Kansas, Arkansas, and Oklahoma; and (*d*) the Southwestern region of Texas.

In general, the coals of this province were not subjected to very high thrust pressure and in consequence they are ordinarily of lower fixed-carbon and higher-volatile matter than those in the Eastern Province. In the Western region, however, in Arkansas, there are areas where mountains rose, and the pressure thus developed transformed the coals to semi-anthracites. This province contains considerable quantities of bituminous coals characterized by their high moisture content, and, in spite of their inferiority to the high-grade coals in the Appalachian region, the development of these fuel resources has done much to make possible the industries centered at Chicago, St. Louis, Kansas City, etc. A suitable grade of blast-furnace coke may be made by mixing Illinois coal with low-volatile Appalachian coals and by the use of specially designed coke ovens.

In general, the coals in this field contain a high percentage of sulfur, are non-coking, and burn with a very smoky flame.

The main contours of the province are shown in Fig. 23.

3. *The Gulf Province* includes (*a*) the Mississippian region, which embraces the coal fields of Louisiana, Mississippi, and Alabama; and (*b*) the Southwestern region, which takes in certain of the Arkansas and Texas fields.

The coals in this province are mainly lignites and have been mined only to a slight extent. Just as in the Interior Province, there were no big disturbances of the earth crust in the Gulf Province, and the relatively low thrust pressures developed gave only the low-grade coals.

4. *The Northern Great Plains Province* includes the coal fields in the Great Plains east of the Front Range of the Rocky Mountains. These are (*a*) the Fort Union region, with the north-

eastern Wyoming field and the lignite fields in the Dakotas and eastern Wyoming; (*b*) the Black Hills region in Wyoming; (*c*) the Judith Basin and the Assiniboine regions of Montana; (*d*) the Denver region in Colorado; and (*e*) the Raton Mesa region of Colorado and New Mexico.

In the eastern part of the Northern Great Plains Province, the coals are mostly lignites. On approaching the mountains in the western part, the earth crust disturbances were somewhat greater, and some of the resultant coals are subbituminous or even higher grade. In North and South Dakota and in northeastern Montana the coals are practically all lignites with high water contents (40 to 50 per cent), but to the south, around Sheridan, Wyo., a fair grade of subbituminous coal (20 to 25 per cent moisture) is mined in considerable quantities. The Raton Mesa region contains the highest-grade coal in the province, and coke made from it is extensively used in the metallurgical industries located in the Rocky Mountain Province.

5. *The Rocky Mountain Province* includes (*a*) the Yellowstone region of Montana; (*b*) the Big Horn, Hams Fork, and Green River regions in Wyoming; (*c*) the Uinta region of Utah and Colorado; (*d*) the San Juan River region of Colorado and New Mexico; and (*e*) the southwestern Utah region.

This province contains coal of all grades from anthracite to lignite, depending mainly on the degree of thrust pressure developed during coal formation. The major part of the resources are subbituminous and bituminous, and these coals are quite extensively mined. The main centers of production are Rock Springs and Kemmerer, Wyo.; Red Lodge, Mont.; Crested Butte and Durango, Colo.; Castlegate and Sunnyside, Utah; and Gallup, N. M.

6. *The Pacific Coast Province* includes the coal fields in California, Oregon, and Washington. The main reserves are in Washington, though small deposits of low-grade fuels are found in both Oregon and California. In Washington, at the points of greatest earth crust disturbance, bituminous and even anthracite coals have been formed. Considerable mining has been done in King County, not only of the high-grade coals near the mountains, but also in certain subbituminous beds which are valuable on account of their nearness to tidewater. Bituminous and higher-grade coals have been mined in Pierce and Kittitas counties. Most of the coal is used locally.

The following figure* (Fig. 24), adapted from a table prepared by Campbell† for the Twelfth International Geological Congress, shows the estimated coal resources in the six provinces.

Coal Resources of Alaska.—Very little field work has been done on the Alaskan coal deposits and only a rough estimate can be given as to the amounts of coal available. The total reserves have been estimated to be about 21,600 million short tons. The coals are mainly lignites and subbituminous, although about 2000 million short tons of anthracite and semi-anthracite are believed to be located in the Pacific Coast region.

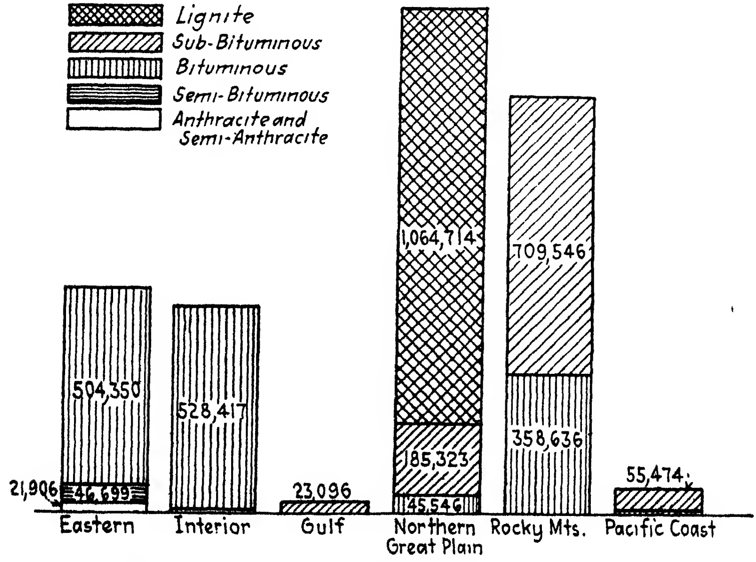

FIG. 24.—Coal resources of the United States. Figures are in millions of net tons.

Coal Resources of Canada.—Practically all the Canadian coal deposits are located in the extreme eastern and extreme western parts of the Dominion. The central portion of the country has very little coal, so that the manufacturing enterprises located in the provinces of Ontario and Quebec are forced to depend on coal imported from the United States. Nova Scotia in the east and Alberta and British Columbia in the west

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* The estimate includes all seams more than 14 in. thick, less than 3000 ft. below the surface.

† The reader is referred to Bibliography, references 1 and 6, for a more complete and detailed exposition of the coal resources of the United States.

have quite large deposits of high-grade coal, and considerable quantities of subbituminous coal and lignite are located in Alberta and Saskatchewan. The estimated coal resources* of the provinces are given in Fig. 25, which is adapted from that prepared by Dowling for the Twelfth International Geological Congress.

Coal Resources of Mexico, Central and South America.—The coal resources of Mexico have not been developed to any great extent and but little is known of them, except in the few instances where coal seams are worked by American or other outside companies in conjunction with various mining ventures. Bituminous coking coals are found in a few districts and a number of coke ovens have been installed.

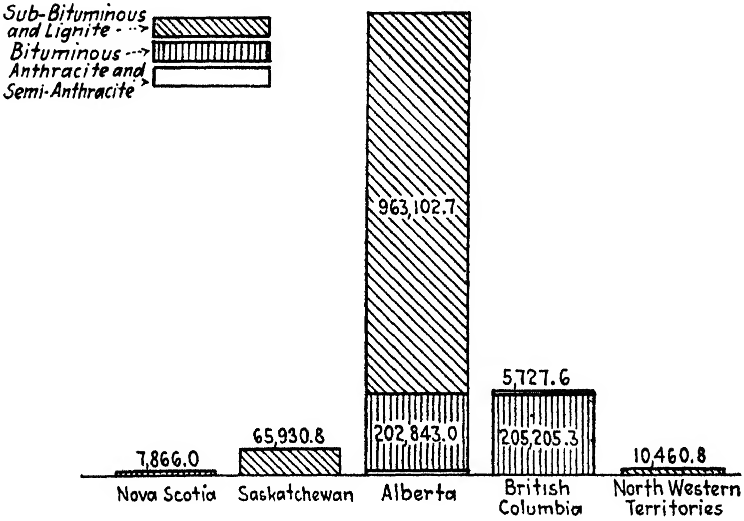

FIG. 25.—Coal resources of Canada. Figures are in millions of net tons.

In Central America, only two countries report coal. These are Honduras and Panama, and although the deposits in Honduras are said to contain about 1,000,000 tons of bituminous coal and 4,000,000 tons of lignite, the coal has never been mined.

The total estimated coal reserves of South America have been placed at 35,285 million short tons, representing only 0.36 per cent of the world's supply. The following countries report coal:

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* The estimate includes all seams more than 1 ft. thick, less than 4000 ft. below the surface.

Colombia, Venezuela, Ecuador, Peru, Bolivia, Argentina, Brazil, and Chile.[6] The coals vary from lignites to anthracites, but very little mining is done, and unless readily accessible large deposits are discovered, South America cannot become a major factor industrially.

Coal Resources of Europe.*—Europe is fairly well supplied with high-grade coal, though the reserves represent only about 15 per cent of the amount of coal in America. The relatively accessible coal resources of Europe occurring in seams more than 1 ft. thick at not more than 4000 ft. below the surface are estimated at about 600,000 million short tons.

Germany's coal reserves are mainly bituminous and lignitic, and this is a major factor in determining the industrial future of the country. During the war the Germans were in possession of the most productive mines of Russia and most of the coal fields in northern France. After the armistice this occupation was, of course, ended and the treaty of peace gave France the coal fields formerly owned by Germany in the Saar Basin, estimated to contain about 18,200 million short tons of coal in seams more than 1 ft. in thickness, less than 4000 ft. from the surface. This new addition to France's coal resources will have an important bearing on her future industrial development. Losing the war also cost Germany the important coal fields of Upper Silesia, which have been ceded to Poland. Thus, although Germany before the war controlled about 40 per cent of the European coal reserves, after the peace treaty[7,8,9] her reserves are less than 30 per cent of the European total.

Great Britain possesses quite large supplies of the higher-grade coals (slightly over 30 per cent of the European coal resources), the most important reserves being bituminous, semi-bituminous, and anthracites. The other countries, with the exception of Poland and Russia, are deficient in coal reserves of sufficient magnitude to permit of free industrial development, and in the future they will have to depend on imports to an increasing extent.

Coal Resources of Asia.—Little accurate information has been obtained with regard to the coal reserves underlying large areas in Asia, and the reported figures give only an approximation of the resources of the continent. By far the largest reserves

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* Information on European coal resources is at present very incomplete. This is largely due to the changed situation following the war.

are found in China (estimated at over 1,000,000 million short tons, of which over 400,000 million tons are reported as anthracite and semi-anthracite). The Chinese coals, in addition to the supplies of anthracite, include considerable quantities of coking coals of good quality. The reserves in Japan are very small (only 8,800 million tons in total), and it is probable that in the future the Japanese will try to control considerable quantities of the coal found on the mainland in order to make possible their own industrial development.

Resources of Africa and Oceania.—The coal resources of Africa have been estimated at 63,755 million short tons, the main coal fields being located in the southern part of the continent. Little field work has been done, however, and it is probable that there are other deposits, as yet undiscovered.

In Oceania, which includes Australia, New Zealand, and Tasmania, there are considerable deposits of high-grade coal, most of which is located in Australia. The latter country is supposed to have about 182,160 million short tons of coal, most of which is bituminous, and has the largest known coal seams in the world. Two of these are reported[6] to be 266 and 227 ft. thick, respectively. According to the estimates on the rate of coal deposition given in Chap. II, it would have taken over 80,000 years to form the vegetal matter to produce a coal seam of this thickness.

COAL PRODUCTION

The world's coal production and the production of the principal coal-producing countries of the world[10] from 1919 to 1922 are shown in Fig. 26, the data for 1912 being included to indicate prewar conditions. The figure shows the large amount of coal produced by the United States, from 35 to 45 per cent of the entire production of the world. Great Britain and the British Empire are the next largest producers, averaging 20 to 25 per cent of the total, the readily accessible nature of a large amount of her coal having made possible her present industrial development. Germany, which before the war controlled greater coal reserves than Great Britain, produces about 20 per cent of the total. The other countries range from 2.5 per cent downwards, and their combined output is in the neighborhood of 15 per cent of the total.

Anthracite.—The production and value (at the mine) of the anthracite mined in the United States[11] from 1918 to 1922 is given in Fig. 27. The data include about 99 per cent of all the anthracite mined, and, as is shown below, this represents about one-fifth the tonnage and about one-third the value of the bituminous coal mined in the United States. It should also be borne in mind that only about 5 to 7 per cent of the world's future supply of solid fuel is anthracite (mostly in Asia) and that the best seams of this coal in the United States will be exhausted comparatively soon.

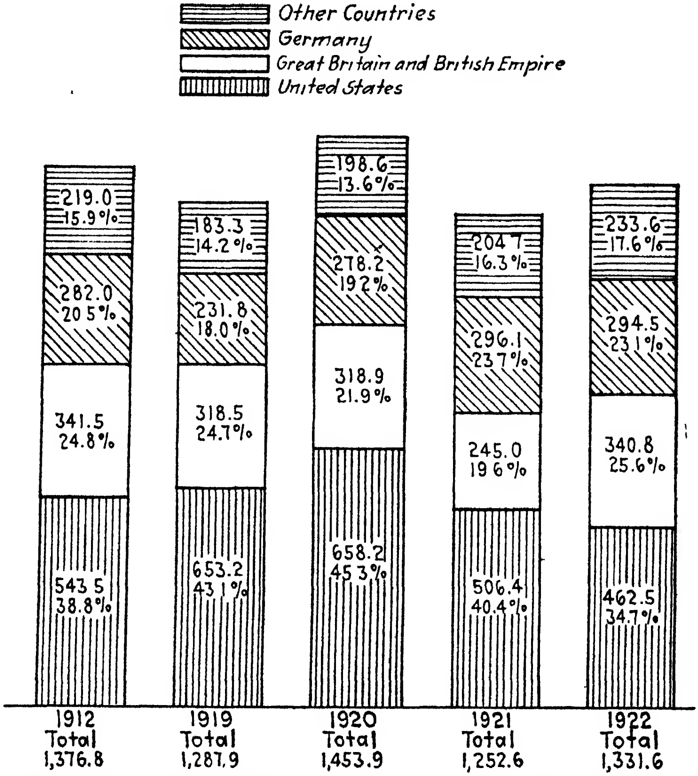


FIG. 26.—World's coal production from 1919 to 1922. Figures are in millions of net tons.

Bituminous Coal.—Bituminous coal, the premier fuel for industrial purposes, represents about 53 per cent of the world's future reserves of solid fuel and about 50 per cent of the future supply is in the United States. The world's production has already been given in Fig. 26, and Fig. 28 is included[11] to give the production of bituminous coal and its value in the United

States. It is interesting to note that the total value of the materials produced by the coal-mining and its associated industries (coke, by-products, briquets, etc.) is about 40 per cent of the value of the entire mineral output of the country.

Lignite.—About one-half of the world's reserve of solid fuel is lignite, and while there is practically no production of this fuel in the United States at present, it will become exceedingly important in the future. The large amounts of lignite produced in Germany and Czecho-Slovakia are shown in Fig. 29, the German production being about one-fourth of the total bituminous coal production of the United States.

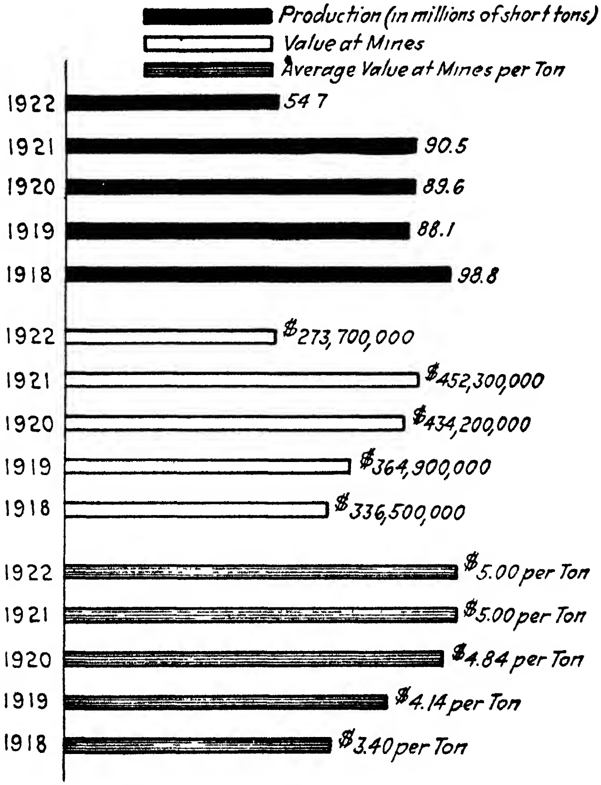


FIG. 27.—Production and value of anthracite coal mined in the United States from 1918 to 1922.

Peat.—Although peat is one of our big fuel reserves and it is destined to become important on the exhaustion of our more readily accessible coal seams, it is relatively unimportant com-

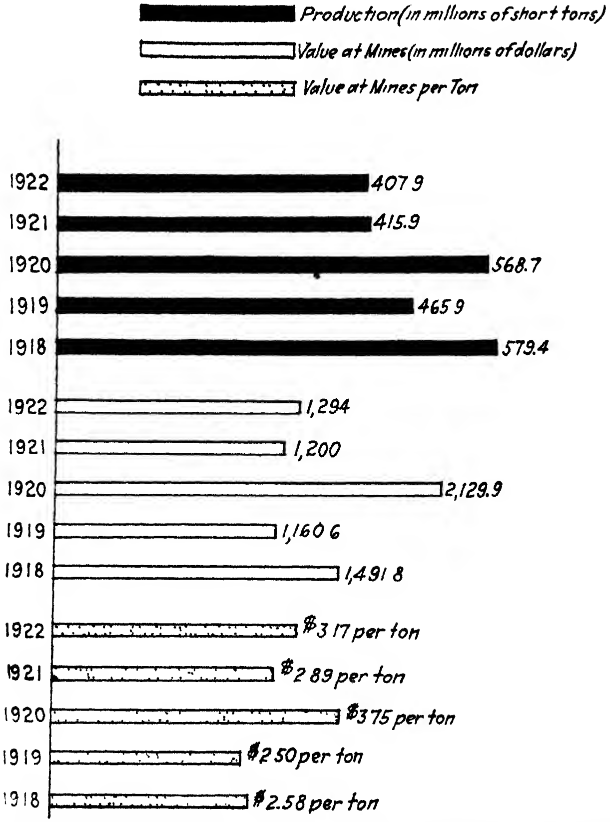


FIG. 28.—Production and value of bituminous coal mined in the United States from 1918 to 1922.

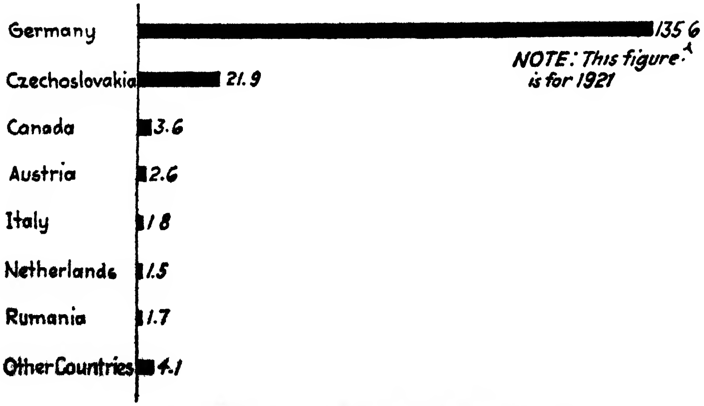


FIG. 29.—World's production of lignite for 1920.

mercially at present and its consumption in the United States is very low. The following table gives production figures for 1917 to 1922:

TABLE XXXII

PEAT PRODUCTION IN THE UNITED STATES, 1917–1922

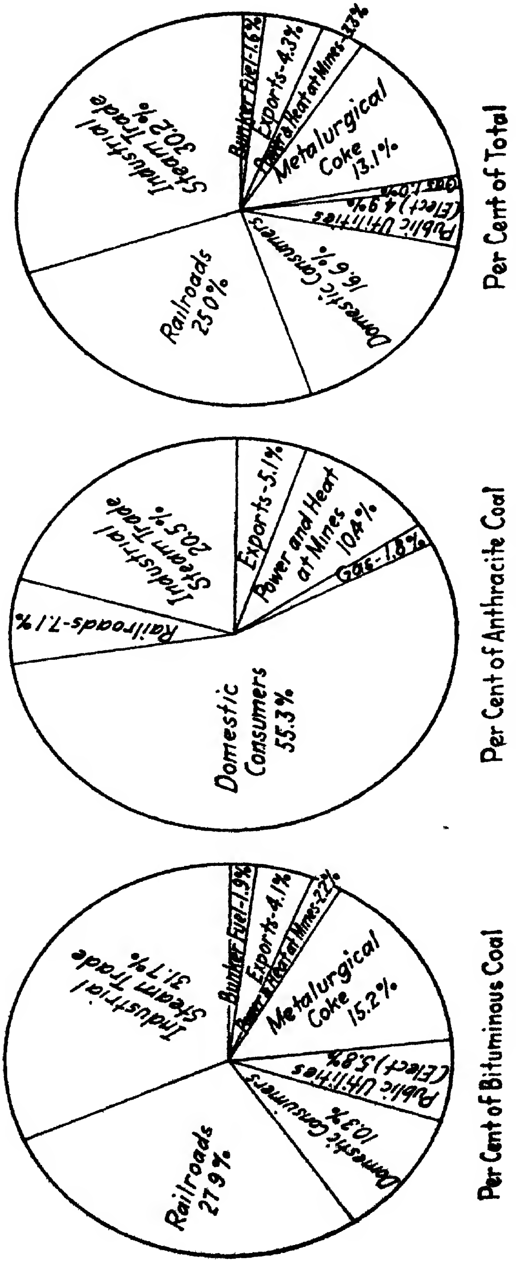
Year	Short tons	Value
1917	97,000	$ 710,000
1918	107,000	1,050,000
1919	69,000	710,000
1920	73,000	920,000
1921	30,000	260,000
1922	61,000	400,000

Coal Production in the United States, by States.—The production and value of coal produced in the main coal areas of the United States has a great bearing on our choice of fuels and location of commercial enterprises. Although over half of all the states mine some coal, over 60 per cent of our total production comes from the three states, Pennsylvania, West Virginia, and Illinois. The following table shows the percentages of the total bituminous output which were mined in the principal coal-producing states from 1920 to 1922.

TABLE XXXIII

BITUMINOUS COAL PRODUCTION IN THE UNITED STATES, 1920–1922 BY STATES

State	Per cent of total		
	1920	1921	1922
Pennsylvania	30.1	27.9	25.8
West Virginia	15.9	17.5	19.9
Illinois	15.6	16.7	14.5
Ohio	8.1	7.7	6.5
Kentucky	6.3	7.6	9.8
Indiana	5.2	4.9	4.2
Alabama	2.9	3.2	4.0
Colorado	2.2	2.2	2.5
Virginia	2.0	1.8	2.7
Other states	11.7	10.5	10.1
Total	100.0	100.0	100.0



Per Cent of Bituminous Coal

Per Cent of Anthracite Coal

Per Cent of Total

FIG. 30.—The utilization of anthracite and bituminous coal in the United States.

Since the mines of West Virginia and Kentucky are largely worked by non-union labor, the change in their percentage is of considerable interest.

COAL CONSUMPTION

At the present time coal serves three purposes: It is used in the production of power, as a source of domestic heat, and as the raw material for coal products. The coal utilized in the production of power for industrial purposes is about equally divided between general manufacturing purposes and the railroads, these two uses consuming about 60 per cent of the total. The production of coke and gas manufacture consume another 15 per cent, while domestic heating uses about 17 per cent. That is, the industrial coal consumption is about five times that of the domestic requirements, and, in consequence, the industrial consumer, by virtue of larger buying capacity, forces the domestic user to bear the brunt of high prices.

Figure 30, adapted from data given by the U. S. Geological Survey,[5] shows that about 90 per cent of the coal produced is consumed by the four purposes of general manufacturing, railroads, domestic heating, and production of coke.

BIBLIOGRAPHY

1. "Coal Resources of the World," Twelfth International Geological Congress, Morang & Co.
2. M. R. Campbell, Bacon and Hamor: in "American Fuels." McGraw-Hill Book Co., Inc. (1922).
3. U. S. Bur. Mines, *Bulls.* 22 (1912); 85 (1914); 123 (1918); 193 (1922); and *Prof. Paper* 100-A.
4. "The Coal Catalog," Keystone Consolidated Publishing Co., Pittsburgh, Pa.
5. "World Atlas of Commercial Geology," U. S. Geol. Survey (1921).
6. E. S. Moore, "Coal," John Wiley & Sons (1922).
7. *Miniera Ital.*, **6**, 254 (1922).
8. Spurr: "Political and Commercial Geology," McGraw-Hill Book Co., Inc. (1920).
9. *Federal Reserve Bull.*, 8 (1922), 1308.
10. *Mineral Industry*, **312**, McGraw-Hill Book Co., Inc. (1922).
11. "Mineral Resources of the United States in 1922," U. S. Geol. Survey (1923).
12. *Nat. Coal Mining News*, Dec. 19 (1923).

CHAPTER VI

PETROLEUM

Petroleum is of particular importance to society. It furnishes gasoline, by far the most important motor fuel, and yields the lubricants so essential to industry. It provides the fuel oil now so widely used in the production of heat, steam, and power, and the gas oil used in producing manufactured gas. It is the source of the kerosene used for lighting purposes outside the cities and for domestic heating, and serves as the raw material for an oil by-products industry which offers great possibilities for future development. Of necessity, the treatment of such a broad subject in a work of this kind is relatively narrow in scope and will include a brief summary of the characteristics of petroleum, its origin and geology, and the types of oil most commonly used as fuel. A resumé of the statistics of petroleum production and distribution has been included in order to show the probable future trend of the industry. Natural gas, similar in origin and so often occurring with petroleum, is discussed in the next chapter. Chapter XVI, on the combustion of liquid fuels, discusses in detail the advantages and disadvantages of petroleum as a fuel.

The Composition of Petroleum.—Crude petroleum is defined[1] as "a naturally occurring mixture, consisting predominantly of hydrocarbons, or of sulfur, nitrogen, or oxygen derivatives of hydrocarbons, which is removed from the earth in liquid state or is capable of being so removed." The crude petroleum falls into three main classes: The first yields on distillation a residue consisting mainly of paraffin wax and is known as a paraffin base crude. The second leaves a black, lustrous residue of asphalt, and is known as an asphalt-base crude. The third, leaving both asphalt and paraffin as a residue, is known as a mixed-base crude.

The hydrocarbons in petroleum all belong to a number of regular families of compounds of the following types.*

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* For a detailed exposition of the various hydrocarbons and hydrocarbon derivaties found in petroleum, the reader is referred to Bibliography, references 2, 3, and 4.

TABLE XXXIV

COMPOUNDS FOUND IN PETROLEUM

1. C_nH_{2n+2}
2. C_nH_{2n}
3. C_nH_{2n-2}
4. C_nH_{2n-4}
5. C_nH_{2n-6}
6. C_nH_{2n-8}
7. C_nH_{2n-10}
8. C_nH_{2n-12}
9. C_nH_{2n-14}

Members of all these series have been isolated from petroleums, as, for example, the gaseous methane (CH_4) and the solid paraffin $C_{30}H_{62}$, both of which are members of the series C_nH_{2n+2}. In addition to these hydrocarbons there are often found small quantities of oxidized bodies, present either as complex acids or phenols.

Nearly all crude petroleums contain nitrogen and sulfur. The nitrogen may vary from only a trace up to 1 per cent and over, and is practically always present in the form of complex organic bases. These compounds are particularly abundant in the California petroleums, and in a number of cases they represent from 10 to 20 per cent of the crude. Sulfur, though rarely absent, is usually present in only small amounts. In oil from the Lima-Indiana district, however, sulfur compounds occurring as sulfides of the paraffin series are present in such amounts as to necessitate special methods of refining. Crystalline sulfur also has been separated from a number of crude petroleums.

The ultimate analysis of a number of petroleums is given in the following table:[3]

TABLE XXXV

ANALYSIS OF REPRESENTATIVE PETROLEUMS

Source	Per cent				Gravity, degrees Baumé
	Carbon	Hydrogen	Oxygen + Nitrogen	Sulfur	
Pennsylvania	84.9	13.7	1.4	. .	28
West Virginia	83.5	13.3	3.2	. . .	30
Ohio	84.2	13.1	2.7	. . . .	28
California, Bakersfield	81.5	10.0	6.9	0.55	15
Texas, Beaumont	84.6	10.9	2.9	1.63	22
Mexico	83.7	10.2	. . .	4.15	22
Russia, Baku	86.6	12.3	1.1	. . . .	17

Origin of Petroleum.—The question of the origin of petroleum has been discussed even more than that of the origin of coal. Numerous theories have been advanced and controversy is still active.* These theories may be divided into (1) the inorganic theories, which consider that the formation of petroleum took place without the aid of living organisms; and (2) the organic theories, which postulate that living organisms, such as bacteria, decomposing vegetal matter, or animal matter, were essential to petroleum formation. A few of the more important theories are discussed below.

Inorganic Theories.—Of the inorganic theories of petroleum formation, that of Mendeleéf has probably received the most attention. It presupposed the existence of iron or other heavy metal carbides in the interior of the earth, to which subterranean waters gain access, forming hydrocarbons. The theory was experimentally supported by a number of researches which demonstrated that hydrocarbons similar to petroleum can be formed by the action of water on carbides. It has never been proved, however, that such carbides existed in the earth, and it is improbable that large quantities of petroleum could have been formed in such a way. Quite recently, Franz Fischer has pointed out, and has shown experimentally, the possibility of forming petroleum hydrocarbons from hydrogen and carbon monoxide of volcanic origin.

Organic Theories.—It is quite certain that hydrocarbons similar to natural gas and petroleum may be derived from either animal or vegetal matter or both,[2,3] and this has led to the two main types of organic theories: (*a*) theories of vegetal origin, and (*b*) theories of animal origin.

(*a*) Theories of Vegetal Origin.—One of the most plausible theories is that petroleum has been formed by the decomposition of immense quantities of gelatinous algae or marine vegetation. These two types of vegetal matter contain little or no cellulosic material, and when they decompose, they form gas (mainly CH_4) and oil which, if formed on sandy soil, is carried away by gravity or by water. If, on the other hand, this material decayed in a peat bog, coal of the cannel or boghead type (see Chaps. II and III) would result. Considerable evidence, such as the formation of petroleum at the present time by the decomposition of sea-

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* For a summary and critical resumé of the various theories see Bibliography, reference 2.

weed in sand, has been advanced in favor of the concept of vegetal origin.

(*b*) Theories of Animal Origin.—Several theories have been advanced which consider that petroleum has resulted from the decay of immense quantities of marine animals, such as fish, mollusks, corals, etc. Such material contains high percentages of nitrogen, and the nitrogen bases of a number of oils (California, in particular) furnish strong evidence of such formation. Furthermore, most oils occur in association with brines which have a composition indicative of marine origin.

In general, it may be said that the evidence is in favor of the organic origin of petroleum, and that any organic substance which is enclosed between sediments may be a source of petroleum. When the resultant petroleum is rich in nitrogen, animal matter was probably the initial material; otherwise, the petroleum most probably was of vegetal origin.

Method of Formation from Organic Matter.—The transformation of organic material into petroleum is considered to have taken place as follows:

1. The accumulation of the organic matter in clays and sands under water along seacoasts or in lakes.
2. The burial of the accumulated material by sediments, which protected it against excessive oxidation.
3. The action of pressure and heat, which brought about a semi-destructive distillation of the organic matter, resulting in the formation of petroleum.

The chemical characteristics of the process have been outlined by Engler[5] somewhat as follows: (1) Putrefaction or fermentation, eliminating albumen, cellulose, etc., leaving fatty material, waxes, etc. (2) Saponification of the glycerides and production of free fatty acids. The residues from certain crude petroleums are probably due to the incompleteness of these actions. (3) Elimination of carbon dioxide, leaving hydrocarbons of high molecular weight containing oxycompounds. (4) Formation of liquid hydrocarbons with violent reaction, and cracking (formation of light or gaseous products).

The Geology of Petroleum.—Petroleum is widely distributed and occurs in commercial quantities in sedimentary rocks. The general requisites for the occurrence of petroleum in commercial quantities are:[6] (1) porous rock to contain the oil, (2) an impervious cover to prevent it from escaping, (3) geological structure which permits the oil from a wide area to collect in a restricted area, and (4) a source from which the petroleum may have been formed.

Usually associated with natural gas or water (or both) the petroleum fills the voids in the porous rock, which is almost always sandstone or limestone. The oil is not usually found where it was formed, its accumulation at a given spot being due to one or more of the following factors:

1. Due to differences in specific gravities of the water, petroleum, and natural gas, separation of these materials takes place in the beds of porous rock, the gas and oil being concentrated in the upper, concave downward portions of the strata (called anticlines or domes). A typical formation is shown in Fig. 31. In certain cases, however, notably in the Appalachian field, the oil does not occur with water and, consequently, owing to the lack of upward pressure due to the water, the oil is found in the bottom dips of the strata. These lower, concave, upward bends are called synclines.

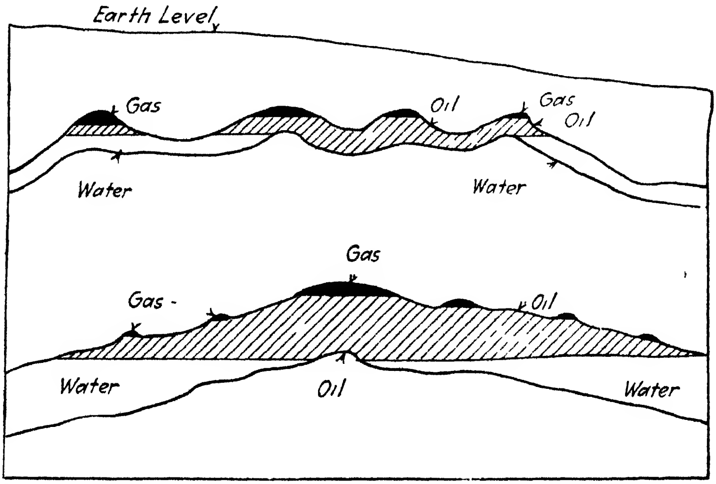

Fig. 31.—Occurrence of petroleum and natural gas in a typical rock formation.

2. The porous bed may be sealed up by an impervious overlying bed which prevents the escape of the oil to higher strata or to the surface.

3. The rock containing the oil may or may not be very porous and this governs the quantity of oil retained and also the freedom of flow.

Oil-well Drilling and Life of Wells.—The crude petroleum is secured by drilling through the non-porous, oil-retaining bed above the porous rock, and allowing the free upward movement of the oil. The gas and oil are under pressure from the water below and, as the impervious layer is pierced, the sudden release of this pressure brings in a gusher. The length of time that this period of high production lasts varies greatly in different wells, some holding up over long periods while others give out in a few hours. The presence of petroleum is indicated by oil springs or seepages, and in such cases its location underground may be determined quite accurately from a consideration of the structural features of the surrounding terrain. The knowledge of oil geology has so far advanced that it is possible to state quite definitely where oil fields may be found and the productive period of a field, once located, may be predicted with reasonable accuracy.

Losses in Petroleum Production.—By no means is all the oil which is underground, even after it has been located by means of a well, raised to the surface. The three main losses are due to the large percentages of oil left in the porous rock underground, the infiltration of water into the well, and the escape of natural gas during the process. The location of the oil depends mainly on the differences in specific gravities between oil and water, and, therefore, if water creeps into an oil well, the location and the ease of recovery of the oil are usually changed. This infiltration of water, which always means a lowered recovery of oil, has ruined many wells, and has seriously affected whole fields. Many wells during drilling enter gas-bearing formations and further work is often suspended until the gas has escaped. The losses resulting from this practice are twofold: first, the escape of the gas itself represents the loss of an enormous quantity of valuable fuel. In addition, there is a much larger waste, essentially the loss of pressure, which, if it were not lost and were exerted on the oil would allow a far greater petroleum recovery than is ordinarily possible. It has been estimated[7] that under present practice from 30 to 90 per cent of the oil is left underground, that, owing to losses from improper storing, fires, and the like, only 25 per cent of the petroleum that is underground reaches the pipe line, and that the overall efficiency of utilization, which takes into consideration the losses due to wasteful use of oil, is only about 10 per cent.

Grades of Petroleum Oils Used as Fuel.—Most crude petroleums, in order to separate the various products, are subjected to fractional distillation. This distillation may consist in "topping," that is, removing only the lighest, lowest-boiling, most volatile constituents, such as gasoline, or, the process may be continued, distilling over gasoline,* kerosene (approximately 44°Bé.), gas oil (approximately 35°Bé.), light fuel oils (18 to 30°Bé.), and lubricating oils, in the order named, a heavy (approximately 10 to 14°Bé.) viscous residue (called residuum) being left behind in the still.

Fuel oil has been defined as "any petroleum product used for the generation of heat in a fire box. In general, any liquid or liquefiable petroleum product may be used as a fuel oil, the controlling factors being cost and adaptability. Fuel oils in common use fall into one of four classes: (1) crude petroleums, (2) distillate fuel oils, (3) residual fuel oils, (4) blended fuels."[1]

1. *Crude Petroleums.*—The crude petroleums which are used as fuel oils are of relatively low commercial value otherwise. Weathered crudes, those which have lost an appreciable quantity of the more volatile constituents by evaporation during storage,[1] often find use as fuel oil. In the early days of the petroleum industry, crudes were extensively used as fuel oil, but owing to the possible recovery of the valuable gasoline, other lighter products and lubricating fractions by distillation, this practice has largely been discontinued and today only those crudes which contain small quantities of valuable material are being used directly as fuel. Untopped crudes (containing small percentages of the light gasoline fractions which lower the flash point) are somewhat more dangerous from the standpoint of fire hazard than residual fuel oils or distillates.

2. *Distillate Fuel Oils.*—Distillate fuel oils are obtained from more complete refining operations than are carried out on the topped crudes. They consist of the fractions distilled intermediate between kerosene and lubricating oils and since they are produced by distillation from the crude petroleum they are free from heavy residual tar and salt. These distillate fuels may vary from 44°Bé. (crude kerosene), through gas oil (about 35°Be.) and the light fuel oils† of 28 to 30°Bé., to somewhat heavier cuts,

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* The products are listed in the order of their recovery, *i.e.*, the lightest, most volatile products first.

† Now being used to an increasing extent as a domestic fuel.

depending on the source of the original crude. Distillates are obtained mainly from the Mid-Continent and Eastern districts and include the gas oils, and certain fractions of the lubricating and cylinder stocks. In general, the viscosity of the lighter grades is quite low.

3. *Residual Fuel Oils.*—Residual fuel oils are topped heavy crudes or the viscous residual products obtained from crude petroleum by the removal of some of the water and an appreciable percentage of the valuable, more volatile hydrocarbons. The topping process, therefore, allows the recovery of the valuable gasoline and other light products, thus raising the flash point, makes for safety in handling and burning the fuel oil, and removes the water. Any water present in residual fuel oil gets in after distillation, from leaky steam coils, rain, and the like. These residual fuel oils come mainly from California, Texas, Mexico, and parts of Oklahoma. They contain very small amounts of ash and salt left from the original petroleum.

4. *Blended Fuels.*—The blended fuels, as the name implies, are mixtures of any or all of the three other classes. Blending with distilled material is often necessary to overcome the high viscosity of the residuums left after the removal of the lighter and lubricating fractions. Blended fuels of this type often can be detected by the small amounts of ash and salt which they contain.

Properties and Specifications of Fuel Oil.—The important properties of fuel oils are: (1) specific gravity, (2) heating value, (3) flash point and fire point, (4) congealing point (cold test), (5) viscosity, (6) sulfur content, (7) moisture and sediment, (8) specific heat and coefficient of expansion. Distillation curves are important only in case the oil is to be used as a motor fuel. Although general specifications for fuel oils cannot be used to advantage, owing to the wide difference in individual requirements, certain points should always be borne in mind in purchasing oil fuel. These points are taken up in the discussion of the properties of fuel oil below.

1. *Specific Gravity.*—Fuel oil is sold by volume and not by weight; accurate specific-gravity measurements are, therefore, an essential.* Users of fuel oil should always bear this in mind in buying, since the storage space required as well as the other

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* Since specific gravity varies with temperature, the temperature should always be given in reporting a specific-gravity determination.

properties must be considered. Figure 32 gives the density of liquids in pounds per gallon plotted against both specific gravity and degrees Baumé (all at 60°F.), the latter being the figure usually used in the petroleum industry. The relation between specific gravity and degrees Baumé (American Petroleum Institute Standard) is:

$$\text{Degrees Baumé (A.P.I. Std.)} = \frac{141.5}{\text{Specific Gravity, }60°/60°\text{F.}} - 131.5.$$

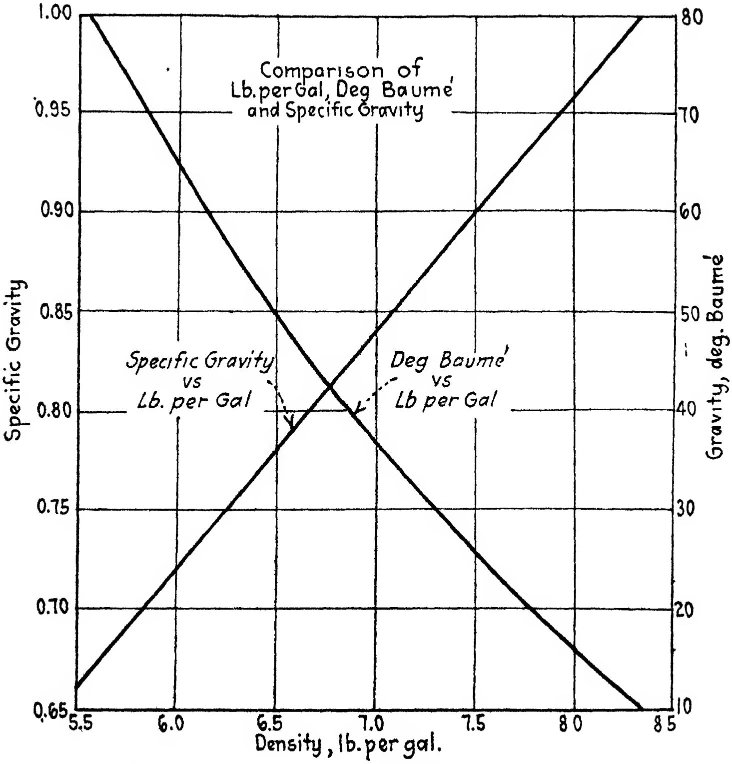

FIG. 32.—Comparison of density in pounds per gal., degrees Baumé, and specific gravity of liquids lighter than water.

2. *Heating Value.*—The *heating value* of most fuel oils is between 18,000 and 19,500 B.t.u. per lb., with about 18,500 B.t.u. as an average. In Fig. 33, the approximate heating values of various fuel oils are plotted against degrees Baumé (°Bé.). This figure shows that, although a crude of high specific gravity (low °Bé.) may have a lower heating value *per lb.* than an oil

of low specific gravity (high °Bé.), the positions are reversed when the heating value per gal. is considered. For example, a crude of 15°Bé. has a heating value of 18,680 B.t.u. per lb., while a 45°Bé. oil has a heating value of 20,400 B.t.u. per lb. On the other hand, the 15°Bé. oil contains 150,500 B.t.u. per gal., as compared with only 136,400 B.t.u. per gal. for the higher-priced 45°Bé. oil. Usually oils that are the cheapest per gal. also have the highest B.t.u. per gal.

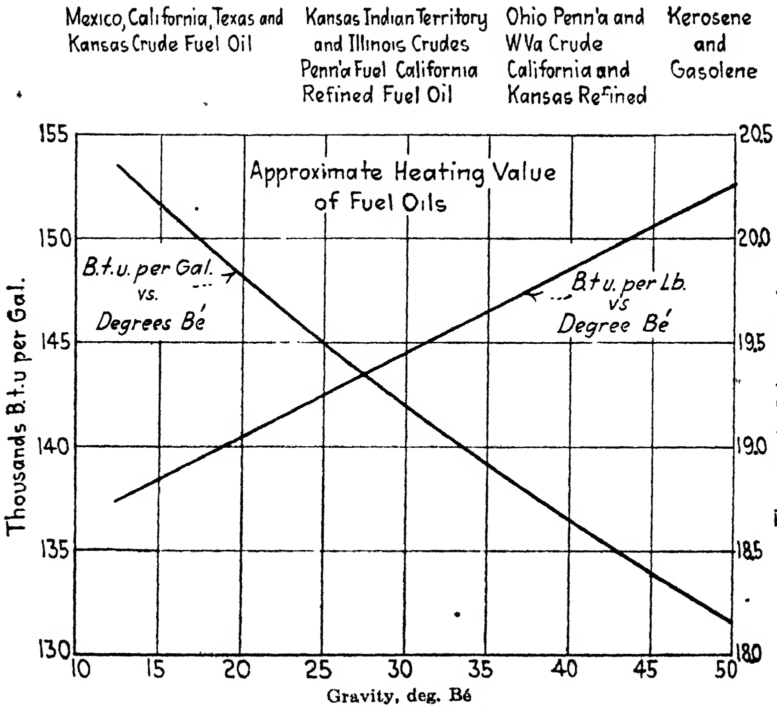

FIG. 33.—Approximate heating values of various fuel oils of different gravity.

3. *Flash Point and Fire Point.*—The *flash point* of an oil is the temperature at which sufficient vapor is given off under certain definite conditions to form a momentary flash (a burning of the mixture of gas and air) when a small flame is brought near its surface. The flash point is determined in a special type of apparatus, the specific form of the apparatus often being prescribed by law. The *fire point* of an oil is the temperature at which the oil gives off enough vapor to burn continuously.

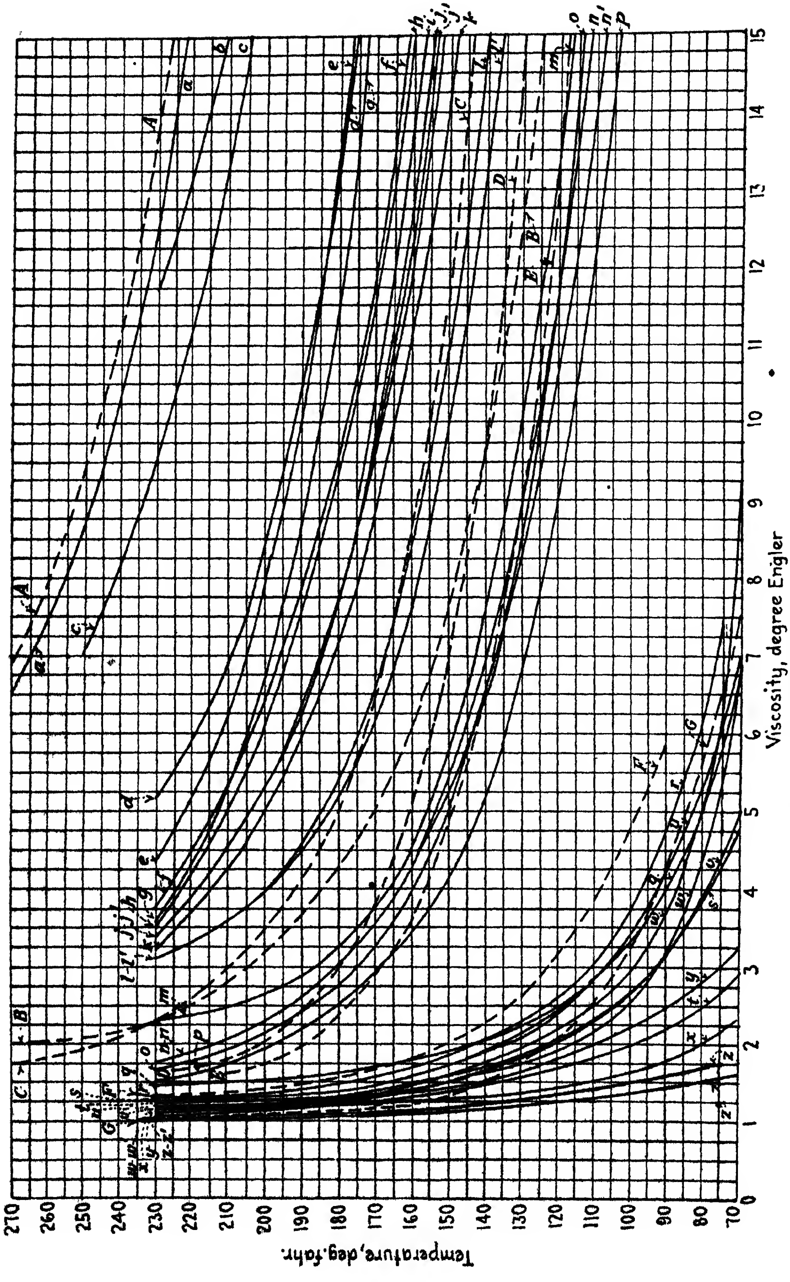

Fig. 34.—Change in viscosity of various oils with temperature.

Legend for Fig. 34.

Curve No.	Type of oil	Gravity		Flash point, °F.
		Specific	°Bé.	
	Solid curves			
a	Mexican residue	1.000	10.0	374
b	"Toltec fuel oil," Inter-Ocean Oil Co., N. Y.	.988	11.7	220
c	"Toltec of Panuco oil," Inter-Ocean Oil Co.	.986	12.0	124
d	"No. 102," Union Oil Co., Bakersfield, Cal.	.980	12.9	280
e	"No. 18," Union Oil Co., Bakersfield, Cal.	.980	12.9	285
f	"Standard" Mexican crude (lot 2)	.964	13.4	202
g	"No. 25," Union Oil Co., Bakersfield, Cal.	.978	13.2	262
h	Mexican crude, Texas Co.	.952	17.3	126
i	Sample No. 3, Anglo-Mex. Pet. Products Co.	.952	17.3	164
j	"Gaviota Refinery," Associated Oil Co., California	.953	17.1	230
j'	Mexican oil, Atlantic torpedo flotilla, March, 1914	.947 .954	18.1 17.0	182 145
k	Standard Mexican crude (lot 1)	.950	17.6	182
l	Mexican oil, U. S. S. *Arethusa*			
l'	"Nos. 1, 2, 3," Anglo-Mexican Pet. Products Co	.955	16.8	188
m	Producers Crude No. 1 fuel oil, Union Oil Co., California	.959	16.1	174
n	"Coalinga Field," Associated Oil Co., Monterey, Cal	.957	16.5	186
n'	"Avon Refinery," Associated Oil Co., Avon, Cal	.953	17.1	168
o	Richmond, California	.953	17.1	228
p	Sun Co., Louisiana	.936	19.8	275
q	"Standard," Illinois	.893	27.3	146
r	Gulf Refining Co., Navy standard oil, U. S. S. *Perkins*	.892	27.5	180
s	"Standard," Indiana	.880	29.6	144
t	"Standard Star," California	.912	23.9	180
u	"Standard," Illinois (lot 4)	.893	27.3	146
v	"Standard," Indiana (lot 4)	.880	29.6	144
w	Gulf Refining Co., Navy contract	.882	29.3	170
w'	"Standard," Lima, Ohio, crude	.876	30.4	149
y	"Star," California	.912	23.9	180
z	Gulf Refining Co., Navy standard oil, U. S. S. *Roe*	.885	28.7	182
z'	Standard Mexican gas oil	.856	34.2	151
*	Indicates test results			
	Dotted curves			
A	Panuca crude, Inter-Ocean Oil Co.	.975	13.7	140
B	Mexican petroleum, Texas Co.	.938	19.5	234
C	Associated Oil Co., California	.971	14.2	257
D	Bakersfield, Cal. pipe line to Port Costa	.970	14.4	260
E	California Standard Oil Co., steamer *Santa Barbara*	.962	15.7	282
F	Beaumont, Tex., Gulf Refining Co.	.907	24.8	222
G	Navy standard oil, Texas Co.	.911 to .900	24 to 26	195 to 220

That is, the oil will burn continuously if it is raised to the fire point. This temperature is usually about 20°F. higher than the flash point. The lower the flash point, the easier it is to ignite oil, but the more hazardous is the oil when used as a combustible material.

4. *Cold Test.*—The *cold test* or congealing point is the temperature at which the oil becomes pasty, usually due to the crystallization of paraffin or other matter. This test is important in showing whether or not an oil will flow freely in pipe lines, etc., and the required cold test depends on the climate.

5. *Viscosity.*—The viscosity of an oil is a measure of the internal friction of the liquid or its resistance to flow. This resistance, which is markedly decreased by raising the temperature, is generally measured by determining the time required for a given quantity of the oil at a constant temperature to flow through a small nozzle, and is reported in a number of different ways,* such as degrees Engler, degrees Saybolt, and the like, depending on the type of instrument used.

As is shown in Chap. XVI, the chief function of an oil burner is to atomize or finely divide the oil, and to do this effectively with a heavy oil the viscosity must first be lowered by heating. Figure 34 shows[8] the changes in viscosity with changing temperature of a number of the common oils. Fuel oils may be divided into two classes on the basis of viscosity:[8]

Class 1 includes asphaltic base crudes, residual fuel oils, or other oils which require preheating to reduce the viscosity to such a point that they may be atomized properly. For proper atomization in mechanical burners the oils must have a viscosity of 4 to 12°Engler and this determines the proper temperature for preheating (see Fig. 34). It is desirable that this temperature should not be above the flash point of the oil when the fire risks are great. It is recommended that the flash point of this class of oils should not be below 150°F. (Navy Specifications).

Class 2 is composed of fuel oils of so low a viscosity as to make preheating before burning unnecessary, 12°Engler at 60°F. being about the maximum viscosity allowable. The distillate fuel oils usually fall in this class. The flash point of such oils should not be below 120°F.

6. *Sulfur Content.*—High sulfur content is objectionable since corrosion results from the action of the sulfur dioxide

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* See references 3 and 10 for description of methods and conversion tables.

(SO_2) formed on burning. Oils of high sulfur content have also been found to corrode tanks and pipe lines and this factor must also be taken into consideration. A sulfur content of 4 per cent or less is sometimes used for general purposes, such as steam raising, but in metallurgical or other work where the sulfur may have a harmful effect the percentage should run much less.

7. *Moisture and Sediment.*—Moisture and sediment are found in practically all fuel oils. They should not, however, be present to the extent of more than 2 per cent by volume (Navy Specifications 1 per cent), as otherwise trouble may be experienced in firing.

8. *Specific Heat and Coefficient of Expansion.*—The specific heat, which is of importance in problems of heater design, varies between about 0.4 and 0.5, while the coefficient of expansion is generally taken as 0.0004 per degree Fahrenheit. In order to find the change in volume with a given change in temperature, substitute in the following equation:

$$V_t = V_o\ (1 + 0.0004t),$$

where V_o is the volume at the initial temperature, V_t is the volume at the final temperature, and t is the degree Fahrenheit temperature *increase*.

PETROLEUM RESOURCES

Petroleum, although widely distributed geographically, occurs, so far as is known at the present time, in commercial quantities in relatively restricted areas, and the major producing fields are few in number. Oil fields of commercial consequence are found in the United States, Mexico, Russia, Galicia, Dutch East Indies, Mesopotamia, Persia, India, and a number of other countries. The first three countries named, however, supply practically nine-tenths of the petroleum in the entire world.

United States Resources.—The oil pools of the United States have been grouped into seven major producing fields or districts. These are (1) the Appalachian district, (2) the Lima-Indiana district, (3) the Illinois district, (4) the Mid-Continent district, (5) the Gulf Coast district, (6) the Rocky Mountain district, and (7) the California district. The locations and boundaries of the districts are shown in the map (Fig. 35) and their production for the period from 1890 to 1924 is shown in Fig. 37. The main

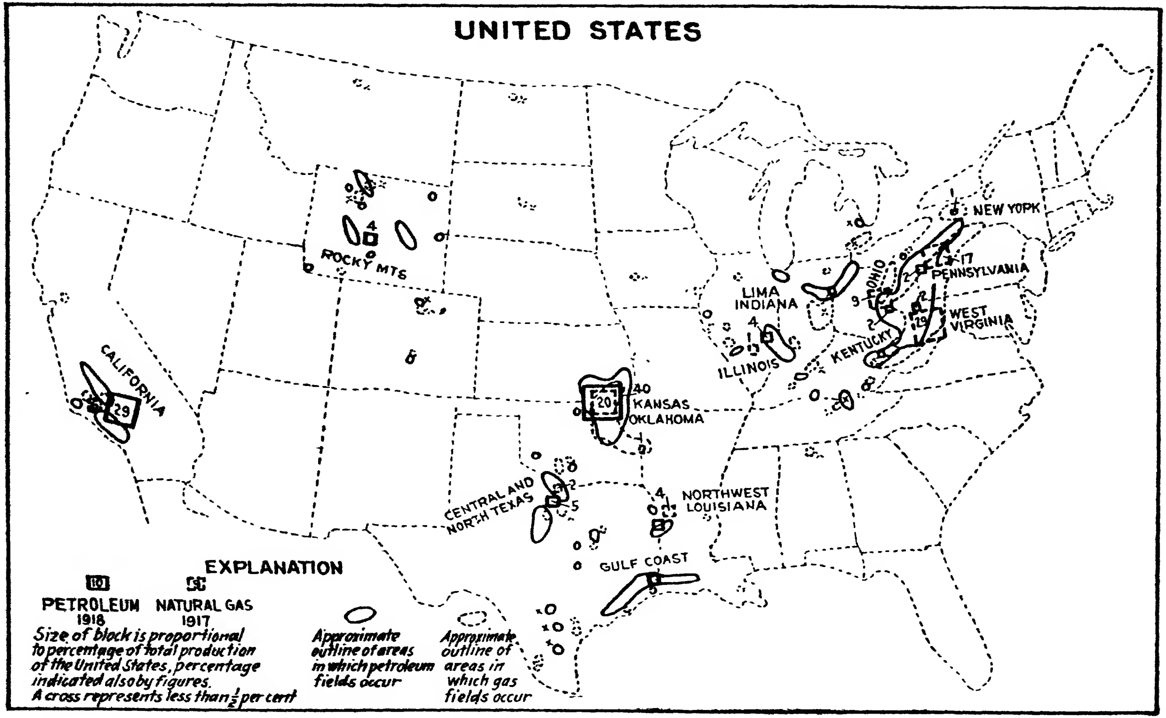

FIG. 35.—Areas within which petroleum and gas fields are known to occur in the United States and the production of petroleum in 1918 and of natural gas in 1917. (*From United States Geological Survey, "World Atlas of Commercial Geology," plate* 16, (1921).)

features of the districts, and the characteristics of the oils* produced in them, are discussed in the following paragraphs.

The Appalachian District.—The Appalachian district takes in the oil fields east of central Ohio and north of Alabama, including those of Pennsylvania, New York, West Virginia, eastern Ohio, Tennessee, and Kentucky. The typical oils are paraffin base, free from asphalt and objectionable sulfur, and rich in the lighter derivatives. The gravity ranges from 39 to 45°Bé., averaging about 43°, and yields on distillation about 35 per cent gasoline, 27 per cent kerosene, 10 per cent gas oil, and 15 per cent lubricating oil. On account of the lighter distillation products and heavy-viscosity cylinder oils in the residuums, these oils are too valuable to be used as fuels. Although no large producer, this field is important to the oil industry, owing to the high quality of oil.

The Lima-Indiana District.—The oil pools in northwestern Ohio and most of those in Indiana are included in the Lima-Indiana district. The oils are paraffin base, of lower quality than those of the Appalachian district, and contain large percentages of sulfur, which must be removed by special treatment. The average gravity is about 39°Bé. and they contain about 35 per cent gasoline, 15 per cent kerosene, 12 per cent gas oil, 9 per cent lubricating oil, and 25 per cent fuel oil.

The Illinois District.—The principal producing areas in the southeastern part of Illinois, as well as scattered pools in the central and western parts of the state, comprise the Illinois district. The oils in the northern part of the district are heavy, asphalt-base crudes with relatively high sulfur. To the south the oils are of better quality, with an average gravity of about 32°Bé. and contain about 20 per cent gasoline, 15 per cent kerosene, 8 per cent gas oil, and 16 per cent lubricating oils.

The Mid-Continent District.—The oil fields in Kansas, Oklahoma, northern and central Texas, and northern Louisiana make up the Mid-Continent district. The crudes vary widely in composition and appearance, ranging from about 21°Bé. to the "gasoline well" product in Oklahoma, which has a gravity of about 55°Bé. A typical oil from this district, of gravity 34°Bé., contains about 22 per cent gasoline, 15 per cent kerosene, 12 per cent gas oil, and 29 per cent fuel oil.

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* The description of the various crudes is abbreviated from that given in references 9 and 12.

The Gulf Coast District.—The Gulf Coast district includes the fields in the Gulf Costal Plain of Texas and Louisiana. The crudes are of asphalt base, averaging about 22°Bé. in gravity, with a high sulfur content. They contain little of the lighter products and yield about 35 per cent gas oil, 45 per cent lubricating oils, and 12½ per cent of residuum asphalt.

The Rocky Mountain District.—The fields in Colorado, Wyoming, and Montana are included in the Rocky Mountain district, the main production being from Wyoming. Two types of oil are found, one of asphalt base, ranging from 15 to 25°Bé., and the other of paraffin base with gravities from 25 to 50°Bé. The latter type is well suited for refining and a crude of 36.5°Bé. gravity yields about 29 per cent gasoline, 15 per cent kerosene, 10 per cent gas oil, and 17 per cent lubricating oils.

The California District.—The oil fields on the sides of the San Joaquin Valley (called the Valley fields) and the pools along the coast (called the Coastal fields) are included in the California district. The crudes are mainly asphalt base, with low sulfur content, of gravities averaging about 21°Bé., and the major part of the production, as a residue after refining, is used as fuel oil. Recent discoveries of large quantities of lighter oils, however, have led to an increasing production of refined oils. A typical crude contains about 22 per cent gasoline, 19 per cent kerosene, 18 per cent lubricating oils, and 21 per cent residuum.

The average crude oil, produced in the United States under present methods,[13] yields 27 per cent gasoline, 10 per cent kerosene, 4 per cent lubricating oil, 46 per cent of gas and fuel oil, and varying amounts of asphalt, wax, etc.

Estimated Reserves in United States Petroleum Areas.—A joint committee of the U. S. Geological Survey and the American Association of Petroleum Geologists[14] has estimated that in the United States, in January, 1922, there were in the ground about 9,000,000,000 bbl. of oil *recoverable by present methods.** Calculations of oil reserves are extremely unreliable, time usually proving them too conservative.[16] The oil known to be in the ground in 1925 still equals the amount estimated in 1922. However, the estimates in the following table,[16] which are conservative, are the most complete and reliable that have been prepared for any oil country or district.

* Petroleum known[16] to be in the ground in the U. S., 26,000,000,000 bbl.

TABLE XXXVI

Estimated Oil Reserves of the United States

(Recoverable by present methods.)

	Millions of Barrels
New York	100
Pennsylvania	260
West Virginia	200
Ohio	190
Indiana and Michigan	70
Illinois	440
Kentucky, Tennessee, northern Alabama, and northeastern Mississippi	175
Missouri, Iowa, North Dakota, Wisconsin, and Minnesota	40
Kansas	425
Oklahoma	1,340
Northern Louisiana and Arkansas	525
Texas, except Gulf Coast	670
Gulf Coast, Texas, and Louisiana	2,100
Colorado, New Mexico, and Arizona	50
Wyoming	525
Montana, Nebraska, and South Dakota	100
Utah, Nevada, Oregon, Washington, and Idaho	80
California	1,850
Eastern Gulf Coastal Plain and Atlantic Coast states	10
Total	9,150

Of the 9,000,000,000 bbl. estimated reserves in 1922, 5,000,000,000 may be termed oil in sight and 4,000,000,000 prospective and possible. Something more than 4,000,000,000 belong to the heavy-oil type and will be recovered mainly in the Pacific Coast, Rocky Mountain, and Gulf states. The reserves of medium- and high-grade paraffin-base oils (as contrasted with the heavier oils) are estimated at 5,000,000,000 bbl.; the reserves of the high-grade oils of the Appalachian states are about 725,000,000 bbl.

Mexico's Petroleum Resources.—There are at present three main producing fields in Mexico: (1) the Panuco district, about 30 miles southwest of Tampico (2) the Dos Bocas-Alamo district, which extends from Dos Bocas to Alamo on the Tuxpam River; and (3) the Tehuantepec district in southern Mexico. The Panuco district crude contains about 27 per cent gasoline, 4 per cent kerosene, 14 per cent gas oil, 12 per cent lubricating oils, and 66 per cent asphalt. It has a specific gravity of about 12°Bé. and is often used as fuel oil after topping. The Dos Bocas-Alamo crude, which after topping is also used as a fuel oil, contains about 8 per cent gasoline, 9 per cent kerosene, 3 per cent gas oil, 1 per cent paraffin, and 74 per cent fuel oil. The average gravity is about 20.5°Bé. Whether an oil is used as a fuel oil or as a raw material for cracking depends on the production of crude oil and on the price of gasoline. The Tehuantepec crudes are of better grade than oils from the other two districts, and contain about 20 per cent gasoline, 25 per cent kerosene, 20 per cent light distillates, 20 per cent lubricating oils, and 15 per cent residuum. The gravity is about 32.5°Bé. There are promising indications of petroleum in other parts of the country.

Since over 85 per cent of the world's present production of petroleum comes from the United States and Mexico, detailed discussion of the petroleum resources of other nations has been omitted.

PETROLEUM PRODUCTION AND THE FUTURE OF THE INDUSTRY

Although petroleum was known to the ancients, and it is certain that the crude oil was long used by the Indians of North America, the modern petroleum industry did not really begin until about 1860, when the production jumped from about 6000 to over 500,000 bbl.* in a single year. The industry has grown phenomenally since that time, until in 1924 the world's production of crude petroleum was over 1,000,000,000 bbl., with a grand total of over 12,000,000,000 bbl. produced since 1857.

Figure 36 shows the petroleum[10] production of the main producing countries over the period from 1890 to 1924.

The figure clearly shows the major part played by the United States, which, with the exception of the period from 1898 to 1901 when Russia took first place, has been the biggest producer of petroleum. In 1923, for example, the United States produced over 730,000,000 bbl. of petroleum. This is an increase of 30 per cent over the United States production for 1922, is more than double the production of the United States in 1918, and is almost 40,000,000 bbl. more than the production of the entire world[11] in 1920.

The figure also shows the decline in Russian production due to internal political unrest, and the increasing production of Mexico. The decrease in Mexican production in 1922 was largely due to the fact that wells ruined by the invasion of salt water were not replaced by new discoveries.

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* 1 bbl. = 42 gal.

One of the factors which has aided the growth of the oil industry in the United States has been the huge pipe lines connecting the producing centers with the refineries and consuming centers. Although a large amount of both the crude and refined products are transported by rail and water, the 60,000 miles of oil pipe lines with capacities totaling over 800,000 bbl. daily have joined widely separated producing areas and allowed production to reach its present magnitude.

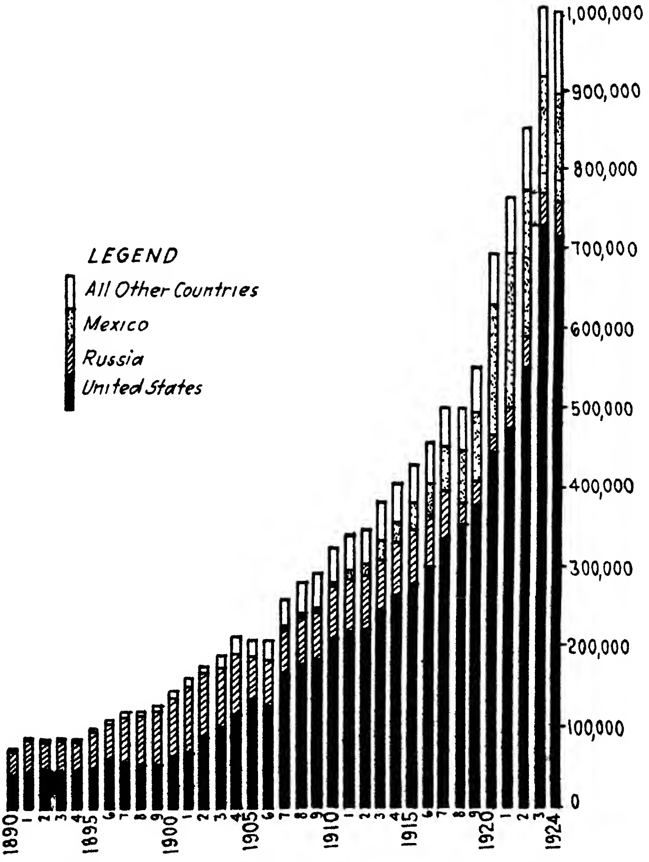

Fig. 36.—Petroleum production of the world from 1890 to 1924 in thousands of barrels. (*From data of U. S. Geol. Survey and Am. Petroleum Inst.*)

Taking the case of the United States, the world's largest producer of petroleum, as typifying the trend of petroleum production and exploitation, it is possible to predict, with a fair degree of accuracy, the future of the industry. In Fig. 37 is plotted the total United States production of petroleum from 1890 to 1924, and the proportion of this total contributed by each of the seven oil districts in the country. As the plot shows,* the Appalachian district is the oldest, as well as the most consistent producer. The American petroleum industry began with the drilling of the discovery well at Titusville, Pa., in 1869. Up until 1900, as the science of oil geology progressed, new discoveries more than offset the exhaustion of the field, but in the

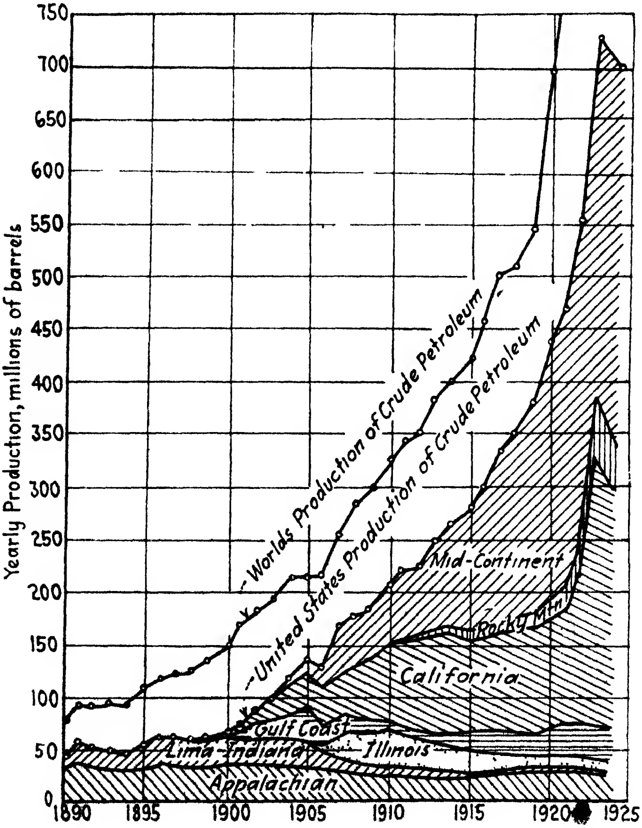

Fig 37.—Petroleum production of the principal districts in the United States from 1890 to 1924 (*From data of U. S. Geol. Survey and the Am. Petroleum Inst.*)

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* For a more complete treatment of the subject refer to "The Petroleum Outlook,"[15] by A. D. Little. The present discussion is based to a certain extent on subject matter in this reference. For a more optimistic view of our present petroleum supplies see reference 16.

last 20 years the situation has been reversed, and with practically the entire district tested (by drilling wells) it is highly improbable that the production in this field will materially increase. It is rather to be expected that the wells will continue in line with their past performances and gradually decline in output. This field, being free from water, has maintained its production by flooding suitable exhausted wells with water to continue the production of the remaining wells.

The second oldest district, the Lima-Indiana, reached its maximum production in 1896, only 10 years after its discovery, and has declined since, until in 1922 the production was less than one-tenth of the peak value. Like the Appalachian, the greater proportion of the oil possibilities have been investigated and increased production of any considerable magnitude cannot be expected in this field.

Development of the Illinois district began in 1904, and in 1908 the maximum production was reached. Since 1910 the decline has been quite rapid, the average rate of decrease being about 10 per cent per year. This field, like the Appalachian and Lima-Indiana, has been quite thoroughly prospected and no major increase is probable.

The Gulf Coast district began its commercial output in 1901 with the drilling of the famous Spindletop well. In 1905 the maximum was apparently reached and the production gradually dwindled, until in 1914 new wells caused increased activity, and, as shown in Fig. 37, the production in 1922 was practically equal to the peak production. Production in this field is erratic, and while there is considerable uncertainty as to the future of the district, there seem to be possibilities of somewhat extending the producing area by deeper drilling or by new discoveries in the Salt Dome region.[9]

The Rocky Mountain district has the largest average production per well in the country. The period of rapidly increasing production began with the discovery of the Salt Creek field in Wyoming, and in 1922 the production was about as large as that of the Appalachian district. There are a number of possibilities of expansion of the producing areas and with the waning of other districts it is practically certain that new pools will be discovered. The possibilities for increasing production in this field are good, and it is to be expected that deeper drilling will bring in additional oil. Production increased considerably in 1923 and 1924.

The California district for a number of years has been the first or second largest producer of petroleum in the country, and the output represents nearly a third of the total for the United States. The addition of new fields near Los Angeles has recently greatly increased the production, so that some of the older producing wells have had to be shut down. Although geologists have expressed the opinion that the producing areas of the district are well defined, the recent discoveries seem to indicate that the peak production has not yet been reached and that the field will continue in substantially its same position of importance for many years to come.

The Mid-Continent district is by far the largest producer of petroleum in the United States, its production of 306,000,000 bbl. in 1922 representing about 55.5 per cent of the total United States production and 36.4 per cent of the petroleum output of the entire world. Since the fields of this district cover such a wide area, and the surface evidence in this case does not permit definite predictions as to the future discoveries, the size and the value of prospective areas are largely matters of conjecture. It is fairly certain, however, that the possibilities of the field have not been entirely explored and that the field will continue in its position of leading producer for some time.

In summarizing the preceding paragraphs it will be seen that three of the seven petroleum-producing districts of the United States have reached the peak of their production and have started to decline. Another, the Rocky Mountain district, offers possibilities of marked expansion, while the other three, although unknown quantities and apparently not yet at their maximum, cannot continue their expansion and increasing production indefinitely. Moreover, it is the opinion of competent geologists that new producing areas of magnitudes corresponding to the present ones cannot be hoped for in the United States, and that the downward trend of the production curve for the Appalachian, Lima-Indiana, and Illinois districts represents the form of the composite curve for the whole country a few years hence. It is interesting to note that the production of petroleum in the United States in 1924 was about 714,000,000 bbl., or 2.5 per cent *less* than in 1923. This is the first decrease in yearly production since 1906.

The reserves shown on page 121 are enough to meet the present requirements for less than 20 years if the oil could be taken out of the ground as fast as it is wanted. The oil will not all be exhausted in 20 years,[14] however, since all the oil pools will not be found within that time, drilling will be spread over many years as the pools are found, and the wells cannot be pumped dry so quickly. Individual wells will continue to give oil for more than 20 years and it is probable that some of the wells will not have been drilled by 1950. On the other hand, the United States is already dependent on foreign sources to supplement her own production, and this dependence will increase as the United States production falls off, except as artificial petroleum is produced from oil shales or the distillation of coal, or some substitute for petroleum becomes commercially successful.

It should also be noted that the above estimates are based on present methods of recovery. However, it is practically certain that new methods, such as "air pressure," "vacuum extraction," etc., will allow the recovery of greater percentages of the oil underground than is possible with present practice.

Furthermore, with the stimulus of advancing prices (in addition to reworking wells now abandoned as unprofitable), it is certain that the vast potential reserve represented by oil-shale deposits will be utilized. At the present time, on account of the high cost of production from shales, oil from this source cannot be produced in the United States at any substantial profit. As prices advance, however, it will become economically possible to obtain petroleum from this source and our workable oil resources will thus be greatly increased. It has been estimated, for example, that a sixfold increase in price would cause an increase of over fifty fold in the petroleum reserves which could be worked profitably.

Having seen the relatively short span of life estimated for the petroleum reserves of the United States, it is interesting to review the world oil situation. It is estimated that the United States has about 12 per cent of the world's recoverable supply, and Mexico 7.5 per cent, most of which is American owned. Outside of these two countries, the United States control about 2 per cent, making the total reserves under American controls about 18 per cent. These resources must be contrasted with those of Great Britain which, up to about 1914, had practically no petroleum reserves. At the present time, through various means, Great Britain controls over 70 per cent of the world's known petroleum areas.

Other Indications of the Future of Petroleum.—In addition to the figures on petroleum resources and production of crude petroleum, the statistics of the refining industry also give an interesting commentary on the future of the industry. In the early days of petroleum refining, the demand for gasoline was very small, and it was essentially a by-product in the manufacture of kerosene. Today, however, the situation has been reversed, and in the petroleum industry, really the motor-fuel industry, gasoline is the most valuable product, and kerosene production has been decreased in order to produce the lighter gasoline by "cracking." About 20 per cent of gasoline can be obtained from the average crude by straight refining methods, while the demand for motor fuel is so great that even at the present time other fractions of the oil have to be cracked. The kerosene fraction cannot be reduced further on account of the necessity for that product. The lubricating fraction is as necessary as the gasoline, thus leaving for further encroachments of the cracking process, first, gas oil, and then fuel oil. The first of these has already taken place and considerable gas oil is being cracked to give gasoline. The same is true of fuel oil and it is to be expected that the process will be continued. With the continually expanding automotive industry, and the possibility of petroleum production not being able to keep step with this advance, it can be seen that questions of petroleum conservation, not the least important of which lies in the proper utilization of fuel oil, should be matters of concern to everyone.

While there is no doubt as to our ability to produce liquid fuels, on the basis of present data it seems probable that the next century will see the production of liquid fuel in large quantities commercially from sources other than petroleum. Whether the distillation of shale, the treatment of coal (carbonization, hydrogenation or via water gas) or some other process will be the most economical cannot be said at the present time.

BIBLIOGRAPHY

1. Am. Soc. Test. Materials, *Rept.* of Subcommittee XX of Committee D-2, Nov. 2 (1923).
2. Clarke: "The Data of Geochemistry," U. S. Geol. Survey, *Bull.* 695 (1920).
3. Bacon and Hamor: "The American Petroleum Industry," McGraw-Hill Book Co., Inc. (1916).
4. Brooks: "Chemistry of the Non-benzenoid Hydrocarbons," Chemical Catalog Co. (1922).

5. *Petroleum,* **1,** 399 (1912).
6. DAY: "Handbook of the Petroleum Industry," John Wiley & Sons (1922).
7. GILBERT and POGUE: "America's Power Resources," the Century Co. (1921).
8. WADSWORTH: "Efficiency in the Use of Oil Fuel," U. S. Bur. Mines (1919).
9. GARFIAS: "Petroleum Resources of the World," John Wiley & Sons (1923).
10. Petroleum in 1922, U. S. Geol. Survey, *Mineral Resources,* Part 2 (1924) (and information from Am. Pet. Inst.)
11. "Preliminary Summary of Statistics of Crude Petroleum for 1923," U. S. Dept. of the Interior, Jan. 30 (1923).
12. "World Atlas of Commercial Geology," U. S. Geol. Survey (1921).
13. *The Lamp,* **6** (October, 1923) the Standard Oil Co. of N. J.
14. *Bull.* Am. Assoc. Petrol. Geol., **6,** 42 (1922).
15. "The Petroleum Outlook," published by A. D. Little, Inc.
16. "American Petroleum, Supply and Demand," a report to the Am. Pet. Inst. McGraw-Hill Book Co., Inc. (1925). A very complete discussion of the entire question of the future demand for petroleum products and how the demand will be met.

CHAPTER VII

OTHER PRIMARY FUELS

In addition to coal and fuel oil, a number of other fuels* are used to a limited extent for industrial heating, or are destined to be used more extensively as fuels as our supplies of petroleum and our best coal seams are depleted. Such fuels as shale oil, natural gas,† wood, various wood products, tanbark, corn, straw, bagasse or extracted sugar cane, briquets, and colloidal fuel, though much less important than either coal or oil, do find a certain application. The properties of these fuels which are important from the viewpoint of combustion are described in this chapter.

Shale Oil.—As pointed out in the preceding chapter, one of the most promising future substitutes for petroleum is the oil derived from oil shales. Oil shales are brown or black sedimentary rocks containing numerous fragments of organic matter, such as fossil plants and animals. This organic matter, on distillation, yields nitrogen compounds and oil.

Oil shale is mined somewhat like coal, is crushed and classified, and fed to stills or retorts. Heat is usually supplied internally to the retorts, part of the products of distillation being used to furnish the necessary fuel. The part of the retort in which oil gases are driven off from the incoming fuel is maintained at not over 900°F. (480°C.). In the waste end of the retort, however, where the oil has been mostly removed from the shale, the temperatures run somewhat higher (1600 to 1750°F., or 872 to 955°C.). A great many different types of retorts have been proposed and used, as, for example, continuous and intermittent vertical retorts, fired internally and externally; horizontal and inclined retorts of various types have also been used. The gases from the retort are condensed, scrubbed, and refined in a manner

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* It will be noted that a number of these are, in reality, secondary or manufactured fuels, and have only been introduced at this point for convenience of presentation.

† The various industrial, manufactured gases are discussed in Chaps. XI, XVIII, XIX, and XX.

similar to that employed in the refining of petroleum. The main products obtained are: (1) gasoline, (2) kerosene, (3) lubricating oil, (4) fuel oil, and (5) considerable quantities of combustible gases. The average yield of crude oil per ton of shale may vary from 20 to over 60 gal. per ton, the average for the Scottish shale industry being about 25 gal. In Scotland and France (the main countries in which large-scale production of oil from shale has so far been economically possible), a large proportion of the nitrogen compounds present in the shale are recovered and marketed as ammonium sulfates. In Scotland, for example, the average yield is about 35 lb. of ammonium sulfate per ton of shale, or 60 per cent of the original nitrogen content. The nitrogen content of American shales is said to be about the same as in the Scottish deposits.*

In spite of the tremendous deposits of shale in the United States there has been practically no successful exploitation on account of the present low prices of petroleum and on account of the time required for developing and putting in operation methods of distillation applicable to American shales. The development of the American oil-shale industry[1] to a point comparable with the present petroleum industry† would require a mining activity comparable in size to the coal-mining industry. The development and the training of such a huge army of workers will, obviously, require a number of years.

Natural Gas.—Natural gas is of similar origin, and usually occurs with petroleum. It consists mainly of methane (CH_4), with smaller quantities of other hydrocarbons, particularly ethane, although carbon dioxide, carbon monoxide, oxygen, and nitrogen are usually present in small amounts. The analyses of natural gas from several different localities are given in Table LXVI (Chap. XI).

Natural gas is an ideal fuel for industrial heating on account of its cleanliness, ease of transportation, high heat content (from 700 to 1400 B.t.u. per cubic foot), and the high flame tempera-

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* For a complete discussion of the subject of Oil Shale, see References 1, 2, 3, and 4.

† Assuming an oil yield of 42 gal. of oil per ton of shale retorted, over 700,000,000 tons of shale would have to be mined yearly to duplicate the petroleum production of the United States in 1924. This is approximately 10 per cent greater than the yearly production of bituminous coal in the United States (658,000,000 tons in 1920).

ture possible. It has been widely used for both industrial and domestic heating, but the declining supply since 1917 (when about 800,000 million cubic feet were produced) has necessitated its replacement by other fuels to an increasing extent. In the future, to eke out the rapidly diminishing supply, natural gas will be used to enrich the lower-heating-value manufactured gases.

Considerable quantities of "casing-head gasoline" are produced from natural gas by removing the hydrocarbons which are liquid at ordinary pressure and temperature, but normally are present as vapor in natural gas and in gas from oil wells. This casing-head gasoline is used mainly for blending with the gasoline-kerosene fractions from petroleum distillation in order to bring heavy fractions up to gasoline standards. The main methods of recovery are: (1) absorption in straw oil, the gasoline being subsequently driven off by heat; (2) direct absorption of the casing-head constituents in naphtha to form the desired blend; (3) separation of the casing-head gasoline by compressing the slightly cooled natural gas; and (4) absorption by highly activated charcoal, the gasoline being removed from the charcoal by a subsequent treatment with superheated steam. The annual recovery of casing-head gasoline from natural gas in the United States is in excess of 500,000,000 gal.

Wood.—Wood is commonly used as a fuel by sawmills and woodworking plants, and has supplied from 10 to 15 per cent of the total fuel used in the United States (*i.e.*, including domestic heating). Industrial wood fuel consists mainly of sawdust, shavings, etc., though the more expensive cord wood has been used to a considerable extent, either in times of coal shortage or in such industries as lime burning, where a long, mellow flame is advantageous.

When freshly felled, wood contains from 40 to 50 per cent of moisture, and seasoning or drying is, therefore, important. Air drying reduces the moisture content to from 15 to 25 per cent and gives a fuel with a heating value* of from 5400 to 6800 B.t.u. per lb., as shown by Table XXXVII.

On the basis of heating value, 1 lb. of coal is the equal of about 2½ lb. of wood. The heating value of sawdust or refuse is, of course, the same as that of the original wood.

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* The heating value of bone-dry wood varies from about 7000 to 9900 B.t.u. per lb.

TABLE XXXVII

WEIGHTS AND COMPOSITIONS OF AIR-DRIED WOODS[5]

Wood	Pounds per cu. ft	Pounds per cord	Per cent					B.t.u. per pound
			Carbon	Hydrogen	Oxygen	Nitrogen	Ash	
Ash	46	3,520	49.18	6.27	43.91	0.07	0.57	5,420
Beech	43	3,250	49.36	6.01	42.69	0.91	1.06	5,400
Birch	45	2,880	50.20	6.20	41.62	1.15	0.81	5,580
Elm	35	2,350	48.99	6.20	44.25	0.06	0.50	5,400
Oak	52	3,850	49.64	5.92	41.16	1.29	1.97	5,460
Pine	30	2,000	50.31	6.20	43.08	0.04	0.37	6,700
Poplar	36	2,130	49.37	6.21	41.60	0.96	1.86	6,660
Willow	25	1,920	49.96	5.96	39.56	0.96	3.37	6,830

Charcoal.—Charcoal is the product left as the residue from the destructive distillation of wood. It is at times used as a fuel where a quick, clean fire is needed and has a heating value of about 11,000 B.t.u. per lb. It ignites easily, burns cleanly and without smoke. Some charcoal is still used for blast-furnace work.

Tanbark.—Tanbark is the fibrous portion of the ground oak or hemlock bark used in tanning. After having been leached in the tanning process, the "spent tan" contains about 65 per cent of moisture. The ultimate analysis[6] of dry tanbark averages: carbon, 51.8 per cent; hydrogen, 6.04 per cent; oxygen, 40.74 per cent; ash, 1.42 per cent; the heating value of the dry bark is about 9500 B.t.u. per lb. Thus a tanbark containing 65 per cent moisture has a heating value of 0.35 × 9500 = 3320 B.t.u. per lb. Before firing, the moisture content of the bark may be reduced by mechanical pressing or by a partial drying by means of the waste heat in the flue gases.

Straw.—Straw is sometimes used as a fuel. The composition[5] is approximately 36 per cent carbon, 5 per cent hydrogen, 38 per cent oxygen, 0.5 per cent nitrogen, 15.75 per cent moisture, and 4.75 per cent ash, with a heating value of 5400 B.t.u. per lb.

Bagasse.—Bagasse is the fibrous refuse left after the grinding or extraction of sugar cane. It represents about 20 per cent of the weight of the original cane and is widely used as a fuel on sugar plantations. The composition and heating value depend somewhat on the amount of fiber and other combustible material present, such as sucrose, gums, and resins contained in the cane. Bagasse is composed of from 35 to 45 per cent fiber, 7 to 10 per cent sucrose and other combustible material, and from 45 to 55 per cent moisture. The average composition of dry bagasse is 45 per cent carbon, 6 per cent hydrogen, 46 per cent oxygen, and 3 per cent ash, with a heating value of about 8300 B.t.u. per lb. On account of the moisture, the heating value of the bagasse as fired is usually around 4000 B.t.u. per lb. and for this reason a number of attempts have been made to reduce the water content by means of flue-gas waste-heat driers.

Bagasse can be burned with good efficiencies at high rates of combustion (from 200 to 300 lb. per sq. ft. of grate area per hr.) and the yield in many cases is sufficient to meet the entire fuel demands of the sugar plantation. At times, however, it has to be supplemented with other fuels, such as wood, coal, or fuel oil.

Briquets.[7]—One of the major problems in coal conservation lies in the utilization of the dust, culm, slack, and other small-sized waste coal produced in mining. These, and various other low-grade fuels, such as lignite and peat, can be best handled and transported in the form of briquets or compressed solid blocks. The briquetting industry is quite extensive in Europe, where coal is scarce and much of it of low grade. Thus, about 10 per cent of Germany's total coal production is marketed in the form of briquets. In the United States, on the other hand, the expense of briquetting makes it difficult for this form of fuel to compete with the relatively cheap, high-grade coals. With the depletion of our best coal seams, however, briquetted fuel will become of more importance.

The term "briquet" includes compressed fuel made from coal, lignite, peat, or coke, either with or without binder. Briquets are mostly machine made, the fine coal being mixed with a binder and the resulting mixture subjected to a high pressure (from 1800 to 3000 lb. per sq. in.) in a roll press. In some instances a part of the volatile matter of the coal and the binder is driven off by a mild baking. This is done to produce a more nearly smokeless fuel and to give a harder, less friable briquet.

Briquets are made in various sizes and shapes. Large rectangular briquets weighing up to 7 lb. are quite extensively used in Europe. They occupy from 10 to 20 per cent less space for storing than lump coal, but have the disadvantage of requiring labor to break them before firing. In the United States 1¾- to 2½-oz. briquets are used for domestic heating and the 2½- to

5-oz. sizes are used for industrial purposes. These briquets are usually either egg-shaped or of the "overstuffed pillow" shape (*i.e.*, with rounded edges) to minimize breakage. The rounded edges have the further advantage of permitting good air circulation, and therefore aiding combustion.

Good briquets should be as hard and break as little during handling as anthracite coal. They must not deteriorate on weathering, and should be essentially waterproof. To meet these requirements the proper binder must be used. Asphalt pitch and pitch made from coal tar or water-gas tar are the customary binders,* and give good results, particularly when the volatile and objectionable cresols are distilled by a preliminary treatment. Various natural hydrocarbons, such as gilsonite, molasses, starch, or dextrin pastes, and sulfite liquor (a by-product in the manufacture of wood pulp by the sulfite process) also have been successfully used. Usually from 5 to 8 per cent of binder is required to make a hard, durable briquet.

Anthracite, semi-bituminous and bituminous coals, and carbocoal (a coke left from the low temperature distillation of coal) have been successfully briquetted in the United States. Lignite and peat briquets, although quite widely manufactured in Europe, have not been produced to any extent in America.

Colloidal Fuel.[8]—Colloidal fuel, an emulsion of powdered solid fuel and oil, was developed by the Submarine Defense Association during the war. The solid fuel may be any of the coals from lignite to anthracite, or peat, coke, or wood, provided at least two-thirds of the dry solid fuel is combustible. It is first pulverized so that 95 per cent will pass through a 100-mesh and 85 per cent through a 200-mesh screen. The solid fuel and the oil are then run into a premix tank, to which is also added a small amount of a "fixateur" to prevent the heavier solid-fuel particles from settling out. The contents of the premix tank are then ground in a ball or tube mill to promote more intimate mixing. Various fuel oils, pressure-still† oil or tar and coal tar have all been successfully used. A lime-rosin grease,‡ or one of the coal tar

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* These binders were used in about 80 per cent of the briquets produced[9] in the United States in 1922.

† Pressure-still oil or tar is the residue left from cracking an oil in a pressure still.

‡ This grease can be prepared by heating 83.5 per cent oil, 10 per cent rosin, 5 per cent lime, and 1.5 per cent water.

distillates, such as creosote, may be used as a fixateur and if the colloidal fuel is properly prepared the solid particles will stay in suspension for several months.

The compositions of two types of colloidal fuel are given in Table XXXVIII.

TABLE XXXVIII

COMPOSITIONS OF TWO TYPES OF COLLOIDAL FUEL

	Grade 3, per cent	Grade 4, per cent
Coal	30.0	30.0
Coal tar	....	12.0
Fixateur	1.5	1.2
Mexican reduced oil	28.8	
Texas Navy oil	8.5	6.8
Pressure-still oil	31 2	50.0
	100.0	100.0

The heating values were 17,200 and 16,670 B.t.u. per lb. for grades 3 and 4, respectively.

Colloidal fuel may be fired in ordinary oil-firing equipment and possesses many of the advantages of oil fuel. A large amount of heat is radiated from the flame of colloidal fuel, owing to the incandescent solid particles. It has the further advantage of conserving our supply of oil fuel.

BIBLIOGRAPHY

1. M. J. GAVIN: "Oil Shale," U. S. Bur. Mines, *Bull.* 210 (1924). (This reference contains an excellent, selected bibliography.)
2. "World Atlas of Commercial Geology," U. S. Geol. Survey (1921).
3. *Chem Met. Eng.*, **32**, 237, 324 (1925).
4. Publications of the Colorado School of Mines.
5. "Steam Boiler Engineering," Heine Safety Boiler Co., *Bull.* 27, p. 473 (1920).
6. D. M. MYERS: "The Economic Combustion of Waste Fuels," U. S. Bur. Mines, *Tech. Paper* 279 (1922).
7. For a detailed discussion of briquetted fuel, the reader is referred to BACON and HAMOR: "American Fuels," pp. 224–266, McGraw-Hill Book Co., Inc. (1922).
8. Colloidal fuel is discussed at length in reference 7 See also J. T. DUNN: "Pulverized and Colloidal Fuel," D. Van Nostrand Co. (1924).

CHAPTER VIII

THE CHEMISTRY OF THE COMBUSTION REACTIONS

There are certain laws of chemistry which deal with the most vital questions in the field of combustion, namely: (1) how much heat will be developed through the operation of any given combustion process or reaction; (2) to what extent will the reaction go, *i.e.*, will all the fuel be consumed if sufficient time be allowed; and (3) how fast will the reaction proceed, or how much fuel will be consumed in a given time under definite conditions? The final answers to these questions are obtained only through the study of the controlling chemical laws, particularly those affecting chemical equilibrium and the rate or velocity at which chemical reactions proceed.* This chapter includes a general discussion of the basic chemical laws, and the application of these laws to the most important individual combustion reactions. The number of chemical reactions of importance are relatively few, those most frequently encountered being as follows:†

$$C + O_2 = CO_2 + 174{,}600 \text{ B.t.u.} \quad (1)$$
$$C + CO_2 = 2CO - 70{,}200 \text{ B.t.u.} \quad (2)$$
$$2C + O_2 = 2CO + 104{,}400 \text{ B.t.u.} \quad (3)$$
$$2CO + O_2 = 2CO_2 + 244{,}800 \text{ B.t.u.} \quad (4)$$
$$2H_2 + O_2 = 2H_2O + 246{,}200 \text{ B.t.u.} \quad (5)$$
$$C + H_2O = CO + H_2 - 70{,}900 \text{ B.t.u.} \quad (6)$$
$$C + 2H_2O = CO_2 + 2H_2 - 71{,}600 \text{ B.t.u.} \quad (7)$$
$$CO + H_2O = CO_2 + H_2 - 700 \text{ B.t.u.} \quad (8)$$

Heats of Reaction.—When substances react, heat is always either given off or absorbed. This heat is called the "heat of

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* For a more extended discussion of this subject see such standard textbooks as Millard[1] or Washburn.[2]

† Heats of reaction at 60°F. and at constant pressure. Data from "Landolt-Bornstein Tabellen." Reaction heats are given in B.t.u. per lb.-mol. and may be converted to calories per gm.-mol. by dividing by 1.8. Another widely used unit is the amount of heat required to heat 1 lb. of water 1°C. This is called the Centigrade heat unit (C.h.u.) or pound Centigrade unit (p.C.u.). See Conversion Factors in Appendix.

reaction" and is here expressed as the British thermal units* of heat evolved or absorbed per lb.-mol† of reacting material. Thus, reaction (1) given above

$$C + O_2 = CO_2 + 174{,}600 \text{ B.t.u.} \qquad (1)$$

not only expresses the fact that one molecule of carbon reacts with one molecule of oxygen to form one molecule of carbon dioxide, but it also indicates that when 1 lb.-mol of carbon (12 lb.) reacts with 1 lb.-mol of oxygen (32 lb.) the heat *liberated* is equal to 174,600 B.t.u. When a reaction takes place with the *absorption* of heat the equation is written with a *minus* sign; thus

$$C + H_2O = CO + H_2 - 70{,}900 \text{ B.t.u.} \qquad (6)$$

indicates that 70,900 B.t.u. are absorbed and must be supplied from an outside source when 1 lb.-mol of carbon (12 lb.) reacts with 1 lb.-mol of water (18 lb.).

The heat of a reaction varies considerably, depending on whether or not the reaction is carried out at constant pressure (as, for example, when a gas is burned in a furnace hearth) or whether the reaction is carried out at constant volume (as in the determination of the heat content of a fuel in a bomb calorimeter). Heats of reaction at constant pressure may be converted to heats of reaction at constant volume by the equation.‡

$$Q_p = Q_v - (\Delta n)RT, \qquad (9)$$

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* A British thermal unit (usually abbreviated to B.t.u.) is one one hundred and eightieth of the amount of heat required to raise 1 lb. of water from 32 to 212°F. For all practical purposes, however, it is sufficiently accurate to regard the B.t.u. as the amount of heat required to raise 1 lb. of water 1°F.

† A pound-mol is the number of pounds of a substance numerically equal to the molecular weight. *For all gases,* 1 *lb.-mol at* 32°*F. and at an absolute pressure of* 760 *mm. occupies* 359 *cu. ft.* This is a most important relationship. The authors are familiar with the fact that at present most engineers do not use the pound-mol as a unit in combustion calculations. However, the use of this unit so simplifies combustion work, and, after the pound-mol is understood clearly, makes the necessary calculations so easy, that they have adopted it throughout. Extensive use of this concept is found in Chap. X.

‡ This equation holds rigidly only when the volume of any condensed phase (such as water formed from the combustion of hydrogen in coal) is negligible and when the gases in question obey the gas laws, *i.e.*, they are perfect gases and are not vapors near the point of condensation.

where Q_p equals the heat of reaction at constant pressure; Q_v equals the heat of reaction at constant volume; (Δn) equals the increase in the number of molecules of gas present after the reaction occurs, R, the gas constant, equals 1.9885, or approximately 2.0; T equals the temperature at which the reaction is carried out in degrees Fahrenheit absolute (degrees Fahrenheit + 460). For the reaction $2CO + O_2 = 2CO_2 + 244{,}800$ B.t.u., Δn equals -1, and the heat of combustion in a closed bomb (or at constant *volume*) is, by Eq. (1), as follows:

$$Q_v = 244{,}800 + (-1)\,(2)\,(60 + 460) = 244{,}800 - 1040 = 243{,}740 \text{ B.t.u. (approx.)}$$

Thus the heat evolved by this reaction at constant volume is 1060 B.t.u. less than if the same amount of CO were burned in a furnace at constant pressure.

Also, the exact value for the heat of reaction depends on the temperature at which the reaction occurs. This is an extremely important point in combustion work, due to the high temperature at which many of the reactions are carried out. The heat of reaction at one temperature may be converted to that at any other temperature by the equation*

$$Q_2 = Q_1 + (b - a) \tag{10}$$

where Q_2 is the heat of reaction in B.t.u. at T_2°F. absolute, Q_1 is the heat of reaction in B.t.u. at the temperature T_1°F. absolute, a is the heat evolved in raising the reacting substances from T_1 to T_2, and b is the heat evolved in raising the products of the reaction from T_1 to T_2. The expression $(b\text{-}a)$ may be evaluated most readily from Fig. 49, Chap. X, showing the heat content of gases at varying temperatures. For example, the heat of the reaction of $2CO + O_2 = 2CO_2$ is 244,800 B.t.u. at 60°F. (15.6°C), but at 2200°F. (1204°C.) the heat of reaction, using the heat content chart (Fig. 49) is

$$Q_{2200} = 244{,}800 + (-)(2)\,(24{,}600) - (-)(3)(15{,}700)$$

or

$$Q_{2200} = 242{,}700 \text{ B.t.u.,}$$

where the minus signs in front of each term within the parentheses are due to the fact that heat is absorbed in raising the CO, O_2, and CO_2 from 60 to 2200°F. (15.6° to 1204°C.) instead of being

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* This equation follows from the first law of thermodynamics, *i.e.*, the energy change is dependent only on the initial and final states and is independent on the path between those states.

evolved; 2 equals number of mols of CO_2, the heat content (above 60°F.) of which at 2200°F. (1204°C.) is 24,600 B.t.u.; and where 3 is the number of mols of reacting substances (2 mols CO plus 1 mol O_2), the heat content of each mol (above 60°F.) being 15,700 B.t.u. at 2200°F. (1204°C). All heats of reaction in this book unless otherwise noted are for constant pressure and at 60°F. (15.6°C.).

TABLE XXXIX

HEATS OF REACTION OF MOST IMPORTANT COMBUSTION REACTIONS
(B.t.u.)

Reaction	1	2	3	4
$C + \frac{1}{2}O_2 = CO$	+ 52,200	+ 53,200	+ 39,400	+ 15,300
$C + O_2 = CO_2$	+174,600	+175,000	+154,800	+106,600
$C + CO_2 = 2CO$	− 70,200	− 68,600	− 95,800	
$C + H_2O = CO + H_2$	− 70,900	− 54,500	− 96,500	
$C + 2H_2O = CO_2 + 2H_2$	− 71,600	− 40,400	−117,000	
$CH_4 = C + 2H_2$	− 36,800	− 49,600	− 69,800	
$CH_4 + 2O_2 = CO_2 + 2H_2O$	+384,000	+340,800	+295,000	+198,600
$C_2H_4 = 2C + 2H_2$	+ 27,000	+ 13,300	− 13,400	
$C_2H_4 + 3O_2 = 2CO_2 + 2H_2O$	+622,400	+578,700	+513,600	+369,000
$2CO + O_2 = 2CO_2$	+244,800	+243,600	+205,200	+157,000
$2H_2 + O_2 = 2H_2O$	+246,200	+215,400	+177,000	+128,800
$CO + H_2O = CO_2 + H_2$	− 700	+ 14,100	− 33,300	

Column 1. Heat evolved at 60°F. (15.6°C.), products cooled to 60°F. (15.6°C.) (water condensed).

Column 2. Heat evolved at 1832°F. (1000°C.) with initial and final compounds at 1832°F. (1000°C.).

Column 3. Heat available above 1832°F. (1000°C.) with initial compounds at 60°F. (15.6°C.), final at 1832°F. (1000°C.)

Column 4. Heat available above 1832°F. (1000°C.), using air instead of O_2.

Data.—Molal heat capacities from total heat chart (Fig. 49, Chap. XI). Reaction heats from "Landolt Bornstein Tabellen" (1923 edition). Molal specific heat of carbon considered equal to 4.2. MC_p of $CH_4 = 7.5 + 0.00277T$, where T is in degrees Fahrenheit absolute. MC_p of $C_2H_4 = 9.380 + 0.0059t$ (from BACON and HAMOR), where t is in degrees Fahrenheit.

Note.—Heat evolved or available indicated by plus (+) sign. Heat absorbed or which must be supplied indicated by a minus (−) sign. Other heats of reaction are given in Table LII (p. 205).

The heats of reaction for the most important combustion reactions under varying conditions are shown in Table XXXIX. In column 1 are given the heats evolved at 60°F. (15.6°C.), products cooled to 60°F. (15.6°C.). In column 2 are given the heats evolved at 1832°F. (1000°C.), the initial and final compounds being at 1832°F. (1000°C.). Columns 3 and 4 show the heats available above 1832°F. (1000°C.), using oxygen and air respectively, the initial compounds being at 60°F. (15.6°C.) and the final at 1832°F. (1000°C.). The table shows the marked differences under the varying conditions.

Reaction Equilibrium.—Many substances, when mixed in definite proportions, combine almost completely. For example, at the temperatures met with in the fuel bed, carbon and oxygen, if present in the proportion of 1 mol of each (12 gm. or lb. of carbon and 32 gm. or lb. of oxygen) react quantitatively to form carbon dioxide according to the reaction, $C + O_2 = CO_2$. This type of reaction is called an *irreversible reaction.* On the other hand, if 1 mol of carbon monoxide and 1 mol of water vapor are mixed, these substances will not react completely according to the equation $H_2O + CO = CO_2 + H_2$; *i.e.,* part of each mol of carbon monoxide and water vapor will be left uncombined. Furthermore, this last reaction is a *reversible* reaction; that is, it will go in either direction, depending on the concentrations of the substances on the right-hand side of the equality sign as compared with the concentrations of the substances on the left-hand side. However, except at the very highest of flame temperatures the majority of combustion reactions are reversible only to a very slight degree.

If more than 1 mol of water vapor is mixed with 1 mol of CO, more CO_2 and H_2 will be formed and *less* CO will be left over than if only 1 mol of H_2O were used. This effect is explained by the *Law of Mass Action,* which states that increasing the concentration of one of the reacting substances increases the amount of product that may be formed. However, in a reversible reaction, such as

$$H_2O + CO = CO_2 + H_2,$$

no matter how much the concentration of the water vapor is increased, there always will be a definite relationship between the concentration of the reacting substances (H_2O and CO) and the products of the reaction (CO_2 and H_2). This relationship can be expressed by the equilibrium equation:

$$\frac{(CO_2)(H_2)}{(H_2O)(CO)} = K, \tag{11}$$

where K is known as the *equilibrium constant.* The numerical value of K for any given reaction depends only on the temperature and the units, and is independent of all other conditions. This means that at any given temperature, after the chemical reaction has taken place and equilibrium between all the reactants and products is reached, the final composition resulting from any initial mixture of CO_2, H_2, CO, and H_2O, in the above example, is fixed. The numerical values for CO_2, H_2, etc. may be either in terms of partial pressures for each gas, as atmospheres or millimeters of mercury, or they may be expressed as concentrations, *e.g.*, as mols of each gas per unit volume of total gas. If, due to the reaction, there is no change in the number of gaseous molecules, then the numercial value of K is independent of the units of pressure or concentration used. Otherwise, changing from partial pressures to concentrations always changes the numerical value of the equilibrium constant K. It is usually convenient to express the numerical values of the components of such a reaction system as per cent by volume. Thus, if CO_2 and H_2 are caused to react together and, after the reaction has ceased, equilibrium having been reached, the composition of the gases is

	Per Cent by Volume
CO_2	47.50
CO	22.82
H_2O	22.82
H_2	6.85
	99.99

then the equilibrium constant may be calculated as

$$K = \frac{(CO_2)(H_2)}{(H_2O)(CO)} = \frac{(47.50)(6.85)}{(22.82)(22.82)} = 0.623.$$

The numerical value of the equilibrium constant in this case i independent of the units used to express the concentration of th reacting substances and products, since in this case the tot number of molecules does not change. Equilibrium values ar often expressed in partial pressures as millimeters of mercury, c as concentrations in mols per cu. ft.,* or in parts per 100 par (as in the example above), and the like.

In calculating the equilibrium constant, if more than on molecule of a given substance enters the reaction the parti pressure or the concentration of that substance must be raise to the power numerically equal to the number of molecules c

* See footnote on p. 138

the substance that enter into the equation. Thus, for the reaction $C + CO_2 = 2CO$, the equilibrium constant is written

$$\frac{(P_{CO})^2}{P_{CO_2}} = K, \tag{12}$$

where K is here expressed in partial pressures. When the chemical reaction involves a solid the partial pressure or concentration of the solid is omitted, since the partial pressure of the solid in the gaseous phase is independent of the amount of solid present. If, for example, a mixture of C and CO_2 reacted together at atmospheric pressure (29.9″ Hg pressure), and after equilibrium was reached the composition of the gas was 74 per cent CO and 26 per cent CO_2, then the equilibrium constant expressed in inches of mercury would be

$$K = \frac{(P_{CO})^2}{P_{CO_2}} = \frac{\left(\frac{74}{100}\,29.5\right)^2}{\left(\frac{26}{100}\right)(29.5)} = 62.$$

Since there is a change in the total number of gaseous molecules when the reaction $C + CO_2 = 2CO$ takes place, it is obvious that K is not the same numerically when the equilibrium constant is expressed in atmospheres or if concentrations are used instead of partial pressures.

The final equilibrium composition of a mixture of gases depends also on the temperature. In other words, the value of the equilibrium constant K for a given system changes with temperature and the equation approximately expressing this change may may be written in the form

$$\log K = \frac{A}{T} + B. \tag{13}$$

A knowledge of the constants A and B for each reaction is highly desirable, since the value of the equilibrium constant at any temperature is then known and the extent to which a mixture of substances can react ultimately may be calculated for the temperature in question. The variation with temperature of the equilibrium constants for the more important combustion reactions is given in the latter part of the chapter, under the specific discussion of the individual reactions.

Rates of Combustion or Reaction Velocities.—Although some of the combustion reactions take place almost instantaneously, others require a measurable amount of time before they are completed. Thus, in addition to considering the equilibrium of a reaction, it is also necessary to know the *velocity or speed of the reaction.* This factor is of extreme importance. If the absolute velocity of a reaction cannot be established, data showing the velocity *relative* to other reactions may be of great value.

Reaction speed or rate may be defined as the change in concentration* dC of the reacting substances in the time dt, divided by that time:

$$\text{Reaction rate} = -\frac{dC}{dt}. \tag{14}$$

From experimental studies it has been determined that the rate of reaction is proportional to the concentrations of the reacting substances raised to the power numerically corresponding to their coefficients in the reaction.† For example, in the general equation representing the reaction of the substances A and B to form the substances M and N:

$$aA + bB \rightleftarrows mM + nN \tag{15}$$

the velocity of the reaction *toward the right* would be

$$-\frac{dA}{dt} = \frac{a}{b}\left(-\frac{dB}{dt}\right) = k_1(A)^a(B)^b, \tag{16}$$

the terms A and B signifying the concentrations of the reacting substances A and B at the time t, a and b being the molecular proportions of A and B respectively, and k_1 the "reaction rate coefficient," which is equal to the change in concentration in unit time if the concentrations of the reacting substances are unity. The actual value of k_1 depends on the specific reaction and on the temperature. If one of the reacting substances is a solid, k_1 also depends on the nature and the amount of the solid surface.

If, as frequently happens, the reaction does *not* go to completion toward the right, the rate of the reaction *toward the left* is important and this rate is expressed as

$$-\frac{dA}{dt} = \frac{a}{m}\left(-\frac{dM}{dt}\right) = \frac{a}{n}\left(-\frac{dN}{dt}\right) = k_2(M)^m(N)^n. \tag{17}$$

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* In the subsequent discussion of reaction rates, concentrations are considered to be in mols per cu. ft.

† The coefficients must correspond to the equation expressing the mechanism of the reaction and not to the equation that merely expresses the proportional relationship between the substances.

The net rate of the *final* reaction *toward the right* is then represented by

$$-\frac{dA}{dt} = k_1(A)^a(B)^b - k_2(M)^m(N)^n. \tag{18}$$

If, however, the reaction goes practically to completion toward the right, then k_2 becomes small in comparison with k_1, and may be neglected and the resultant equation for reaction rate is equation (16).

If the reaction is allowed to proceed until a state of equilibrium between the substances is reached, then the rate at which A is being *formed* just equals the rate at which A is being *decomposed*, and the net rate of change in the concentration of A becomes zero.

Therefore, at equilibrium:

$$-\frac{dA}{dt} = 0 = k_1(A)^a(B)^b - k_2(M)^m(N)^n \tag{19}$$

or

$$k_1\,(A)^a(B)^b = k_2\,(M)^m\,(N)^n \tag{20}$$

and

$$\frac{k_1}{k_2} = \frac{(M)^m(N)^n}{(A)^a(B)^b} = K \text{ (the equilibrium constant).} \tag{21}$$

That is, the equilibrium constant K numerically equals the rate of reaction constant in one direction divided by the rate of reaction constant in the opposite direction, or,

$$K = \frac{k_1}{k_2}. \tag{22}$$

Velocity of First-order Reactions.—When the concentration of only one reacting substance is changing, then only one molecule of the substance is required to take part in the reaction and the reaction is said to be of the first order. The general equation showing the velocity of this reaction then becomes:

$$-\frac{dA}{dt} = k(A) \tag{23}$$

or, calling C_o the concentration of A when the reaction started and C the decrease in concentration of A during time t, then at any time the concentration of A becomes $C_o - C$ and the rate is represented by:

$$\frac{dC}{dt} = k(C_o - C). \tag{24}$$

For example, the thermal decomposition of methane,

$$CH_4 = C + 2H_2,$$

is a first-order process, since the concentration of only one reacting substance, methane, is changing and only one molecule of methane appears in the equation. The rate of this reaction may be expressed:

$$\frac{d(C_{\mathrm{CH_4}})}{dt} = kC_{\mathrm{CH_4}} = k(C_o - C) \tag{25}$$

where C_o and C are, respectively, the initial concentration of methane and the mols of methane (per cu. ft.) decomposed during the time t.

It will be noted that this equation and subsequent equations dealing with rate of reaction are not given in integrated form. The integration has been omitted because the result would apply under only one special set of conditions, the integrated forms usually given referring only to constant volume, constant pressure, and constant temperature. In dealing with combustion reactions, although they may take place at constant pressure, the temperature usually changes enormously as the reaction progresses and there is frequently a change in volume. These factors make integration difficult and must always be borne in mind in using reaction-rate expressions.

Velocity of Second-order Reactions.—When two molecules* react and the concentrations of both decrease, the reaction is said to be of the second order. For example, when water vapor reacts with CO in space according to the reaction†

$$H_2O + CO = CO_2 + H_2$$

the rate of the reaction is

$$-\frac{dH_2O}{dt} = k\ (H_2O)(CO) \tag{26}$$

or

$$-\frac{d\,H_2O}{dt} = -\frac{dCO}{dt} = k(C_{\mathrm{H_2O}} - C)(C_{\mathrm{CO}} - C), \tag{27}$$

where $C_{\mathrm{H_2O}}$ and C_{CO} equal the initial concentrations of H_2O and

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* If two molecules of the *same* substance take part in the reaction, as, for example, in the dissociation of CO:

$$2CO \rightarrow C + CO_2,$$

the reaction is also of the second order, although the concentration of only one reacting substance is changing.

† For purposes of illustration, assuming for the moment that the concentrations are such that the composition is far from equilibrium and, therefore, the reaction acts as if it were irreversible.

CO, and C equals the amount of H_2O or CO consumed at any time t.

For the general second-order irreversible reaction

$$A + B \rightarrow AB$$

the rate is expressed by

$$\frac{dC}{dt} = k(C_A - C)(C_B - C). \tag{28}$$

Velocity of Third-order Reactions.—For a third-order reaction, *i.e.*, one in which *three* reacting molecules are involved, as in the reaction

$$aA + bB + eE = \ . \ . \ . \ ,$$

the general expression for the velocity or rate of reaction is:

$$\frac{dC}{dt} = k(C_A - C)^a(C_B - C)^b(C_E - C)^e \quad . \quad , \tag{29}$$

where C_A, C_B, C_E are the initial concentrations of the reacting substances, A, B, and C. For example, the combustion reaction

$$2CO + O_2 \rightarrow 2CO_2$$

is trimolecular, at least between 1652 and 2732°F., (900 and 1500°C.) and since this reaction is essentially irreversible at these temperatures the reaction rate is written:

$$-\frac{dCO}{dt} = k(CO)^2(O_2). \tag{30}$$

Effect of Temperature on the Rate of Reaction (Combustion).—The velocity of all combustion reactions increases with temperature. In most chemical reactions taking place at room temperature the rate in general about doubles for each 18°F. (10°C.) rise in temperature. The increase in velocity of most of the combustion reactions, however, is far less, since the increase in rate of reaction for a given increase in temperature falls off markedly at high temperatures. At temperatures of about 1800°F. (1000°C.), for example, it usually requires an increase of about 180°F. (100°C.), to double the rate. An empirical equation for change in the specific reaction rate with temperature is

$$\frac{d \log k}{dT} = \frac{A}{T^2} + B, \tag{31}$$

where A and B are constants that must be determined for each reaction and T is the absolute temperature. In integrated form the equation becomes

$$2.3 \log k = -\frac{A}{T} + BT + C, \tag{32}$$

where C is the constant resulting from the integration. If the values of the constants of A, B, and C are known, it is possible to calculate the velocity constant k for any desired temperature. Also, if the values of k are known for three different temperatures, the values of A, B, and C can be determined and from them it is possible to calculate the value of k for any other temperature.

Data on the velocities of the major combustion reactions are given in the latter part of the chapter.

Influence of Diffusion on Reaction Rate.—In many cases it has been proved that under given temperature conditions the velocity of a reaction between a solid and a fluid is solely dependent on the rate of diffusion. In such cases the reaction proceeds only as rapidly as the reacting substance can diffuse up to the solid surface. For many of the combustion reactions we have to deal with the interaction of a solid and a fluid (usually a gas). In the reaction

$$CO_2 + C = 2CO$$

the rate of diffusion of CO_2 to the carbon surface, at the temperatures of the fuel bed, is greater than the *velocity of the chemical combination* and therefore chemical reaction rate and *not* diffusion is the controlling factor. On the other hand, the velocity of the chemical reaction

$$C + O_2 = CO_2$$

is so great that the diffusion of oxygen to the carbon surface is the slower and therefore is the factor controlling the speed with which this reaction proceeds in the fuel bed of a furnace.

At all solid surfaces there is a relatively motionless gas film adjacent to the solid,* and this film offers the major resistance to diffusion. The thickness of this film may be decreased by increasing the velocity of the gas past the solid surface, the decrease in film thickness being proportional to a fractional power (usually 0.6 to 0.8) of the velocities (see Appendix I). Increase in velocity is often used as a means of increasing diffusion, and

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* For a detailed exposition of the subject of Fluid Films, see "Principles of Chemical Engineering," by Walker, Lewis, and McAdams.

hence the rate of reaction or combustion where diffusion controls (as in the reaction $C + O_2 = CO_2$).

The Influence of Surface on Reaction Rate.—When a reaction takes place between two phases, as, for example, between a liquid and a gas, or between a solid and a gas, the rate of reaction is influenced by the area of the boundary between the two phases and by the rate of diffusion of the reacting substances. Thus, when gaseous oxygen and solid carbon react, the velocity of the reaction is limited by the area of the carbon surface exposed to oxygen. In Chap. IV it was pointed out that finely divided coal will ignite spontaneously much more quickly than coarse lumps of the same coal. This is due to the greater surface and consequent greater amount of oxygen that can react with the fine coal in a given time.

On account of the influence of the solid surface, in combustion reactions dealing with solid carbon it is desirable to refer to the solid phase, rather than to the gas, as was done in the previous discussion of specific reaction rates. It is, therefore, convenient to define the *specific surface reaction rate* k_s as the *pound-mols of carbon reacting per unit of time per unit volume of container per sq. in. of solid surface* when the concentrations of the reacting gas are equal to one mol per cu. ft. That is, $k_s = \frac{k_1}{A}$ where A is the area of carbon surface per unit volume of container,* and k is the specific reaction rate toward the right.

For a given temperature k_s is independent of the *size* of particle, but is greatly influenced by the *character* of the surface. For example, k_s for the reduction of CO_2 by lignite or charcoal is much greater than for the same reduction by graphite or diamond,

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* If A is defined as the square inches of carbon surface per 1000 cu. in. of container volume, then

A = (number of particles, n, per 1000 cu. in. of container volume) × (area of each particle).

$n = \frac{1070}{d^3}$ (see reference 4) and the area of each particle $= \pi d^2$, where d is the diameter of the particle in inches and the particles are assumed spherical.

Then

$$A = \frac{1070}{d^3} \times \pi d^2 = \frac{3360}{d} \tag{33}$$

and,

$$k_s = \frac{k_1}{A} = \frac{dk_1}{3360} = 0.00298\, dk_1. \tag{34}$$

since the surface of these latter forms of carbon is relatively inert.

The main use of the conception of a specific *surface* reaction rate is to determine the change of specific reaction rate from one size particle to another. Since k_s is a constant independent of particle size,

$$k_s = 0.00298 d_1 k_1 = 00.00298 d_2 k' \tag{35}$$

or

$$k' = \frac{d_1}{d_2} k_1 \tag{36}$$

where d_1 and d_2 are the diameter in inches of the particles. For example, if the specific reaction rate for a 2-in. particle is known, and it is desired to determine the rate for a 0.5-in. particle, then the specific rate of reaction for 0.5-in. particles is four times as great, since

$$k_{0.5} = \frac{2}{0.5} k_{2.0} = 4 k_{2.0}. \tag{37}$$

Catalysis.—All gas reactions are greatly speeded up by contact with hot surfaces, *i.e.*, hot surfaces act as catalysts. It is generally accepted that the function of such catalysts is to adsorb (that is, concentrate on the surface of the solid) the reacting gases. The reaction in the adsorbed film of concentrated gas in accordance with the principles discussed above under *Reaction Velocity* is much more rapid than in the main body of the gas where the concentrations of the reacting substances are lower. Gas reactions which take place slowly in space proceed much more rapidly when the reactions are carried out in contact with the hot walls of the containing vessel. In addition to the general speeding up of combustion reactions by contact catalysis, which influences all gas reactions that take place in contact with hot surfaces, it is well known that materials such as platinum, nickel, ferric oxide, etc. act as *specific* catalysts for certain reactions, particularly at lower temperatures (in general, up to about 950°F., or 500°C.). At high (incandescent) temperatures this selectivity disappears and one surface is as good a catalyst as another. Fire brick, for example, is a good catalyst at temperatures of incandescence.

The presence of water is essential for the progress of a number of reactions. Thus CO and O_2, *if dry*, will not explode, though the mixture becomes explosive as soon as a trace of water is admitted. The presence of this small amount of moisture apparently acts as a kind of catalytic percussion cap in starting the reaction. The mechanism of the process is, however, but little understood.

THE EQUILIBRIUM AND THE VELOCITY OF THE COMBUSTION REACTIONS

The remainder of this chapter is devoted to a discussion of the individual combustion reactions tabulated on page 137. The discussion of each reaction is divided into two parts, (*a*) the equilibrium of the reaction, and (*b*) the reaction velocity, or speed of the reaction (combustion).

1. Reaction $C + O_2 = CO_2 + 174{,}600$ B.t.u.—(*a*) *Equilibrium.* The equilibrium constant *K* for this reaction is represented by:

$$K = \frac{(CO_2)}{(O_2)}$$

and varies with the temperature according to the equation*

$$\log_{10} K = \frac{36{,}980}{T} - 0.300 \log T + 7.89 \times 10^{-5}T - 7.5 \times 10^{-9}\,T^2 + 0.893 \qquad (38)$$

where *K* is in terms of partial pressures of CO_2 and O_2 in atmospheres (where one atmosphere equals 14.7 lb. per sq. in. or a column of mercury 760 mm. high) and *T* is the temperature in degrees Fahrenheit absolute. Calculated for 1650°F. (898°C.) *K* equals approximately:

$$K = \frac{(CO_2)}{(O_2)} = 3.5 \times 10^{17}.$$

That is, in the equilibrium mixture at this temperature, there is 3.5×10^{17} as much carbon dioxide as oxygen present. In other words, since only an infinitesimal amount of oxygen, namely

$$(O_2) = \frac{(CO_2)}{3.5 \times 10^{17}} = 2.9 \times 10^{-18}(CO_2)$$

can exist in contact with carbon at 1650°F. (898°C.), this reaction

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* This and subsequent equations showing the variation of *K* as a function of temperature are expressed in terms of degrees Fahrenheit absolute. If it is desired to express these equations in terms of degrees Centigrade absolute the following procedure should be used: (1) terms in which the reciprocal of *T* appears should be multiplied by 1.8; (2) terms in which *T* to the first power appears should be divided by 1.8; (3) terms in which T^2 appears should be divided by $(1.8)^2$, and (4) terms in the form (Constant × log. T°F. absolute) should be changed to (Constant × log T°C. abs. + Constant × log 1.8). These equations are calculated from the data given by LEWIS and RANDALL.[8]

may be considered to be irreversible at the temperatures occurring in the fuel bed of a furnace.

(b) *Velocity.**—The expression for the velocity at which this reaction (disappearance of oxygen) takes place may be written

$$\frac{dC}{dT} = k(C_o - C), \tag{39}$$

where C_o represents the initial concentration of oxygen and C the oxygen consumed at the time t.

Information on the speed or velocity of this reaction is not complete, although, as near as can be told, *the reaction is substantially instantaneous* at temperatures above 1800°F. (982°C.). From the data of Rhead and Wheeler[5] relative velocities at *low* temperatures are:

Temperature	Relative Velocity
662°F. (350°C.)	1
752°F. (400°C.)	10
932°F. (500°C.)	400

In the above table the speed of the reaction at 662°F. (350°C.) is taken as equal to 1 and the speeds at the other temperatures are relative to the speed at 662°F. (350°C.). The rapid increase is apparent and the view that this reaction is substantially instantaneous at 1800°F. (982°C.) or above does not seem unreasonable.

2. Reaction $C + CO_2 = 2CO$ — 70,200 B.t.u.—(a) *Equilibrium.*—The equilibrium constant K for this reaction, written

$$K = \frac{(CO)^2}{CO_2},$$

varies with temperature according to the equation

$$\log_{10} K = -\frac{16,060}{T} + 2.46 \log T - 6.0 \times 10^{-4}T + 3.43 \times 10^{-8}T^2 + 2.15, \tag{40}$$

where T is the temperature in degrees Fahrenheit absolute, and K is in terms of the partial pressures in atmospheres of CO and CO_2 in the equilibrium mixture. The values of K from Rhead and Wheeler's[5] data, for temperatures from 1500 to 2200°F. (815 to 1204°C.) are plotted in Fig. 38. The percentages of CO and CO_2 in the *equilibrium* mixtures at temperatures ranging

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* For further information regarding the velocity of this reaction see Chap. XII.

from 800 to 2000°F. (427 to 1093°C.) are plotted in Fig. 39, showing the large amount of CO_2 at 800°F. (427°C.) and the progressive decrease in CO_2 content of the gases until at 1750°F. (954°C.) only 1.3 per cent of CO_2 can remain when the gases finally attain equilibrium. The change is seen to be particularly rapid between 1100 and 1400°F. (593 and 760°C.).

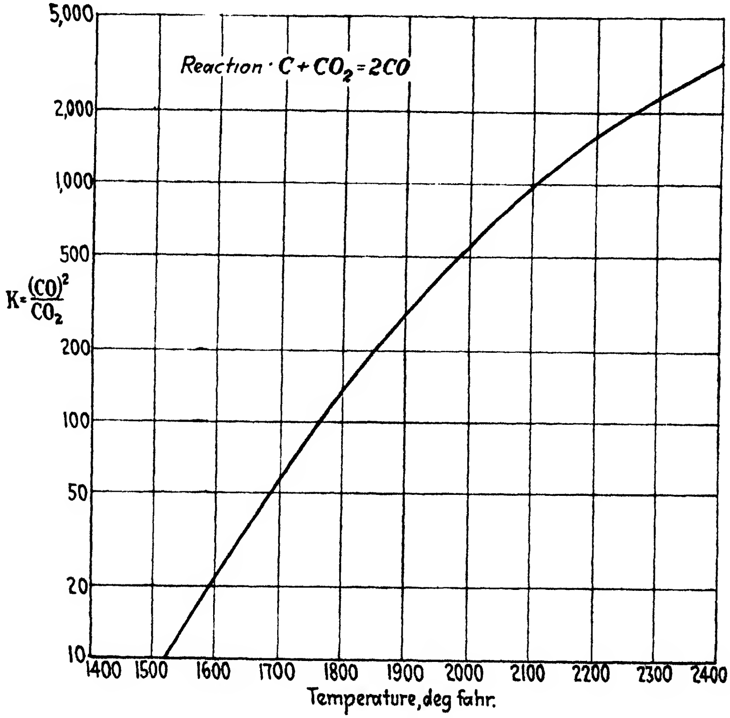

FIG. 38.—Change with temperature of the equilibrium constant for the reaction $C + CO_2 = 2CO$. Carbon as graphite.

(*b*) *Velocity.*—From Fig. 39 it can be seen that the reaction is not complete over a wide range in temperature, 850 to 1750°F. (454 to 954°C.). The upper part of this range is particularly important in combustion work. In this temperature range the CO_2 in contact with carbon is not *completely* consumed, due to the fact that *two* reactions are taking place, the reduction of CO_2,

$$CO_2 + C \rightarrow 2CO, \qquad (a)$$

which is quite rapid, but which takes place at a measurable rate, and the dissociation of CO into C and CO_2,

$$2CO \rightarrow C + CO_2, \tag{b}$$

which, at these temperatures, is slow compared to the rate of reduction (reaction (*a*)). For example, Rhead and Wheeler's[6] experiments have shown that at 1562°F. (850°C.) reaction (*a*) goes 166 times as fast as reaction (*b*).

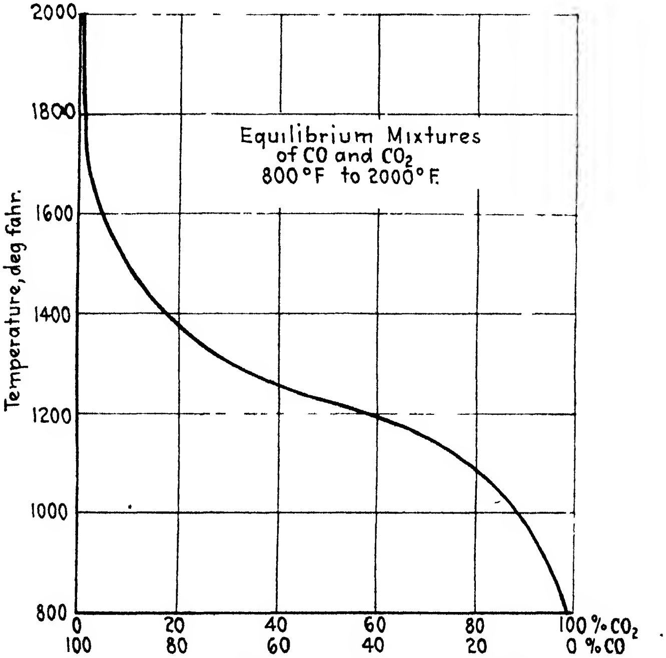

FIG. 39.—Percentages of CO and CO_2 in mixtures of these two gases in equilibrium with carbon at various temperatures.

The rate of reduction of CO_2 by carbon to CO is extremely important commercially and its rate has been determined by Clement, Adams, and Haskins,[7] who used solid carbon in the form of coke, charcoal, and anthracite. The reaction rates, k_1 and k_2 for reaction (*a*) above, may be written:*

$$\frac{d(CO)}{dt} = k_1(CO_2) - k_2(CO)^2, \tag{41}$$

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* Small k_1 is used in this chapter as the velocity constant for a reaction going in the direction of left to right as written. Small k_2 is the velocity constant for the reaction going in the reverse direction, *i.e.*, from right to left.

where (CO) and (CO_2) are the concentrations of carbon monoxide and dioxide in pound-mols per cu. ft. at the time t; k_1 and k_2 are the specific reaction rates toward the right and left respectively. The values of k_1 *for* 5-*mm.* solid particles of carbon at different temperatures are given in the following table:

TABLE XL

VALUES OF k_1 FOR REACTION $C + CO_2 = 2CO$ WITH DIFFERENT FORMS OF CARBON

Temperature	k_1 (for particles of 5 mm. dia.)		
	Charcoal	Coke	Anthracite
1470°F. (800°C.)	0.021		
1650°F. (900°C.)	0.159	0.0023	
1830°F. (1000°C.)	0.629	0.023	
2010°F. (1100°C.)	1.530	0.134	0.119
2190°F. (1200°C.)	.....	0.410	0.237
2370°F. (1300°C.)	.....	1.48	0.579

An example of the method of calculating the amount of reduction of CO_2 that may be expected under a given set of conditions is given below:

Suppose it is desired to estimate the percentage of CO formed by keeping pure CO_2 in contact with 5-mm. coke particles for 2 sec. at 2190°F.(1200°C.). At this temperature the velocity of the reaction

$$2CO \rightarrow C + CO_2$$

is extremely slow and k_2 therefore may be neglected. The reaction rate expression then becomes:

$$\frac{d(CO)}{dt} = k_1(CO_2). \tag{42}$$

Now let

x = the fraction of CO in the gas at time t.
m = the fraction of CO_2 in the gas at the start ($t = 0$).
k_1 = the specific reaction rate* = 0.410 at 2190°F.(1200°C.).
t = the time in seconds.

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* Numerical values of k_1 *for this reaction* are independent of the units in which the concentration of CO_2 is expressed if the same units are used for CO_2 *and* CO.

Then the reaction-rate equation becomes:*

$$\frac{dx}{dt} = k_1\left(m - \frac{m+1}{2}x\right). \tag{43}$$

On integration this expression becomes:

$$k_1t = \frac{2.3}{\frac{1+m}{2}} \log\left(\frac{1}{m - \frac{1+m}{2}x}\right) + C, \tag{44}$$

where C is a constant of integration. The numerical value of C may be determined by substituting in the equation the proper values when $t = 0$. In the case under consideration m equals 1 and $x = 0$ when $t = 0$. Then, by substitution,

$$k_1 \times 0 = \frac{2.3}{\frac{1+1}{2}} \log \frac{1}{1 - \frac{1+1}{2}0} + C$$

$$0 = 2.3 \log 1 + C$$

and since $\log 1 = 0$, $C = 0$.

The general expression, therefore, is:

$$k_1t = \frac{2.3}{\frac{1+m}{2}} \log\left(\frac{1}{m - \frac{1+m}{2}x}\right). \tag{45}$$

Solving for x when $t = 2$ sec.

$$0.410 \times 2 = \frac{2.3}{\frac{1+1}{2}} \log\left(\frac{1}{1 - \frac{1+1}{2}x}\right) = 2.3 \log \frac{1}{1-x}$$

$$\frac{0.410 \times 2}{2.3} = \log \frac{1}{1-x} = 0.356$$

$$\frac{1}{1-x} = 2.27$$

$$x = 0.56.$$

That is, if CO_2 is passed through a bed of 5-mm. coke particles at 2190°F. (1200°C.) at such a gas velocity that the time of contact is 2 sec., the resultant gas will contain† 56 per cent of CO.

The values of k_1 given in the table above vary with temperature according to the equation:‡

$$2.3 \log k_1 = -\frac{A}{T} + BT + C, \tag{32}$$

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* See U. S. Bur. Mines, *Bull.* 7, p. 22 (1911) for derivation.

† The actual value observed in the Bureau of Mines experiments under these conditions was 55 per cent CO.

‡ See p. 148.

where T is expressed in degrees Fahrenheit absolute; the approximate numerical values for the constants of this equation are given in Table XLI.

TABLE XLI

VARIATION WITH TEMPERATURE OF CONSTANTS FOR DETERMINING RATE OF REACTION $C + CO_2 = 2CO$ FOR 5 MM. PARTICLES

Constant	Charcoal	Coke	Anthracite
A	$-91,700$	$-85,100$	$+57,400$
B	$-\ 0.0113$	-0.00539	$+\ 0.0126$
C	$+65.38$	$+45.60$	-56.61

With the aid of these values of A, B, and C, it is, of course, possible to calculate k_1 for any desired temperature and then from the actual concentrations of CO_2 and CO, calculate the rate of reduction.

Since the above values for k_1 were all obtained for the same size particles (5-mm. diameter) the ratios of k_1 for the different modifications of carbon will give directly the ratios of the values of k_s, and hence the relative ease with which these carbonaceous materials will reduce CO_2 to CO. The values of these ratios at 1750 and 2000°F. (955 and 1094°C.) are given in the following table:

TABLE XLII

RELATIVE RATES OF REDUCTION OF CO_2 WITH COKE, ANTHRACITE, AND CHARCOAL

	Relative rate of reduction	
	1750°F. (955°C.)	2000°F. (1094°C.)
Coke	1	1
Anthracite	29	1
Charcoal	69	12

These differences in the rates of reduction are also brought out in Fig. 40, in which, for each of the three types of carbon, the per cent of CO (Bureau of Mines experiments[7]) formed from CO_2 in 1 sec. is plotted against the temperature. The plot shows the large percentages of CO produced from charcoal at temperatures at which coke and anthracite have relatively little reducing power.

It was shown on page 145 that the ratio of the two specific reaction rates k_1 and k_2 is equal to the equilibrium constant K, *i.e.*,

$$\frac{k_1}{k_2} = K \text{ (the equilibrium constant).}$$

From this it would appear possible to combine the values of k_1 determined by Clement, Adams, and Haskins with the values of the equilibrium constant K determined by Rhead and Wheeler,

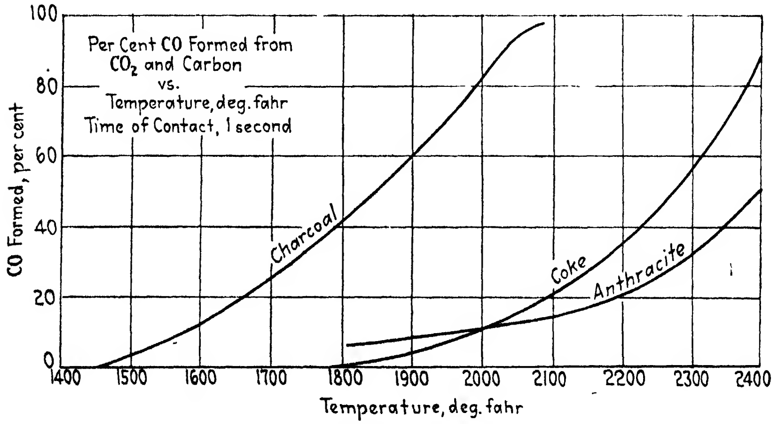

FIG. 40.—Percentages of CO formed from CO_2 in contact with various types of carbon for 1 sec. at temperatures from 1400 to 2400°F.

thus calculating k_2 the rate of decomposition of CO. This is impossible, however, because the decomposition of CO yields only one kind of carbon (lampblack) and the values of k_1 have been shown to vary greatly according to the type of carbon used. Since k_1 and k_2 in this case would refer to different forms of carbon, the calculated values of the reaction constant k_2 would be meaningless.

3. Reaction $2C + O_2 = 2CO + 104{,}400$ B.t.u.—This equation is often used to express the final result of two combustion reactions.

First stage	$C + O_2 = CO_2$	(1)
Second stage	$CO_2 + C = 2CO$	(2)
Final result	$2C + O_2 = 2CO$	(3)

The direct formation of CO from carbon does not take place in most combustion work, although CO may be formed to a very limited extent in certain cases. This reaction will not be discussed further, since it is usually the result of reaction (1) followed by reaction (2).

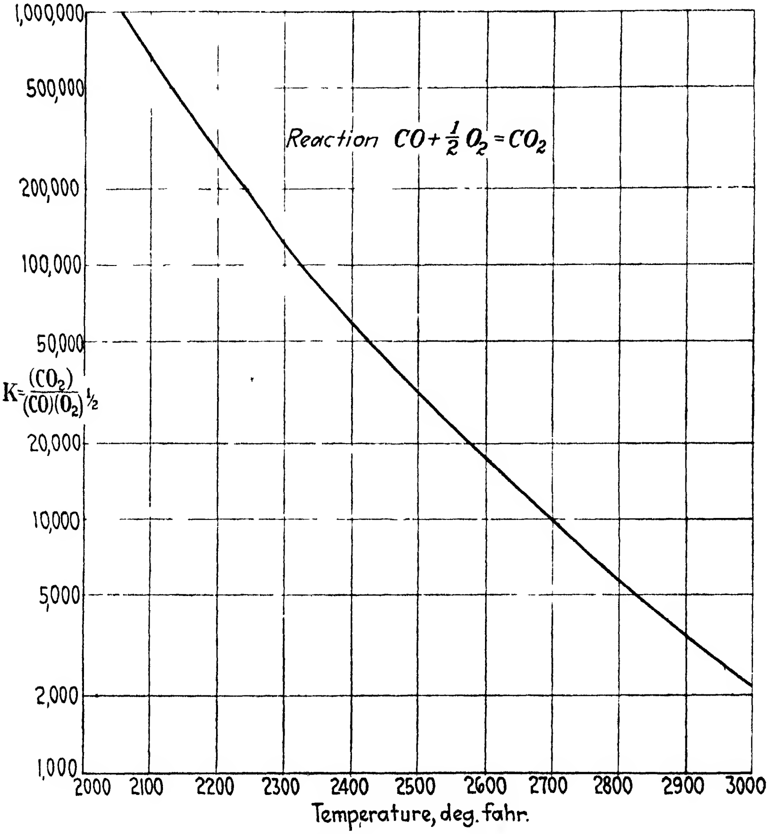

FIG. 41.—Change with temperature of the equilibrium constant for the reaction $CO + \frac{1}{2}O_2 = CO_2$.

4. Reaction $2CO + O_2 = 2CO_2 + 244{,}800$ B.t.u. (*a*) *Equilibrium.*—This reaction is extremely important commercially, and is further discussed in Chap. IX. The equilibrium constant may be expressed by

$$K' = \frac{(CO_2)^2}{(CO)^2 \times (O_2)},$$

or, extracting the square root of both sides of the equation,

$$K = \frac{(CO_2)}{(CO)(O_2)^{1/2}}.$$

Values of the equilibrium constant* in this form are plotted against temperature in Fig. 41. The concentrations of the gases in this case are also expressed as partial pressures of the gases in atmospheres. The equilibrium constant K varies with temperature according to the equation

$$\log K = \frac{26{,}520}{T} - 1.38 \log T + 3.39 \times 10^{-4}\, T, \\ - 2.09 \times 10^{-8} T^2 - 0.586, \qquad (46)$$

where T is in degrees Fahrenheit absolute.

Figure 41 clearly shows the preponderance of CO_2 in the equilibrium mixture until very high temperatures are reached. That is, the dissociation of CO_2 into CO and O_2 is very small, except at very high temperatures (0.014 per cent at 2050°F. and 0.4 per cent at 2800°F.). A flue gas containing 10 per cent CO_2 would be only about 3 per cent dissociated at 3000°F. (1648°C.), if there were no excess O_2. In other words, in most cases encountered in combustion work there can be little or no free CO if there is any excess O_2 (unless the CO and O_2 have been chilled to a low temperature before combustion was completed and equilibrium reached).

(*b*) *Velocity.*—This reaction takes place with extreme rapidity, and although its rate relative to the rates of other combustion reactions is fairly well known, absolute values of the specific reaction rate have not been determined. The general expression for the rate of the reaction is

$$-\frac{dCO}{dt} = k_1(CO)^2(O_2),$$

since it has been shown[9] that the reaction is trimolecular between 1650°F. (900°C.) and 2730°F. (1500°C.) (*i.e.*, temperatures inside combustion chambers). This reaction is slower than either:

$$C + O_2 = CO_2 \qquad (1)$$

or

$$2H_2 + O_2 = 2H_2O. \qquad (5)$$

The ratio of the rate of combustion of CO to the rate of combustion of H_2 is approximately $\frac{k_{CO}}{k_{H_2}} = 0.35$. In other words, CO burns only about one-third as fast as H_2.

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* From the data collected by LEWIS and RANDALL.[8]

5. Reaction $2H_2 + O_2 = 2H_2O$ + 246,200 B.t.u.—(*a*) *Equilibrium.*—The equilibrium constant for this reaction is written:

$$K' = \frac{(H_2O)^2}{(H_2)^2(O_2)}$$

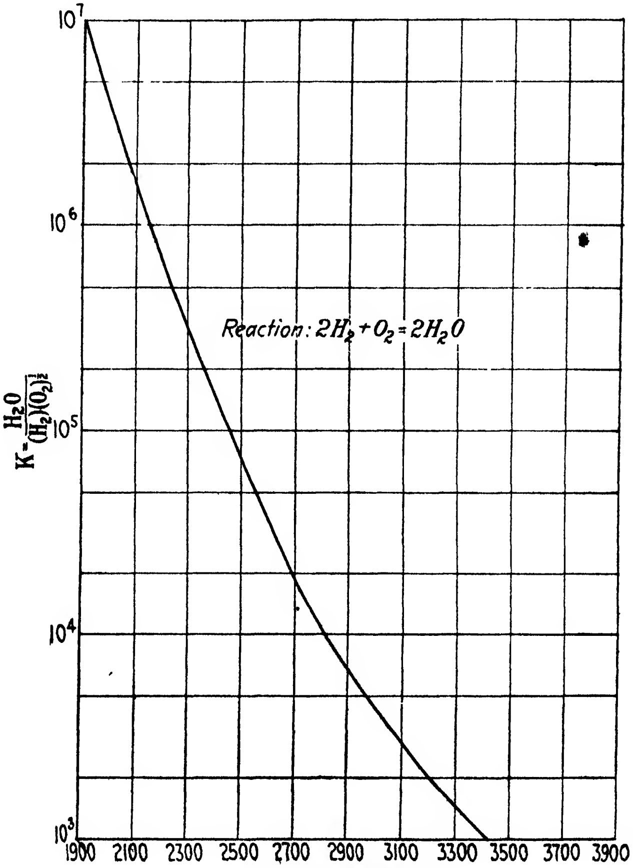

FIG. 42.—Change with temperature of the equilibrium constant for the reaction $H_2 + \frac{1}{2}O_2 = H_2O$.

or, extracting the square root of both sides of the equation,

$$K = \frac{(H_2O)}{(H_2)(O_2)^{\frac{1}{2}}}.$$

The variation[8] of values of the equilibrium constant K with temperature is shown by the plot (Fig. 42). The concentrations of the gases are again expressed as partial pressures in atmospheres. The constant K varies with temperature according to the equation

$$\log K = \frac{22,550}{T} - 0.472 \log T - 2.00 \times 10^{-4}T + 2.49 \times 10^{-8}T^2 - 0.735, \quad (47)$$

where T is the temperature in degrees Fahrenheit absolute.

As in the preceding reaction, only small amounts of hydrogen and oxygen are present in equilibrium mixtures at temperatures below 3000°F. (1650°C.). That is, water vapor is very little dissociated into gaseous hydrogen and oxygen except at very high temperatures. For example, at 3190°F. (1755°C.), it is only 0.37 per cent dissociated.

(*b*) *Velocity.*—Although absolute values of the velocity of this reaction have not been determined, it is known that the reaction is extremely rapid. As already mentioned, it has been found[9] that the combustion of hydrogen* takes place 2.86 times as fast as the combustion of CO. At low temperatures the velocity of this reaction is greatly increased by the presence of hot solid surfaces.

6. Reaction† $C + H_2O = CO + H_2$ − 70,900 B.t.u.—(*a*) *Equilibrium.*—The value of the equilibrium constant K for this reaction is written

$$K = \frac{(CO)(H_2)}{(H_2O)}$$

and values[8] of K over the temperature range from 1200 to 2700°F. (649 to 1483°C.) are shown in Fig. 43. Over the temperature range covered in the plot, K varies approximately according to the equation

$$\log K = -\frac{11,800}{T} + 7.13, \quad (48)$$

where T is the temperature in degrees Fahrenheit absolute. The concentrations of the gases are given in partial pressures in atmospheres. There is a rapid decrease in the amount of undecomposed water in equilibrium mixtures at the high temperatures, particularly above 1800°F. (981°C.).

(*b*) *Velocity.*—This reaction is important in the formation of water gas but below 1650°F. (900°C.) is so slow as to be

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* See also Chap. IX.

† See also reaction (7) below.

practically non-existent.[10] At temperatures above 1650°F. (900°C.), however, the reaction is relatively fast. If the velocity constant of this reaction is taken as $k = 1$, the velocity constant of the reaction,

$$C + CO_2 = 2CO \qquad (2)$$

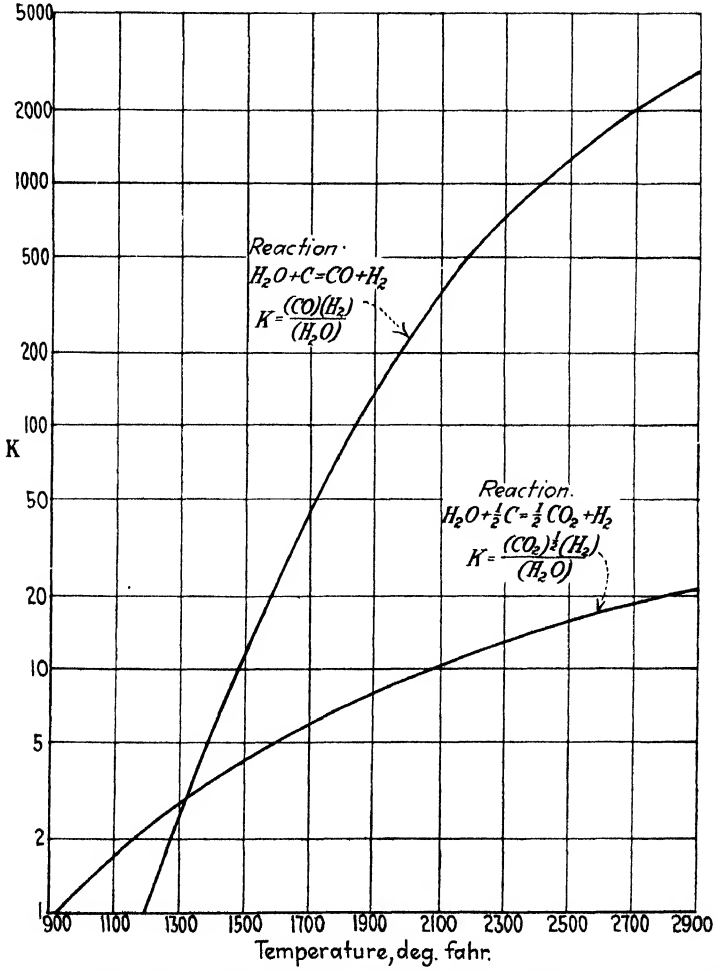

Fig. 43.—Change with temperature of the equilibrium constants for the reactions $C + H_2O = CO + H_2$ and $\frac{1}{2}C + H_2O = \frac{1}{2}CO_2 + H_2$. Carbon as graphite.

is 2.18, or the reduction of CO_2 by a given carbon takes place approximately 2.18 times as fast as the reduction of water vapor to CO and H_2 by the same carbon.

7. Reaction $C + 2H_2O = CO_2 + 2H_2 - 71{,}600$ B.t.u.—This reaction always occurs with reaction (6) and the two must, therefore, be considered together.

(*a*) *Equilibrium.*—The equilibrium constant for this reaction is written:

$$K' = \frac{(CO_2)(H_2)^2}{(H_2O)^2}$$

or, extracting the square root of both sides,

$$K = \frac{(CO)^{1/2}(H_2)}{H_2O}.$$

Values of K in the latter form are shown[8] in Fig. 43, for temperatures from 900 to 2700°F. (482 to 1483°C.). The concentrations of the gases are expressed as partial pressures in atmospheres. At low temperature the equilibrium constant for this reaction is much greater than the equilibrium constant for reaction (6).

(*b*) *Velocity.*—Both reactions (6) and (7) take place at the carbon surface and are, in general, monomolecular (of the first order), *i.e.*, the rate is proportional to the first power of the concentration of the water vapor.* The velocity of reaction (7) is measurable at temperatures above 750°F. (400°C.), and is fast compared to reaction (6) up to 1650°F. (900°C.). Above 1650°F., the rates of reactions (6) and (7) are about equal.

When steam reacts with carbon the following reactions go on practically simultaneously.

$$C + H_2O = CO + H_2 \qquad (6)$$

$$C + 2H_2O = CO_2 + 2H_2 \qquad (7)$$

and

$$C + CO_2 = 2CO. \qquad (2)$$

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* Several theories have been advanced to explain why reaction (7) is monomolecular when apparently *two* molecules of water take part in the reaction. For example, reaction (7) is equal to reaction (6) followed by reaction (8):

$$\begin{array}{lr} C + H_2O = CO + H_2 & (6) \\ CO + H_2O = CO_2 + H_2 & (8) \\ \hline \end{array}$$

Adding, $C + 2H_2O = CO_2 + 2H_2.$

Reaction (8) is very susceptible to catalysis and takes place most readily on the surfaces of solids. It is possible that CO from reaction (6) may be adsorbed on the carbon surface. Interaction may then take place between the adsorbed CO and H_2O.

Consequently, all three must be considered together. To sum up the relative velocity data concerning these reactions, it may be said that considering the *rate of reaction* (7) *as unity*, at temperatures below 1650°F. (900°C.), reaction (6) is practically non-existent, and the velocity constant of reaction (2) is 2.0; at temperatures above 1650°F. (900°C.), the relative velocity constants of reactions (6) and (7) are 1.0 and that of reaction (2) is 2.18.

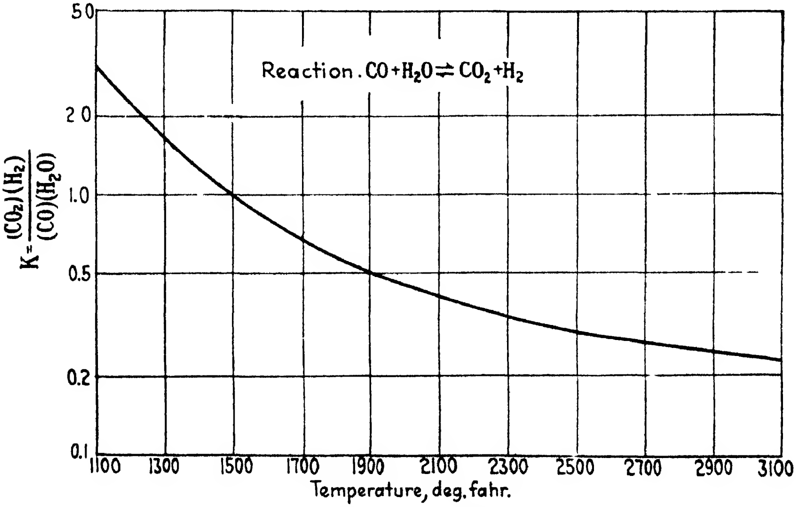

FIG. 44.—Change with temperature of the equilibrium constant for the reaction $CO + H_2O = CO_2 + H_2$.

8. Reaction $CO + H_2O \rightleftharpoons CO_2 + H_2$ − 700 B.t.u.—(*a*) *Equilibrium.*—The equilibrium constant for this reaction* is expressed:

$$K = \frac{(CO_2)(H_2)}{(CO)(H_2O)}$$

and varies with temperature[8] as shown in Fig. 44. Over the temperature range shown in the plot, K varies according to the equation

$$\log_{10} K = \frac{3967}{T} - 0.91 \log T + 5.39 \times 10^{-4}T - 4.58 \times 10^{-8}T^2 + 0.114, \quad (49)$$

where T is degrees Fahrenheit absolute. This figure shows: (1) that increasing temperature increases the proportion of CO in

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* Often called "the water-gas reaction."

the equilibrium mixture, and (2) that an excess of steam will tend to depress CO formation.

(*b*) *Velocity.*—This reaction is practically instantaneous above 2700°F., (1492°C.) but is relatively slow below 1650°F. (900°C.). In between these temperatures "frozen equilibria" may be obtained, *i.e.*, after the four substances CO, H_2O, CO_2, and H_2 have reacted at a high temperature and have come to equilibrium, the proportions of the various components only change with temperature very slowly if at all, so that on quickly cooling down an equilibrium mixture the proportions of the different gases undergo little change.

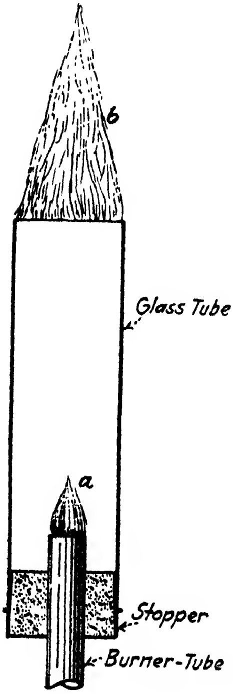

FIG. 45.—Smithel's lamp.

This point is illustrated[11] by analyses taken from the various portions of a Bunsen flame separated as shown in Fig. 45. The inner cone *a* burns quietly at the top of the burner tube, and the outer cone *b* burns at the opening of the glass tube.

The analysis of gas taken from the space between the two cones shows that the value of

$$K = \frac{(CO_2)(H_2)}{(CO)(H_2O)}$$

is the same, irrespective of the position from which the sample is withdrawn. Therefore, although the gases cool several hundred degrees between *a* and *b*, the equilibrium reached in the inner cone of the flame readjusts itself so slowly that the composition of the gas is unchanged during the cooling period. *The rate of this reaction is very susceptible to catalysis,* particularly by materials containing iron oxides. This point is of considerable technical importance, both in the manufacture of hydrogen and in the production of water gas.

Review of the Individual Combustion Reactions.—The major reactions involved in combustion processes are:

$$C + O_2 = CO_2 + 174{,}600 \text{ B.t.u.} \quad (1)$$
$$CO_2 + C = 2CO - 70{,}200 \text{ B.t.u.} \quad (2)$$
$$2C + O_2 = 2CO + 104{,}400 \text{ B.t.u.} \quad (3)$$
$$2CO + O_2 = 2CO_2 + 244{,}800 \text{ B.t.u.} \quad (4)$$

$$2H_2 + O_2 = 2H_2O + 246{,}200 \text{ B.t.u.} \tag{5}$$
$$C + H_2O = CO + H_2 - 70{,}900 \text{ B.t.u.} \tag{6}$$
$$C + 2H_2O = CO_2 + 2H_2 - 71{,}600 \text{ B.t.u.} \tag{7}$$
$$CO + H_2O = CO_2 + H_2 - 700 \text{ B.t.u.} \tag{8}$$

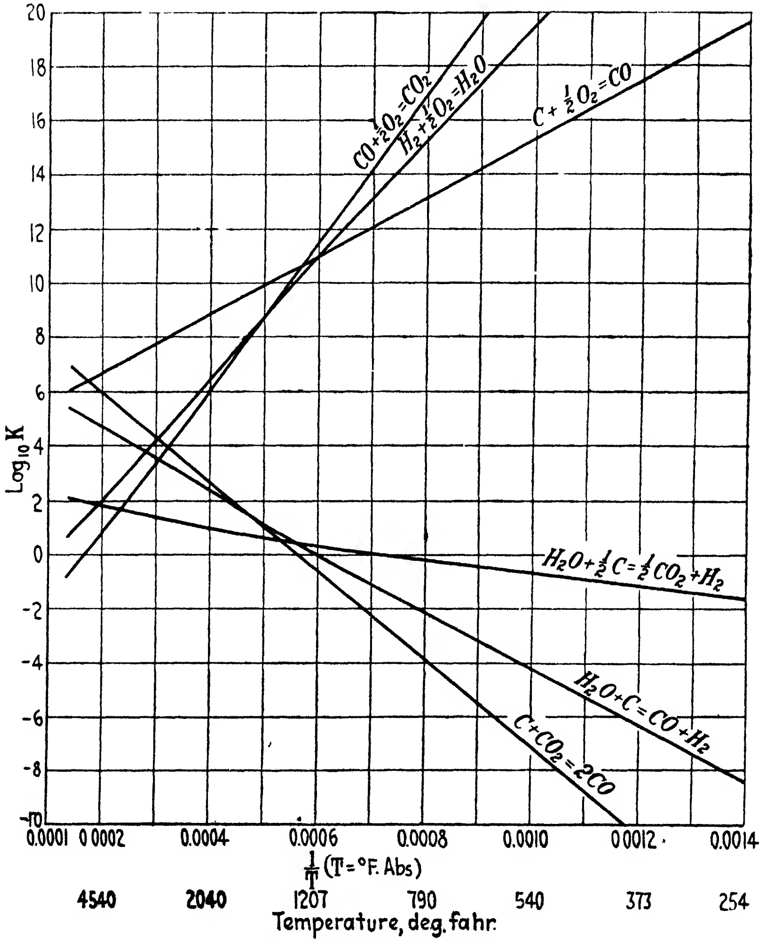

FIG. 46.—Change with temperature of the equilibrium constants of the most important combustion reactions. Carbon as graphite. Partial pressures in atmospheres.

In any reaction which liberates heat, the equilibrium is displaced to the *left* with increasing temperature. This can be seen from an inspection of Fig. 46, in which the variations of the equilibrium constants of the various reactions are plotted against temperature.*

Reaction (1) goes practically to completion to the right at the ordinary temperatures in the fuel bed of a furnace. CO_2, however, at the high temperatures of flames (above 2700°F.) (1483°C.) is slightly dissociated into CO and O_2 (reverse of reaction (4)); this dissociation increases with increasing temperature. CO is very stable and is only slightly dissociated into C and O_2 even at the exceedingly high temperature of 4500°F. (2482°C.). Reaction (2) goes almost completely to the right above 1800°F. (982°C.); that is, at high temperatures the equilibrium mixture is practically all CO, although at low temperatures (around 900°F. (482°C.)) CO_2 reacts with carbon only slightly. Water begins to dissociate into hydrogen and oxygen at about 2400°F. (1315°C.), but the dissociation is small until after the temperature of 3100°F. (1705°C.) is reached. Reactions (6) and (7) always go together, and neither reaction takes place alone; the CO_2 formed by this latter reaction interacts with excess carbon according to reaction (2). Reaction (8) is important in the combustion of gaseous fuels, and with high temperatures tends to form greater proportions of CO.

BIBLIOGRAPHY

1. E. B. Millard: "Physical Chemistry for Colleges," McGraw-Hill Book Co., Inc. (1921).
2. E. W. Washburn: "Principles of Physical Chemistry." McGraw-Hill Book Co., Inc.
3. Bacon and Hamor: "American Fuels," McGraw-Hill Book Co., Inc. (1921).
4. G. Martin: *Combustion*, **8**, 192 (1923).
5. Rhead and Wheeler: *J.* Chem. Soc., **101**, 831, 846 (1912).
6. Rhead and Wheeler: *J.* Chem. Soc., **97**, 2189 (1910).
7. Clement, Adams and Haskins: "Essential Factors in the Formation of Producer Gas," U. S. Bur. Mines, *Bull.* 7 (1911).
8. Lewis and Randall: *J.* Am. Chem. Soc., **36**, 1969 (1914), and **37** (1915); *Cf.* also Lewis and Randall: "Thermodynamics," McGraw-Hill Book Co., Inc. (1923).

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* It should be noted that, in order to plot values of K against temperature for all the reactions, it was necessary to give the logarithm of K and to use the reciprocal of the absolute temperature. Thus a change of one unit in log K represents a tenfold change in K; also, high temperatures are found at the *left* of the plot. In this plot K for the reactions involving carbon is calculated for carbon as graphite.

9. R. T. Haslam: "The Simultaneous Combustion of Hydrogen and Carbon Monoxide," *J. Ind. Eng. Chem.*, **15**, 679 (1923).
10. Haslam, Hitchcock, and Rudow: "The Water-gas Reactions." *J. Ind. Eng. Chem.*, **15**, 115 (1923)
11. Haber: "Thermodynamics of Technical Gas Reactions." p. 308

General

Walker, Lewis and MacAdams: "Principles of Chemical Engineering," McGraw-Hill Book Co., Inc. (1923).

CHAPTER IX

THE COMBUSTION OF THE ELEMENTARY FUELS

As pointed out previously, the various fuels differ greatly in both physical and chemical characteristics. Not only are there solid, liquid, and gaseous fuels, but each fuel within these general groups is made up of many components. Although the number of compounds existing in coal or petroleum is uncertain, it is known that they number in the hundreds. Fortunately, however, from the standpoint of combustion these various compounds can be divided into three classes and the individual compounds partake of the characteristics of the class. For example, while all hydrocarbons burn somewhat differently from one another, the heaviest and most unsaturated having the greatest tendency to decompose on heating, still, from the standpoint of combustion, such a great similarity exists that all hydrocarbons may be considered as belonging to a single class. In the same way we have different solid carbonaceous fuels, such as charcoal and coke from various sources, but all these may be considered in the same class. Here again minor differences occur that are peculiar to each fuel; for example, charcoal burns much more freely than coke.

Therefore, all the various complex fuels may be considered as mixtures of three simple or elementary fuels: (1) solid carbon, (2) gaseous hydrocarbons, and (3) a mixture of carbon monoxide plus hydrogen. Sometimes the complex fuel is a simple mixture of the elementary fuels, as coal gas, which is essentially a mixture of the two elementary fuels, (2) gaseous hydrocarbons and (3) carbon monoxide plus hydrogen. With other fuels, however, the three simple or elementary fuels come into existence only when the fuel is burned, but the decomposition of the complex fuel into the elementary fuels occurs *before actual combustion* starts. Coal is an example of this latter type of fuel. When coal is fed into a fire it is first distilled, giving off gaseous hydrocarbons, carbon monoxide, and hydrogen, leaving behind a residue of solid carbon, coke. The combustion of coal is, therefore, first, the separation of the complex coal into its elementary fuels and, second, the combustion of all three elementary fuels. Another example of the transformation of a complex fuel into an elementary fuel is encountered in the burning of fuel oil, which volatilizes into the elementary fuel "gaseous hydrocarbons" before actual combustion. When, however, the fuel oil is burned with a smoky flame an additional process takes place to a greater or less extent; the complex fuel oil is decomposed ("cracked") into the elementary fuels (1) solid carbon and (2) hydrogen, previous to combustion. The separation of the complex fuel into the elementary fuels may take place either before the final combustion of the elementary fuels or simultaneously with it. Carbon monoxide and hydrogen have been grouped together primarily because industrially they usually occur together and, as will be shown later, burn similarly.

In addition to the three elementary fuels, almost all of the complex fuels contain inert material. For example, coal contains ash and moisture, while producer gas has a large percentage of nitrogen (N_2) and a smaller amount of carbon dioxide (CO_2). In almost all cases the character and the amount of these inert materials greatly affect the suitability of the fuel and its method of handling.

After the complex fuels are separated into the three elementary fuels, it is easy to see that the number of important chemical reactions by which they may burn are relatively few. The elementary fuels either (1) burn directly with the oxygen of the air, (2) change their character by reacting with water vapor, or (3) decompose thermally into other elementary fuels before final or complete combustion. The first type of combustion reaction, direct burning, is illustrated by the equation expressing the combustion of hydrogen with air,

$$2H_2 + O_2 + 3.76N_2 = 2H_2O + 3.76N_2.$$

The second type, change of character by interaction with water vapor, is illustrated by the reaction of steam on hot coke, changing the elementary fuel, solid carbon, over to the elementary fuel, CO plus H_2,

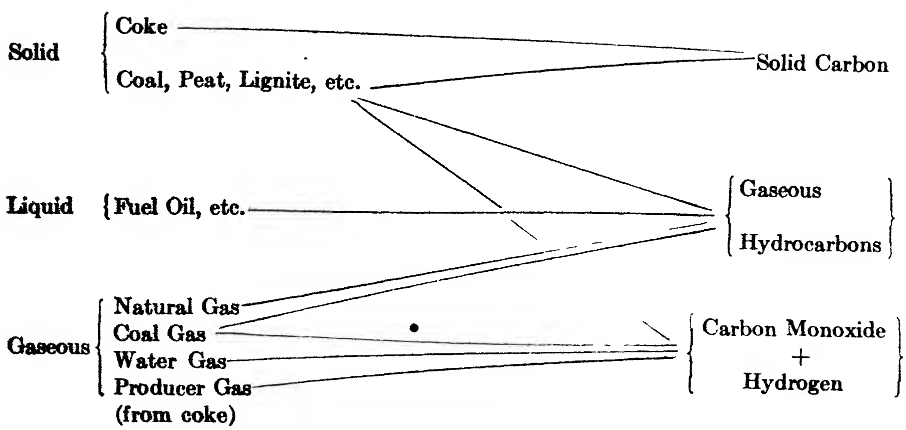
$$C + H_2O = CO + H_2.$$

The third type of reaction, thermal decomposition of an elementary fuel into other elementary fuels, is illustrated by the reaction for the cracking, or thermal decomposition, of ethylene,

$$C_2H_4 + \text{heat} = 2C + 2H_2.$$

TABLE XLIII

PRODUCTION OF ELEMENTARY FUELS FROM COMPLEX FUELS



	Complex Fuel	Elementary Fuel	Combustion Reactions
Solid	Coke Coal, Peat, Lignite, etc.	Solid Carbon	(1) $C + O_2 = CO_2$ (2) $C + CO_2 = 2CO$ (3) $C + \frac{1}{2}O_2 = CO$ (6) $C + H_2O = CO + H_2$ (7) $C + 2H_2O = CO_2 + 2H_2$
Liquid	Fuel Oil, etc.	Gaseous Hydrocarbons	(9) $CH_4 = C + 2H_2$ (10) $CH_4 + 2O_2 = CO_2 + 2H_2O$ (11) $C_2H_4 = 2C + 2H_2$ (12) $C_2H_4 + 3O_2 = 2CO_2 + 2H_2O$
Gaseous	Natural Gas Coal Gas Water Gas Producer Gas (from coke)	Carbon Monoxide + Hydrogen	(4) $2CO + O_2 = 2CO_2$ (5) $2H_2 + O_2 = 2H_2O$ (8) $CO + H_2O = CO_2 + H_2$

In all cases the final result usually desired is the complete and economical interaction of the elementary fuels with oxygen from the air.

The breaking down of the complex into the elementary fuels and their combustion are shown in Table XLIII, together with the essential chemical equations by which these changes are brought about.

THE MECHANISM OF THE COMBUSTION OF (1) SOLID CARBON (2) GASEOUS HYDROCARBONS, AND (3) CO PLUS H_2

Since from the standpoint of combustion all the complex fuels may be considered as mixtures of the three elementary fuels, (1) solid carbon, (2) gaseous hydrocarbons, and (3) carbon monoxide and hydrogen, the various theories that have been advanced to explain the mechanism of the combustion of these three elementary fuels and a summary of the most widely accepted opinions are of value.

1. The Mechanism of the Combustion of Solid Carbon.—During the past 50 years the three main theories that have been advanced to explain the way in which solid carbon reacts with oxygen have been: (*a*) that CO_2 is the first product of combustion of carbon, (*b*) that CO is the first product, and (*c*) the modern view that a complex compound of carbon and oxygen is formed, which breaks down to give both CO_2 and CO, the relative amounts being determined by the temperature.

(*a*) *The Carbon Dioxide Theory.*—One of the early views regarding the mechanism by which solid carbon burns was that CO_2 is the *initial* product of combustion, according to the equation:

$$C + O_2 = CO_2. \tag{1}$$

Any CO formed was supposed to result as a secondary product from the reaction of the dioxide with carbon

$$C + CO_2 = 2CO. \tag{2}$$

This theory was based mainly on the fact that a diamond (pure carbon) when strongly heated burns to CO_2 without any visible flame and was apparently further substantiated by the discovery of Smith[1] that moist oxygen which has been absorbed by carbon at 10°F. (−12°C.) is evolved as CO_2 when the carbon is heated to 212°F. (100°C.).

(*b*) *The Carbon Monoxide Theory.*—Widespread doubt as to the validity of the carbon dioxide theory began when Sir Lowthian Bell[2] brought out evidence tending to show that CO is the first product of the reaction between carbon and oxygen. Bell based this concept mainly on the fact that samples of gas taken just above the tuyères (where air is admitted) of a blast furnace contain much more CO than CO_2. For example, two of the samples analyzed by Bell had the following composition:

TABLE XLIV

Analyses of Gas from Tuyères of a Blast Furnace

	Sample 1, per cent	Sample 2, per cent
CO_2	0.76	0.8
CO	37.60	33.3
H_2	......	1.0
N_2	61.64	64.9
	100.00	100.0

According to this concept, CO is the first product of the reaction, and any CO_2 is the result of subsequent combustion of the CO with oxygen.

Further evidence in support of this view was brought out by C. J. Baker.[3] He repeated the experiments of Smith[1] and found that, although *moist* oxygen absorbed by charcoal at 10°F. (−12°C.) is evolved at 212°F. (100°C.) as CO_2, carefully *dried* oxygen absorbed by *dry* charcoal is driven off only very slowly below 842°F. (450°C.). Above this temperature the gases evolved are mainly CO. This CO could not have been formed by the reduction of initially formed CO_2, since CO_2 cannot be reduced to CO by carbon at this temperature.

H. B. Baker[4] found that when pure carbon is heated in a closed vessel with *dry* oxygen, the carbon does not glow while combining with dry oxygen as it does if moist oxygen is used. Furthermore, the gases resulting from such combustion always contain both CO and CO_2, even if a large excess of oxygen is present. Baker also found that pure, dry CO_2 is not reduced by dry charcoal, even at red heat. On the other hand, this same charcoal will give off CO principally if it is heated in dry oxygen. The analyses of some of the gases obtained in this experiment are given in the following table:

TABLE XLV

COMPOSITION OF GASES EVOLVED FROM CHARCOAL PREVIOUSLY EXPOSED TO DRY OXYGEN

Composition of gases	Charcoal heated to redness in oxygen dried over P_2O_5 for	
	1 week	2 weeks
O_2	55.0 per cent	58.1 per cent
CO	40.0 per cent	39.5 per cent
CO_2	5.0 per cent	1.7 per cent

From those experiments Baker concluded: (1) that moisture catalyzes the oxidation of carbon and (2) that carbon burns first to the monoxide.

Further evidence tending to support this same view was brought out by Dixon[5] and his collaborators. These investigators found that in the burning of *gaseous* carbon, CO is formed before CO_2. For example, when *gaseous* carbon in the form of cyanogen (C_2N_2) is burned, the rate of flame propagation is fastest with a mixture $C_2N_2 + O_2$ (in the proportion to form CO). Furthermore, photographic evidence proved that, if $C_2N_2 + 2O_2$ is exploded, one-half the oxygen is inert in the explosion wave itself and as the gases cool the second molecule of O_2 then combines with the CO initially formed in the explosion wave. The actual values of the rate of detonation of various gaseous carbon (cyanogen) and oxygen mixtures are given in the following table:

TABLE XLVI

RATE OF DETONATION IN MIXTURES OF CYANOGEN, OXYGEN AND NITROGEN

Experiment No.	Initial mixture	Products of combustion	Rate of detonation, feet per sec.
1	$C_2N_2 + O_2$	$2CO + N_2$	8951
2	$C_2N_2 + O_2 + O_2$	$2CO_2 + N_2$	7615
3	$C_2N_2 + O_2 + N_2$	$2CO + 2N_2$	7868

The rate of detonation is greatest with equal volumes of C_2N_2 and O_2, although more heat is evolved in forming CO_2 than in the formation of CO. Also, in experiment 2 the excess oxygen present acts like the nitrogen in experiment 3, in retarding the rate of explosion. In fact, excess O_2 slows down the rate at which the detonating wave travels even more than the equivalent amount of N_2.

These results, coupled with the observation by Smithels and Ingle,[6] that gases in the center of a cyanogen flame contain CO and CO_2 in the ratio 2:1, seemed to show definitely that CO is the first product when *gaseous* carbon burns in air. The CO_2, under these circumstances, is formed by a subsequent reaction of the monoxide with oxygen.

It was felt by many that the conclusive evidence that gaseous carbon burns to CO supported the concept of Bell and Baker that solid carbon also burns first to CO and that CO_2 is the product only of the secondary oxidation of the monoxide.

(*c*) *The "Carbon-complex" Theory.*—In 1912 Rhead and Wheeler[7] pointed out that neither the carbon dioxide nor the carbon monoxide theory could explain all the known facts. These authorities reinvestigated the problem of carbon combustion, and on the basis of their results evolved the explanation that is now widely accepted. Briefly, this theory states that the first step in the combustion of carbon is the formation of a loosely bonded physicochemical compound of carbon and oxygen of the general formula C_xO_y. This unstable compound later breaks up into CO and CO_2 in various proportions, depending on the temperature and other conditions.

Rhead and Wheeler's first experiments, on the relative rates of the various reactions between C, CO, CO_2, and O_2, showed that in the combustion of carbon *both* oxides are produced simultaneously, and that neither oxide is the initial product. For example, CO is one of the products of combustion at 932°F. (500°C.), at which temperature the rate of the reaction

$$C + CO_2 = 2CO \tag{3}$$

is practically *nil*, so that the monoxide, not being formed from this reaction, must have been formed *directly* from carbon and oxygen.

In 1913, Rhead and Wheeler published the results of a series of experiments in which air was circulated over purified wood charcoal at temperatures ranging from 212 to 1652°F. (100 to 900°C.). The more important results of these experiments may be summarized briefly as follows:

1. If carbon (in the form of charcoal) is heated to 1652°F. in vacuo and is then cooled to 392 to 750°F. (200 to 400°C.) and exposed to the air, the charcoal absorbs oxygen (see[7] Fig. 47).

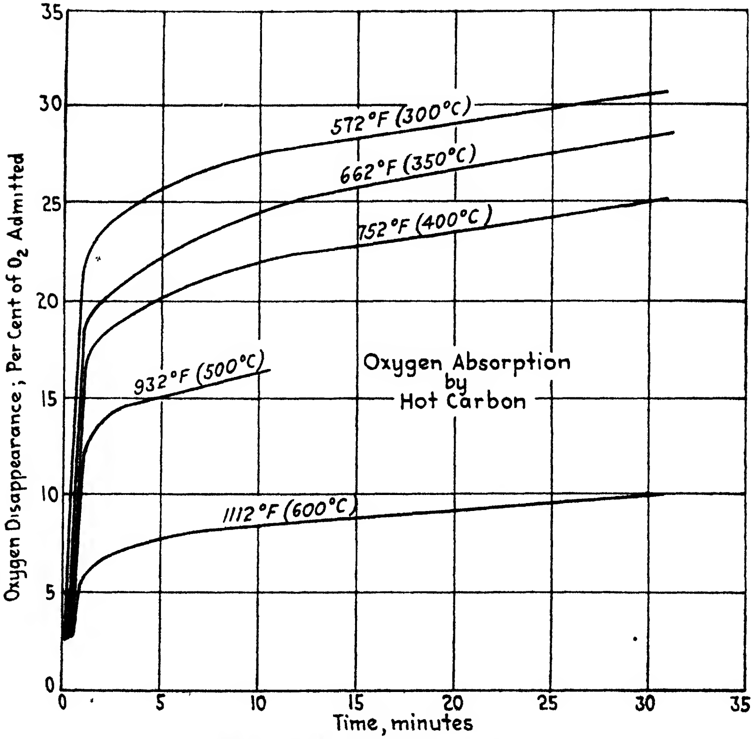


Fig. 47.—Rate of absorption of oxygen by hot carbon.

2. This oxygen cannot be withdrawn by the application of vacuum alone.

3. To break up the product initially formed by the exposure of charcoal to air, *considerable heat and vacuum* are both required. Even then the oxygen comes off not as oxygen, but as a *mixture* of *carbon monoxide* and *dioxide*.

4. In order to drive off most of the oxygen as CO and CO_2, a vacuum and a temperature of about 1650°F. (900°C.) are *both* required.

From this, Rhead and Wheeler came to the conclusion that when solid carbon unites with oxygen or air three products are formed, CO_2, CO, and a complex compound of carbon and oxygen. The formula C_xO_y was assigned to this complex compound, since the proportion of oxygen to carbon varies with the temperature, kind of carbon, etc. This compound C_xO_y is a physicochemical complex, *physical* since it has no definite composition, and *chemical* since it cannot be broken up by strictly physical means, such as evacuation. The reaction between carbon and oxygen does not seem to be straight adsorption because the oxygen combined in this way cannot be removed as *oxygen*, but always comes off as a mixture of CO and CO_2.

Rhead and Wheeler's concept has received strong support from the work of more recent investigators of the problem. Langmuir,[8] for example, investigated the reaction between oxygen and highly graphitized carbon at various temperatures but at extremely low pressures (0.004 to 0.007 mm.). The hot carbon consisted of an electrically lighted graphitized filament of a lamp, the bulb of which was immersed in liquid air. As oxygen was admitted, any CO_2 initially formed was frozen out on the cold walls of the lamp bulb. By using very low partial pressures of oxygen and by providing a means for condensing any CO_2 to the solid state (thus reducing the partial pressure of the CO_2 to a negligible quantity) Langmuir prevented any secondary reaction between the carbon lamp filament and any CO_2 that would be formed. At 1740°F. (950°C.) the oxygen disappears rapidly at the start of the experiment, but the rate of disappearance decreases to a constant rate, until the oxygen finally disappears altogether. At this temperature, as the oxygen disappears, CO_2 (but no monoxide) is formed. The oxygen appearing as CO_2, however, is not equivalent to the total oxygen admitted, showing that a certain amount has combined with the carbon to form the complex C_xO_y even at this high temperature. At 2600°F. (1430°C.) some CO is formed (probably due to a reduction of CO_2 by carbon), although even at this high temperature a certain amount of the carbon complex still exists. A difference between the results of Rhead and Wheeler on the one hand and of Langmuir on the other should be noted. According to Rhead and Wheeler, the C_xO_y complex breaks down to CO and CO_2, whereas Langmuir's experiments seem to show conclusively that the complex decomposes to CO_2 only, any CO formed coming from the secondary reaction of CO_2 with carbon. Langmuir also studied the action of a small amount of CO_2 on a hot metallized carbon filament at low pressure. He found that the CO_2 was reduced to CO but that the volume of CO formed was not double that of the CO_2 used, thereby showing that the reaction was

$$CO_2 + C = (C + \text{absorbed or combined } O_2) + CO$$

indicating that a C_xO_y complex is also formed when CO_2 reacts with carbon.

In general, oxygen combining with carbon at low temperatures (200 to 300°F.) goes almost wholly to CO; the major part of the dioxide which appears at these low temperatures probably results from the reaction

$$2CO + O_2 = 2CO_2,$$

which reaction Rideal[21] has proved to be markedly catalyzed by surfaces. At high temperatures, as proved by the work of Langmuir, the complex C_xO_y is still an intermediate product, but the equivalent oxygen appears initially as CO_2, any CO that may be present usually being formed as a secondary product by the reduction of CO_2 to CO by hot carbon.

Therefore, the most probable and generally accepted concept of the mechanism of the combustion of solid carbon is that oxygen from the air, on uniting with hot carbon, forms CO_2, some CO, and a physicochemical complex C_xO_y. This last compound, in turn, breaks down to form CO_2 and some CO. Thus the C_xO_y is an intermediate compound, but nevertheless a real one. The products *given off* are CO_2 (and some CO at the lower temperatures); the CO_2 will be reduced or the CO burned to CO_2 as a *secondary* reaction depending on the time of contact, the temperature, and whether there is *an excess of carbon or an excess of oxygen.*

This explanation is not only the most satisfactory working hypothesis so far evolved, but also serves to explain the facts on which the two earlier theories of carbon combustion were based. For example, the formation of CO_2 by the action of oxygen on charcoal at 212°F. (100°C)., and the appearance of CO_2 and CO in the gases given off by coal which has absorbed oxygen at 212°F. (100°C.), are explained by the fact that at low temperatures, although the complex C_xO_y breaks down mainly into the monoxide, the surfaces present catalyze the reaction:

$$2CO + O_2 = 2CO_2.$$

Similarly, the results of H. B. Baker, who found that both CO and CO_2 are evolved from dry charcoal and dry O_2 at red heat, would be expected, since at this temperature C_xO_y breaks up to form both gases and the dry oxygen cannot burn the CO to CO_2.

2. The Mechanism of the Combustion of Gaseous Hydrocarbons.—Three theories have been advanced to explain the way in which hydrocarbons burn: (*a*) that the hydrogen of the hydrocarbon burns with oxygen in preference to or before the carbon, (*b*) that the carbon burns in preference to the hydrogen, and (*c*) that there is an initial association of oxygen with the hydrocarbon molecule, forming intermediate, "hydroxylated" compounds, which, in turn, burn or are broken down thermally.

(*a*) *Theory of Preferential Combustion of Hydrogen.*—Until about 1892 it was almost universally believed that hydrogen is the more combustible part of the hydrocarbon and, on burning a hydrocarbon, is "singed off" the molecule, leaving free carbon, as indicated by the reaction

$$2C_2H_2 + O_2 = 4C + 2H_2O. \tag{4}$$

Thus, H. B. Dixon,[9] in 1884, speaking of the combustion of hydrocarbons said: "There is a race for the oxygen of the air between the two constituents of ethylene (for example), and the hydrogen, being the fleeter of the two, gets to the oxygen first, and is burnt to water."

(*b*) *Theory of Preferential Combustion of Carbon.*—In 1892 the older concept of the preferential combustion of hydrogen had to be discarded, since Dixon[10] found that a mixture of equal volumes of ethylene and oxygen yields on detonation almost exactly twice its volume of hydrogen and CO, according to the equation

$$C_2H_4 + O_2 = 2H_2 + 2CO. \tag{5}$$

This led to the explanation that the carbon of the hydrocarbon is preferentially burned to CO, leaving the hydrogen thus liberated to combine with excess oxygen.

This theory, brought out originally[11] in 1861, apparently received additional proof when Smithels and Ingle[12] found that hydrogen is present in the inner parts of aerated hydrocarbon flames.

(*c*) *The Hydroxylation Theory.*—Neither of the preferential-combustion concepts, however, explains all the known facts. A number of investigations have been carried out which indicate that the combustion of gaseous hydrocarbons takes place by an initial interaction of the hydrocarbon with oxygen, forming an intermediate, unstable, "oxygenated" or "hydroxylated" compound.

The principal researches on which this view is based were carried out by Bone[13] and his collaborators.* The experiments were of three general types: (1) static experiments, in which the gas or hydrocarbon was heated in bulbs with varying proportions of oxygen, at temperatures from 482 to 662°F. (250 to 300°C.), the contents of the bulbs after definite periods of time being cooled and analyzed; (2) dynamic experiments, in which the gaseous mixtures were recirculated over a surface of porous porcelain kept at a constant temperature in a furnace, cooling and scrubbing arrangements being provided for the removal of intermediate products; (3) explosion experiments, in which the hydrocarbons were exploded with varying amounts of oxygen, (and in certain cases mixtures of oxygen and hydrogen or oxygen and CO), and the products analyzed. These experiments were run with (1) the paraffin hydrocarbons (*i.e.*, of the formula C_nH_{2n+2}) from methane to hexane, (2) the olefines (*i.e.*, of the formula C_nH_{2n}) up to butylene, (3) acetylene and trimethylene, and (4) independently with an ordinary commercial kerosene.[14]

The results of experiments along this line may be summarized as follows:

1. There is a combination of the hydrocarbon with oxygen preliminary to final combustion. Hydrocarbons combine with oxygen to form alcohols and aldehydes previous to, and as a preliminary step in, burning to CO, CO_2 or water. *This holds true at all temperatures.*

2. Contrary to the ideas expressed in earlier theories, there is no "selective combustion" of either the carbon or the hydrogen of the hydrocarbon.

3. The initial addition of oxygen to the hydrocarbon molecule forms an alcohol with which, in turn, more oxygen reacts, forming an aldehyde. This aldehyde, in many cases, breaks down into intermediate combustible gases, CO and H_2, or the aldehyde may burn completely to CO_2 and H_2O. This process is termed hydroxylation because the first addition of oxygen to the hydrocarbon molecule gives the hydroxyl grouping (—OH) characteristic of alcohols.

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* The recent researches of Blair and Wheeler[15] have brought out extremely valuable additional proof of the hydroxylation theory practically as it was originally formulated by Bone *et al.*

4. As much as 92 per cent of the theoretical formaldehyde possible from the oxidation of ethane has been isolated, thus conclusively showing that the mechanism of hydrocarbon combustion is essentially as just given.

5. Initial combustion or hydroxylation of the hydrocarbon molecule starts with kerosene-air mixtures[14] at about 392°F. (200°C.), and the start of combustion or hydroxylation of benzene begins at about 550°F. (290°C.).

6. If sufficient oxygen is present to transform all the hydrocarbon to formaldehyde, combustion may be carried out without the formation of soot or free carbon. The minimum amount of oxygen required to burn unsaturated hydrocarbons of the olefine type without soot or free carbon is given by the following reaction:

$$(CH_2)_x + \frac{x}{2} O_2 = xCO + xH_2. \qquad (6)$$

It is to be noted from the above equation that the products of combustion, with the minimum oxygen that will prevent soot, are CO and H_2. On the other hand, similar mixtures $(C_xH_{(2x+2)} + \frac{x}{2} O_2)$ of a saturated paraffin hydrocarbon and oxygen yield *carbon*, CO and H_2, H_2O, so that, to prevent the formation of soot with the paraffin hydrocarbons, it is necessary to use at least the following ratio of oxygen to the hydrocarbon:

$$C_xH_{(2x+2)} + \frac{x+1}{2} O_2 = xCO + xH_2 + H_2O. \qquad (7)$$

7. At the high temperatures of flames it was found that in mixtures of hydrocarbons, hydrogen, and CO (with insufficient oxygen for complete combustion) more oxygen combines with the hydrocarbons than with either hydrogen or CO. Thus, when a mixture of methane and electrolytic gas is exploded in the proportions $CH_4 + 2H_2 + O_2$, the reaction is practically:

$$CH_4 + O_2 + (2H_2) = CO + H_2 + H_2O + (2H_2). \qquad (8)$$

Actually, over 95 per cent of the oxygen reacts with the methane and less than 5 per cent combines with the hydrogen originally in the mixture. Practically the same thing holds true for the mixture C_2H_4 (ethylene) $+ O_2 + 2H_2$. Similar experiments with mixtures in the proportions of $CH_4 + 2CO + O_2$ show that *even less* than 5 per cent of the oxygen in the mixture is consumed by the CO.

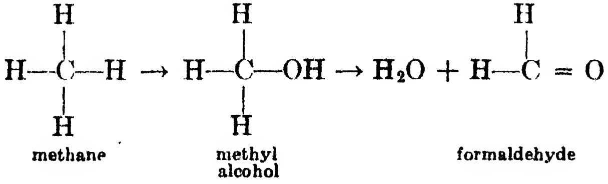
The relative rates of combustion of hydrocarbons, hydrogen, and CO are also indicated by the composition of furnace gases. For example, the following series of analyses[25] of the combustible gases taken at various distances above the top of bituminous coal fire clearly show the rapidity with which methane disappears (as compared with CO and hydrogen) as the gas from a coal fire burns in a furnace:

Sample taken at	Per cent		
	CH_4	H_2	CO
Surface of fuel bed	1.5	5.9	9.7
11 in. above fuel bed	0.1	1.8	4.1
5 ft. above fuel bed	0.0	0.2	0.6
8½ ft. above fuel bed	0.0	0.0	0.1
18½ ft. above fuel bed	0.0	0.0	0.0

This table shows for this case that the methane is practically all consumed before the gases have gone 1 ft. from the fuel bed, the hydrogen practically at the end of 5 ft., while some monoxide is left even at 8½ ft. from the surface of the fuel bed. The relative speeds with which these fuels burn are as follows: (1) hydrocarbons, (2), hydrogen, and (3), CO, *i.e.*, hydrocarbons burn the fastest and CO the slowest.

The following reactions are given to illustrate the steps occurring in the combustion or hydroxylation of the various hydrocarbons. The oxygen required for each successive hydroxylation is not shown.

1. At temperatures below the ignition point, the hydroxylation of methane is as follows:



$$H-\overset{H}{\underset{H}{C}}-H \rightarrow H-\overset{H}{\underset{H}{C}}-OH \rightarrow H_2O + H-\overset{H}{C}=O \qquad (9)$$

Depending on the amount of oxygen present, the formaldehyde thus formed may either:

(*a*) Break down thermally to CO and H_2:

$$CH_2O \rightarrow CO + H_2. \tag{10}$$

(*b*) React with oxygen to form CO and H_2O:

$$CH_2O + \tfrac{1}{2}O_2 = CO + H_2O. \tag{11}$$

(*c*) React with sufficient oxygen for complete combustion to CO_2 and H_2O:

$$CH_2O + O_2 = CO_2 + H_2O. \tag{12}$$

2. The combustion of a chain hydrocarbon is similar, *i.e.*, one of the CH_3 groups is oxidized in the same manner. The following diagram shows the mechanism for the slow combustion of ethane at temperatures below its ignition point;

$$\underset{\text{ethane}}{H_3C.CH_3} \rightarrow \underset{\substack{\text{ethyl}\\\text{alcohol}}}{CH_3.CH_2OH} \rightarrow [CH_3.CH\ (OH_2)] \rightarrow \underset{\text{acetaldehyde}}{CH_3.CHO} +$$

$$H_2O \rightarrow CO + H_2 + \underset{\substack{\text{formal-}\\\text{dehyde}}}{H.CHO} \rightarrow \underset{\substack{\text{formic}\\\text{acid}}}{HCOOH} \rightarrow \underset{\substack{\text{carbonic}\\\text{acid}}}{CO(OH)_2} \rightarrow$$

$$CO_2 + H_2O \tag{13}$$

3. The combustion of an unsaturated hydrocarbon is similar to the process for the saturated hydrocarbons. The hydroxylation of a typical unsaturated hydrocarbon, ethylene, is as follows:*

$$\underset{\text{ethylene}}{H_2C{:}CH_2} \rightarrow \underset{\text{vinyl alcohol}}{H_2C{:}CHOH} \rightarrow [HO.HC{:}CH.OH] \rightarrow \underset{\text{formaldehyde}}{2CH_2O} \tag{14}$$

As in the combustion of methane shown above, the formaldehyde may decompose thermally to CO and H_2, or it may burn to H_2O and CO_2.

4. In passing from natural gas to gasoline and to the typical fuel oils, the hydrocarbons present become heavier and their structure becomes more complex. In the case of these higher hydrocarbons, where R represents a long-chain hydrocarbon radical, the mechanism may be:

(*a*) $RCH_2.CH_3 \rightarrow RCH_2.CH_2OH \rightarrow RCH_2CHO + H_2O$ (15)
$RCH_2.CHO \rightarrow R.CHOH.CHO \rightarrow RCHO + CH_2O \rightarrow RCHO + H_2 + CO$

(*b*) $RCH_2CH_3 \rightarrow RCH_2CH_2OH \rightarrow H_2O + \underset{\text{unsaturated}}{(R.CH = CH_2)}$ (16)

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* It can be seen that this reaction readily explains the data (reaction (5)) brought forward by Dixon as evidence in favor of the view of preferential combustion of carbon (see p. 180).

The heavier hydrocarbons usually undergo reactions other than simple hydroxylation, and the same is true of the lighter hydrocarbons if conditions are unfavorable for the formation of hydroxylated compounds. The heavy hydrocarbons may be "cracked" to give saturated and unsaturated lighter hydrocarbons, or they may be decomposed completely into carbon and hydrogen.

The lightest, saturated hydrocarbons are "cracked" much more slowly than the heavier hydrocarbons. Thus methane[13] when heated at about 2400°F. (1300°C.) is *not completely* decomposed at the end of an hour according to the reaction.

$$CH_4 = C + 2H_2. \qquad (17)$$

Ethane, the next hydrocarbon above methane in the paraffin series, is less stable, however, and can be decomposed very rapidly by heating in a tube at about 2100°F. (1150°C.).

With regard to the speed at which the still heavier hydrocarbons are decomposed to carbon and hydrogen, it has been found in commercial practice that the rate doubles for each 20 to 90°F. (11 to 50°C.) rise in temperature, and that at the end of 1 min. at 1650°F. (900°C.) about 40 per cent of the hydrocarbon is decomposed to form carbon. With these data as a basis, Katz[17] has estimated that only 0.036 sec. are required for complete decomposition of heavy hydrocarbons at 2732°F. (1500°C.), not an unreasonable temperature when considering coal and oil fires. At 3300°F. (1800°C.), about the maximum temperature in combustion work, the time required is much less.

Thus in the ordinary combustion of hydrocarbons, there is a race between thermal decomposition and hydroxylation. If the conditions favor hydroxylation (such as preheating the hydrocarbons and air, and allowing time for the entrance of oxygen into the hydrocarbon molecule), there will be no soot. If, however, conditions favor cracking, as, for example, if the hydrocarbons and oxygen from the air are not thoroughly mixed together, the heat from the combustion of part of the hydrocarbon decomposes or cracks the remainder.

3. The Mechanism of the Combustion of Hydrogen and CO.—All of our industrial fuels either contain CO and H_2, (producer gas, water gas, coke-oven gas, etc.), or else they contain hydrocarbons (coal, fuel oil, etc.), which may, and usually do, give off CO and H_2 in the initial or early stages of combustion. The preceding discussion has shown that any hydrocarbons present first "hydroxylate" and then decompose either wholly or in part into hydrogen and carbon monoxide. Furthermore, it has just been shown that in the explosion of hydrocarbons such as methane, etc., the hydrocarbons are all consumed or transformed over to CO and H_2 before there is any appreciable combustion of either of these gases. Thus, since the combustion of all the H_2 and CO, either present initially or as a result of thermal decomposition, is the last stage in the combustion of fuels, a knowledge of the mechanism of the combustion of these two gases is quite essential. The general literature indicates that there is some doubt as to the exact nature of these two reactions.

With regard to the combustion of hydrogen, Bone and his collaborators[18] have shown that the reaction between H_2 and O_2 is catalyzed by surfaces at all temperatures, and that at high temperatures all surfaces are equally good. These investigators also found that the rate of combustion is proportional to the partial pressure of the hydrogen (not proportional to the square of the (H_2) concentration as if the reaction were trimolecular). From this it was concluded that either: (1) the hydrogen on burning to water first forms hydrogen peroxide, according to the equation

$$H_2 + O_2 = H_2O_2 \text{ (bimolecular)}, \qquad (18)$$

which, in turn, is decomposed to H_2O and oxygen, or (2) that the rate of combustion is proportional to the rate at which hydrogen diffuses through the stationary film on the outside of the catalyst. The latter view is the more probable.

Bone and Haward[19] found that when hydrogen is *exploded* with oxygen the reaction was trimolecular, as shown by the following equation:

$$2H_2 + O_2 = 2H_2O \text{ (trimolecular)}. \qquad (19)$$

Concerning the combustion of CO, Falk,[20] from experiments on ignition temperatures, concluded that two carbon monoxide molecules react directly with one oxygen to form CO_2 in the way the equation is usually written:

$$2CO + O_2 = 2CO_2 \text{ (trimolecular)}. \qquad (20)$$

Falk's reasoning was that if the ignition temperatures of mixtures of a gas with varying proportions of oxygen are determined, the lowest temperature of ignition is obtained when the gases are mixed in their combining volumes. Thus, if the reaction is as just written, a mixture of two volumes of CO with one of oxygen will ignite at a lower temperature than mixtures of the two gases in any other proportions. It was from such experiments that Falk concluded the reaction was trimolecular.

On the other hand, Rideal's[21] experiments, in which mixtures of hydrogen and CO were passed over an iron oxide catalyst at temperatures from 212 to 750°F. (100 to 400°C.), and Bone and Haward's experiments on the explosion of hydrogen-carbon monoxide-oxygen mixtures pointed toward a dimolecular reaction.

SUMMARY OF VIEWS AS TO COMBUSTION OF CARBON MONOXIDE AND HYDROGEN

Stoichiometric reaction	Mechanism	Investigator	Type of experiment
$2H_2 + O_2 = 2H_2O$	$2H_2 + 2O_2 = 2H_2O_2$ $2H_2O_2 = 2H_2O + O_2$	Falk[20]	Ignition temperature
		Rideal[21]	Low temperature. Catalytic Combustion
	$2H_2 + O_2 = 2H_2O$	Bone & Wheeler[22]	Catalytic at 450°C.
	$2H_2 + O_2 = 2H_2O$	Bodenstein[23]	Combustion at 600°C. Rate proportional to H_2 pressure
		Rowe[24]	Recalculation of Bodenstein
		Bone[18]	Explosion of $H_2 + CH_4$
		Bone & Haward[19]	Explosion of $H_2 + CO$
$2CO + O_2 = 2CO_2$	$2CO + O_2 = 2CO_2$	Bone & Wheeler[22]	Explosion of $CH_4 + XCO$
		Falk[20]	Ignition temperature
	Proportional to 1st power of CO	Rideal[21]	Catalytic combustion
	$H_2 + \frac{1}{2}O_2 = H_2O$ $H_2O + CO = CO_2 + H_2$	Bone & Haward[19]	Explosion of $H_2 + CO$

These different views indicate the strong possibility that the two combustion reactions under discussion take place differently under varying conditions.[22,23,24] Furthermore, the investigations just summarized were not carried out under conditions existing inside combustion chambers. These conditions are: (1) quiet burning under approximately atmospheric pressure, (2) a temperature range of 1300 to 2700°F. (700 to 1500°C.), (3) combustion occurring in free space uncatalyzed by surfaces, and (4) CO and hydrogen burning simultaneously with an excess of oxygen.

Combustion of Hydrogen and CO under Furnace Conditions.—In order to obtain results that meet the above conditions it is necessary to burn these gases and to take out samples for analysis from point to point along the path of combustion. Two series of experiments along this line are available.

First, Kreisinger, Augustine, and Ovitz[25] burned several varieties of coal under varying conditions of excess air and rate of firing, analyzing the resultant flue gases as they passed along a 40-ft. flue. In their experiments, the conditions were exactly those outlined above, both gases burning simultaneously along a flue in which excess oxygen was present. Similar results have been obtained from a smaller furnace fired with producer gas.[26] The actual conditions in the two sets of experiments were as follows:

	Length, feet	Fired with	Temperature range
Furnace 1[25]....	40	Coal	2192–2732°F. (1200–1500°C.)
Furnace 2[26]....	10	Producer gas	1652–2012°F. (900–1100°C.)

The range of concentration of the gases over which the data were obtained was as follows:

TABLE XLVII

VARIATION IN GAS COMPOSITION IN INVESTIGATIONS OF SIMULTANEOUS COMBUSTION OF CO AND H_2

	Concentration of gases in parts per 100 parts of N_2		
	H_2	CO	O_2
Maximum, per cent..............	13.45	24.9	9.63
Minimum, per cent..............	0.12	0.12	2.09
Change.........................	110 fold	200 fold	5 fold

Table XLVIII shows the calculated ratio of $\frac{k_{CO}}{k_{H_2}}$ for the most probable values of m, n, r, and s in the equation

$$\frac{dCO}{dH_2} = \frac{k_{CO}}{k_{H_2}} \frac{(CO)^m(O_2)^n}{(H_2)^r(O_2)^s}, \qquad (19)$$

where the symbols m, n, r, and s are used, since the order of the reactions is uncertain.

TABLE XLVIII

Run, number	$\frac{k_{CO}}{k_{H_2}}$ when				
	$n = s$ $m = 1$ $r = 1$	$n = s$ $m = 2$ $r = 2$	$n = s$ $m = 1$ $r = 2$	$n = s$ $m = 2$ $r = 1$	$n = \frac{1}{2}$ $s = 1$ $m = 1$ $r = 1$
Bureau of Mines[25]					
161	1.47	0.46	0.76	0.93	2.71
134	0.88	0.35	0.72	2 74	1.53
181	1.16	0.37	0.53	1.18	3.30
180	0.89	0.30	0.41	0.66	1.76
172	0.76	0.30	0.37	0.71	1.94
193	1.69	0.45	2.94	0.99	3.89
197	0.79	0.30	0.70	0.42	1.59
199	0.72	0.29	0.76	0.52	1.12
192	3.04	0.35	0.86	1.58	4.59
177	1.02	0.34	0.51	0.74	1.92
178	1.71	0.28	0.31	0.67	5.11
195	0.90	0.34	0.91	1.59	1.47
189	0.95	0.29	0.70	0.55	2.13
Franklin, Tomlinson, and Widell:[26]					
3	0.46	0.41	0.24	0.79	0.90
4	1.07	0.39	0.56	0.76	2.04
Average...................	....	0.35			

Inspection of this table shows the most uniform ratio of $\frac{k_{CO}}{k_{H_2}}$ when $m = r = 2$ and $n = s$. All the other combinations vary considerably from run to run, showing the assumed values of m, n, r, and s, to be incorrect, with the exception of the particular values just mentioned. This indicates that under ordinary combustion conditions the two reactions are:

$$2CO + O_2 = 2CO_2 \text{ and } 2H_2 + O_2 = 2H_2O$$

i.e., both reactions are trimolecular. The ratio of the reaction rate constants $\frac{k_{CO}}{k_{H_2}} = 0.35$, showing that over the temperature range studied the ratio of the rate of combustion of hydrogen to that of CO is 1:0.35 or that hydrogen burns approximately 2.86 times as fast as CO.

This line of reasoning is applicable only in case the velocities of both reactions change with temperature to about the same extent. In other words, with a given change in temperature the per cent change in the velocity of each reaction should be nearly equal. That the velocities of the reactions $2H_2 + O = 2H_2O$ and $2CO + O_2 = 2CO_2$ change to a similar extent with a given change in temperature is shown by the temperature coefficients $\left(\frac{k_{t\,+\,18°F.}}{k_t}\right)$ of these two reactions as determined by Falk.[20]

TABLE XLIX

Temperature Coefficient of Rate of Combustion of Hydrogen and of Carbon Monoxide

	Temperature coefficient of reaction velocity for each 18°F. (10°C.) rise at		
	1472°F. [800°C.]	1652°F. [900°C.]	1832°F. [1000°C.]
Combustion of Hydrogen	1.31	1.13	....
Combustion of Carbon Monoxide	....	1.24	1.14

Falk[20] also has evidence, based on ignition temperatures, that when *both* CO and hydrogen burn simultaneously both reactions are trimolecular. When the reactions take place separately, however, Falk concluded that the combustion of hydrogen with oxygen was bimolecular, while the combustion of CO with oxygen was trimolecular. There is the possibility, therefore, that the mechanism by which CO and hydrogen burn simultaneously (*i.e.*, as they burn in furnaces) may differ from the mechanism by which they burn separately. This view is also supported by the work of Bone and Haward.

Summary.—The mechanism of the combustion of the three elementary fuels is as follows:

1. *Solid Carbon.*—Hot carbon unites with air passing over it, forming carbon dioxide, some carbon monoxide, and an intermediate physicochemical complex C_xO_y. This last compound breaks down to form CO_2 and CO, which are given off and more C_xO_y is formed. At low temperatures the products of combustion may consist largely of CO. At high temperatures, however, such as are encountered in the fuel bed of a furnace, the initial product of combustion is CO_2. This CO_2 may, either wholly or in part, be reduced to CO, the extent of this secondary reaction being dependent on the relative amount of carbon and oxygen, the time of contact between the initially formed CO_2, the reactivity of the carbon surface and the temperature.

2. *Gaseous Hydrocarbons.*—When hydrocarbons burn, there is first an addition or association of oxygen with the hydrocarbon molecule, producing unstable hydroxylated compounds which, in turn, form aldehydes. These aldehydes break down or are oxidized until formaldehyde is produced. Depending on the amount of oxygen present, the formaldehyde may break down thermally to form CO and hydrogen, or it may burn to either CO and water or to CO_2 and water. This process of hydroxylation is very rapid under favorable conditions; it is so rapid, in fact, that in mixtures of methane and hydrogen or methane and CO the hydrocarbon burns faster than either hydrogen or CO. If sufficient oxygen is present so that all the hydrocarbon may be transformed to formaldehyde no soot or smoke will be produced.

Hydrocarbons, especially the higher ones, also may decompose into carbon (soot) and hydrogen by "cracking." This reaction is very rapid at high temperatures.

Under ordinary combustion conditions there is a race between thermal decomposition or "cracking" of the hydrocarbons and the process of hydroxylation. If the conditions (such as preheating hydrocarbon and air, and allowing time for entrance of oxygen into the hydrocarbon molecule) favor hydroxylation, there will be no soot. On the other hand, if the conditions favor "cracking" the hydrocarbon will decompose thermally into carbon and hydrogen, and will burn with a smoky flame.

3. *Carbon Monoxide and Hydrogen.*—Under the conditions of most importance in ordinary combustion work, mixtures of CO and hydrogen burn as if the reactions were

$$2H_2 + O_2 = 2H_2O$$

and

$$2CO + O_2 = 2CO_2$$

These important conditions are (1) quiet burning under approximately atmospheric pressure, (2) at temperatures between 1300 and 2700°F. (700 and 1500°C.), (3) with combustion occurring in free space uncatalyzed by surfaces, and (4) CO and hydrogen burning simultaneously with an excess of oxygen. Under these conditions hydrogen burns about 2.86 times as fast as CO.

BIBLIOGRAPHY

1. Angus Smith: *Proc.* Roy. Soc. (London), **12,** 424 (1863).
2. Sir Lowthian Bell: "The Chemical Phenomena of Iron Smelting," D. Van Nostrand Co. (1872).
3. C. J. Baker: "*J.* Chem. Soc., **51,** 249 (1887).
4. H. B. Baker: *Phil. Trans.*, **179, A,** 571 (1888).
5. Dixon, Strange, and Graham: *J.* Chem. Soc., **59,** 759 (1891).
6. Smithels and Ingle: *J.* Chem. Soc., **61,** 216 (1892).
7. Rhead and Wheeler: *J.* Chem. Soc., **97,** 2178 (1910); **99,** 1140 (1911); **101,** 831 and 846 (1912); **103,** 461 and 1210 (1913).
8. Irving Langmuir: *J.* Am. Chem. Soc., **37,** 1154 (1915).
9. H. B. Dixon: Cantor Lectures on "The Use of Coal Gas" (1884).
10. H. B. Dixon: *Phil. Trans.*, **159** (1893).
11. O. Kersten: *J. prakt. Chem.*, **84,** 290 (1861).
12. Smithels and Ingle: *J.* Chem. Soc., **61,** 204–16 (1892).
13. Bone and collaborators: *J.* Chem. Soc., **81,** 555 (1902); **83,** 1074 (1903); **85,** 693 (1904); **85,** 1659 (1904); **87,** 1248 (1905); **89,** 660 (1906).
14. F. T. Whitworth, R. Abrams and H. D. Kinsey: Mass. Inst. Tech. *Thesis* (1922).
15. Wheeler and Blair: *J.* Soc. Chem. Ind., **61,** 81T, 303T, 331T (1922); **62,** 84T, 87T, 260T, 343T, 415T, 491T (1923).
17. S. H. Katz: "New Views of the Combustion of Volatile Matter in Coal," U. S. Bur. Mines, *Tech. Paper* 183 (1918).
18. Bone and collaborators: *Phil. Trans.*, **215A,** 275 (1915).
19. Bone and Haward: *Proc.* Roy. Soc. (London), **100A,** 67 (1921).
20. Falk: *J.* Am. Chem. Soc., **29,** 1536 (1907).
21. Rideal: *J.* Chem. Soc., **115,** 993 (1919).
22. Bone and Wheeler: *Phil. Trans.*, **206** (1906).
23. Bodenstein *et al.*: *z. physik. Chem.*, **29** (1899); *ibid.* **53,** 166 (1906).
24. Rowe: *Ibid.*, **59,** 41 (1907).
25. Kreisinger, Augustine, and Ovitz: "Combustion of Coal and the Design of Furnaces," U. S. Bur. Mines *Bull.* 135 (1917).
26. Franklin, Tomlinson, and Widell: Mass. Inst. Tech. *Thesis* (1921).
27. Haslam: "The Simultaneous Combustion of Carbon Monoxide and Hydrogen," *J. Ind. Eng. Chem.*, **15,** 679 (1923).

CHAPTER X

COMBUSTION CALCULATIONS

For purposes of design, testing, and a thorough understanding of combustion, certain calculations are indispensable. The accumulation of data is useless unless the data are used to produce results. The efficiency of a combustion process can seldom be measured directly and calculations of some type are almost invariably required. In many cases these calculations are relatively simple and in only a very few cases are they at all difficult, although often they are made up of a series of simple calculations which may appear complex.

Use of Calculations.—From an engineering point of view, perhaps the most important use of such calculations lies in the determination of quantities which would be difficult or almost impossible, to measure directly. For instance in the blast furnace the quantities of air used and gas made almost invariably must be determined by calculation. For this purpose the data required are the amounts and analyses of coke and stone used, the pig made, and an analysis of the blast-furnace gas. All of these are available, with the possible exception of the gas analysis, and this can be secured easily. The actual measurement of such large quantities of gas and air is, on the other hand, both difficult and expensive.

One of the most general applications of such calculations is in the determination of the air used for combustion in a boiler furnace. Any direct measurement of the air is impossible, since it enters the furnace at many points, but the total amount may be calculated readily from easily secured data.

Fundamental Considerations in Combustion Calculations.—All combustion calculations are based on a relatively few simple laws, and on a number of concepts concerning the application of these laws to combustion work. These considerations are discussed in the following paragraphs.

In their simplest form the four fundamental laws on which all combustion calculations depend are:

1. *Conservation of Matter.*—Matter cannot be destroyed and cannot be created.
2. *Conservation of Energy.*—Energy cannot be destroyed and cannot be created.
3. *The Gas Law.*—The volume of a gas is directly proportional to its absolute temperature and inversely proportional to its absolute pressure.
4. *The Law of Combining Weights.*—All substances combine *only* in accordance with a definite, simple relationship as to relative weights.

Law of Conservation of Matter.—The law of conservation of matter, in simple terms, states that over a sufficient period of time, whatever material enters a process must also leave it. There can be no destruction or production of matter during the process, although the form or the combinations usually change. Thus the materials entering a boiler furnace are air and fuel. Part of the fuel leaves in the cinder* and part goes up the stack as flue gas. The air all leaves as flue gas. The total material in the cinder and flue gas must equal the total material in the fuel and air used. A similar balance also exists for each of the elements present. For example, the carbon entering in the fuel may separate, so that a part remains in the cinder, part is in the flue gas as carbon monoxide (CO) and as carbon dioxide (CO_2), and part is deposited as soot. The total carbon output, however, must equal the total carbon entering in the fuel, that is, the carbon entering in the fuel over a suitable test period must equal the carbon leaving either as cinders or as flue gas.

Under certain circumstances when the rate of flow of material through the process is not strictly continuous, it is possible for a temporary accumulation of material to take place in the furnace If data are obtained over a sufficiently long period of time, however, this accumulation will be negligible in comparison with the total material that has passed through the process during the test period. This may be illustrated by an efficiency test on a boiler where the amount of coal on the grate is kept as near constant as possible, but, in order to eliminate any serious error from this source, the test is run over a long period, usually 24 hr. or more. An accumulation of even 500 lb. of coal on the grate

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* In order to avoid confusion between the ash in the original coal and the "ash" as residue from combustion, the term *cinder* is used to denote the latter.

of a furnace using only 1000 lb. of coal per hr. is then but 2 per cent of the total heat input. If the data were obtained over an 8-hr. period with the same possible error in the amount of coal on the grate, the deviation would be as high as 6 per cent. Therefore, it is essential, in using a balance of this type, to ascertain the possible accumulation during the time period over which the test is made. This is usually the main factor that determines the length of time over which an efficiency test should be run.

Law of Conservation of Energy.—The conservation of energy is the same principle applied to energy instead of to matter. In the boiler example above, the heat in the fuel is distributed so that some goes into the water to form steam, some remains undeveloped in the carbon of the cinder, some is lost up the stack as sensible heat in the gases and possibly as undeveloped heat in CO, and some is lost by radiation from the furnace and flues. The total heat energy, utilized and lost, must, however, over a suitable period of test, equal the heat energy in the fuel used.

The Gas Law.—The fundamental gas law is expressed by the equation

$$PV = NRT$$

Where

R is the gas constant.
P is the absolute pressure.
V is the volume.
N is the number of *mols*.
T is the absolute temperature of the gas.

To use this equation to its fullest extent, in fact, to simplify greatly the whole system of combustion calculations, it is necessary to have a clear understanding regarding the mol. A mol of a substance is the number of grams, kilograms, or pounds of the substance equivalent to the molecular weight. If the unit of weight is grams, this quantity of material is called a gram-mol, while if the unit of weight is pounds, it is called the pound-mol. *In this chapter a mol always refers to the pound-mol** (for example, the molecular weight of carbon is 12 and, therefore, 1 mol of carbon equals 12 lb.). When the chemical composition of a gas is known, its molecular weight can readily be determined. The accompanying table shows the molecular weights for the common gases.

*See foot-note on p. 138.

TABLE L

MOLECULAR WEIGHT OF GASES

Gas	Symbol	Molecular weight
Hydrogen	H_2	2.02
Oxygen	O_2	32.0
Carbon monoxide	CO	28.0
Carbon dioxide	CO_2	44.0
Methane	CH_4	16.0
Ethylene	C_2H_4	28.0
Nitrogen	N_2	28.0
Water vapor	H_2O	18.0

Avogadro's hypothesis says that equal numbers of molecules of perfect gases occupy the same volume under identical conditions. Thus, *one molecular weight* of any gas in pounds (the pound-mol) will occupy a *definite volume* at standard conditions, these conditions being a temperature of 32°F. (0°C.) and an absolute pressure of 1 atmosphere (760 mm. or 14.7 lb. per sq. in.). This volume has been established as 359 cu. ft., and consequently a pound-mol (2.02 lb.) of hydrogen, a pound-mol (32 lb.) of oxygen, or a pound-mol (16 lb.) of methane at 32°F. and 760 mm. of mercury (Hg) pressure will each occupy 359 cu. ft.*

This hypothesis is true only for perfect gases, and is not exact in all cases. Any gas which is at a temperature far above its boiling point acts as a perfect gas. Thus, for industrial gases, except for water vapor, the hypothesis is sufficiently exact, while even in this case the deviation, from the standpoint of combustion calculations, is slight.

This hypothesis holds not only for all perfect gases but also for mixtures of gases. For example, ⅓ mol of hydrogen, ⅓ mol of oxygen, and ⅓ mol of nitrogen must together occupy 359 cu. ft. Thus, the mixture of gases acts as if it were a single gas with a molecular weight equal to the *average* of the molecular weights of its components. In the case just mentioned the average molecular weight would be

$$(\tfrac{1}{3} \times 2) + (\tfrac{1}{3} \times 32) + (\tfrac{1}{3} \times 28) = 20.67,$$

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* In fact, the molecular weight of a gas, such as O_2, N_2, etc., may be defined as the weight in pounds of 359 cu. ft. of the gas under standard conditions (32°F. and an absolute pressure of 1 atm).

and therefore 20.67 lb. of this gaseous mixture at 32°F., and 1 atmosphere pressure will occupy 359 cu. ft. By this means it is very simple to calculate the weight per cu. ft. of any known mixture of gases. Atmospheric air contains 78.14 per cent nitrogen, 20.92 per cent oxygen, 0.04 per cent CO_2, and 0.90 per cent argon and its average molecular weight is 28.97. For ordinary combustion calculations, since argon appears in the analysis as nitrogen, the composition of air may be taken as 79 per cent nitrogen and 21 per cent oxygen, and its molecular weight as 29.0.

If in the equation $PV = NRT$, P is expressed in pounds per sq. in. absolute (14.7 + pounds per sq. in. gage pressure), V in cubic feet, N in pound-mols, and T in degrees Fahrenheit absolute (460 + °F.), the value of R is 10.72. It is seldom necessary to use the equation in this form, as its greatest value is in calculating, by means of ratios, the effect of a change in temperature or pressure on gas volume. Thus, if a gas is at pressure P, volume V, and temperature T, and the conditions are changed so that the pressure is P_2, volume V_2, and temperature T_2, the following relationship is true:

$$\frac{P_2V_2}{P_1V_1} = \frac{T_2}{T_1} \text{ or } \frac{V_2}{V_1} = \frac{T_2P_1}{T_1P_2}.$$

The volume of gas is *directly proportional* to the absolute temperature and *inversely proportional* to the *absolute pressure.* For example, 3000 cu. ft. of gas at 2 atmospheres gage pressure (29.4 + 14.7 = 44.1 lb. per sq. in. absolute pressure) and 800°F. (800 + 460 = 1260°F. absolute temperature.) is equivalent to

$$3000 \times \frac{460 + 32}{1260} \times \frac{44.1}{14.7} = 3510 \text{ cu. ft.}$$

at standard conditions (14.7 lb. per sq. in. or 760 mm. pressure and 32°F.) or this volume of gas is equal to $\frac{3510}{359} = 9.78$ mols.

Similarly, the change in the volume of a mol (359 cu. ft.) brought about by changing the pressure and the temperature from standard conditions is as follows:

$$359 \times \frac{\text{Final Absolute Temperature of Gas}}{\text{Absolute Temperature of Gas under Standard Conditions}} \times \frac{\text{Standard Pressure (Absolute)}}{\text{Final Pressure (Absolute)}}$$

$$= 359 \times \frac{(T_2 + 460)}{(32 + 460)} \times \frac{(14.7)}{(14.7 + P_2)},$$

where T_2 is the final temperature of the gas in degrees Fahrenheit and P_2 is the final pressure of the gas in pounds per sq. in. gage pressure.

Law of Combining Weights.—The law of combining weights states that the elements combine in simple and constant proportions to form definite compounds. These proportions are always in a simple ratio to the molecular weights, so that all reactions may be expressed on a mol basis by using whole numbers to denote the number of mols of each substance taking part in the reaction. The use of this law requires a knowledge of the relative proportion in which substances combine. These amounts are shown by chemical equations, in which the chemical symbols of the substances involved represent 1 mol unless otherwise designated by a number placed before the symbol.

The combustion reactions most commonly encountered in furnace calculations, written to include the nitrogen from the air which almost invariably furnishes the necessary oxygen for combustion, are as follows:

$$C + O_2 + 3.76N_2 = CO_2 + 3.76N_2 \quad (1)$$
$$2C + O_2 + 3.76N_2 = 2CO + 3.76N_2 \quad (2)$$
$$2CO + O_2 + 3.76N_2 = 2CO_2 + 3.76N_2 \quad (3)$$
$$2H_2 + O_2 + 3.76N_2 = 2H_2O + 3.76N_2 \quad (4)$$
$$CH_4 + 2O_2 + 7.52N_2 = CO_2 + 2H_2O + 7.52N_2 \quad (5)$$

The first four of these reactions are the ones most commonly encountered. Reaction (1) shows that 1 mol of carbon (12 lb.) unites with 1 mol of oxygen (32 lb. or 359 cu. ft. under S. C.*) to form 1 mol of CO_2 (44 lb. or 359 cu. ft. S. C.). Reaction (2) shows that 2 mols of carbon (24 lb.) unite with 1 mol of oxygen (32 lb.) to form 2 mols of CO (56 lb. or 718 cu. ft. S. C.), which may then, in accordance with reaction (3), unite with another mol of oxygen to form 2 mols of CO_2. Reaction (4) shows that 2 mols of hydrogen (4 lb.) unite with 1 mol of oxygen (32 lb.) to form 2 mols of water (36 lb. or 718 cu. ft. S. C.). It is on the basis of the combining proportions as represented in these equations that most combustion calculations are founded.

If, in accordance with reaction (5), we wish to calculate the air required to burn completely 1 mol of methane, we find that 1 mol of oxygen is required for the carbon (reaction 1) plus

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* S. C. refers to the "Standard Conditions" of 32°F. and 1 atmosphere absolute. (760 mm. Hg or 14.7 lb. per sq. in.).

1 mol of oxygen for the 2 mols of hydrogen (reaction 4). Thus the *air* required is $2\frac{(100)}{(21)}$, or 9.52 mols air per mol of methane, since there is only 21 per cent of oxygen in the air. Since molal relationships are volume relationships, 9.52 cu. ft. of air would be required also to burn 1 cu. ft. of methane.

The Choice of Units.—Before applying these four principles to the solution of problems, the units in which the problem is to be worked out should be specified. Undoubtedly, the metric system has distinct advantages in many cases, but the final answer to most engineering problems must be expressed in English units. For this reason the English unit is used here throughout. Data in the literature are often given in metric units and for facilitating the conversion to English units a table of corresponding values is given below:

TABLE LI*

CONVERSION FACTORS FROM ENGLISH TO METRIC UNITS

ENGLISH UNIT	METRIC EQUIVALENT
1 in.	= 2.54 cm.
1 ft.	= 0.305 m.
1 cu. ft.	= 0.0283 cu. m. = 28.3 liters
1 lb.	= 0.454 kg. = 454 gm.
1°F.	= 0.555°C.
Degrees Fahrenheit	= 9/5 (degrees Centigrade) + 32
1 B.t.u.	= 0.252 cal.
1 B.t.u. per lb.	= 0.555 cal. per gm.
1 hp.	= 0.746 kw.

* For more complete tables see Appendix IV.

Use of Ratios.—The use of ratios greatly simplifies the solution of combustion problems. Fundamentally, a ratio has no units and is, therefore, independent of the system employed. If the weights of two materials are in a definite ratio, then this relationship holds for any unit of weight. Thus if one substance has twice the specific gravity of another, then for any definite volume the first will weigh twice as much as the second, irrespective of whether the weight is expressed in grams, ounces, or pounds. If 10 lb. of water are evaporated in a boiler per lb. of coal fired, the same relationship may also be expressed as 10 gm. of water evaporated per gm. of coal or 10 tons of water per ton of coal. The same is true of volume relationships. If the air required for combustion of a gas is 5 cu. ft. per cu. ft. of gas, it is also 5 cu. m. per cu. m. of gas, or, since a pound-mol is a definite volume, it is 5 mols of air per mol of gas. By similar reasoning an *analysis of a gas being on a volume basis gives directly the number of mols of each constituent per 100 mols of gas.* Since specific heats and specific gravities are merely ratios, these values for a given material are the same in any units.

The Selection of a Basis for Calculation.—For the solution of combustion problems it is necessary to choose some convenient basis of calculation (*e.g.*, 100 lb. of coal fired, etc.) in order that the various steps in the calculation may be comparable. The basis used in calculation must be chosen at the outset and rigidly adhered to throughout and care must be taken to keep all relationships on the same basis. To compare directly the per cent carbon in producer gas with the per cent carbon in the flue gas is obviously inaccurate. In this case a basis such as 100 mols of flue gas or 100 mols of producer gas may be used and the corresponding volume of the other gas must be calculated, usually from a carbon balance. The amount of carbon in this calculated volume may then be compared directly with the carbon in 100 mols of the gas originally taken as a basis. Similarly, deduction of the carbon per 100 lb. of cinder from the carbon per 100 lb. of coal would have no meaning. In this case 100 lb. of coal may be the basis and the amount of cinders determined by an ash balance. The carbon in this corresponding amount of cinders may then be deducted from the carbon per 100 lb. of coal without error. The choosing of a basis for calculations and the relating of all quantities to this basis are absolutely essential for accuracy in solving combustion problems.

Analysis.—All combustion calculations necessitate the use of analyses for the determination of the quantities of various elements. Analyses are fully as important as absolute weight measurements in most cases, and frequently relative weights, as calculated from analyses, are sufficient for the purpose intended. For example, the per cent stack and cinder losses from a boiler may be quite accurately determined from analyses and temperatures alone, while actual weight or volume measurements usually cannot be made.

Analyses of solids, such as coal, are always reported on a *weight percentage basis.* In order to convert these to a mol basis, it is necessary to divide each constituent by its molecular weight. Thus a coal analyzed as 72 per cent carbon, 4 per cent hydrogen, and 6 per cent oxygen would contain 6 mols of carbon ($^{72}\!/_{12}$ =

6), 2 mols of hydrogen (4/2 = 2), and 0.187 mols of oxygen (6/32 = 0.187) per 100 lb. of coal. In most instances it is most advantageous to use this expression of mols per 100 lb. rather than a complex mol per cent, because the ultimate result will be expressed in pounds.

Analyses of gases, obtained with the Orsat apparatus or some modification of it, are on a *volume percentage basis*. Thus an Orsat analysis of air would show 21 per cent oxygen and 79 per cent nitrogen. Since the mol is a volume unit, corresponding to 359 cu. ft. at standard conditions, *gas analyses give molal compositions directly*. For example, 100 mols of air (100 × 359 = 35,900 cu. ft.) contain 21 mols of oxygen (21 × 359 = 7539 cu. ft.) and 79 mols of nitrogen (79 × 359 = 28,361 cu. ft.). The realization of this point greatly simplifies calculations.

Due to the analytical method employed, gas analyses as obtained and reported (from the Orsat determination) are always on the dry basis, *exclusive* of the moisture present. Gas analyses are usually made over water, but all the measurements are made at the same temperature and therefore with the same partial pressure of water vapor in the gas. Therefore, if 10 per cent of the total gas (a gas containing 10 per cent CO_2, for example) is absorbed by some reagent, 10 per cent of the water vapor originally present will condense out, so that the partial pressure of water vapor remains unchanged. Thus the final result is the same as if dry gas were used.

Volume Relationships.—Much use is made of gas analyses in combustion calculations in determining relative volumes. Thus if the analysis of a fuel gas and of the flue gas formed from it are known, the relative amounts of fuel gas, air, and flue gas may be determined by use of the fundamental concept of material balances. Consider, for example, the following analyses obtained from a furnace using methane as fuel:

Fuel gas		Flue gas			
Constituent	Per cent	Constituent	Per cent	Mols C	Mols O_2
CH_4	100	CO_2	5.5	5.5	5.5
		O_2	11.1		11.1
		N_2	83.4		
		Total	100.0	5.5	16.6

Air—Flue Gas Ratio and Carbon Balance.—The relative volumes of fuel gas and flue gas may be determined from a carbon balance (*i.e.* carbon in fuel gas equals carbon in flue gas). Taking as a basis 100 mols of dry flue gas, there are 5.5 mols C (as CO_2), 16.6 mols total O_2 (5.5 mols as CO_2 and 11.1 as O_2), and 83.4 mols of N_2. But all the carbon came from the fuel gas, so that 5.5 mols of methane (since there is 1 mol of carbon per mol of CH_4) were used to form 100 mols of dry flue gas. Thus the ratio of dry flue gas to fuel gas is $\frac{100}{5.5}$, or 18.2, mols dry flue gas per mol fuel gas. Since molal relationships are also volume relationships, there were also 18.2 cu. ft. of dry flue gas for every cubic foot of dry methane measured under the same conditions.

Dry Flue Gas—Air Ratio and Nitrogen Balance.—By the use of a nitrogen balance (*i.e.*, nitrogen in air = nitrogen in flue gas) the ratio of dry flue gas to dry air may be calculated. In this example there are 83.4 mols of N_2 per 100 mols of dry flue gas. This N_2 came from air which contains 79.0 per cent N_2 (79 mols of N_2 per 100 of air), so that to obtain 83.4 mols of N_2, $83.4 \times \frac{100}{79.0}$ mols of air were necessary. The ratio $\frac{\text{dry flue gas}}{\text{dry air}}$ is thus $\frac{100}{83.4 \times \frac{100}{79.0}} = \frac{79.0}{83.4} = 0.948$ mols of dry flue gas per mol of air, or 0.948 cu. ft. of dry flue gas per cu. ft. of air used for the combustion of methane.

Fuel Gas-Air Ratio.—The ratio of air to fuel gas, methane, may now be determined from the above relationships. Since there were 18.2 cu. ft. of flue gas per cu. ft. methane and 0.948 cu. ft. of dry flue gas per cu. ft. of air, the ratio of air to methane is $\frac{18.2}{0.948}$, or 19.2 cu. ft. air used per cu. ft. of methane.

Excess Air.—The term *per cent excess air* is used to denote the per cent of air in excess of that theoretically required for *complete* combustion. Thus, for methane, taking 100 mols as the basis, we have: 100 mols CH_4 are equivalent to 100 mols of C, and 200 mols of H_2. The carbon requires 100 mols of O_2 for combustion, and the H_2 also requires $^{200}/_{2} = 100$ mols of O_2. The corresponding quantity of air is $(100 + 100)\ ^{100}/_{21} = 952$ mols. In other words, 1 mol (or 1 cu. ft.) of methane requires 9.52 mols (or cu. ft.) of air for combustion. In the example given above as shown by the fuel gas–air ratio, the amount actually used was 19.2 cu. ft. of air per cu. ft. of gas, so that the per cent excess air is $\frac{19.20 - 9.52}{9.52} \times 100 = 101.7$ per cent. Since the N_2 in the gas analysis indicates the total air used while the free O_2 indicates the excess, the excess air may be calculated from the flue-gas analysis without the fuel analysis, *provided the original fuel contains little or no nitrogen.* Thus in the above example the O_2 corresponding to 83.4 mols of N_2 is 22.17 mols, while the free O_2 in the flue gas is 11.1 mols. The O_2 required is then 11.07 mols and the excess air in the above example is, therefore, $\frac{22.17 - 11.07}{11.07} \times 100 = 100.3$ per cent. This value does not check exactly with that obtained by the previous method, owing to a slight error in the gas analysis.

When CO is present in the flue gas the amount of oxygen necessary to burn it must be deducted from the free oxygen in the gas before determining the excess air. For every mol of CO, ½ mol of O_2 is necessary (reaction (3)). The excess oxygen is the free O_2 minus one-half the CO. If the analysis of a flue gas from combustion of a nitrogen-free fuel is 12 per cent CO_2, 5 per cent O_2, 2 per cent CO, and 81 per cent N_2, the *total* O_2 corresponding to the N_2 is $81 \frac{(21)}{(79)} = 21.5$ (where $\frac{21}{79}$ is the ratio of O_2 to N_2 in air) and the *excess* O_2 is $5 - \frac{(2)}{(2)} = 4$. The per cent excess air for this case, therefore, is $\frac{4}{21.5 - 4} \times 100 = 22.85$ per cent.

Net Hydrogen.—The net hydrogen in a fuel (see Chap. II, p. 44), or the hydrogen in excess of that which can combine with the oxygen present, may also be calculated from the flue-gas analysis. The oxygen of the air is used to burn carbon and hydrogen, with an excess which appears in the analysis as free oxygen. Since the water formed from the combustion of the net hydrogen does *not* appear in the analysis (because the water formed was condensed during analysis), it must be calculated. This may be done most readily by the use of an oxygen balance. Thus in the example above for the case of methane, the total oxygen from the air, corresponding to the 83.4 mols of N_2 formed in the flue gas, was 22.17 mols, of which 16.6 appeared as CO_2 and O_2 in the gas analysis. Then $22.17 - 16.6 = 5.57$ mols of O_2, corresponding to 11.1 mols of H_2O or H_2 (since 1 mol of O_2 is equivalent to 2 mols of H_2O or H_2), were formed for each 100 mols of flue gas. There are 5.5 mols of carbon per 100 of flue gas, so that the ratio of net hydrogen to carbon in the fuel gas is calculated as $\frac{11.1}{5.5}$, or 2 mols of H_2 per mol of C.

Moisture in Flue Gas.—The moisture present in the flue gas is the sum of the water from free and combined moisture in the fuel, from combustion of the hydrogen and from the moisture in the air. This latter factor is often so small that it may be neglected. When the analysis of the fuel is known, the moisture in the flue gas (from moisture in the fuel plus that from combustion of H_2) may be calculated directly. Thus if dry methane were used the volume of water vapor would be 2 mols per mol of *fuel* gas, or $\frac{2}{18.2} \times 100 = 11.1$ mols per 100 mols of dry *flue* gas. Thus, for each 100 volumes or cu. ft. of dry flue gas there would be 111 volumes or cu. ft. of wet flue gas. When the analysis of the fuel is not known, the water vapor from the net hydrogen alone may be calculated from the flue-gas analysis. In the above example there were 11.1 mols of water per 100 mols of dry flue gas, which agrees with the value calculated from the fuel analysis, since there was no oxygen present in the fuel. Since the net hydrogen in most fuels is not a very large proportion of the total hydrogen, and the accuracy of analysis of both solid and gas for H_2 and O_2 is susceptible to error, the calculation of moisture in the gas from the flue-gas analysis is of sufficient accuracy for most purposes.

Heat Energy.—The energy involved in combustion calculations is heat energy. It is, therefore, necessary to know the heat effects attending any changes in state which take place in the process under consideration. The important change taking place in most cases is the oxidation of the carbon and hydrogen in the fuel with the oxygen in the air. A knowledge of the heat attending this reaction, or the heat of combustion of the fuel, is essential for all efficiency calculations.

Heats of Combustion.—The heat of combustion of the pure elementary fuels is shown in Table LII, which gives values of the heat developed for gas measured under different conditions.

The heat of combustion of pure *carbon* is 14,544 *B.t.u. per lb.*

Although the B.t.u. per mol or per cu. ft. at 32°F. and 29.92 in. Hg (1 atmosphere) are the most convenient for purposes of cal-

TABLE LII

Combustion Constants of Gases

←cu. ft. per cu. ft. fuel gas or mol per mol→

Gas	Formula	Molecular weight	Weight per cu. ft. (dry)	Specific gravity (air = 1.0)	B.t.u. per cu. ft. high (gross) 60°F.; 30" sat. H_2O	B.t.u. per cu. ft. low (net) 60°F.; 30" sat. H_2O	B.t.u. per cu. ft. high (gross) 60°F.; 30" dry	B.t.u. per cu. ft. low (net) 60°F.; 30" dry	B.t.u. per cu. ft. high (gross) 32°F.; 30"	B.t.u. per cu. ft. low (net) 32°F.; 30"	B.t.u. per mol high (gross)	B.t.u. per mol low (net)	Required for combustion.		Flue products			Theoretical flame temperature, °F.
													Oxygen	Air	CO_2	H_2O	N_2	
Hydrogen	H_2	2.02	0.00531	0.0695	318.8	269.1	324.5	274.5	343.0	290.0	123,100	104,100	0.5	2.38	....	1.0	1.88	4010
Carbon monoxide	CO	28.0	0.07381	0.9665	317.1	317.1	322.6	322.6	341.0	341.0	122,400	122,400	0.5	2.38	1.0	.....	1.88	4475
Carbon dioxide	CO_2	44.0	0.11598	1.5188														
Oxygen	O_2	32.0	0.08435	1.1046														
Nitrogen	N_2	28.0	0.07385	0.9671														
Paraffins																		
Methane	CH_4	16.0	0.04226	0.5534	995	896	1012	912	1069	963	384,000	346,000	2.0	9.52	1.0	2.0	7.52	3750
Ethane	C_2H_6	30.0	0.07920	1.0371	1731	1582	1762	1612	1862	1703	668,300	611,300	3.5	16.67	2.0	3.0	13.17	3820
Propane	C_3H_8	44.1	0.11615	1.5210	2465	2266	2509	2309	2652	2440	952,000	876,000	5.0	23.80	3.0	4.0	18.80	3840
Butane	C_4H_{10}	58.1	0.15309	2.0047	3204	2955	3261	3010	3445	3180	1,237,000	1,142,000	6.5	30.95	4.0	5.0	24.45	3870
Olefines																		
Ethylene	C_2H_4	28.0	0.07389	0.9676	1613	1514	1641	1541	1734	1631	622,400	584,400	3.0	14.29	2.0	2.0	11.29	4250
Propylene	C_3H_6	42.0	0.11083	1.4513	2313	2164	2354	2204	2487	2328	893,000	836,000	4.5	21.43	3.0	3.0	16.93	4090
Butylene	C_4H_8	56.1	0.14778	1.9352	3033	2834	3087	2887	3262	3050	1,171,000	1,095,000	6.0	28.58	4.0	4.0	22.58	4030
Acetylene	C_2H_2	26.0	0.06858	0.8981	1456	1396	1483	1433	1567	1514	562,000	543,000	2.5	11.90	2.0	1.0	9.40	4770
Aromatics																		
Benzene	C_6H_6	78.0	0.2057	2.6940	3658	3509	3722	3577	3933	3774	1,413,000	1,356,000	7.5	35.70	6.0	3.0	28.20	4110
Toluene	C_7H_8	92.1	0.2427	3.1778	4364	4165	4441	4241	4693	4481	1,685,000	1,609,000	9.0	42.87	7.0	4.0	33.87	4050
Xylene	C_8H_{10}	106.1	0.27962	3.6616	5064	4815	5153	4902	5446	5181	1,955,000	1,860,000	10.5	50.00	8.0	5.0	39.50	4010
Naphthalene	$C_{10}H_8$	128.1	....	....	....	....	....	....	....	....	2,219,000	2,143,000	12.0	57.10	10.0	4.0	45.00	4100
Ammonia	NH_3	17.0	0.04489	0.5878														
Hydrogen sulfide	H_2S	34.1	0.08986	1.1767	......	......	......	......	......	......	....	....	1.5	7.14	SO_2 1.0	1.0	5.64	
Water	H_2O	18.0	0.04749	0.6219														

Notes:

(1) Molecular weights to nearest tenth.
(2) Weight per cu. ft. of gases, and densities from "Combustion," Am. Gas. Assoc. (1924), p. 18.
(3) Heating values per mol from weighted mean of values given in Landolt Börnstein Tabellen, 1923 ed.
(4) Low heating values calculated from gross by subtracting heat of vaporization of water at 60°F. (15.6°C.).
(5) Flue products figured with theoretical air (21 per cent O_2, 79 per cent N_2).
(6) Flame temperatures not corrected for dissociation.

culation, it so happens that the standard values used in the gas industry are those at 60°F., 30 in. of Hg, and saturated with water vapor. The values at 32°F. and 29.92 in. Hg may be calculated from those at 60°F., 30 in. Hg, saturated, by multiplying by the factor 1.073.

Gross and Net Heating Values.—The heating values given in the Table LII include the net or low and the gross or high values. When hydrogen is burned to form water and the products cooled to the initial temperature the heat liberated is called the gross or high heating value and includes the heat of condensation of the water. This value is obtained by ordinary calorimetric measurements (for example, by burning a gas in a Junker's calorimeter). The net or low heating value assumes the products cooled to the initial temperature, but the water *not* condensed. The difference in these values is equivalent to the heat of vaporization of water at the initial temperature. The low heating values of the various gases may be obtained by subtracting 19,000 B.t.u. per mol, or 50 B.t.u. per cu. ft. at 60°F., 30 in. Hg and saturated with H_2O vapor for each mol of hydrogen present in a mol of the original gas.

Use of Net Heating Value.—Obviously, the efficiency of a furnace will differ depending upon whether the high or the low heating value is used. This discrepancy is not great when fuels of low hydrogen content are used, but in fuels with high hydrogen content the difference may be quite appreciable. When a fuel such as coke-oven gas containing 1.20 mols of hydrogen per mol of gas is used, the high or gross heat of combustion may be 600 B.t.u. per cu. ft., while the net heating value is only 540 B.t.u. per cu. ft., or a difference of over 10 per cent. If the gross heating value is used the efficiency of the process will be 10 per cent lower than if calculated on the low heating value, as the per cent heat loss in the stack will be higher, since the heat of vaporization of the water must be considered as a loss when the higher heating value is used. Since it is impossible to recover this heat of vaporization in any industrial furnace, it is not proper to consider it as a valuable heat input to the system. The use of the gross value in comparing a low-hydrogen and a high-hydrogen fuel may lead to erroneous conclusions. For example, although the gross heating value of hydrogen is the same as that of CO, the net value is 15 per cent lower. The rational basis for the comparison of various fuels and for the calculation of efficiencies is the net heating value. This value cannot easily be determined experimentally, so that the gross heating value is almost invariably reported and used in practice.

Specific Heat.—The specific heat of a substance is the ratio of the heat required to raise 1 lb. of the material in question 1°F. to the heat required to raise 1 lb. of water 1°F. It may be considered as the B.t.u. necessary to raise 1 lb. of the material 1°F., since 1 B.t.u. will raise 1 lb. of water* 1°F. The heat necessary to raise a substance to any given temperature is thus the product of the weight in pounds, the temperature rise in degrees Fahrenheit, and the mean specific heat over the temperature range in question.

Specific Heat and Heat Capacity of Gases.—Specific heats of most materials, especially gases, may be more simply expressed as *molal* specific heats, that is, the heat required to raise 1 *mol* of the substance 1°F. The relation between the molal specific heats of gases at constant pressure, MC_p, which is the specific heat commonly used, is very simple. The molal heat capacities, or molal specific heats, of the permanent diatomic gases (2 atoms per molecule, such as O_2) are identical. Likewise, the molal heat capacities of the triatomic gases, CO_2 and SO_2, are similar. The molal heat capacity varies greatly with temperature, and the equations expressing this relationship for the most important gases are shown in the following table. These are the values given by G. N. Lewis and M. Randall.[1]

TABLE LIII

MOLAL SPECIFIC HEATS OF GASES

(Where T equals °F. absolute or °F. + 460.)

N_2, O_2, CO, NO	$MC_p = 6.50 + 0.00055T$
H_2	$MC_p = 6.50 + 0.00050T$
Cl_2, Br_2, I_2	$MC_p = 7.4 + 0.00056T$
H_2O, H_2S	$MC_p = 8.81 - 0.00105T + 0.000000685T^2$
CO_2, SO_2, N_2O	$MC_p = 7.0 + 0.0039T - 0.000000574T^2$
CH_4	$MC_p = 7.5 + 0.00277T$
NH_3	$MC_p = 8.04 + 0.000388T + 0.000001577T^2$

For all but very exact work, or unless pure hydrogen is being considered, the molal heat capacity of hydrogen may be considered the same as the molal heat capacity of the other diatomic gases (N_2, O_2, etc.). This has been done in Fig. 49.

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* See footnote on p. 138.

The specific heats of gases and the heat capacity of gases (*i.e.*, the total heat content of the gas above some convenient temperature level) are extremely important in combustion work. A knowledge of these properties is essential for determining the amount of heat carried out the stack by flue gases, for calculations of flame temperature (as shown in the next chapter), etc. General considerations applying to specific gases have just been given, and a series of formulas tabulated with which it is possible to calculate the specific heat of any of the common gases at any desired temperature.

Thus, the equation for the molal specific heat of CO_2 is

$$MC_p = 7.0 + 0.0039T - 0.000000574T^2$$

where MC_p is the molal specific heat (B.t.u. required to raise one molecular weight in pounds 1°F.) at constant pressure, and T is in degrees Fahrenheit absolute (degrees Fahrenheit + 460). For example, the molal specific heat of CO_2 at 60°F. may be calculated by means of this formula as follows:

$$\begin{aligned} MC_{p_{60°\text{F.}}} &= 7.0 + 0.0039\,(460 + 60) - 5.74 \times 10^{-7}\,(460 + 60)^2 \\ &= 7.0 + 2.03 - 0.155 = 8.88. \end{aligned}$$

In combustion calculations, however, the specific heat at a single temperature is seldom desired. Usually it is necessary to know the amount of heat required to raise a gas to a given temperature, and for this purpose the mean specific heat over the given temperature range must be known. The mean specific heat is determined by integrating the formulas for specific heat, thus determining the total heat in the gas, and then dividing by the temperature change to find the mean. For example, the total heat equation for CO_2 (obtained by integrating the above equation for specific heat) shows the total heat in the gas above −460°F. (−273°C.) and is as follows:

$$\text{Total heat in } CO_2 \text{ above } -460°\text{F.} = 7.00T + 1.95 \times 10^{-3}T^2 - 1.91 \times 10^{-7}\,T^3,$$

where the total heat is expressed in B.t.u. per lb.-mol and T is in degrees Fahrenheit absolute. At 60°F. the total heat above −460°F. (−273°C.) for 1 mol of CO_2 then is:

$$\begin{aligned} \text{Total heat of } CO_2 \text{ at } 60°\text{F.} &= 7.00\,(460 + 60) + 1.95 \times 10^{-3}\,(460 + 60)^2 - 1.91 \times 10^{-7}\,(460 + 60)^3 \\ &= 3640 + 527 - 27 = 4140 \text{ B.t.u. per mol,} \end{aligned}$$

and, similarly, the total heat at 2040°F. (1116°C.), above −460°F. (−273°C.), is:

$$\text{Total heat of } CO_2 \text{ at } 2040°\text{F.} = 7.00\,(460 + 2040) + 1.95 \times 10^{-3}\,(460 + 2040)^2 - 1.91 \times 10^{-7}\,(460 + 2040)^3$$
$$= 17{,}500 + 12{,}180 - 2980 = 26{,}700 \text{ B.t.u. per mol.}$$

The total heat above 60°F. (15.6°C.) of 1 mol of CO_2 at 2040°F. (1116°C.) then is 26,700 − 4140 = 22,560 B.t.u., and the mean molal specific heat over the temperature range from 60°F. (15.6°C.) to 2040°F. (1116°C.) is $\frac{22,560}{(2040 - 60)} = 11.39$ B.t.u. required, on the average, to raise 1 mol (44 lb.) of CO_2 through each 1°F. over the specified temperature range. This value can, of course, be converted to the specific heat per lb. by dividing by the molecular weight: thus $\frac{11.39}{44} = 0.259$, the number of B.t.u. required to raise 1 lb. of CO_2 1°F. over the temperature range specified.

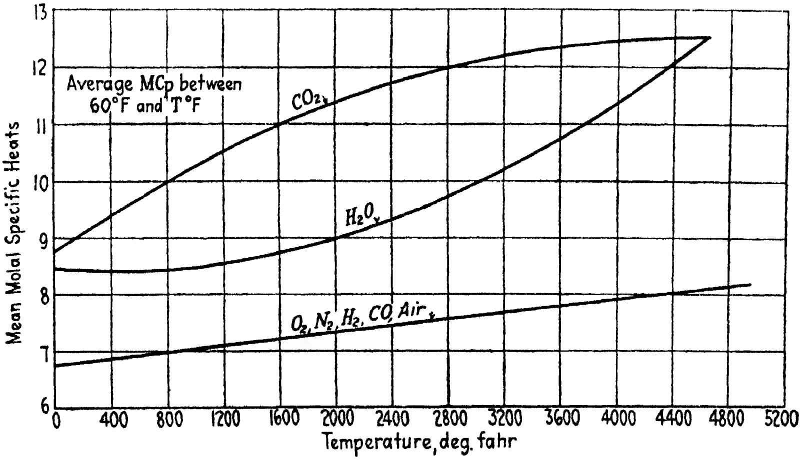

FIG. 48.—Mean molal specific heats of CO_2, H_2O, CO, N_2 and O_2 between 60°F. and temperatures up to 4600°F. The accuracy of values above 3800°F. is doubtful.

Values of the mean molal specific heats of the common gases from 60°F. (15.6°C.) up to 4500°F. (2482°C.) calculated in this way have been plotted in Fig. 48. For most combustion calculations it is sufficiently accurate to read from this plot the mean molal specific heats over the desired temperature range and use this directly. For example, suppose it is desired to determine the amount of heat required to heat 2.5 mols of air (*i.e.*, 2.5 × 29 = 72.5 lb.) from 60°F. (15.6°C.) to 2500°F. (1371°C.). Reading from Fig. 48 it is found that the mean molal specific heat over this temperature range is about 7.45; the heat required is then calculated to be (2.5) (7.45) (2500 − 60) = 45,450 B.t.u.

Part of this last calculation may be solved graphically with the aid of Fig. 49 which shows the total heat contents above 60°F.

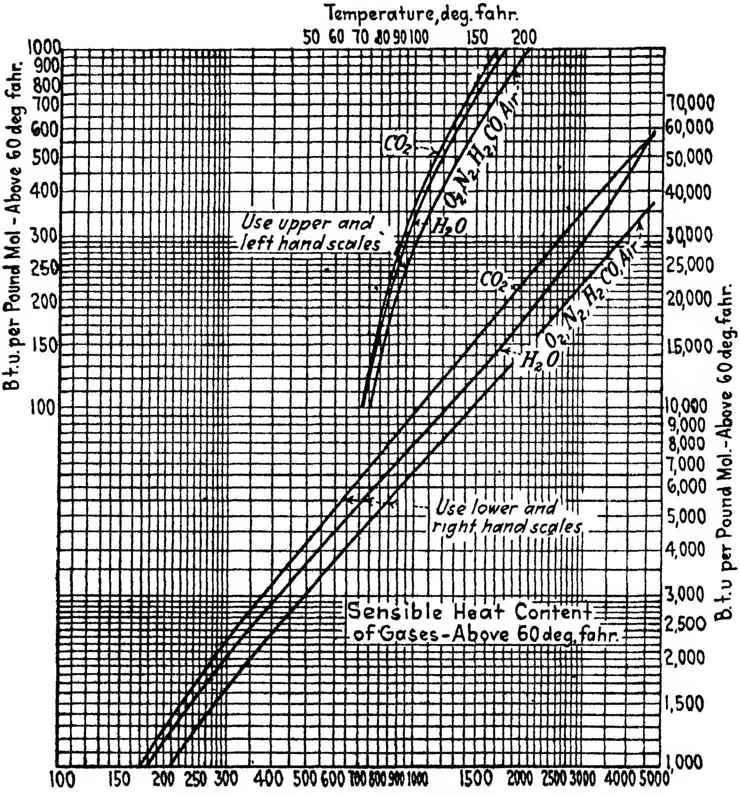

FIG. 49.—Sensible heat content of gases above 60°F.

The accuracy of values above 3800°F is doubtful. (For more accurate values of total heat in H_2 use equation on p. 207.)

(*i.e.*, the mean specific heat multiplied by the temperature rise) for the common gases. For example, again determining the amount of heat required to heat 2.5 mols of air from 60 to 2500°F., the total heat above 60°F. per mol at 2500°F. is read directly from Fig. 49 as approximately 18,200 B.t.u., which multiplied by 2.5 gives 45,500 B.t.u., thus agreeing quite well (within approximately 0.1 per cent) with the previously calculated value of 45,450 B.t.u. Figure 49 is of particular value for determining the amount of heat required to raise a gas from one temperature level to another. Thus to find the amount of heat required to raise 1 mol of CO_2 from 500°F. (260°C.) to 3000°F. (1649°C.), the total heats at the two temperatures are read from Fig. 49 and the difference between them gives the amount of heat required. At 3000°F. the total heat of CO_2 per mol is about 35,200 B.t.u., and at 500°F. is 4500 B.t.u., and 34,200 − 4500 = 29,700 B.t.u. is the amount of heat which must be used to heat 1 mol of CO_2 from 500 to 3000°F.

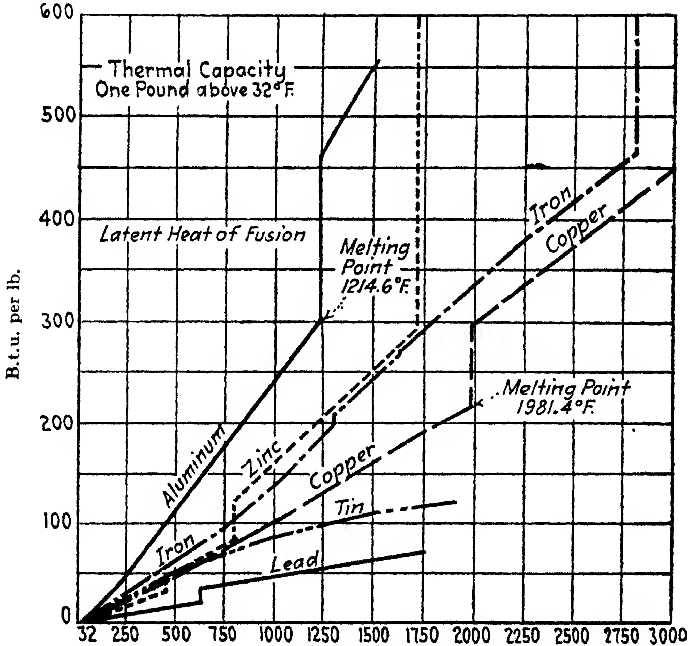

FIG. 50.—Total heat content of metals in B.t.u. per pound above 60°F. (*Adapted from "Combustion," Am. Gas Assoc.* (1924).)

Specific Heat and Heat Capacity of Liquids and Solids.—The molal heat capacity of liquids and solids usually cannot be simplified as readily as for gases. For accurate results, the values must be determined individually for each substance. Table LIV gives the specific heat of several elements over various temperature ranges. A portion of these data are shown

TABLE LIV

SPECIFIC HEAT OF CHEMICAL ELEMENTS*

Element	Symbol	Temperature range†		Specific heat	
		Degrees Fahrenheit	Degrees Centigrade	B.t.u. per lb. per °F.	B.t.u. per mol per °F.
Aluminum	Al	60.8–212	16– 100	0.2122	5.74
Aluminum	Al	60.8–579	16– 304	0.2250	6.10
Aluminum	Al	932	500	0.2739	7.43
Antimony	Sb	59	15	0.0489	5.88
Antimony	Sb	212	100	0.0503	6.05
Antimony	Sb	392	200	0.0520	6.25
Arsenic	As	32– 212	0– 100	0.0822	6.16
Bismuth	Bi	68– 212	20– 100	0.0302	6.31
Carbon (graphite)	C	51.8	11	0.160	1.92
Carbon (graphite)	C	1791	977	0.467	5.60
Carbon (graphite)	C	3146	1730	0.500	6.00
Chromium	Cr	32	0	0.1039	5.40
Chromium	Cr	212	100	0.1121	5.82
Chromium	Cr	1112	600	0.1872	9.73
Cobalt	Co	932	500	0.1452	8.51
Copper	Cu	212	100	0.0942	5.97
Copper	Cu	683	362	0.0997	6.36
Copper	Cu	1652	900	0.1259	8.00
Gold	Au	32– 212	0– 100	0.0316	6.22
Iron:					
Cast	Fe	68– 212	20– 100	0.1189	6.63
Wrought	....	59– 212	15– 100	0.1152	6.43
Wrought	....	932	500	0.176	9.81
		1832–2192	1000–1200	0.1989	11.09
Hard drawn	....	32– 392	0– 200	0.1175	6.85
		32– 932	0– 500	0.1338	7.46
		32–1472	0– 800	0.1597	8.90
		32–1652	0– 900	0.1644	9.18
		38–1832	0–1000	0.1557	8.68
		32–2012	0–1100	0.1534	8.56
Lead	Pb	212	100	0.0311	6.44
		572	300	0.0338	7.00
Magnesium	Mg	68– 212	20– 100	0.2492	6.05
Manganese	Mn	68– 212	20– 100	0.1211	6.65
Mercury	Hg	212	100	0.0328	6.58
Molybdenum	Mo	68– 212	20– 100	0.0647	6.21
Nickel	Ni	212	100	0.1128	6.62
		572	300	0.1403	8.24
		932	500	0.1299	7.61
Platinum	Pt	68–2372	20–1300	0.0359	7.01
Silver	Ag	62.6–945	17– 507	0.0600	6.47
Tin (cast)	Sn	69.8–228	21– 109	0.0551	6.85
Tungsten	W	32– 212	0– 100	0.0336	6.18
		4354	2400	0.045	8.28
Zinc	Zn	68– 212	20– 100	0.0931	6.08
		572	300	0.1040	6.80

---

* Adapted from Smithsonian Physical Tables.

† When one temperature is given, the specific heat at that temperature is indicated, otherwise the mean specific heat.

graphically in Fig. 50, where the heat capacity in B.t.u. per lb. is plotted against the temperature. A plot of this type is more serviceable than the tabulated values, since it includes the heats of transformation and of fusion, in addition to the specific heats. Thus, the heat capacity of copper at 2000°F. (1093°C.) is 300 B.t.u., of which 78 B.t.u. is the heat of fusion at 1982°F. (1083°C.) and the remainder is sensible heat of the solid and liquid copper.

TABLE LV

SPECIFIC HEAT OF SOME COMMON LIQUIDS AND SOLIDS*

Material	Temperature range		Mean specific heat
	Degrees Fahrenheit	Degrees Centigrade	
Alloys:			
Bell metal	59–208	15– 98	0.0858
Brass, red	32	0	0.0899
Brass, yellow	32	0	0.0883
80% Cu; 20% Sn	57– 208	14– 98	0.0862
88.7% Cu; 11.3% Al	68– 212	20– 100	0.1043
German silver	32– 212	0– 100	0.0946
Rose's alloy	68– 192	20– 89	0.0552
Lead-tin	Approx.	Approx.	0.042
Antimony-lead	Approx.	Approx.	0.040
Glass:			
Crown	50– 122	10– 50	0.161
Flint	50– 122	10– 50	0.117
Brickwork	Approx.	Approx.	0.20
Clay, sand, stone	Approx.	Approx.	0.20
Coke (5% Ash)	68– 500	20– 260	0.240
	68–1000	20– 538	0.303
	68–1500	20– 815	0.338
	68–2000	20–1093	0.363
(25% Ash)	68– 500	20– 260	0.234
	68–1000	20– 538	0.289
	68–1500	20– 815	0.320
	68–2000	20–1093	0.342
Petroleum	Approx.	Approx.	0.50
Benzol	Approx.	Approx.	0.40
Wood	Approx.	Approx.	0.60
Water	Approx.	Approx.	1.00

* Adapted mainly from American Gas Association.[3]

Approximate values of the specific heats of some of the more common liquids and solids are given in Table LV, and the heats of fusion and transformation, and the temperatures at which these changes occur, are shown in Table LVI.

TABLE LVI

THERMO-PHYSICAL DATA OF CHEMICAL ELEMENTS*

Element	Melting point		Heat in solid above 32°F., at melting point B.t.u. per lb.	Latent heat of fusion, B.t.u. per lb.	Total heat in liquid at melting point, B.t.u. per lb.	Specific heat of liquid, B.t.u. per lb. per °F.
	Degrees Fahrenheit	Degrees Centigrade				
Aluminum	1214	656	301.5	163.5	365.0	0.308
Antimony	1166	630	61.2	72.5	133.7	0.0605
Bismuth	516	269	16.2	21.6	37.8	0.0363
Chromium	2712	1489	634	128	762	0.24
Copper	1982	1083	214	78	292	0.156
Iron:	2795	1535	461	122	583	0.20
First transformation point.	1346	730				
Latent heat of transformation 9.4 B.t.u. per lb.						
Second transformation point.	1652	900				
Latent heat of transformation 10.8 B.t.u. per lb.						
Lead	620	327	20.9	10.8	31.7	0.0402
Manganese	2205	1207	505	72.5	578	0.357
Nickel:	2642	1450	398	123	521	0.20
Transformation point	446	230				
Latent heat of transformation 8.4 B.t.u. per lb.						
Silver	1763	962	117	44	161	0.0748
Tin	450	232	25.8	24.9	50.7	
Zinc	786	419	81.5	40.6	122.1	0.179

* Data from RICHARDS.[3]

Although it is not possible to determine the specific heat of all solids accurately, *Kopp's law* offers a simple method of approximating the correct values when no more accurate data are available. This law states that the formal heat capacity (total heat in one formular weight of the material) of a solid may be determined by a summation of the atomic heat capacities of the constituents. The values of the atomic heat capacities at room temperature are C = 1.8; H = 2.3; B = 2.7; Si = 3.8; O = 4.0; P = 5.4; F = 5.4, and for all other elements 6.2. The heat capacity of $CaCO_3$ per mol then should be,

$$\overset{Ca}{6.2} + \overset{C}{1.8} + 3 \times \overset{O}{4} = 20.0.$$

The specific heat per lb. should be 20.0 divided by the formula weight, or $20.0 \times \frac{1}{100} = 0.200$. The actual value experimentally determined is 0.202. Errors as high as 10 per cent may be found at times in calculating the specific heat in this manner, but the method may be used where more accurate data are not available.

Balances.—The solution of all combustion problems may be visualized as the determination of a series of balances. Thus, as outlined above, the ratios of air to flue gas, flue gas to fuel, and air to fuel may be calculated by equating the carbon or nitrogen put into the furnace with what goes out. The per cent excess air, the net hydrogen in the fuel, and the moisture in the flue gas are determined from an oxygen balance. Element balances of this type, and total material and heat balances, form the nucleus of most calculations. Where each of the streams entering and leaving a furnace may be measured and analyzed, such balances are ideal for ascertaining the accuracy of measurement. When one or more of the important pieces of data are missing, the balance method is valuable for determining them.

The application of this method to the calculation of results from furnace tests is shown by the following section of examples of combustion calculations.

Examples of Combustion Calculations.—The remainder of this chapter is devoted to an outline and a discussion of the calculations involved in interpreting the data obtained in tests on a boiler furnace, a gas producer, and a lime kiln. The examples given illustrate the application of the principles discussed in the earlier part of the chapter, and the methods employed may be adapted to all types of combustion problems. Since graphical short cuts may be used for the rapid and reasonably accurate solution of many combustion problems, a short section on this subject has been included.

Boiler-furnace Calculations. Example I.—The following data were obtained from a boiler plant:

DATA

Coal used	4,980 lb. per hr.
Water fed to boilers	41,400 lb. per hr.
Feed-water temperature	202°F.
Steam temperature	564°F.
Steam pressure (gage)	245 lb. per sq. in.
Flue-gas temperature	700°F.

Coal analysis (as fired), per cent		Flue-gas analysis, per cent		Cinder analysis, per cent	
C	76.3	CO_2	13.0	Combustible	28.5
H_2	4.8	O_2	5.7	Ash	71.5
O_2	4.0	CO	0.6		
N_2	0.6	N_2	80.7		100
S	1.2				
Ash	9.7		100		
Moisture	3.4				
Total	100				
B.t.u. per lb.	13,400				

From the data given the magnitude of the various items of heat entering and leaving the boiler should be calculated and the total heat input compared with the total heat output. Such a comparison is called a *heat balance*. The items of the heat balance are as follows:

Heat input:
1. Heat in the coal.
2. Heat in the feed water.

Heat output:
3. Heat in the steam.
4. Heat lost in cinders (unburned fuel).
5. Heat loss in stack gases:
   (*a*) As sensible heat in the dry gas.
   (*b*) As the heat carried by water vapor from moisture and from hydrogen in the coal.
   (*c*) As undeveloped heat (unburned CO, H_2, or CH_4).
6. Radiation, with which are usually grouped all unaccounted-for losses.

In this particular case the *basis* of calculation will be taken as 1 hr., and the amount of heat above 60°F. (our datum line) that flows into or out of the furnace will be calculated.

Heat-balance Calculations*

Basis: 1 hr. Datum line, 60°F.

Input:

1. Heat in the coal:

Weight of coal multiplied by B.t.u. per lb.		Lb. coal	B.t.u. per lb.
	4980	x 13,400	= **66,730,000 B.t.u.**

2. Heat in feed water:

Pounds of feed water multiplied by rise in temperature above datum line, and by specific heat.		Lb. water	Temp. °F. above 60	Specific heat
	41,400	x (202 − 60)	x 1	= **5,879,000 B.t.u.**

Total input.................................... **71,609,000 B.t.u.**
2

Output:

3. Heat in steam:

Determine B.t.u. per lb. from steam tables for saturated steam at 245 lb. per sq. in. Add heat due to superheat. Temperature corresponding to 245 lb. per sq. in. is 405°F. Therefore, superheat is 564 − 405 = 159°F.

	Heat in sat. steam	Degrees Superheat	Specific heat
1202	+ 159	x 0.58	= 1294.7 B.t.u. per lb.

---

* The legend above the number in the space indicates the units of the number. This method of indicating the steps in a computation is a modification of that used by Walker, Lewis and McAdams, "Principles of Chemical Engineering," McGraw Hill Book Co. Inc., 1923. The headings used by them indicated the results accomplished by all operations up to the point in question but did not describe the factors employed. Since some of the readers of this book will be unfamiliar with chemical methods of computation it has been deemed wiser to indicate first the result to be accomplished in the calculation and then above each step to describe the character of the factor employed. Much more explanation has been given than is absolutely required, and the calculations have been broken down into simple steps in order to acquaint the reader readily with this method.

Correct for difference between datum line chosen and datum line of steam tables

B.t.u. per lb.	B.t.u. per lb. of feed water above steam-table datum line	
1294.7	− 28.1	= 1266.6 B.t.u. per lb.

Multiply B.t.u. per pound by pounds of steam.

B.t.u. per lb.	Lb. steam	
1266.6	× 41,400	= **52,437,000 B.t.u.**

4. Loss due to unburned fuel in cinders:

Calculate ash in original coal.

Lb. coal	Lb. ash per lb. coal	
4980	× 0.097	= 483 lb.

Calculate pounds of carbon* corresponding to this ash.

Lb. ash	Lb. cinders per lb. ash	Lb. carbon per lb. cinders	
483	× $\frac{100}{71.5}$	× $\frac{28.5}{100}$	= 193 lb.

Multiply pounds of carbon by heating value per lb.

Lbs. carbon	B.t.u. per lb. carbon	
193	× 14,544	= **2,807,000 B.t.u.**

5. Heat loss in stack gases:

(a) Sensible heat in stack gases:

Calculate mols of carbon in stack gases (carbon balance)

Carbon in coal = 4980 × 0.763 = 3800 lb.
Carbon in ash (see item 4) = 193 lb.

Carbon in gas (by difference) = 3607 lb.
Mols carbon in gases = 3607 ÷ 12 = 301 mols

Calculate mols of dry stack gases.

Mols of carbon	Mols gas per mol carbon	
301	× $\frac{100}{13.6}$	= 2210 mols dry gas †

---

*Although probably not strictly true, the combustible in the ash may be considered as carbon.

† Since gas analyses are on a dry basis, this calculation does not include moisture in stack gases.

Calculate mols of each gas.

Gas	Mols of gas		Mols CO_2 per mol gas	
CO_2	2210	×	0.130	= 287 mols

O_2, N_2, CO .... 2210 × 0.870 = 1923 mols

From sensible-heat chart (Fig. 49) read values of heat in CO_2, O_2, N_2, CO, and H_2O at 700°.

Gas	B.t.u. per Mol.
CO_2	6300
O_2, N_2, CO	4450
H_2O	5400

Multiply mols by B.t.u. per mol.

Gas	Mols CO_2		B.t.u. per mol	
CO_2	287	×	6300	= 1,808,000 B.t.u.

O_2, N_2, CO .... 1923 × 4450 = 8,557,000 B.t.u.

Total heat in dry stack gases .... **10,365,000 B.t.u.**

(*b*) Heat carried by water vapor from moisture and from hydrogen in coal.

H_2O from Moisture in Coal

Calculate mols of moisture in stack gases.

Lb. coal		Lb. H_2O per lb. coal	
4980	×	0.034	= 169 lb.

H_2O from Hydrogen in Coal

Lb. coal		Lb. H_2 per lb. coal		Lb. water per lb. H_2	
4980	×	0.048	×	$\frac{18}{2}$	= 2150 lb.

Total = 2319 lb.

Calculate sensible heat in water vapor.

Mols H_2O		B.t.u. per mol		
$\frac{2319}{18}$	×	5400	= 697,000 B.t.u.	= 4

Calculate heat of vaporization of water vapor.*

Lb. H_2O | B.t.u. per lb.

2319 | × 1057 | = 2,455,000 B.t.u.

Total heat (sensible plus latent) in water vapor **3,152,000 B.t.u.**

(*c*) Undeveloped heat:

Determine mols of CO.

Mols gas | Mols CO per mol gas

2210 | × 0.006 | = 13.3 mols

Multiply mols of CO by B.t.u. per mol.

Mols CO | B.t.u. per mol

13.3 | × 122,400 | = **1,628,000 B.t.u.**

Total stack losses = (*a*) + (*b*) + (*c*) = **15,145,000 B.t.u.**

6. Radiation and unaccounted-for losses:

Determine by difference—(Items 1 + 2)−(Items 3+4+5).

Total input.........	72,609,000 B.t.u.	
Accounted-for output	70,389,000 B.t.u.	
Unaccounted for	2,220,000 B.t.u.	**2,220,000 B.t.u.**
Total output....................		**72,609,000 B.t.u.**

The results of the above calculations may now be assembled in the form of a heat balance and the percentage of each item calculated.

HEAT BALANCE

Basis: 1 hr. Datum line, 60°F.

	B.T.U. PER HOUR	PER CENT
Input:		
1. Heat in coal..............................	66,730,000	91.9
2. Heat in feed water..........................	5,879,000	8.1
Total..................................	72,609,000	100.0
Output:		
3. Heat in steam..............................	52,437,000	72.2
4. Heat loss in cinder..........................	2,807,000	3.9

---

* Since the high heating value of the fuel was used in the heat input, heat of vaporization of the water must be included in the output.

5. Heat loss in stack gases		15,145,000	20.9
(*a*) Sensible heat in dry gases	14.3%		
(*b*) Heat in water vapor	4.4%		
(*c*) Undeveloped heat in CO	2.2%		
6. Radiation and unaccounted-for losses		2,220,000	3.0
Total		72,609,000	100.0

It is advantageous many times, especially in comparing boilers operating under different conditions as to feed-water temperature, to rearrange the heat balance so that percentages represent fractions of the heat in the coal alone, and not that in the coal plus feed water. In this case, the heat output in the steam is heat above feed temperature, and is numerically equal to item 3 minus item 2. The heat balance on this basis is as follows:

HEAT BALANCE

Basis: 1 hr. Coal only input. Datum line, 60°F.

		B.T.U. Per Hr.	Per Cent
Input:			
1. Heat in coal		66,730,000	100.0
Output:			
3*a*. Heat in steam minus heat in feed water		46,558,000	69.8
4. Heat lost in cinder		2,807,000	4.2
5. Heat lost in stack gases		15,145,000	22.7
(*a*) Sensible heat in dry gas	15.6%		
(*b*) Heat in water vapor	4.7%		
(*c*) Undeveloped heat in CO	2.4%		
6. Radiation and unaccounted-for losses		2,220,000	3.3
Total		66,730,000	100.0

In addition to the heat balance, the excess air used and the ratio of air to coal should also be calculated. The methods are as follows:

Excess air:

Basis: 100 mols dry stack gas

From N_2, calculate total O_2 in the air used (N_2 balance).

$$\overset{\text{Mols } N_2}{80.7} \times \overset{\text{Mols } O_2 \text{ per mol } N_2}{\frac{21}{79}} = 21.5 \text{ mols}$$

From flue-gas analysis determine excess oxygen (O_2 balance).

Free O_2 in gas = 5.7 mols

O_2 necessary to burn CO. = 0.6 × ½ = 0.3 mols

Excess O_2 = 5.4 mols

O_2 necessary for combustion.

$$\overset{\text{Total}}{21.5} - \overset{\text{Excess}}{5.4} = 16.1 \text{ mols}$$

Excess air equals (excess O_2 divided by necessary O_2) × 100.

$$\frac{5.4 \times 100}{16.1} = \mathbf{33.5 \text{ per cent}}$$

Air–coal ratio:

Basis: 1 lb. coal

From N_2 calculate total mols air used (N_2 balance). 2210 mols dry flue gas per 4980 lb. coal (5*a*).

$$\overset{\text{Mols gas per lb. coal}}{\frac{2210}{4980}} \times \overset{\text{Mols } N_2 \text{ per mol gas}}{0.807} = 0.359 \text{ mol } N_2$$

$$\overset{\text{Mols } N_2}{0.359} \times \overset{\text{Mols air per mol } N_2}{\frac{100}{79}} = 0.454 \text{ mol air}$$

Convert mols to pounds.

$$\overset{\text{Mols air}}{0.454} \times \overset{\text{Lb. air per mol air}}{29} = \mathbf{13.2 \text{ lb. of air per lb. of coal}}$$

and then to cubic feet at 60°F.

$$\overset{\text{Mols air}}{0.454} \times \overset{\text{Cu. ft. per mol}}{359} \times \overset{\text{Temperature correction}}{\frac{460 + 60}{460 + 32}} = \mathbf{172 \text{ cu. ft. air per lb. of coal}}$$

It is sometimes convenient to approximate the efficiency of a boiler by a much simpler process if precise results are not necessary. By determination of the stack and cinder losses, and assuming a proper per cent for radiation and unaccounted-for losses, the efficiency may be calculated. Although this method obtains the efficiency by difference, the major losses are measured directly, and the amount and the distribution of the losses are often of more importance from the point of view of improvement than the efficiency itself. The following example illustrates the method:

Example II.

DATA

Flue-gas analysis, per cent		Cinder analysis, dry basis, per cent		Coal analysis, dry basis, per cent	
CO_2	11.0	Carbon	20.0	Carbon	75.0
O_2	8.0	Ash	80.0	Ash	10.0
CO	0.0			Moisture	3.0
N_2	81.0		100.0	B.t.u. per lb.	14,000
	100.0				

Stack-gas temperature, 600°F.

CALCULATIONS

Basis: 100 lb. of dry coal. Datum line, 60°F.

Input:

1. Total heat in coal:

Multiply pounds of coal by B.t.u. per lb.

Lb. coal		B.t.u. per lb. coal	
100	x	14,000	= 1,400,000 B.t.u.

2. Heat in feed water:

This item is not included as heat input, in this method, and only the net heat in the steam (above feed-water temperature) is considered as an output in item 3. Thus, percentages represent direct fractions of the heat in the coal as in the second heat balance above.

Output:

3. Heat in steam above feed-water temperature:

—obtained by difference after item 6.

4. Loss due to unburned fuel in cinder:

From the lb. of ash per 100 lb. coal calculate the lb. of carbon in the ash and then the heat in this amount of carbon.

Lb. coal		Lb. ash per lb. coal	
100	x	0.10	= 10 lb. ash

Lb. ash		Lb. cinder per lb. ash		Lb. carbon per lb. cinder	
10	x	$\frac{100}{80}$	x	$\frac{20}{100}$	= 2.5 lb. carbon in ash

Lb. carbon	B.t.u. per lb. carbon	
2.5	× 14,000	= 35,000 B.t.u.

5. Heat loss in stack gases:

(*a*) Sensible heat in dry stack gases:

Calculate mols carbon in stack gases (carbon balance) and then mols dry stack gases.

	Lb. coal	Lb. carbon per lb. coal	
Carbon in coal =	100	× 0.75	= 75.0 lb.

Carbon in ash = (see item 4) 2.5 lb.

Carbon in gas (by difference) 72.5 lb.

	Lb. carbon	Mols per lb.	
Mols carbon in gas =	72.5	× $\frac{1}{12}$	= 6.04 mols

	Mols carbon	Mols gas per mol carbon	
Mols dry stack gas =	6.04	× $\frac{100}{11}$	= 54.9 mols

	Mols gas	Mols CO_2 per mol gas	
Number of mols of CO_2 =	55	× 0.11	= 6.04 mols

Number of mols of O_2, N_2 = 55 × 0.89 = 48.9 mols

From Fig. 49 read total heat content of each gas at 600°F.

GAS	B.T.U. PER MOL AT 600°F.
CO_2	5200
O_2, N_2	3780
H_2O	4550

Multiply mols by B.t.u per mol.

	No. mols	Total heat per mol	
Heat in CO_2 =	6.04	× 5,200	= 31,400 B.t.u.

	No. mols	Total heat per mol	
Heat in O_2, N_2 =	48.9	× 3,780	= 185,000 B.t.u.

Total heat in dry stack gases................ 216,400 B.t.u.

(*b*) Heat in water vapor:

(1) H_2O from moisture in coal:

Calculate mols of moisture in stack gases.

Lb. coal | Lb. water per lb. coal

100 × 0.03 = 3 lb. water

or 3 ÷ 18 = 0.17 mol H_2O

(2) Water from burning net hydrogen in fuel:*

Basis: 100 mols dry flue gas

From an O_2 balance calculate moisture equivalent to the net H_2 of the coal.

Mols N_2 | Mols O_2 per mol N_2

81 × $\frac{21}{79}$ = 21.5 mols O_2 used

Mols CO_2 | Mols O_2

11 + 8.0 = 19.0 mols O_2 in gas

By difference (O_2 disappearance) } 2.5 mols O_2 used to burn net H_2

Mols O_2 | Mols H_2 per mol O_2

2.5 × 2 = 5.0 mols H_2 per 100 mols gas
or 0.05 mol H_2 per mol gas

Convert the net H_2 to coal basis.

Mols H_2 per mol gas | Mols gas per 100 lb. coal (see 5a)

0.05 × 55 = 2.75 mols H_2 per 100 lb. coal

Total H_2O in gas equals

H_2O from moisture....................	0.17 mol
H_2O from net H_2....................	2.75 mols
Total H_2O in gas..........................	2.92 mols

---

* This calculation neglects H_2O formed from O_2 in coal. See p. 229 for method of calculating this item.

Calculate sensible heat in water vapor.

Mols H_2O		B.t.u. per mol	
2.92	×	4550	= 13,300 B.t.u.

Calculate heat of vaporization of water vapor.

Mols H_2O		Lb. per mol		B.t.u. per lb.	
2.92	×	18	×	1057	= 55,500 B.t.u

Total heat in water vapor . . . . **68,800 B.t.u.**

(*c*) Undeveloped heat:

No combustible gases. . . . . 0 B.t.u.

Total stack losses [(*a*) + (*b*) + (*c*)] . . . . **285,200 B.t.u.**

6. Radiation and unaccounted-for losses (assumed equal to 5 per cent):

Lb. coal		B.t.u. per lb.		Per cent loss	
100	×	14,000	×	0.05	= **70,000 B.t.u.**

3. Heat in steam:

Determined by difference

Item 1 − Items (4 + 5 + 6).

Item 1 = 1,400,000 B.t.u.

Items (4 + 5 + 6) = 390,200 B.t.u.

1,009,800 B.t.u. **1,009,800 B.t.u.**

ASSEMBLY OF ABOVE DATA INTO HEAT BALANCE

Basis. 100 lb. coal. Datum line, 60°F.

		B.T.U.	PER CENT
Input:			
1. Heat in coal		1,400,000	100.0
Output:			
3. Heat in steam above feed-water temperature		1,009,800	72.1
4. Heat loss in cinder		35,000	2.5
5. Heat loss in stack gas		285,200	20.4
(*a*) Sensible heat in dry gas	15.5%		
(*b*) Heat in water vapor	4.9%		
(*c*) Undeveloped heat in CO	0.0%		
6. Radiation and unaccounted-for losses (assumed)		70,000	5.0
Total		1,400,000	100.0

In both of the above examples of boiler-furnace calculations the ultimate analysis of the coal was used. For all accurate calculations it is necessary, as in these two cases, to have the percentage of total carbon in the coal. In many short plant tests, however, only the proximate analysis is available, and for calculating the results of such tests it is necessary to calculate the per cent of carbon in the coal by means of the formulas given on page 257. The remainder of the calculations are the same as those outlined in the two examples given. The calculations made on the basis of the empirically determined carbon content, although sometimes subject to considerable error, may be very valuable for quick and reasonably accurate determinations of plant efficiency.

*Producer-gas Calculations.**—The calculations from producer-gas data are of distinct importance, including several items not found in boiler calculations. The following example shows the method employed in connection with a producer-gas fired furnace.

Example III.

DATA

Coal analysis, dry basis, per cent		Producer-gas analysis, per cent		Cinder analysis, dry basis, per cent		Analysis of furnace-flue gas, per cent	
C	76.0	CO_2	7.5	C	15	CO_2	10.7
O_2	5.8	O_2	0.2	Ash	85	O_2	8.3
H_2	5.4	CO	20.5		—	CO	0.0
Ash	9.8	H_2	12.5		100	N_2	81.0
N_2	1.5	C_2H_4	0.5				——
S	1.5	CH_4	3.0				100.0
B.t.u. per lb.	14,400	N_2	55.8	Producer-gas temperature = 1350°F.			
Moisture, 4.0 per cent.		B.t.u. per cu. ft. gross, 60°F., 30 in. Hg and saturated = 142.7.					

From the above data, the following results should be determined:

1. Cubic feet of gas (measured under standard conditions) per lb. of coal.
2. Cubic feet of air per lb. of coal.
3. Steam decomposed per lb. of coal.
4. Total steam used per lb. of coal.

---

* Certain steps in this calculation follow directly from the discussion of producer gas in Chap. XVIII. It is suggested that Chap. XVIII be read before these calculations are taken up.

5. Air required for combustion of gas.
6. Excess air used in combustion of gas.
7. *Heat balance* on the producer, showing:
   (*a*) Input:
      (1) Heat in coal.
      (2) Heat in steam.
   (*b*) Heat Output:
      (3) Undeveloped heat in gas.
      (4) Sensible and latent heat in gas.
      (5) Loss in cinder.
      (6) Radiation and unaccounted-for losses

CALCULATIONS

Basis: 1 lb. dry coal. Datum line, 60°F.

1. Cubic feet of gas per lb. of coal:

(*a*) Dry gas (by carbon balance):

	Lb. coal		Lb. carbon per lb. coal	
Carbon in coal.	1	x	0.760	= 0.760 lb.

	Lb. coal		Lb. ash per lb. coal		Lb. cinder per lb. ash		Lb. carbon per lb. cinder	
Carbon in cinder.	1	x	0.098	x	$\frac{100}{85}$	x	$\frac{15}{100}$	= 0.0173 lb.

Carbon in gas (by difference) 0.743 lb.

	Lb. carbon		Mols carbon per lb. carbon	
Total mols carbon in gas	0.743	x	$\frac{1}{12}$	= 0.062 total mols of carbon in producer gas

Mols carbon per mol of producer gas (from analysis).

From CO_2		From CO		From CH_4		From C_2H_4	
0.075	+	0.205	+	0.03	+	0.01	= 0.32 mol carbon per mol of producer gas

From total mols of carbon in gas and mols carbon per mol gas get mols gas per lb. of coal.

Mols carbon per lb. coal | Mols gas per mol carbon

$0.062 \times \frac{1}{0.32} =$ 0.194 mol gas per lb. of coal

Convert mols to cubic feet.

Mols gas | Cu. ft. per mol | Temperature correction

$0.194 \times 359 \times \frac{520}{492} =$ **73.6 cu. ft. of dry gas at 60°F. and 29.92 in. Hg**

Change from dry gas to gas saturated at 60°F. (vapor pressure H_2O at 60°F. = 0.52 in. Hg).

Cu. ft. dry gas | Correction for water vapor

$73.60 \times \frac{29.92}{29.92 - 0.52} =$ **75.0 cu. ft. gas saturated with water at 60°F. and at 29.92 in Hg**

Convert to a pressure of 30 in. Hg.

Cu. ft. gas | Pressure correction

$75.0 \times \frac{29.92}{30.00} =$ **74.8 cu. ft. gas at 60°F., 30 in. Hg, and saturated with water vapor**

(*b*) In addition to the dry gas, there is a certain amount of water vapor from the moisture and oxygen in the coal, and from undecomposed steam. The water vapor from the coal may be calculated as follows:

From moisture in coal.

Lb. coal | Lb. H_2O per lb. coal

$1 \times 0.04 =$ 0.04 lb.

From oxygen in coal.

Lb. coal | Lb. oxygen per lb. coal | Lb. H_2O per lb. oxygen

$1 \times 0.058 \times \frac{18}{16} =$ 0.065 lb.

Total.................................... 0.105 lb.

Equivalent mols H_2O

$$\overset{\text{Lb. }H_2O}{0.105} \times \overset{\text{Mols per lb.}}{\frac{1}{18}} = 0.00584 \text{ mol } H_2O \text{ per lb. of coal}$$

Convert to mols H_2O per mol of dry gas

$$\overset{\text{Mols }H_2O\text{ per lb. coal}}{0.00584} \times \overset{\text{Lb. coal per mol dry gas}}{\frac{1}{0.194}} = 0.03 \text{ mol } H_2O \text{ per mol dry gas}$$

(c) The undecomposed steam, has not been included, and unless the steam is metered, it cannot be determined directly. The *total* water vapor in the gas may be calculated approximately from the gas analysis, using the equation (Chap. XVIII) as follows:

$$\text{Total mols } H_2O \text{ per 100 mols of dry producer gas} = \frac{(CO_2)(H_2)}{(CO)(0.096L)}$$

where L is the depth of fuel bed in feet.

In this case, assuming a 5-ft. fuel bed, it would be:

$$\text{Total mols } H_2O = \frac{(7.5)(12.5)}{(20.5)(0.48)} = 9.5 \text{ mols } H_2O \text{ per 100 mols of gas.}$$

The mols of water per lb. of coal equals

$$\overset{\text{Mols }H_2O\text{ per mol gas}}{\frac{9.5}{100}} \times \overset{\text{Mols gas per lb. of coal (See 1a)}}{0.194} = 0.0184 \text{ mol } H_2O \text{ per lb. coal.}$$

The total mols of producer gas per lb. of coal is the sum of the mols of dry gas (Item 1(a)), 0.194 and the mols of water vapor, 0.0184, or a total of 0.212 mols. If the temperature and pressure of the producer gas are known this figure may, of course, be converted to the actual volume of gas leaving the producer.

The undecomposed steam is then $\overset{\text{Total mols }H_2O}{\frac{9.5}{100}} - \overset{\text{Mols }H_2O\text{ from coal}}{0.03} = 0.065$ mol per mol of gas

$$\text{or, } \overset{\text{Mols }H_2O\text{ per mol gas}}{0.065} \times \overset{\text{Lb. per mol }H_2O}{18} \times \overset{\text{Mols gas per lb. coal}}{0.194} = 0.227 \text{ lb. } H_2O \text{ per lb. coal}$$

2. Cubic feet of air per lb. of coal:

From mols of gas per lb. of coal (see 1 (*a*)) and per cent N_2 in gas calculate mols of air per lb. of coal (N_2 balance).

Mols gas per lb. coal burned	Mols N_2 per mol gas	Mols air per mol N_2	
0.194	x 0.558	x $\frac{100}{79}$	= 0.137 mol of air

Convert mols to cubic feet.

Mols air	Cu. ft. per mol	Temperature correction	
0.137	x 359	x $\frac{520}{492}$	= 52.0 cu. ft. dry air at 60°F.

3. Steam decomposed per lb. of coal:

Method (*a*). By H_2 balance. From the total H_2 in the producer gas subtract the *net* H_2 in the coal.

Determine net H_2 per lb. of coal

Total H_2 in coal. .... 0.054 lb.

H_2 corresponding to O_2 in coal = $\frac{0.058}{8}$ ... 0.007 lb.

Net H_2 in coal....... 0.047 lb.

Convert pounds H_2 to mols.

Lb. H_2	Mols H_2 per lb. H_2	
0.047	x $\frac{1}{2}$	= 0.0235 mol net H_2 in coal

Determine total H_2 in producer gas from analysis.

From H_2	From CH_4	From C_2H_4	
0.125	+ 2 x 0.030	+ 2 x 0.005	= 0.195 mol total H_2 per mol gas

Mols H_2 per mol gas	Mols gas per lb. coal	
0.195	x 0.194	= 0.0378 mol H_2 in gas per lb. of coal

Calculate H_2 from steam (by difference).

Total H_2	Net H_2 from coal	
0.0378	− 0.0235	= 0.0143 mol H_2 from steam

Calculate pounds of steam.

Mols H_2	Mols H_2O per mol H_2	Lb. H_2O per mol	
0.0143	× $\frac{1}{1}$	× 18	= 0.258 lb. of steam per lb. of coal

Method (*b*). By O_2 balance. From the total O_2 in the producer gas subtract the O_2 from the air to get the oxygen from the steam.

By O_2 balance calculate O_2 from air.

Mols air per lb. coal [from item 2]	Mols O_2 per mol air	
0.137	× $\frac{21}{100}$	= 0.0288 mol O_2 from air per lb. of coal

Determine total O_2 in producer gas from analysis.

From CO_2	From O_2	From CO	
0.075	+ 0.002	+ $\frac{0.205}{2}$	= 0.180 mol total O_2 per mol producer gas

Mols O_2 per mol gas	Mols gas per lb. coal	
0.180	× 0.194	= 0.0349 mol total O_2 in gas per lb. coal

Calculate O_2 from steam (by difference).

Total O_2	O_2 from air	
0.0349	− 0.0288	= 0.0061 mol O_2 from steam

Calculate pounds of steam.

Mols O_2	Mols H_2O per mol O_2	Lb. per mol H_2O	
0.0061	× $\frac{2}{1}$	× 18	= 0.220 lb. of steam per lb. of coal

Average* amount decomposed $= \frac{0.258 + 0.220}{2} =$ 0.239 lb. of steam per lb. of coal.

---

* The difference in results obtained by the two methods of calculation is due to slight inaccuracies in the gas or coal analyses. A small error in either develops a relatively large error, since the results are obtained by differences. The average value gives the best approximation.

4. Total steam used per lb. of coal:

Sum of decomposed and undecomposed steam (items 1(c) + 3).

$$\overset{\text{Undecomposed steam}}{0.227} + \overset{\text{Decomposed steam}}{0.239} = 0.466 \text{ lb. total steam per lb. coal}$$

The per cent decomposed is $\dfrac{0.239 \times 100}{0.227 + 0.239} = 51.3$ per cent

5. Air required for combustion of producer gas:

Basis: 100 mols dry producer gas

Determine O_2 required to burn each constituent.

Gas	Mols	Mols O_2 Required for Combustion
CO_2	7.5	0.0
CO	20.5	10.25
H_2	12.5	6.25
Ill. (C_2H_4)	0.5	1.5
CH_4	3.0	6.0
Total		24.0

Subtract O_2 present and multiply by air-oxygen ratio.

$$\overset{\text{Net mols } O_2 \text{ required}}{(24.0 - 0.2)} \times \overset{\text{Mols air per mol } O_2}{\frac{100}{21}} =$$

113 mols air or 1.13 mols air per mol gas, or by direct ratio 1.13 cu. ft. air per cu. ft. of gas (measured under same conditions)

6. Excess air used in combustion of gas:

Basis: 100 mols dry producer gas

Method (*a*). Using value of air required as above:

Calculate mols of flue gas (carbon balance).

Mols carbon in producer gas ($CO_2 + CO + 2C_2H_4 + CH_4$) = 32.0

$$\overset{\text{Mols carbon}}{32} \times \overset{\text{Mols flue gas per mol carbon}}{\frac{100}{10.7}} = 299 \text{ mols dry flue gas}$$

Calculate air used (N_2 balance).

$$\text{Mols } N_2 \text{ in flue gas} = \overset{\text{Mols flue gas}}{299} \times \overset{\text{Mols } N_2 \text{ per mol flue gas}}{0.81} = 242.0 \text{ mols}$$

Mols N_2 from producer gas.................. = 55.8 mols

Mols N_2 from air (by difference)............ = 186.2 mols

$\overset{\text{Mols N}_2\text{ from air}}{186.2} \times \overset{\text{Mols air per mol N}_2}{\frac{100}{79}}$ = 236 mols of air used in burning the producer gas

Calculate per cent excess air.

Air used = 236 mols
Air required (see 5) = 113 mols

Excess air (difference) = 123 mols

or $\frac{123 \times 100}{113}$ = **109 per cent excess air**

Method (*b*). Not using value of air required:

Calculate O_2 from air in flue gas

Mols N_2 in flue gas from air (as in method (*a*)) = 186.2 mols

O_2 from air $= \overset{\text{Mols N}_2}{186.2} \times \overset{\text{Mols O}_2\text{ per mol N}_2}{\frac{21}{79}}$ = 49.5 mols O_2

Determine excess O_2 from flue-gas analysis.

O_2 in flue gas = 8.3 mols per 100 mols flue gas

$\overset{\text{Mols O}_2\text{ per 100 mols flue gas}}{8.3} \times \overset{\text{Mols flue gas per mol producer gas}}{\frac{299}{100}}$ = 24.8 mols excess O_2 in flue gas per 100 mols producer gas

Determine O_2 consumed (difference)

$\overset{\text{Total O}_2\text{ from air}}{49.5} - \overset{\text{Excess O}_2\text{ in flue gas}}{24.8}$ = 24.7 mols O_2 utilized for combustion

Calculate per cent excess air

$\frac{\text{Excess } O_2 \times 100}{\text{Required } O_2} = \frac{24.8 \times 100}{24.7}$ = **101 per cent excess air**

Average per cent excess air $\frac{109 + 101}{2}$ = **105 per cent**

The slight discrepancy between the results of the two methods is due to the usual inaccuracies in the gas analysis. For example the apparent ratio of carbon to hydrogen in the flue gases is not quite accurate, as is shown by the following calculation.

Basis: 100 mols producer gas

Calculate carbon-hydrogen ratio in producer gas itself

Total C in producer gas = 32.0 mols.
Total H_2 in producer gas = 19.5 mols

$$\therefore \frac{C}{H_2} = \frac{32.0}{19.5} = 1.64.$$

Calculate the apparent carbon-hydrogen ratio in the flue gases and compare the two ratios.

Calculate total mols O_2 in flue gas

Mols O_2 in flue gas from air per 100 mols producer gas (see item 6 (*b*))	= 49.5 mols
Mols O_2 in flue gas from producer gas ($CO_2 + O_2 + \frac{1}{2}CO$)	= 18.0 mols
Total mols O_2 in flue gas per 100 of producer gas.....	67.5 mols

Apparent O_2 in flue gas per 100 mols of producer gas by carbon ratio

Total mols O_2 per 100 mols flue gas; Total mols Carbon in producer gas; Total mols carbon in flue gas

$$19 \times 32 \times \frac{1}{10.7} = 57.0 \text{ mols}$$

Calculate O_2 used to form H_2O.

Total O_2; Apparent O_2

$$67.5 - 57.0 = 10.5 \text{ mols } O_2 \text{ (disappeared) used to form } H_2O$$

Calculate H_2 corresponding to this O_2.

Mols O_2; Mols H_2O per mol O_2

$$10.5 \times 2 = 21.0 \text{ mols } H_2O \text{ or mols } H_2$$

Calculate carbon-hydrogen ratio in flue gas.

$$\frac{C}{H_2} = \frac{32.0}{21.0} = 1.53$$ calculated from the flue gas as compared with 1.64 determined directly

7. Calculation of heat balance:

HEAT BALANCE

Basis: 1 lb. of dry coal. Datum line, 60°F

Input:

1. Heat in coal:

Lb. coal; B.t.u. per lb. coal

$$1 \times 14{,}400 = \mathbf{14{,}400 \text{ B.t.u.}}$$

2. Heat in steam:

Total steam used (from item 4) = 0.466 lb. steam

Steam pressure = 60 lb. per sq. in.

B.t.u. per lb. steam at 60 lb. per sq. in. = 1181 B.t.u. (from steam tables)

B.t.u. per lb. water at 60°F. (datum line) = 28 B.t.u. (from steam tables)

Heat in steam equals weight of steam per lb. of coal times B.t.u. per lb.

Lb. steam per lb. coal	Heat in steam per lb.	
0.466	× (1181 − 28)	= 536 B.t.u.

Total heat input.......... .... **14,936 B.t.u.**

Output:

3. Undeveloped heat in gas:

74.8 cu. ft. of producer gas at 60°F., 30 in. Hg (saturated) from item 1.

Cu. ft.	B.t.u. per cu. ft.	
74.8	× 142.7	= 10,670 B.t.u.

4. Sensible and latent heat in gas:

Total mols dry gas (see p. 229) = 0.194 mol

Total mols H_2O (see p. 230) = 0.0184 mols

Gas	Mols of Gas	B.t.u. per Mol at 1350°F.	Sensible Heat (Mols Gas × B.t.u. per Mol), B.t.u.
CO_2..........	0.194 × 0.075	13,700	200
O_2, N_2, H_2, CO	0.194 × 0.890	9,200	1,590
$CH_4 + C_2H_4$*.	0.194 × 0.035	13,700	93
H_2O.........	0.0184	11,000	203

Total sensible heat in gas = 2,086 B.t.u.

Heat of vaporization of water (latent heat)

Mols H_2O per lb. coal	Lb. per mol	B.t.u. per lb.	
0.0184	× 18	× 1057	= 350 B.t.u.

Total sensible and latent heat = **2,436 B.t.u.**

---

* The molal heat capacity of CH_4 and C_2H_4 is not given on the chart (Fig. 49) and is assumed to be equal to that of CO_2. Since the percentage of these constituents is small, the error is negligible.

5. Heat loss in cinders:

Pounds carbon in cinder (p. 228) = 0.017

Lb. carbon		B.t.u. per lb. carbon		
0.017	×	14,500	=	**246 B.t.u.**

6. Radiation and unaccounted-for losses:

Input	14,936 B.t.u.	
Accounted for output (Items 3 + 4 + 5)	13,352 B.t.u.	
Radiation and unaccounted-for losses	1,584 B.t.u.	**1,584 B.t.u.**

The results of the above producer-gas calculations may be summarized as follows:

1. 74.8 cu. ft. of gas at 60°F., 30 in. Hg (saturated) produced per lb. of coal.
2. 52.0 cu. ft. of dry air at 60°F. used per lb. of coal.
3. 0.239 lb. of steam decomposed per lb. of coal.
4. 0.466 lb. of steam supplied per lb. of coal.
5. 1.13 cu. ft. of air required for combustion per cu. ft. of producer gas.
6. 105 per cent excess air used in combustion of gas.
7. The heat balance is:

HEAT BALANCE

Basis: 1 lb. of coal. Datum line, 60°F.

Input:	B.T.U.	PER CENT
1. Heat in coal	14,400	96.4
2. Heat in steam	536	3.6
Total	14,936	100.0
Output:		
3. Undeveloped heat in gas (cold efficiency)	10,670	71.5
4. Sensible heat in gas	2,436	16.3
5. Heat loss in cinders	246	1.6
6. Radiation and unaccounted-for losses	1,584	10.6
	14,936	100.0

The cold efficiency of the producer is item 3 or 71.5 per cent. The hot efficiency, 87.8 per cent, is the sum of items 3 and 4.

Lime-kiln Calculations.—In a combustion process where the material being heated is in direct contact with the products of combustion, a slightly different method of attack may be necessary. The ratio of flue gas to fuel cannot be directly determined from a carbon balance when some carbon comes from sources other than the fuel. The method to be employed in such a case is illustrated by the calculations of the results from a test on a lime-burning plant, given below.

Example IV.

DATA ON LIME-BURNING PLANT TEST

Analysis of coal, dry basis, per cent		Analysis of cinder, dry basis, per cent		Analysis of producer gas, per cent		Analysis of flue gas, per cent	
C	80.35	Moisture	13 3	CO_2	7.9	CO_2	22.4
H_2	7.59	Combustible	11.5	O_2	1.0	O_2	5.6
S	1.5	Ash	88.5	CO	18.4	CO	0.3
N_2	1.5			H_2	12.1	N_2	71.7
O_2	3.56			CH_4	3 7		
Ash	5.50			N_2	56.9	Total	100.0
Total	100 00			Total	100.0		

Moisture in coal, 5.4 per cent; B.t.u. per lb., 14,315.
Per cent carbonate in limestone (calculated as $CaCO_3$), 97.5 per cent.
Per cent calcium oxide in lime, 97.0 per cent.

The following calculations show the variety of results which may be obtained from a series of analyses with *no direct weight measurements.*

CALCULATIONS

1. Producer gas-coal ratio:

Basis: 1 lb. of dry coal

By a carbon balance get the total quantity of carbon going into the producer gas. Then from the total carbon per 100 mols of producer gas, as shown by the gas analysis, obtain the mols of producer gas and convert to cubic feet.

Carbon balance.

$$\text{Carbon in 1 lb. of coal} \qquad 1 \times \frac{80.35}{100} \qquad = 0.8035 \text{ lb.}$$

$$\text{Carbon in cinder}\ \overset{\text{Lb. ash per lb. coal}}{\frac{5.5}{100}} \times \overset{\text{Lb. cinder per lb. coal}}{\frac{100}{88.5}} \times \overset{\text{Lb. carbon per lb. cinders}}{\frac{11.5}{100}} = \underline{0.0071 \text{ lb.}}$$

$$\therefore \text{Carbon in gas} \ldots\ldots\ldots\ldots \quad 0.7964 \text{ lb.}$$

Converting to mols, $\frac{0.7964}{12} = 0.0664$ mol.

From the gas analysis determine mols carbon per 100 mols of producer gas

	Mols
From CO_2	7.9
From CO	18.4
From CH_4	3.7
	30.0 mols carbon per 100 mols of producer gas

∴ Mols producer gas per lb. of coal $\overset{\text{Mols carbon per lb. coal}}{0.0664} \times \overset{\text{Mols producer gas per mol carbon}}{\frac{100}{30}} = 0.221$ mol.

Converting to cubic feet gas per lb. coal as fired. $\overset{\text{Mols gas per lb. coal}}{0.221} \times \overset{\text{Cu. ft. per mol}}{359} \times \overset{\text{Temperature correction}}{\frac{520}{492}} \times \overset{\text{Moisture correction}}{\frac{29.92}{29.40}} \times \overset{\text{Pressure correction}}{\frac{29.92}{30.00}}$ = **85.1 cu. ft. of gas at 60°F., 30 in., saturated per lb. dry coal**

Allowing for moisture in coal. $85.1 \times \frac{1}{1.054}$ = **80.6 cu. ft. of producer gas per lb. coal as fired**

2. Flue gas-producer gas ratio:

Basis: 100 mols dry producer gas

(a) From the producer-gas analysis get the nitrogen equivalent of the oxygen necessary for combustion of 100 mols of *producer gas* and add to it the N_2 from 100 mols of producer gas. (b) From the flue-gas analysis get the nitrogen equivalent of the oxygen used in excess of that necessary for combustion. (c) Subtracting this quantity from the nitrogen in the flue gas gives, per 100 mols of *flue gas*, the nitrogen equivalent to the oxygen necessary for combustion plus the nitrogen from the producer gas. (d) Therefore, the ratio of (a) to (c) is the ratio of flue gas to producer gas.

(a) N_2 equivalent of O_2 necessary for combustion per 100 mols *producer gas.*

Constituent	Mols per 100 of Producer Gas	Oxygen Necessary for Combustion
CO	18.4	9.2
H_2	12.1	6.05
CH_4	3.7	7.4
O_2	1.0	-1.0
Total O_2 required		**21.65 mols**

Mols O_2 | Mols N_2 per mol O_2

N_2 equivalent of O_2, 21.65 | $\times \frac{79}{21}$ | $= 81.5$ mols N_2

N_2 in 100 mols producer gas .... 56.9 mols N_2

Total N_2 in flue gas per 100 mols of producer gas.. . 138.4 mols N_2

(*b*) N_2 equivalent to the excess oxygen in the *flue gas* per 100 mols flue gas.

Mols free O_2 | O_2 req. for CO

5.6 | $- \frac{0.30}{2}$ | $= 5.45$ mols O_2

Mols O_2 | Mols N_2 per mol O_2

5.45 | $\times \frac{79}{21}$ | $= 20.5$ mols N_2

(*c*) N_2 equivalent of O_2 necessary for combustion of producer-gas plus N_2 from producer gas equivalent to 100 mols of flue gas.

Mols N_2 in flue gas | Mols N_2 corresponding to excess O_2

71.7 | $- 20.5$ | $= 51.2$ mol. N_2 equivalent to theoretical O_2 per 100 mols *flue gas*.

(*d*) Ratio total N_2 in flue gas per 100 mols of producer gas (theoretical) (*a*) to mols N_2 equivalent to theoretical O_2 per 100 mols of flue gas (*b*) = ratio mols flue gas to mols of producer gas.

$$\frac{a}{c} = \frac{138.4}{51.2} = 2.7$$ **mols of flue gas per mol of producer gas.**

or 2.7 cu. ft. of flue gas per cu. ft. of producer gas.

or 270 mols of flue gas per 100 mols producer gas.

3. Lime-coal ratio (fuel-ratio):

Carbon balance.

Basis: 100 mols dry flue gas.

Determine carbon from limestone by difference between total carbon in flue gas and carbon from producer gas.

Total carbon in flue gas (CO_2+CO)...... 22.7 mols

Mols producer gas per mol flue gas | Mols carbon per mol producer gas

Carbon from producer gas $= \frac{100}{270} \times \frac{30.0}{100} \times 100 = 11.1$ **mols**

Since 1 mol of carbon comes from 1 mol of limestone, then

Mols $CaCO_3$ decomposed equals 11.6 mols per 100 mols flue gas

Pounds of stone decomposed

Mols $CaCO_3$	Lb. $CaCO_3$ per mol	Lb. stone per lb. $CaCO_3$	
11.6	$\times$ 100	$\times \frac{1}{0.975}$	= 1190 lb. limestone per 100 mols flue gas

Pounds of lime formed (1 mol CaO = 1 mol $CaCO_3$)

Mols CaO	Lb. CaO per mol	Lb. lime per lb. CaO	
11.6	$\times$ 56.1	$\times \frac{1}{0.97}$	= 668 lb. lime per 100 mols flue gas

Coal used equals

Calculate pounds of coal per 100 mols flue gas from ratios of producer gas to flue gas and coal to producer gas.

Mols producer gas per mol flue gas	Lb. coal per mol producer gas	
$\frac{100}{270}$	$\times \frac{1}{0.221}$	$\times$ 100 = 167.5 lb. dry coal per 100 mols flue gas

Pounds wet coal per lb. dry coal equals

Lbs. dry coal + lbs. moisture = lbs. wet coal.

Dry coal	Moisture	
1.00	+ 0.054	= 1.054

Pounds coal as fired per 100 mols flue gas = 167.5 × 1.054 = 176.6 lb. coal per 100 mols dry flue gas.

Pounds of lime per lb. of coal fired

Divide pounds of lime per 100 mols flue gas by pounds of coal per 100 mols flue gas to get the fuel ratio.

Lb. of lime per mol flue gas	Mols flue gas per lb. of coal	
$\frac{668}{100}$	$\times \frac{100}{176.6}$	= 3.78 lb. of lime per lb. coal as fired

Divide pounds of stone decomposed per 100 mols of flue gas by the pounds of coal per 100 mols of flue gas.

Lb. stone per mol flue gas	Mols flue gas per lb. coal	
$\frac{1190}{100}$	$\times \frac{100}{176.6}$	= 6.74 lb. of stone per lb. coal as fired

Obtain stone ratio by dividing pounds of lime by pounds of stone.

Lb. lime per lb. coal

Lb. of coal per lb. stone

$3.78 \times \frac{1}{6.74}$ = **0.561 lb. of lime per lb. of stone**

In addition to the material-balance results as calculated above, the excess air used and the cubic feet of flue gas per lb. of coal may be determined largely from the previously calculated data.

4. Excess air:

Basis: 100 mols dry flue gas

O_2 required per 100 mols producer gas = 21.65 mols.

Mols O_2 per 100 mols producer gas

Mols producer gas per mol flue gas

$21.65 \times \frac{100}{270}$ = 8.01 mols O_2 per 100 mols flue gas

Excess O_2 in flue gas = $5.60 - \frac{1}{2}(0.3) = 5.45$ mols

$$\text{Excess air (per cent)} = \frac{\text{Excess } O_2 \times 100}{\text{Necessary } O_2} = \frac{5.45 \times 100}{8.01}$$

= **68.1 per cent excess air**

5. Flue gas per lb. of coal:

Basis: 100 mols dry flue gas.

From item 3, above, 176.6 lb. of coal (as fired) per 100 mols flue gas, then

$= 100 \times \frac{1}{176.6}$ = 0.566 mol dry flue gas per lb. coal

Converting the dry gas to cubic feet at 60°F. and 30 in. pressure, saturated.

Mols dry flue gas per lb. of coal

Cu. ft. per mol

Temperature correction

Moisture correction

Pressure correction

$0.566 \times 359 \times \frac{520}{492} \times \frac{29.92}{29.40} \times \frac{29.92}{30.00}$ = **218.1 cu. ft. flue gas at 60°F., 30 in. Hg, saturated per lb. coal**

If the temperature of the flue gas is 580°F. the volume of the wet (actual) flue gas at this temperature is calculated (*a*) by determining the moisture in the coal and from the hydrogen in the producer gas (for the sake of simplicity neglecting the moisture passing through the producer undecomposed); (*b*) adding to dry gas to get total flue gas; and then (*c*) converting to the cubic feet at 580°F. per ton of lime.

(*a*) Moisture in flue gas:

From H_2 in producer gas:

Mols H_2 per mol producer gas	Mols producer gas per mol flue gas	
19.5	$\times \frac{100}{270}$	= 7.22 mols H_2 per 100 mols flue gas or 7.22 mols water per 100 mols flue gas

From moisture in coal:

Lb. H_2O per lb. coal	Mols flue gas per lb. coal	Mols H_2O per lb. H_2O	
0.054	$\times \frac{1}{1.766}$	$\times \frac{1}{18}$	$\times$ 100 = 0.17 mols water per 100 mols flue gas

Total moisture: $7.22 + 0.17 = 7.39$ mols per 100 mols flue gas

(*b*) Total flue gas:

Mols dry gas	Mols wet gas per mol dry gas	
0.566	$\times \frac{107.39}{100}$	= 0.609 mol wet flue gas per lb. of coal

(*c*) From mols wet gas per *lb. of coal* calculate cubic feet of wet flue gas at 580°F. per *ton of lime.*

Mols wet flue gas per lb. coal	Cu. ft. per mol.	Temperature correction	Pressure correction	Lb. coal per lb. lime	Lb. per ton	
0.609	359	$\frac{460 + 580}{492}$	$\frac{29.92}{30.00}$	$\frac{1}{3.78}$	2000	= 243,000 cu. ft. wet flue gas at 580°F. and 30 in. Hg. pressure per ton of lime

This type of problem also may be solved by an algebraic method. Using a basis of 100 mols of flue gas, there are three unknowns, the quantities of producer gas, air, and stone. These may be represented by the symbols P, A, and S respectively and three equations set up and solved. The method is as follows:

Basis 100 mols flue gas.

P = mols producer gas, A = mols air, S = pounds of stone.

Carbon balance:

$$\text{Carbon in stone} = \overset{\text{Lb. stone}}{S} \times \overset{\frac{\text{Lb. CaCO}_3}{\text{Lb. stone}}}{0.975} \times \overset{\frac{\text{Mols carbon}}{\text{Lb. CaCO}_3}}{\frac{1}{100}} = 0.00975S$$

Carbon in producer gas (1)	$0.30P$
Carbon in air	0
Carbon in flue gas (3)	22.7

$$0.00975S + 0.30P = 22.7 \qquad (a)$$

N_2 balance:

N_2 in stone	0
N_2 in producer gas	$0.569P$
N_2 in air	$0.79A$
N_2 in flue gas	71.7

$$0.569P + 0.79A = 71.7 \qquad (b)$$

O_2 balance:

$$O_2 \text{ in stone } \overset{\text{Lb. stone}}{S} \times \overset{\frac{\text{Lb. CaCO}_3}{\text{Lb. stone}}}{0.975} \times \overset{\frac{\text{Mols O}_2}{\text{Lb. stone}}}{\frac{1}{100}} = 0.00975S$$

$$O_2 \text{ in 1 mol producer gas} = 0.079 + 0.010 + \frac{0.184}{2} = 0.181$$

$$O_2 \text{ for } H_2 \text{ per mol of producer gas equals } \frac{0.121}{2} + 0.037 = 0.0975$$

which must be subtracted since it does not appear in flue gas.

$O_2 = P(0.181 - 0.0975) =$	$0.0835P$
O_2 in air	$0.21A$
O_2 in flue gas $22.4 + 5.6 + \frac{0.3}{2} =$	28.15

$$0.00975S + 0.21A + 0.0835P = 28.15 \qquad (c)$$

These three equations, (*a*), (*b*) and (*c*), may be solved readily to give the following results:

A = 64.1 mols of air per 100 mols flue gas.
P = 37.1 mols of producer gas per 100 mols flue gas, or
$\frac{100}{37.1}$ = 2.70 mols flue gas per mol producer gas.
S = 1190 lb. of stone per 100 mols of flue gas.

The required ratios may be determined from these values as in the previous method. For those persons familiar with the algebraic methods, this type of solution is much simpler and more easily visualized than the solution based on the nitrogen and theoretical air previously given.

The results of the above calculations on the lime-burning plant test are summarized as follows:

1. Fuel ratio: 3.78 lb. of lime produced per lb. of coal as fired.
2. Stone-coal ratio: 6.74 lb. of stone per lb. of coal as fired.
3. Lime-stone ratio: 0.56 lb. of lime produced per lb. of stone.
4. Producer gas-coal ratio: 80.6 cu. ft. of gas (60°F., 30 in. Hg, saturated) made per lb. of coal, as fired.
5. Excess air used for burning: 68.1 per cent.
6. Volume of flue gas per ton lime: 243,000 cu. ft. at 580°F., 30 in. mercury.

From the above ratios, several others may be calculated directly. In addition, various factors in the producer operation may be determined, as shown under the previous example on producer calculations. This last example shows the large number of results which may be obtained by analysis without direct measurement of any quantities. The ratio of lime to coal is an important one, and can be determined readily by the above method, where measurement of the quantities involved might be difficult.

The solution of most combustion problems and the calculation of results from tests are based on the methods outlined above. The carbon balance is used to determine flue-gas-fuel ratios, and the nitrogen balance is used to determine flue-gas-air ratios. When these ratios are known, the remaining calculations are simply further balances of materials, elements, or energy, to obtain the results desired. In the producer-gas problem, for example, the amount of steam decomposed was calculated by hydrogen and oxygen balances. In lime-kiln tests, the ratio of stone to lime was determined by a calcium balance. The *material* or *element* selected as a basis for comparing quantities may vary in different problems, but the *method* of calculation outlined is applicable to all cases.

Graphical Methods in Combustion Calculations.—It is possible in many cases, with the aid of charts, to determine quickly, and with a fair degree of accuracy, the per cent of excess air and the stack loss from a knowledge of the stack gas analysis and the stack temperature. Several charts for this purpose and the limitations restricting their use are given below.

Excess-air and Stack-loss Charts for Typical Fuels.—Since excess air is defined as the air present in excess of that required for combustion, the value may be determined from the flue-gas analysis for all cases where there is little or no nitrogen in the fuel. In such cases, the excess air may be written as

$$\text{Per cent excess air} = \frac{O_2 - \frac{1}{2}CO}{N_2 \times {}^{21}\!/\!_{79} - (O_2 - \frac{1}{2}CO)} \times 100,$$

where $[O_2 - \frac{1}{2}CO]$ represents the oxygen in excess of that required for combustion. The term $[N_2 \times {}^{21}\!/\!_{79}]$ represents the total O_2 in the air used, and $[N_2 \times {}^{21}\!/\!_{79} - (O_2 - \frac{1}{2}CO)]$ represents the O_2 necessary for combustion.

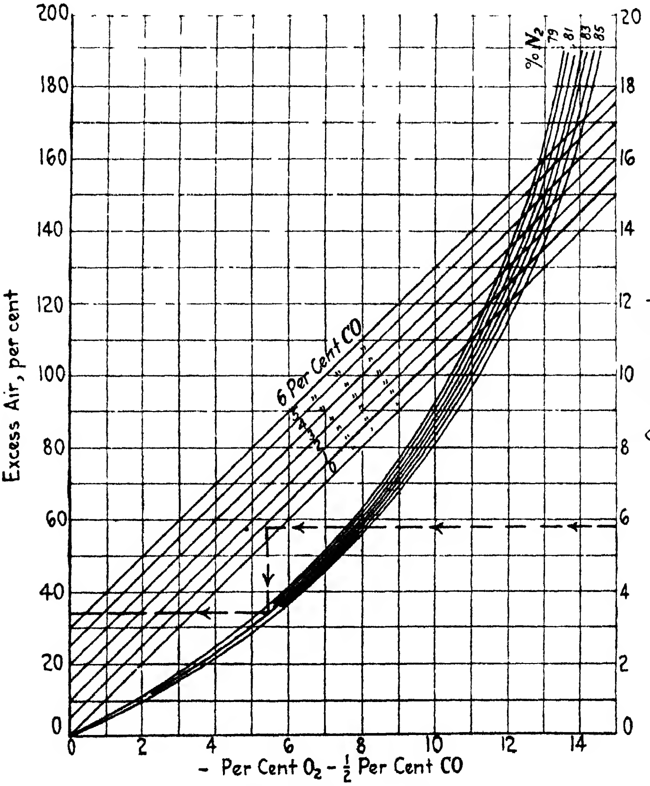

Fig. 51.—Chart for graphical determination of excess air from flue-gas analysis.

In Fig. 51 the per cent excess air is plotted as ordinates with $(O_2 - \frac{1}{2}CO)$ as abscissae. The nitrogen in the flue gas affects the value of the denominator in the above equation, so that a series of curves for varying percentages of nitrogen result. By drawing a series of straight lines on the same plot, with per cent O_2 as ordinates, it is possible to determine the value of (O_2 − ½CO) without actual subtraction. The method of using this chart is to start with the per cent O_2 on the right, pass horizontally to the proper CO line, and then vertically down to the per cent N_2, and horizontally to the per cent excess air. Thus, if the analysis of

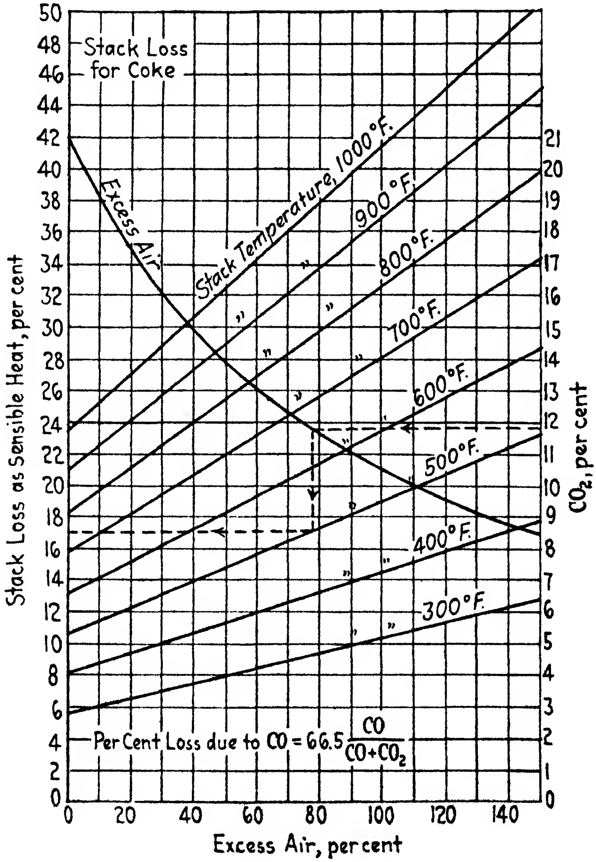

FIG. 52.—Chart for graphical determination of per cent excess air and per cent stack loss in a coke-fired furnace.

a flue gas were CO_2, 13 per cent; O_2, 5.7 per cent; CO, 0.6 per cent; N_2, 80.7 the excess air would be found, by following the dotted line, to be 34 per cent. This value read directly from the chart checks with the value of 33.5 per cent excess air calculated on page 222.

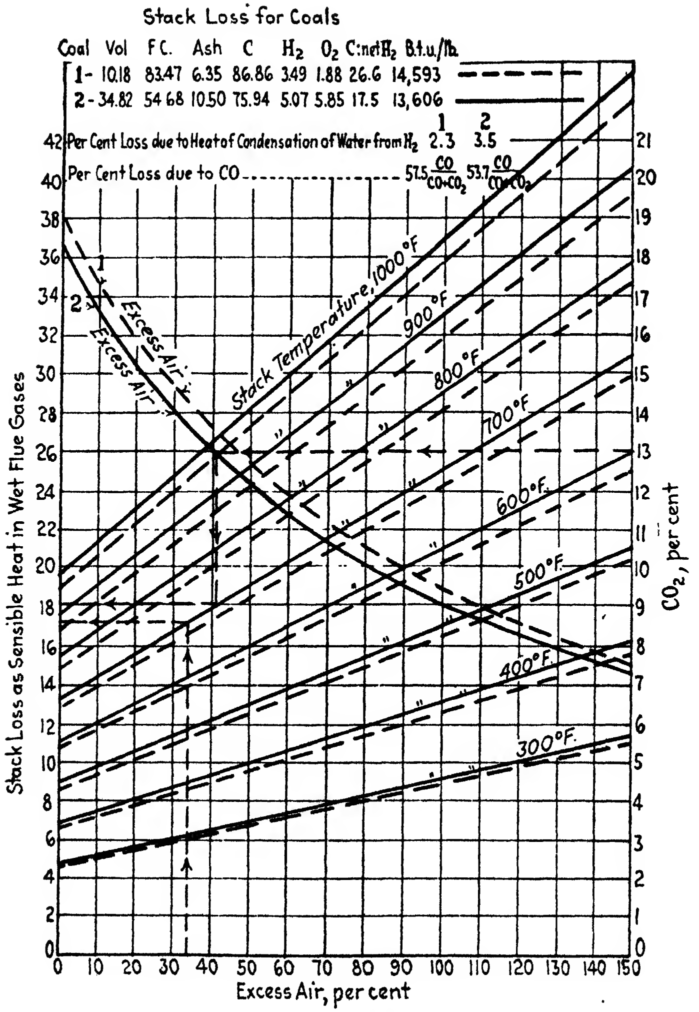

FIG. 53.—Chart for graphical determination of per cent excess air and per cent stack loss in a furnace fired with a high-volatile or low-volatile bituminous coal.

This chart covers all cases where the nitrogen in the fuel is negligible, which is true for most cases except when producer or blast-furnace gas is used. If the hydrogen-carbon ratio in the fuel is known, it is possible to develop a simpler chart in which the excess air may be determined from the CO_2 content of the gas alone, provided no CO is present. In such a case, the N_2 and the O_2 are direct functions of the CO_2, so that the CO_2 shows the excess air directly. In Figs. 52, 53, 54, 55, and 56 the excess-air and stack losses as functions of the CO_2 in the gas and its temperature are plotted for typical solid, liquid, and gaseous fuels. These figures are read by taking the percentage of CO_2 in the flue gases, reading across (from right to left) to the excess-air line, then reading up or down to the stack temperature, and then across (to the left) to determine the per cent heat loss as sensible heat in the stack gases. Thus, for a coke-fired furnace, if the flue gases contained 11.8 per cent CO_2 and the stack temperature

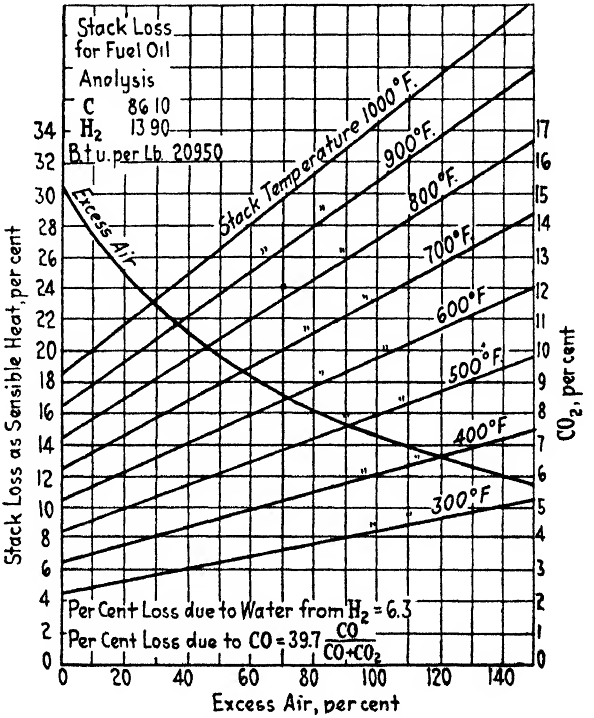

FIG. 54.—Chart for graphical determination of per cent excess air and per cent stack loss in a furnace fired with fuel oil.

were 500°F., by following the dotted line, the excess air is found to be 77 per cent, and the sensible heat in the stack gases to be 17 per cent of the heat in the coke.

It is interesting to compare results obtained from the charts with the calculated values obtained in the illustrative problems

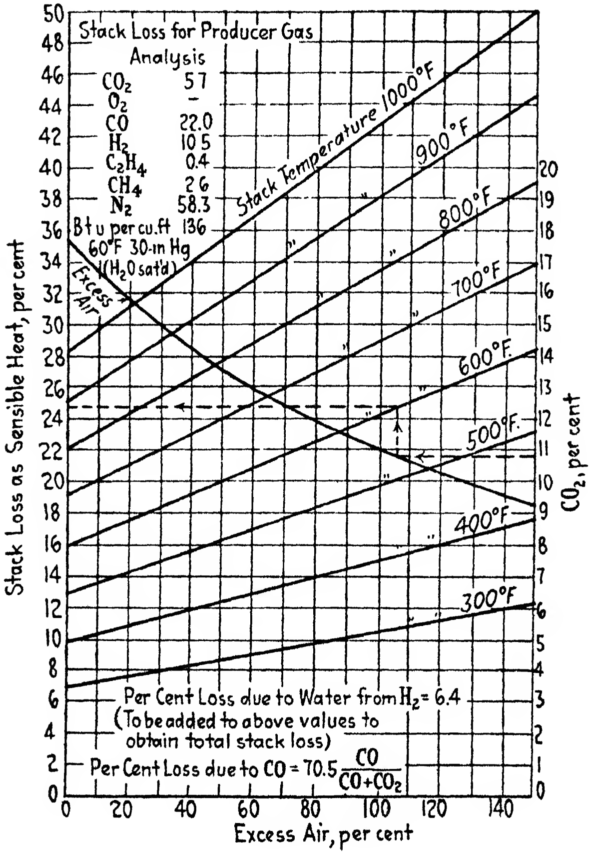

FIG. 55.—Chart for graphical determination of per cent excess air and per cent stack loss in a furnace fired with producer gas.

previously given. For example, in the first problem on boiler-furnace calculations (p. 216) the stack gas contained 13.0 per cent CO_2, 0.6 per cent CO, and was at a temperature of 700°F. Reading across to the left from 13.0 per cent CO_2 in Fig. 53 (with the dotted line), the excess air is found to be 41.5 per cent (as against 33.5 per cent calculated), the total sensible heat in the wet stack gases* at 700°F. is found to be 18.1 per cent, the loss due to heat of condensation of water (from legend on Fig. 53)

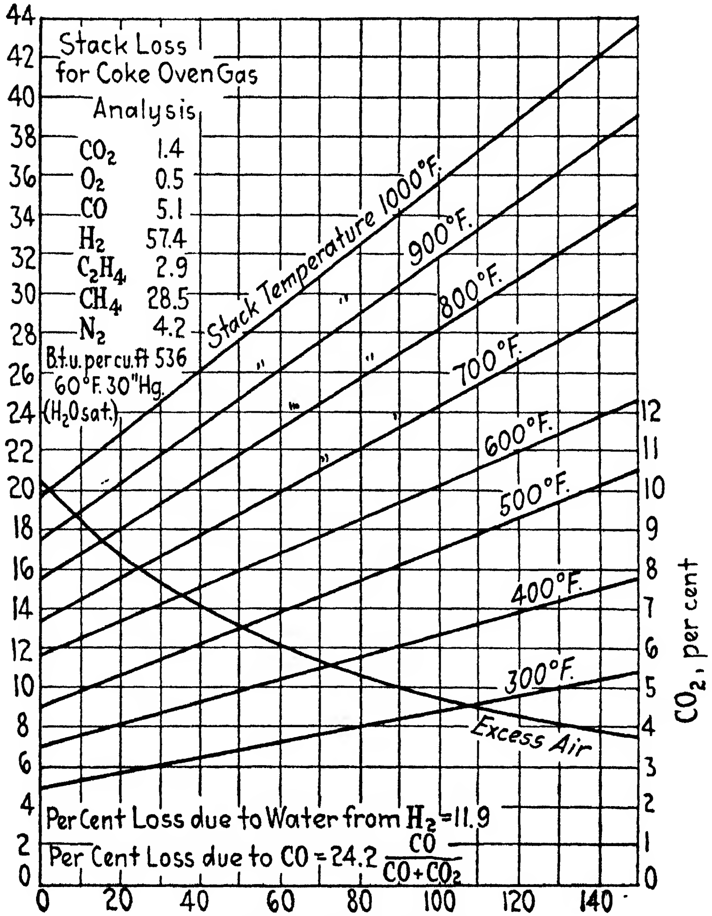

FIG. 56.—Chart for graphical determination of per cent excess air and per cent stack loss in a furnace fired with coke-oven gas.†

* That is, including the sensible heat in the water vapor from the combined water in the coal, but *not* including moisture in the coal, since the chart is drawn up from calculations on a dry coal.

† The method for calculating the loss due to unburned CO as given in the legend of Figs. 52, 53, 54, 55, 56 is based on the following calculation:

$$\frac{\left(\frac{CO}{CO + CO_2}\right) \times (\text{mols Carbon per lb. Coal}) \times 122{,}400}{\text{B.t.u. per lb. Coal}} = \text{Per cent Heat Loss in Unburned CO.}$$

is found to be 3.5 per cent, and the loss due to unburned CO (substituting in the formula given in the legend) is found to be $53.7 \times \frac{0.6}{0.6 + 13.0} = 2.4$ per cent. The total stack loss, therefore, is $18.1 + 3.5 + 2.4 = 24.0$ per cent (as against 22.7 per cent calculated). This discrepancy between the values read from the chart and the calculated values is due to the fact that the excess-air line in Fig. 53 is drawn up on the assumption that combustion is complete (*i.e.*, there is no CO in the gas). The figures obtained for stack loss as sensible heat in the wet flue gases and for the loss due to heat of condensation of the water vapor would be quite accurate if the gas contained 13 per cent CO_2 and no CO. Even the presence of as little as 0.6 per cent CO throws off the excess-air curve, and, therefore, for cases where CO is present in the gas it is necessary first to determine the excess air from Fig. 51. By using Fig. 51, which gives the per cent excess air accurately, and then using on Fig. 53 the value for excess air thus determined, it is possible to obtain much more accurate values. This may be illustrated for the same stack gas by repeating the graphical determination of stack loss, using the value of 34 per cent excess air previously read from Fig. 51. Reading up from 34 per cent excess air in Fig. 53 (with the dotted line) the stack loss as sensible heat in the wet flue gases is found to be 17.2 per cent and the total stack loss, therefore, is $17.2 + 3.5 + 2.4 = 23.1$ per cent as against 22.7 per cent calculated. Then, assuming 5 per cent as radiation and unaccounted-for losses, and taking the calculated value* of 4.2 per cent (p. 218) for the cinder loss, the total losses are $23.1 + 5 + 4.2 = 32.3$ per cent. The efficiency is 100 per cent minus the losses, or $100 - 32.3 = 67.7$ per cent as against the calculated value of 69.8 per cent. The comparison of the calculated and graphically determined values for the first problem shown in the following table.

---

* Per cent cinder loss $= \frac{\text{Pounds of Ash}}{\text{Pounds of Coal}} \times \frac{\text{Pounds of Combustible}}{\text{Pounds of Cinder}} \times \frac{14{,}544}{\text{B t.u. of Coal}} \times 100$, or, for this example, $= 0.097 \times \frac{28.5}{71.5} \times \frac{14{,}544}{13{,}400} \times 100 = 4.2$ per cent.

TABLE LVII

Comparison of Calculated and Graphically Determined Results of Boiler Test

(In per cent of heat in coal.)

Item	Calculated values	Graphical values
Sensible heat + heat of condensation of water in wet stack gases	20.3	20.7
Undeveloped heat in CO	2.4	2.4
Radiation and unaccounted for	3.3	5.0 (assumed)
Heat lost in cinder (same calculation for both)	4.2	4.2
Total losses	30.2	32.3
Efficiency	69.8	67.7

A similar comparison between calculated and graphically determined results, for the second boiler-furnace problem (p. 223), is given in the following table.

TABLE LVIII

Comparison between Calculated and Graphically Determined Results of Boiler Test

Data:

CO_2 in stack gases	11.0 per cent
Temperature of stack gases	600°F.
Cinder loss	2.5 per cent
Excess air (from Fig. 51)	60.5 per cent

Item	Calculated, per cent	Graphically determined, per cent
Sensible heat in *wet* stack gases	16.4	17.0
Heat loss due to heat of condensation of H_2O vapor	4.0	3.5
Radiation + unaccounted-for losses (assumed)	5.0	5.0
Heat lost in cinder	2.5	2.5
Total losses	27.9	28.0
Efficiency	72.1	72.0

The agreement between the efficiency values calculated from complete test data and those obtained by the shorter graphical method clearly shows the value of such charts as Figs. 52 to 56 for quick, fairly accurate plant tests. They are particularly valuable for rough approximations where the only data available are the percentage of CO_2 from a recorder and the stack temperature. These charts are subject to quite high errors: (1) where the carbon-hydrogen ratio of the fuel is markedly different from those used in preparing the charts, (2) where either or both the CO and cinder losses are large, and (3) where the moisture content of the fuel is high. The charts offer, however, a convenient and rapid method for the approximation of efficiencies from readily available plant data.

Methods for Approximating Heating Value and Analysis of Fuels.—In testing work it is often necessary to know the heating values or ultimate analyses of fuels when the apparatus required for such determinations is not available. A number of empirical methods have been evolved for this purpose, and, though often inaccurate, it may be necessary to use them when actual data are missing.

Dulong's formula gives the most satisfactory method for calculating the heating value of solid fuels. The formula is:

$$\text{B.t.u. per lb.} = 14{,}544C + 61{,}500\left(H - \frac{O}{8}\right) + 4500S$$

where C = the fraction of carbon in the coal; H is the fraction of hydrogen; O is the fraction of oxygen; and S is the fraction of sulfur. The probable error using this formula is about 2 per cent except in the case of cannel coals and coals of high oxygen content, such as lignites and midwest bituminous coals, in which the error may run higher. With oil fuels the per cent error may be as much as 8 per cent. This method, in addition to the error introduced, has the disadvantage of requiring the lengthy ultimate analysis.

Figure 57, adapted from Kent,[5] has been utilized to determine (approximately) the heating value of coals from a knowledge of the proximate analysis. The figure shows the relationship between the volatile matter as the per cent of the total combustible, and the B.t.u. per lb. of combustible. Thus, if a coal has 30.0 per cent volatile, 60 per cent fixed carbon, and 10 per cent ash, the volatile matter is $\frac{30}{30 + 60}$, or 33.3 per cent of the combustible and the B.t.u. per lb. of combustible is 15,300. The heating value of the coal is, then, 15,300 × 0.9, or 13,770 B.t.u. per lb.

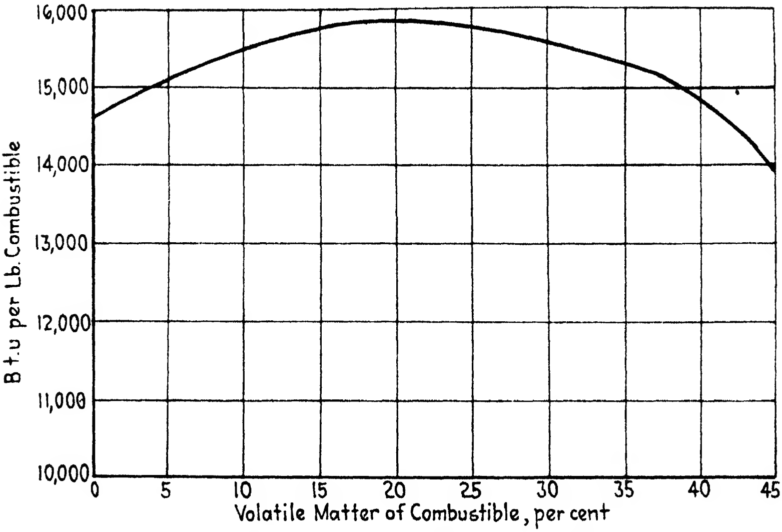

FIG. 57.—Relation between heating value and percentage of volatile matter in the combustible of coal. (*Adapted from William Kent: "Steam Boiler Economy," p.* 59 (1915). *John Wiley and Sons.*)

The heating value of fuel oils may be determined approximately by means[6] of the equation:

$$\text{B.t.u. per lb.} = 18{,}650 + 40\,(\text{degrees Baumé} - 10).$$

Values obtained with this formula have been plotted as a function of the degrees Baumé in Fig. 33, Chap. VI.

The heating value of a gaseous fuel may be determined from the analysis by adding together the products of the percentage and heating value of each constituent. The method is outlined in Table LIX.

TABLE LIX

CALCULATIONS OF HEATING VALUE OF GAS

Constituent	Per cent	Heating value (high) B.t.u. per cu. ft., 60°F., 30in., sat.	$\frac{(\text{Heating value})\ (\text{per cent})}{100}$
CO_2	2.0	0.0	0.0
O_2	0.4	0.0	0.0
C_2H_4	2.6	1613.0	41.9
CO	5.4	317.1	17.1
H_2	55.4	318.8	176.6
CH_4	29.5	995 0	293.5
N_2	4.7	0.0	0.0
Total	100.0		529.1 B.t.u.(high) per cu. ft., 60°F., 30 in. Hg, saturated with H_2O

The net heating value may be calculated in the same manner using the values for the lower heats of combustion given in Table LII. The heating value at 32° and 29.22 in. Hg may be calculated from the above value by correcting for temperature and moisture as follows:

$$\overset{\text{B.t.u. 60°F. 30 in. Hg saturated}}{529.1} \times \overset{\text{Temperature correction}}{\frac{520}{492}} \times \overset{\text{Correction for water vapor}}{\frac{30.00}{30.00 - 0.52}} \times \overset{\text{Pressure correction}}{\frac{29.92}{30.00}} = 529.1 \times 1.073$$

or 529.1 × 1.073 = 567.6 B.t.u. per cu. ft. at 32°F., 29.92 in. Hg.

As stated earlier, it is often necessary to know the carbon content and the ultimate analysis of a coal, when only the proximate analysis and the heating value are available. Although the actual ultimate analysis is essential for accurate work, it is possible to estimate the ultimate from the proximate analysis by means of a series of equations derived by Calderwood.[7] These equations, which were derived from the consideration of a large number of Bureau of Mines coal analyses, are purely empirical, but often offer the only available method for estimating the carbon content of a coal. The five equations are:

FUNDAMENTAL FACTORS IN THE COMBUSTION OF GASEOUS FUELS

In burning a gas with air, it is first necessary to have a mixture that will ignite, and then to raise this mixture to the temperature of ignition. In order to carry out this procedure with efficiency and safety, it is necessary to know the speed with which such a mixture burns. From the standpoint of quality and efficiency, it is also highly important to have a knowledge of the amount of heat radiated directly from the flame, the temperature of the flame, the amount of air required, and the effect of various factors, such as contact with surfaces, on the speed with which the gaseous combustion will take place. These points are discussed individually in the subsequent paragraphs. A brief descriptive treatment of regeneration and recuperation has also been included on account of the indispensability of these methods of heat recovery in many important industrial operations.

Ignition Temperatures.—If a combustible gas-air mixture is heated gradually, the velocity of combustion at first increases regularly but fairly slowly as the temperature is raised. Finally, however, the ignition temperature is reached, chemical combination is not dependent on heat from external sources, and practically instantaneous combustion occurs. The *ignition temperature* is the lowest temperature at which this rapid combustion will take place, and may be defined as the temperature at which heat is generated by combustion faster than heat is lost to the surroundings, and combustion thus becomes autogenous or self-propellent. Below this temperature the gas-air mixture will not burn freely and continuously unless heat is supplied. The actual ignition temperature of a gas is, therefore, seen to depend on surrounding conditions, but the term as ordinarily applied is used to signify the temperature at which rapid combustion occurs in ordinary air.

Two general methods of ignition-temperature measurement have been used to prevent errors due to "pre-ignition" or slow combustion previous to actual ignition: (1) by exploding the gaseous mixtures by adiabatic compression, and (2) by preheating independently each stream of gas and air before mixing, gradually increasing the temperature of each stream until self-propellent combustion takes place on mixing. In each of these methods the time possible for the preliminary interaction after mixing is negligible.

The method used by Tausz and Schulte[1] for determining the ignition temperatures of liquids and solids consisted in introducing drops of the liquid or the finely pulverized solid into a cup through which was passed a controlled amount of preheated air or oxygen. The temperature of the cup was raised well above the ignition point by external heating, and was then allowed to fall gradually until a temperature (the ignition point) was reached below which ignition was no longer possible. It was found that ignition points determined in this way could be more closely duplicated than by gradually raising the temperature and noting where ignition first took place.

The gaseous ignition temperatures observed by Dixon and Coward,[2] who used the method of separately preheating streams of gas and air mentioned above, and the ignition temperatures of various liquids and solids determined by Tausz and Schulte[1] are given* in Table LXI.

The ignition temperature of a substance under a given set of conditions is quite definite, and the range indicated for the gaseous substances in the above table is due to the impossibility of obtaining thorough mixing in all the experiments carried out by Dixon and Coward. Dixon and Coward's values give a good idea, however, of the temperatures required for the ignition of the common gases. The ignition temperatures are, in general, lower in oxygen than in air, though the values for hydrogen and carbon monoxide are not apparently affected. The ignition temperature of hydrocarbons of the paraffin series is lowered with increasing molecular weight.

The ignition point of most substances is lowered with increasing pressures. This is of great importance in gas engines, in which the explosive mixture is compressed before ignition. This compression increases the temperature of the gas and it is, therefore, necessary to choose a fuel which will withstand this compression without preignition. Increasing pressure also lowers the ignition point of the paraffin hydrocarbons, mineral oils, etc., although this decrease is not much over 180°F. or 100°C. at pressures below 150 lb. per sq. in. At higher pressures the decrease becomes more marked, and in the case of a petroleum,[1] for example, increasing the pressure from about 40 to about 385 lb. per sq. in. gage pressure lowered the ignition temperature from

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* The ignition temperatures of several types of coal are given in Chap. IV, p. 76.

TABLE LXI

Ignition Temperatures (at Atmospheric Pressure) of Gases, Liquids, and Solids

Substance	Ignition temperature in air		Ignition temperature in oxygen	
	Degrees Fahrenheit	Degrees Centigrade	Degrees Fahrenheit	Degrees Centigrade
Hydrogen (H_2)	1076–1094	580–590	1076–1094	580–590
Carbon monoxide (CO)	1191–1216	644–658	1179–1216	637–658
Methane (CH_4)	1202–1382	650–750	1033–1292	556–700
Ethane (C_2H_6)	968–1166	520–630	968–1166	520–630
Propane (C_3H_8)	....	....	914–1058	490–570
Ethylene (C_2H_4)	1008–1018	542–547	932– 966	500–519
Acetylene (C_2H_2)	763– 824	406–440	781– 824	416–440
Hexane* (C_6H_{14})	909	487	514	268
Decane ($C_{10}H_{22}$)	865	463	396	202
Benzol (C_6H_6)	1364	740	1224	662
Toluol	1490	810	1026	552
Phenol	1319	715	1065	574
Aniline	1418	770	986	530
Methyl alcohol	....	....	1031	555
Ethyl alcohol	1036	558	797	425
Propyl alcohol	941	505	833	445
Isopropyl alcohol	1094	590	954	512
n-Butyl alcohol	842	450	725	385
Amyl alcohol	768	409	734	390
Ethyl ether	649	343	352	178
Glycerine	932	500	777	414
Acetone	1292	700	1054	568
Sugar	725	385	712	378
Cylinder oil	783	417	608	320
Pennsylvania crude	601	367	468	242
Gas oil	637	336	518	270
Kerosene	563	295	518	270
Acetaldehyde	365	185	284	140
Benzaldehyde	356	180	334	168

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* Ignition temperatures for hexane and for other substances in the remainder of table are for dry air and dry oxygen.

788°F. (420°C.) to 392°F. (200°C.). The amount of moisture in the air or oxygen used for ignition also exerts a small effect, usually not more than 90°F. (45°C.), the tendency of the moisture, in general, being to increase the ignition temperature.

Limits of Inflammability.—The explosive limits or limits of inflammability may be defined as the limiting composition of a combustible gas and air mixture beyond which the mixture will not ignite and continue to burn. In other words, a gas-air mixture of a composition outside these limits cannot generate heat at a sufficient rate to make the combustion self-propellent under the conditions of the experiment. The lower limit of inflammability represents the smallest proportion of the gas which, when mixed with air, will burn without the continuous application of heat from an external source. Above the upper limit the large amount of combustible gas present acts as a diluent and again combustion cannot become self-propellent.

In the ordinary combustion of a gas, the temperature of the whole gas-air mixture is usually not up to the ignition point. Inflammation is started at a single point by some external source of heat, and the heat developed by combustion serves to raise the rest of the mixture up to the ignition temperature. For this to take place, it is necessary that the proportions of gas and air in the mixture be within the limits of inflammability; otherwise no combustion will take place. The question of explosive limits may also be important in plant work, from the standpoint of recognizing that an explosive mixture is being handled.

The range of the limits of inflammability of the various gases in air is given[3,4,5] in Table LXII.

Mason and Wheeler[6] have shown that increased pressure above atmospheric narrows both the upper and lower limits of inflammability of H_2 and CO. Thus, at high pressure more CO or H_2 is necessary to form a lower-limit mixture, and more air is necessary to form an upper-limit mixture. Similar results were obtained with lower-limit methane mixtures, but in upper-limit mixtures increasing pressure extended the upper limit of inflammability. Payman and Wheeler[7] have shown that ethane, propane, butane, and pentane act like methane. The influence of pressure on the upper limits of inflammability of the hydrocarbons studied by Payman and Wheeler is shown in Fig. 58. The influence of reduction in pressure[5] on the limits of inflammability of CO-air mixtures is shown in Fig. 59. The limits of inflam-

TABLE LXII

LIMITS OF INFLAMMABILITY OF GASES IN AIR

(At ordinary temperatures and at atmospheric pressure)

Gas	Limits as per cent by volume in air	
	Per cent lower-limit mixture	Per cent upper-limit mixture
Hydrogen (H_2)	6.2	71.4
Carbon monoxide (CO)	16.3	71.2
Coal gas	7	21
Coke-oven gas	7	21
Blue water gas	12	67
Blast-furnace gas	36	65
Methane (CH_4)	5.8	13.3
Ethane (C_2H_6)	3.3	10.6
Ethylene (C_2H_4)	3.4	14.1
Ethyl alcohol (C_2H_5OH)	3.7	13.7
Benzene (C_6H_6)	1.4	5.5
Pentane (C_5H_{12})	1.3	4.9
Ether ($(C_2H_5)_2O$)	1.6	7.7

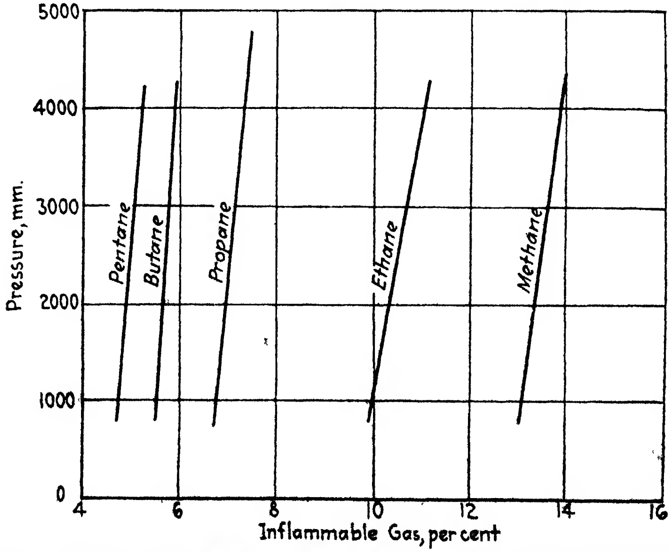

FIG. 58.—Influence of increase in pressure on the upper limits of inflammability of hydrocarbon-air mixtures.

mability are narrowed as the pressure is decreased until finally a critical pressure is reached, below which the gas-air mixture will not ignite.

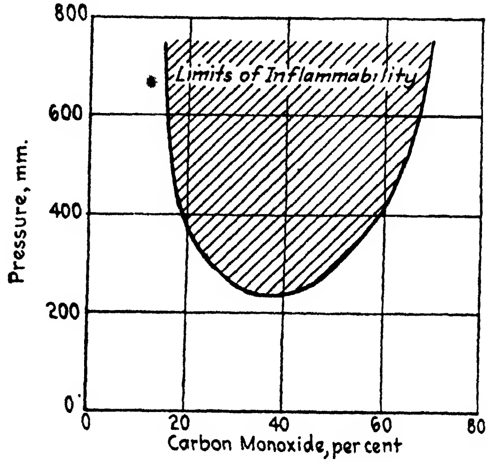

FIG. 59.—Effect of reduction in pressure on the limits of inflammability of carbon monoxide.

An increase in temperature tends to decrease the lower limit of inflammability, as is shown[10] for the case of CO in the following table:

TABLE LXIII

EFFECT OF TEMPERATURE ON LOWER LIMIT OF INFLAMMABILITY OF CO

TEMPERATURE	LOWER LIMIT, PER CENT CO
752°F. (400°C.)	14.2
914°F. (490°C.)	9.3
1112°F. (600°C.)	7.4

Rate of Flame Propagation.—The rate of flame propagation, or the speed with which a flame will travel through a gas-air mixture, is of great importance, particularly from the standpoint of gas-burner design. For example, in a burner of the Bunsen type (in which part of the air for combustion is injected into a mixing tube by the momentum of gas flowing through an orifice), if the linear velocity of the gas-air mixture flowing through the burner ports is not greater than the velocity of flame propagation, back-firing will result.

Flame in a gas-air mixture, depending on the conditions under which the mixture is ignited, can take place in three ways: (1) with a relatively slow, uniform motion, (2) a somewhat faster, accelerating, vibratory motion, which may develop into (3) a very rapid explosion wave, which may be, and often is, accompanied by detonation. The so-called vibratory movement, as far as practical considerations go, is less important than either the speed of uniform movement or the velocity of the explosion wave.

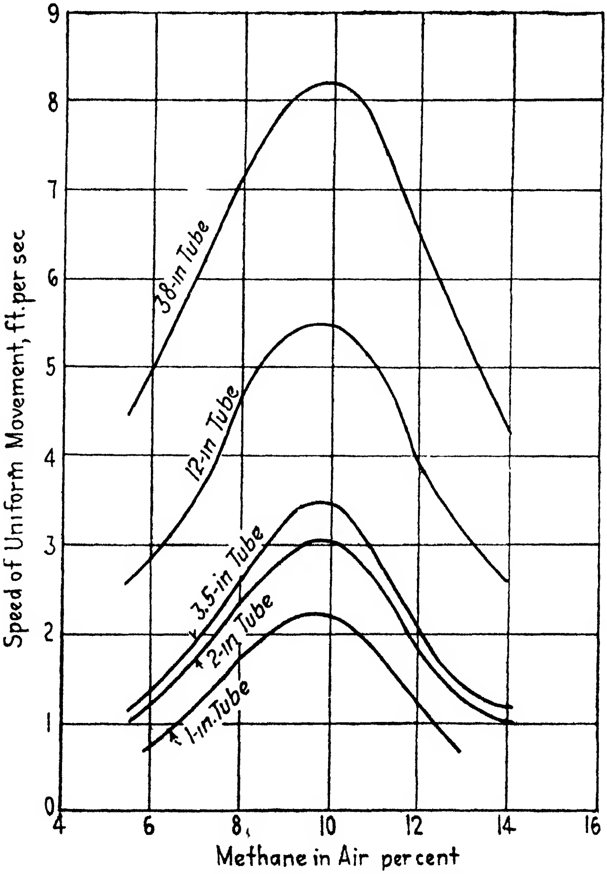

FIG. 60.—Speed of uniform movement of methane flames in air. For tubes of different sizes.

Slow flame propagation is influenced by the concentration of oxygen and combustible gas in the mixture, and by the size and shape of the pipe or container in which the flame propagation takes place. This is illustrated[6] by Fig. 60, which shows the effect of varying tube diameter on the flame velocity in mixtures of methane and air. It is evident from the curves that the speed of flame propagation is greatly decreased as the size of the tube is decreased due, in part at least, to the proportionally greater cooling effect exerted by the smaller tubes. Payman and Wheeler,[9] working with coal gas and air, found that flame propagation was impossible through tubes smaller than 0.08-in. diameter (2 mm.), with any mixture of coal gas and air. Similarly, with methane, no flame travel was observed in tubes of 0.142-in. diameter (3.6 mm.). It is interesting to note that flame travel with hydrogen, which has a much higher rate of flame propagation than either methane or coal gas, was reported by Mallard and Le Chatelier[10] through a tube as small as 0.035-in. diameter (0.9 mm.). The recognition of the effect of tube diameter on flame propagation is, obviously, of major importance in problems of burner design.

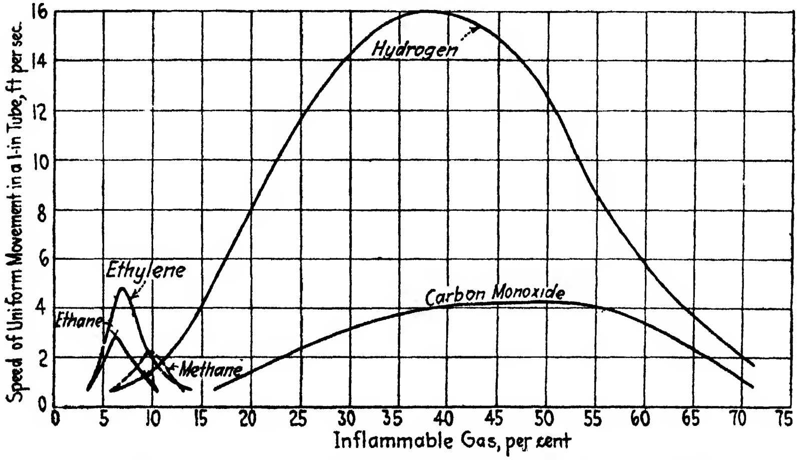

FIG. 61.—Speed of uniform movement of flames of various pure gases in air.

The maximum flame speed in tubes is usually obtained in mixtures containing about 15 per cent more combustible gas than is required for complete combustion. With methane, however, maximum speed occurs when the methane and the oxygen are in their combining proportions.

Reported values of speed of uniform movement apply only to the specific conditions under which the determination was carried out. The relative flame speeds of varying mixtures usually remain approximately the same even though the absolute values vary.

As already illustrated by Fig. 60, when the flame speed is plotted against the percentage of combustible gas in the gas-air mixture (giving a so-called "speed-percentage" curve), a bell-shaped line is obtained. At the lower limit of inflammability, the flame speed is low, gradually increasing, however, to a maximum as the proportion of combustible gas in the mixture is increased. Beyond the maximum-speed point, the addition of combustible gas decreases the flame speed, until finally, at the upper limit of inflammability, the same low speed is found as for the lower-limit mixture. This is illustrated[9] by Fig. 61, which shows the speeds of uniform movement of flame (in a 1-in. diameter tube) observed for various mixtures of air and the simple gases H_2, CO, CH_4, C_2H_6, and C_2H_4. Figure 62 gives[9] the observed flame speeds for mixtures of equal volumes of CO and H_2 with varying proportions of air, and also shows the flame speeds of coal-gas-air and producer-gas-air mixtures.

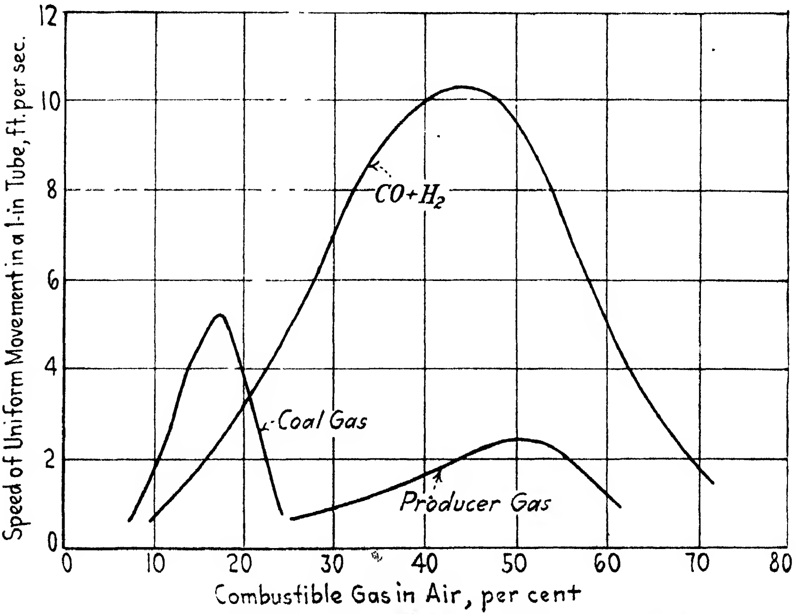

FIG. 62.—Speed of uniform movement of flames of various gases in air.

The observed flame speeds in the maximum-speed mixtures, as well as the composition of the limit mixtures (*i.e.*, the upper and the lower limits of inflammability) for these gases, are given in Table LXIV.

TABLE LXIV

Maximum-speed and Limit Mixtures of Common Gases in Air

(All in 1-in. diameter tube)

Gas	Per cent of gas in maximum-speed mixture	Speed (feet per sec.) in maximum-speed mixture	Per cent of gas in	
			Lower-limit mixture	Upper-limit mixture
Hydrogen (H_2)	38.5	15.9	6.2	71.4
Carbon monoxide (CO)	50.0	4.1	16.3	71.2
Methane (CH_4)	9.9	2.2	5.8	13.3
Ethane (C_2H_6)	6.3	2.8	3.3	10.6
Ethylene (C_2H_4)	7.1	4.66	3.4	14.1
CO + H_2 (equal volumes)	44.0	10.3	9.2	71.3
Coal gas	17.0	5.2	7.2	24.3
Producer gas	51.0	2.4	24.7	61.6

The speed of uniform movement of *all the limit* (*upper or lower*) *mixtures, simple or complex, is approximately the same,* namely 0.65 ft. per sec. under the conditions specified above.

The figures and the table both emphasize the much greater flame speed of hydrogen, compared with either CO or the hydrocarbons, and the data in the table for equal mixtures of CO and H_2 show how the admixture of hydrogen raises the flame speed of the mixture.

The speed of uniform movement or flame propagation in a mixture of two or more of the simple gases, CO, H_2, CH_4, C_2H_6, C_2H_4, with air can be calculated with a fair degree of approximation. As Payman and his co-workers have stated,[9] the combustion of complex gaseous mixtures can be regarded as involving the simultaneous but independent burning of a number of simple mixtures of the individual gases with air, in which the proportions of inflammable gas and air are such that each mixture, if burning alone, would propagate flame with the same speed (under the same conditions of experiment) as does the complex mixture. Thus a mixture of CO, H_2, and air which has a flame speed of 2 ft. per sec. may be regarded as consisting of separate mixtures of CO and air, and H_2 and air, both mixtures being of such composition that, if burned separately, they would give a flame speed of 2 ft. per sec. Similarly, any complex limit mixture has the composition and the characteristics that would be obtained by mixing the simple limit mixtures (provided they all be of the same limit, *i.e.*, upper or lower) of the individual gases. In the same manner a maximum-speed complex gas-air mixture may be considered as consisting of a mixture of maximum-speed simple gas-air mixtures of the individual combustible gases. These relationships may be used to calculate the composition of the limit mixture for a complex gas-air mixture, to determine the speed and composition of the maximum-speed mixture, and to estimate graphically the flame speed of mixtures other than the maximum speed or limit. Methods for making these calculations (as outlined by Payman and Wheeler[9]) and the limitations for the coal gas and producer gas having the flame speeds shown in Fig. 62 are given in the following paragraphs.

Suppose, for example, that it is desired to calculate the composition of the limit mixtures (both upper and lower) and of the maximum-speed mixture of the coal gas having the observed flame speed shown in Fig. 62. This gas had the following analysis:

	Per Cent
Carbon monoxide	9.6
Hydrogen	49.2
Unsaturated hydrocarbons	3.7
Methane	30.9
Ethane	3.0
Inerts	3.6
	100.0

The composition of the various limit mixtures can then be calculated by means of the following formula:*

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* An abbreviated derivation of this expression for the limit mixtures (adapted from Payman) is as follows: Assume a limit mixture composed of a per cent of the simple gas A, b per cent of the simple gas B, and c per cent of the simple gas C, the respective limit mixtures of these simple gases being L_a, L_b, and L_c. Then the complex limit mixture is composed of:

$$aA + bB + cC + \ . \ . \ . + 100 - (a + b + c + \ . \ . \ . \ ) = \text{gas} + \text{air}.$$

From the hypothesis that the complex limit mixture may be regarded as

$$L_x = \frac{a + b + c + \cdots}{\frac{a}{L_a} + \frac{b}{L_b} + \frac{c}{L_c} + \cdots}, \tag{1}$$

where L_x is the required limit* of inflammability of the gas mixture; a, b, c, etc. are the percentages of the simple gases as shown by the above analysis (*e.g.*, a, or per cent CO in the above analysis, equals 9.6 per cent); and L_a, L_b, L_c, etc. are corresponding limit percentage of the simple gases, already given in the table on page 270.

The compositions of the various mixtures for the coal gas are then calculated as follows:

1. Lower-limit mixture:

$$L_x = \frac{9.6 + 49.2 + 3.7 + 30.9 + 3.0}{\frac{9.6}{16.3} + \frac{49.2}{6.2} + \frac{3.7}{3.4} + \frac{30.9}{5.8} + \frac{3.0}{3.3}}$$

= 6.2 per cent coal gas in the lower-limit mixture.

2. Upper-limit mixture:

$$L_x = \frac{9.6 + 49.2 + 3.7 + 30.9 + 3.0}{\frac{9.6}{71.2} + \frac{49.2}{71.4} + \frac{3.7}{14.1} + \frac{30.9}{13.3} + \frac{3.0}{10.6}}$$

= 26.2 per cent coal gas in the upper-limit mixture.

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a mixture of simple limit mixtures, it follows that the limit mixture is composed of:

$$a + \left(\frac{100 - L_a}{L_a}\right)a,\ b + \left(\frac{100 - L_b}{L_b}\right)b,\ c + \left(\frac{100 - L_c}{L_c}\right)c,$$

Since these terms are expressed as percentages,

$$a + \left(\frac{100 - L_a}{L_a}\right)a + b + \left(\frac{100 - L_b}{L_b}\right)b + c + \left(\frac{100 - L_c}{L_c}\right)c + \cdots = 100,$$

which on simplification becomes:

$$\frac{a}{L_a} + \frac{b}{L_b} + \frac{c}{L_c} = 1.$$

Now, since $a + b + c = L_x$, where L_x is the percentage of the complex gas in the limit complex mixture, it follows that:

$$\frac{a + b + c}{L_x} = 1 = \frac{a}{L_a} + \frac{b}{L_b} + \frac{c}{L_c}$$

or

$$L_x = \frac{a + b + c}{\frac{a}{L_a} + \frac{b}{L_b} + \frac{c}{L_c}}$$

A similar process leads to Eq. (2) for the speed of flame in maximum-speed mixtures.

* Where it is desired to determine the composition of the maximum speed mixture, the word "maximum" should be substituted for "limit" in the explanation of the formula.

3. Maximum-speed mixture:

$$L_x = \frac{9.6 + 49.2 + 3.7 + 30.9 + 3.0}{\frac{9.6}{50.0} + \frac{49.2}{38.5} + \frac{3.7}{7.1} + \frac{30.9}{9.9} + \frac{3.0}{6.3}}$$

= 17.2 per cent coal gas in the maximum-speed mixture. These calculated values should be compared with the observed values shown in Fig. 62. Thus the observed lower-limit mixture was 7.2 per cent coal gas in air as against 6.2 per cent calculated; the observed maximum-speed mixture was 17 per cent coal gas in air as against 17.2 per cent calculated; the observed upper-limit mixture contained 24.3 per cent coal gas in air, as against 26.2 per cent calculated. These values agree fairly closely, particularly in the maximum-speed mixture.

The relative speed of the limit mixtures, as has already been stated, is about 0.65 ft. per sec. for all gases. The speed of flame movement in maximum-speed mixtures may be calculated as follows: The law of speeds (as it applies to maximum-speed mixtures) may be restated: "A maximum-speed complex gas-air mixture is composed of a mixture of maximum-speed simple gas-air mixtures." The speed in the maximum-speed mixture is calculated by the formula:

$$S = \frac{a \cdot S_a + b \cdot S_b + c \cdot S_c + \ . \ . \ .}{a + b + c + \ . \ . \ .}, \tag{2}$$

where S is the speed of the complex gas-air mixture required; a, b, c, etc. are the amounts present of each simple maximum-speed mixture with air; and S_a, S_b, S_c, etc. are the speeds in those mixtures. Values of S_a, S_b, S_c are given in the table on page 270, or may be read from Fig. 61. Values of a, b, c, etc. must be calculated. It has been shown that the maximum-speed mixture contained 17.2 per cent coal gas and, therefore, the proportion of each individual gas in the maximum-speed gas-air mixture will be:

PERCENTAGE OF

CO = 9.6 per cent of 17.2 = 1.65 per cent CO

H_2 = 49.2 per cent of 17.2 = 8.41 per cent H_2

C_2H_4 = 3.7 per cent of 17.2 = 0.64 per cent C_2H_4

CH_4 = 30.9 per cent of 17.2 = 5.31 per cent CH_4

C_2H_6 = 3.0 per cent of 17.2 = 0.52 per cent C_2H_6

N_2 = 3.6 per cent of 17.2 = 0.67 per cent N_2

17.20 per cent

Since the maximum-speed mixture may be considered as a proportional mixture of maximum-speed individual gas-air mixtures, then, 100 volumes of the maximum-speed mixture of CO contains 50 volumes of CO (see table on p. 270) and the coal-gas mixture will contain $1.65 \times {}^{100}\!/_{50}$, or $a = 3.30$ volumes of the CO maximum-speed mixture. Similarly, b (for H_2) $= \dfrac{8.41 \times 100}{38.5}$ $= 21.8$ volumes of H_2 maximum-speed mixture.

c (for C_2H_4) $= \dfrac{0.64 \times 100}{7.1} = 9.0$ volumes of C_2H_4 maximum-speed mixture

d (for CH_4) $= \dfrac{0.64 \times 100}{9.9} = 53.6$ volumes of CH_4 maximum-speed mixture

e (for C_2H_6) $= \dfrac{0.52 \times 100}{6.3} = 8.3$ volumes of C_2H_6 maximum-speed mixture

Then substituting in Eq. (2):

$$S = \frac{3.3 \times 4.1 + 21.8 \times 15.9 + 9.0 \times 4.66 + 53.6 \times 2.2 + 8.3 \times 2.8}{3.3 + 21.8 + 9.0 + 53.6 + 8.3}$$

= 5.7 ft. per sec., the speed of the maximum-speed mixture. This calculated maximum flame speed of 5.7 ft. per sec. agrees fairly well with the observed value of 5.2 ft. per sec. shown in Fig. 62.

In the case of the coal gas just cited, the small amount of nitrogen present had practically no effect on the flame speed. This is not so in the limit mixtures of a gas containing a large proportion of nitrogen, as, for example, in the producer gas having the flame speeds shown in Fig. 62. This gas had the following analysis:

	Per Cent
CO_2	5.0
CO	21.3
H_2	12.6
CH_4	3.1
N_2	58.0
	100.0

or, on a N_2- and CO_2-free basis,

	Per Cent
CO	57.6
H_2	34.0
CH_4	8.4
	100.0

A gas containing these percentages of CO, H_2, and CH_4 (with no N_2 or CO_2) has as its maximum-speed mixture the composition 34.7 per cent of combustible gas (CO + H_2 + CH_4 in the above proportions) and 65.3 per cent of air.

$$L_x = \frac{57.6 + 34.0 + 8.4}{\frac{57.6}{50} + \frac{34.0}{38.5} + \frac{8.4}{9.9}} = 34.7 \text{ per cent.}$$

If nitrogen is added to this mixture to give a gas of the same composition as the producer gas (assuming the CO_2 content to have the same effect as the same amount of nitrogen), the result is a mixture containing 21.7 per cent inflammable gas. (The amount of N_2 required is $34.7 \times {}^{63}\!/\!_{37} = 59.1$ volumes of N_2 per 34.7 volumes of inflammable gas. The amount of inflammable gas in 100 volumes of the mixture then is $34.7 \times \frac{100}{100 + 59.1} =$ 21.7 per cent.) This calculated value of 21.7 per cent agrees not at all with the observed value (Fig. 62) of 51 per cent, and impresses the fact that calculations of the above nature are of practically no value for gases containing large amounts of inert matter, such as nitrogen.

Velocity and Characteristics of Explosion Waves.—If the third type of flame propagation (rapid explosion wave) takes place, the rate of propagation is very much faster than the speed of uniform flame movement and ranges from 6500 to 10,000 ft. per sec. The following facts regarding the behavior of explosion waves have been established from the extensive researches of Dixon[11] and others.

1. The velocity of the explosion wave increases rapidly up to a maximum, after which the rate becomes practically uniform.
2. Increase in pressure slightly increases the velocity.
3. Increase of temperature slows down the velocity slightly.
4. Inert gases act as a diluent and slow down the velocity (large excesses of the combustible gases usually act similarly).
5. Combustion usually is incomplete in an explosion wave.
6. Water vapor, up to 5 per cent in the gaseous mixture, increases the velocity of the explosion of CO. Further increases in the moisture content of the gas mixture retard the velocity.

Radiation from Flames.—Although most of the heat from flames is utilized by convection or conduction, a considerable proportion of it also may be transferred by radiation. Thus, in the ordinary Bunsen flame, from 12 to 18 per cent of the total heat of combustion is radiated away and consequently does not permit the attainment of maximum flame temperature. The amount of heat radiated from flames depends to a large extent on the degree of luminosity. Non-luminous flames lose a certain amount of heat by radiation (usually as much as 10 per cent), this radiation being intimately connected with the chemical reactions which are taking place. Luminous flames, in addition to the heat radiated in the process of chemical reaction (as in non-luminous flames), radiate an extra amount of heat by virtue of the highly incandescent particles of carbon* which impart the luminosity to the flame. The subsequent discussion summarizes certain information regarding radiation from non-luminous flames and points out a number of features of the extra radiation from flames of high luminosity.

The first-recorded measurements of radiation from flames were made by von Helmholtz[12] in 1890. He determined the absolute values of radiation from flames of various gases when burning in air. Helmholtz found that the amount of radiation per unit of gas burned depended upon the kind of burner used, the size of the flame, the amount of air mixed with the gas before burning, the temperature of the gases before combustion, and the kind of gas used. His values for the amount of energy radiated, as per cent of the latent heat in the gas, ranged from 5 to 15 per cent. The amount of radiation was found to decrease with the aeration, especially with luminous flames, and also to decrease with increased temperature of the mixture before combustion. The amount of radiation from different gases varied over rather wide limits. Upon these experimental results Helmholtz based his theory of flame radiation, according to which the radiation comes only from the products of combustion. Helmholtz postulated that the formation of a mol of water vapor in a flame will radiate a definite amount of energy, which may be determined by measurements of the radiation of a hydrogen flame. Furthermore, the formation of a mol of CO_2 was considered to result in a definite amount of energy, which could be determined by measuring the radiation from a CO flame. On this basis a flame of methane, for instance, which burns with the formation of 1 mol of CO_2 and 2 mols of water per mol of methane, will radiate

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* The presence of carbon particles in luminous flames is demonstrated by the fact that a beam of light is polarized by passing through the flame; this is characteristic of the effect of solid surfaces.

an amount of energy equal to the radiation for the combustion of a mol of CO plus that for 2 mols of H_2. Comparison of such calculated values with experimental data should check the validity of the theory, and the experimental basis upon which Helmholtz's theory rests is the agreement between his calculated and observed values, as shown in Table LXV.

TABLE LXV

VON HELMHOLTZ'S DATA ON RADIATION FROM FLAMES

Gas	Mols product formed per mol gas burned		Radiation in relative units			Per cent of theoretical air*
			Luminous	Non-luminous		
	H_2O	CO_2	Observed	Observed	Calculated	
H_2	1	0	.....	74	...	0
CO	0	1	.....	177	...	0
CH_4	2	1	391	327	325	55
C_2H_4	2	2	1140†	510	502	141
Illuminating	1.2	0.5	310	181	179	30

* Calculated from data as given for non-luminous flame. Not given in original table.

† Given elsewhere as 1200.

This table, copied from Helmholtz' paper, gives, in the first three columns, the gas used and the products of combustion. The fourth and fifth columns give the observed radiation, in relative units. The CO flame gave 2.4 times the amount of radiant energy that the H_2 flame did. The sixth column gives the radiation, also in the same relative units, calculated for the gas in question according to the theory. The last column, which does not appear in the original work, gives the per cent of theoretically required air which is premixed with the gas before burning in the free flame. These values were calculated from information given elsewhere in the published account. From the data given, it is impossible to find the radiation corresponding to any one given aeration, with the exception of zero primary air (which corresponds in many cases to a luminous flame). Complete aeration data are given only for illuminating gas, and in this case the relative radiation for a 6-mm. flame varies from 100 to 30 over a range of primary aeration of from about 30 to 150 per cent. Hence the selection of any particular value must, of necessity, be decidedly arbitrary. On the other hand, the amount of primary aeration is so important a factor that all comparisons should be based upon a comparable aeration. In Helmholtz' experiments, however, as is evident from the table, the per cent of primary aeration varies from 0 to 141 per cent. There is also an unknown amount of secondary aeration. The criterion of aeration upon which the values chosen were determined seems to be the production of a flame with just enough air so that it is free from incandescent carbon particles. Evidently, this value will vary from zero primary air, in the case of H_2 and CO, to a very high value for acetylene, for instance. In addition, this criterion will depend upon the type of burner used and the air conditions and currents, as well as the kind of gas. Such a criterion of comparable aeration, therefore, seems to be without any basis whatsoever. If, on the other hand, the comparison of calculated and observed values as above is based on aeration as per cent of theoretical air, the only reasonable basis, it is evident that the agreement observed by Helmholtz will no longer be obtained.

Callendar[13] measured radiation from flames of coal gas burning in a modified Méker burner. He found that large flames radiated more per unit area than small ones, and that the radiation increased with the depth of flame behind the unit area according to an exponential law. Callendar also found that in the Méker burner flames the radiation dropped off, as more primary air was admitted to the base of the burner. His values of radiation were about 15 per cent of the latent heat in the gas burned for flames from 1- to 4-in. diameter.

These results, coupled with and interpreted in the light of data obtained in a more recent investigation,[14] show that the radiation of heat from a non-luminous flame, although definite in amount under given conditions, varies greatly with the condition of the flame. The radiation drops off as the amount of primary air is increased beyond that for complete combustion, and with less air the radiation is also decreased. For methane, with theoretical air, the radiation is 54,500 B.t.u. per lb.-mol of gas burned and with 140 per cent of theoretical air the radiation is decreased to 49,000 B.t.u. per lb.-mol. For CO the radiation for these flame compositions are 12,200 and 11,340 B.t.u. respectively, and for illuminating gas the values are 28,300 and 27,000 B.t.u. These values do not agree with the theory propounded by von Helmholtz. The above values show that the radiation from flames of different gases burned with air varies greatly; the radiation with theoretical air, expressed as percentages of the latent heat which is lost by radiation, has the values of 14.9 per cent for methane, 10.4 per cent for CO, and 13.8 per cent for illuminating gas.

There appears to be no simple relationship between the amount of radiation and the composition of the gas or the products of combustion. The radiation per unit area for a flame is found to increase with increasing depth of flame, up to 40 in. in depth, according to an exponential law of the form $R = K(1 - e^{kx})$, where K and k are constants depending upon the composition of the flame and the temperature of the gases before combustion. The radiation from a flame under conditions of premixing is decreased as the air for combustion is preheated.*

In luminous flames, as was intimated above, in addition to the radiation due to chemical reaction (electronic rearrangement), common to both luminous and non-luminous flames, there is also the radiation from the incandescent carbon particles which have been deposited in the flame by cracking. The generally accepted theory is that the carbon particles act as carriers or forwarders of the heat. These tiny bits of carbon, deposited in the flame by cracking, are instantly heated by the hot gases of the flame to a state of incandescence, in which condition they radiate heat to the cooler surroundings, exactly the same as would a solid body (see Appendix II). While in non-luminous flames experimental results seem to indicate that the radiation is to a large extent independent of the temperature, a consideration of the above statement shows that this is obviously not true for that part of the radiation from the luminous flame which is due to the carbon particles, for if the radiation from the carbon particles is the same as that from solid bodies, which there is every reason to believe is the case, then the radiation from the carbon particles will follow the Stefan-Boltzman law,

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* These is considerable evidence by Paschen and others indicating that the radiation from the non-luminous flame is due to thermal radiation of the hot products of combustion, carbon dioxide and water vapor. However, this theory offers no satisfactory explanation as to the cause of decreasing radiation when the air for combustion is preheated.

and will be a function of the fourth power of the absolute temperature of the flame.

While it is generally known that luminous flames radiate more strongly than non-luminous flames, the existing data on radiation from luminous flames are comparatively meager. Von Helmholtz also measured the radiation from luminous as well as from non-luminous flames. His results show that the radiation from the luminous flame is from 20 to 120 per cent greater than that from non-luminous flames. The greatest increase (120 per cent) noted in his work is with ethylene, which is the least saturated hydrocarbon he used. This is interesting, because, with the same amount of air, the less saturated the hydrocarbon the greater will be the cracking and deposition of carbon particles, thus giving proof of the importance of the incandescent carbon particles in the radiation from the luminous flame.

Investigations have been carried out in Germany by Lent and Thomas,[15] in which they used a blast-furnace gas flame to which benzol had been added to make the flame strongly luminous. The conditions were so regulated as not to increase the flame temperature. They found the radiation from the flame in this experiment was four times that from a non-luminous flame, and was practically the same as from a black body at that temperature.

In general, it can be said that the radiation from luminous flames amounts to from 10 to 40 per cent of the potential heat of the gas, depending upon the degree of luminosity and the flame temperature. This shows the importance of radiation wherever luminous flames come into consideration.

Flame Temperature.—The question of the temperatures of the flames obtained in the combustion of gases is highly important. The thermal efficiency of a process is measured by the ratio $\frac{T_2 - T_1}{T_2}$, where T_2 is the highest temperature encountered (*i.e.*, the flame temperature) and T_1 is the lower, or stack, temperature. The difference between T_2 and T_1 represents the amount of heat used up in the process and T_2 is a measure of the heat input. It can be seen that the higher the flame temperature (T_2) the greater becomes the possible efficiency, because T_1 usually is fixed. Furthermore, the rate of heating an object is proportional to the difference in temperature between the flame and the object. Thus a gas having a flame temperature of

3300°F. (1816°C.) will heat an object to 2600°F. (1427°C.) much faster than a gas with a flame temperature of 2800°F. (1538°C.).

It is very difficult to measure the actual temperature of a flame, or even to calculate its temperature with any degree of accuracy. Consequently, it is customary to speak of the "theoretical flame temperature," which is the temperature that would be obtained if the combustion took place instantaneously, completely, and without loss of any heat to the surroundings; that is, if all the heat of combustion were used to heat the gaseous products of combustion. Thus the theoretical flame temperature is found by dividing the sum of the potential and sensible heat of a given quantity of gas and air mixture by the volume of the products of combustion of that same quantity of gas and air times its specific heat:

Theoretical Flame Temperature =

$$\frac{\text{(Heat of Combustion + Sensible Heat in Fuel + Sensible Heat in Air)}}{\text{(Total Quantity of Combustion Products)} \times \text{(Their Mean Specific Heats)}}.$$

The flame temperatures evaluated in this way are, of course, in error, since *all* the heat of combustion is not available for heating the products of combustion, but the calculated temperatures are nevertheless of great value for the purpose of comparing different fuels and for determining the effects of excess air, the preheating of gas and air, etc.

It will be seen from the above formula, that, in order to calculate the theoretical flame temperature, it is necessary to know the heat of combustion of the fuel gas, the amount of sensible heat in the gas and air, and the specific heats of the various gases involved. Heat of combustion has already been discussed in Chap. X, page 204, and the heats of combustion of the important simple gases have been given for various conditions in Table LII, on page 205. The heats of combustion and the various combustion constants for representative manufactured and natural gases are given in Table LXVI. In flame-temperature calculations it is always necessary to use the low or net heating value of the gas, because the latent heat of condensation of the water vapor (which is included in the high or gross heating value) cannot be used to heat the products of combustion on account of the high temperature. In other words, only sensible or potential heat can be used to heat the gases to the flame temperature, since the latent heat of vaporization cannot be utilized unless the gases are cooled below 212°F. (100°C.).

TABLE LXVI

COMPOSITION AND COMBUSTION CONSTANTS OF REPRESENTATIVE MANUFACTURED AND NATURAL GASES

	Methane (CH_4), per cent	Ethane (C_2H_6), per cent	Propane (C_3H_8), per cent	Ethylene (C_2H_4), per cent	Carbon monoxide (CO), per cent	Carbon dioxide (CO_2), per cent	Hydrogen (H_2), per cent	Oxygen (O_2), per cent	Nitrogen (N_2), per cent	B.t.u. per cu. ft., high (gross) 60°F., 30 in. Hg, satd. H_2O	B.t.u. per cu. ft., low (net) 60°F., 30 in. Hg, satd. with H_2O	Cubic feet air required per cu. ft. gas	Cubic feet CO_2 per cu. ft. of gas burned	Cubic feet H_2O per cu. ft. of gas burned	Cubic feet N_2 per cu. ft. of gas burned	High (gross) B.t.u. per cu. ft. gas-air mixture	Theoretical flame temperature, degrees Fahrenheit
Coal gas	34.0	....	....	6.6	9.0	1.1	47.0	....	2.3	634	560	5.50	0.573	1.282	4.36	97.5	3910
Coke-oven gas	28.5	....	....	2.9	5.1	1.4	57.4	0.5	4.2	536	476	4.65	0.420	1.205	4.71	95.0	3430
Coke-oven gas	33.9	....	....	5.2	6.1	2.6	47.9	0.6	3.7	600	538	5.28	0.529	1.260	4.21	95.6	3860
Blue water gas	....	....	....	....	43.4	3.5	51.8	....	1.3	310	285	2.28	0.469	0.518	1.81	94.5	4167
Carburetted water gas	14.8	....	....	12.8	33.9	1.5	35.2	....	1.8	578	529	4.85	0.758	0.904	3.85	98.8	4090
Oil gas	27.0	....	....	2.7	10.6	2.8	53.5	....	3.4	516	461	4.25	0.458	1.129	3.39	98.3	3970
Producer gas	2.6	....	....	0.4	22.0	5.7	10.5	....	58.8	136	128	1.08	0.311	0.165	1.44	65.3	3050
Blast-furnace gas	....	....	....	....	26.2	13.0	3.2	....	57.6	93	91.6	0.70	0.392	0.032	0.55	54.7	2660
Natural gas, Follansbee, W. Va.	....	31.8	77.7	....	....	....	....	....	0.5	2469	2268	23.80	2.967	4.062	18.81	99.6	3840
Natural gas, Follansbee residual	....	79.4	20.0	....	....	....	....	....	0.6	1868	1711	17.97	2.188	3.182	14.21	98.5	3830
Natural gas, McKean County, Pennsylvania	32.3	67.0	....	....	....	....	....	....	0.7	1482	1350	14.25	1.663	2.756	11.29	97.3	3770
Natural gas, Sandusky, Ohio	83.5	12.5	. .			0.2	.	.	3.8	1047	946	10.04	1.087	2.045	7.98	94.7	3740

Notes.—1. Heating values calculated from values given in Table LII, p. 205.

2. Products of combustion and theoretical flame temperatures figured with theoretical air (21 per cent O_2, 79 per cent N_2).

3. Theoretical flame temperatures not corrected for dissociation.

4. Analysis of 600 B.t.u. coke-oven gas and blue water gas from "Combustion," Am. Gas Assoc. (1924).

5. Coal gas, coke-oven gas, carburetted water gas, and oil gas contain small amounts of benzene and oxygen.

6. Cu. ft. of air required, cu. ft. of products of combustion, and B.t.u. of gas-air mixture based on use of theoretical air.

Specific heats in general, were discussed in Chap. X (p. 207). From the chart given (Fig. 49) it is possible to read directly the heat required to raise a mol of the common gases to any desired temperature. Heating values of gases, however, are usually given in terms of B.t.u. per cu. ft. and the use of Fig. 49 requires the calculation of the number of mols of gas involved.

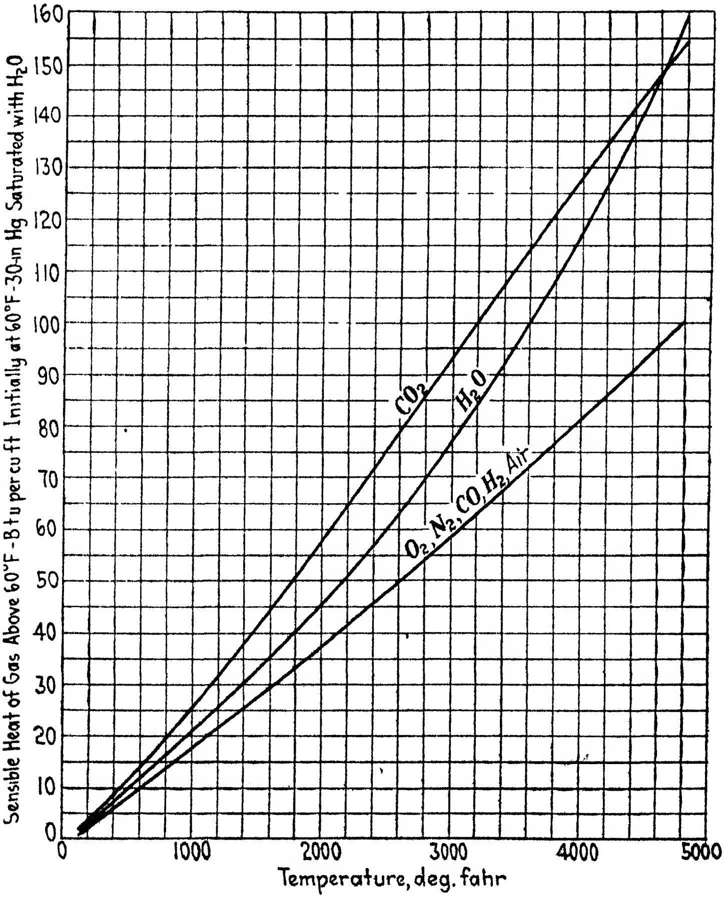

FIG. 63.—Sensible-heat contents of CO_2, H_2O (vapor), N_2, O_2, air, and CO above 60°F. The accuracy of values above 3800°F. is doubtful. Use equation on p. 207 for more accurate values for H_2.

It, obviously, would be much simpler for flame-temperature calculation if the total heats were given in terms of B.t.u. per cu. ft., and Fig. 63 has been drawn up on this basis. For example, suppose it is desired to calculate the theoretical flame temperature (with theoretical air) of CO, the heating value of which has been given in Table LII as 317.1 B.t.u. per cu. ft. at 60°F., 30 in. Hg pressure, and saturated with water vapor (*i.e.*, on the same basis as the total heats given in Fig. 63). Then assuming that the CO and the air for combustion are both at 60°F. initially, the problem resolves itself into a question of how high the amount of heat developed by the combustion can heat the gaseous products of combustion (given in Table LII as 1 cu. ft. of CO_2 and 1.88 cu. ft. of N_2 per cu. ft. of CO burned). This problem is solved most readily by trial and error: First assume a flame temperature of 4500°F. (2482°C.); the total heat for the CO_2 at this temperature, read from Fig. 63, is 144.5 B.t.u. and for N_2 is 93 B.t.u. Multiplying 93 by the number of cubic feet (93×1.88) gives 174.8 B.t.u., and the total heat in the products of combustion of the CO is found to be $174.8 + 144.5 = 319.3$, or, in other words, *more* than the heating value of the original CO; evidently, therefore, the assumed flame temperature is too high. Repeating this process for 4475°F., the sensible heat in the CO_2 is found to be 143.5 B.t.u., and in the N_2 is $1.88 \times 92.2 = 173.5$ B.t.u., the sum of these being $143.5 + 173.5 = 317.0$ B.t.u., or substantially equal to the heat of combustion of the CO. The theoretical flame temperature of CO, therefore, is 4475°F. (2469°C.).

Suppose, for a second example of this method of calculation, that it is desired to calculate the theoretical temperature of the flame of the producer gas given in Table LXVI, when this gas and the air for combustion are both preheated to 1000°F. and 25 per cent excess air is used. The theoretical flame temperature under these conditions is the temperature at which the sensible heat in the products of combustion (*i.e.*, the sensible heat in 0.311 cu. ft. of CO_2 + 0.165 cu. ft. of H_2O + 1.44 cu. ft. N_2 from the theoretical air + 0.25×1.08 cu. ft. of excess air, the values given in Table LXVI) just equals the sum of the potential heat in the producer gas (128 B.t.u. per cu. ft.) plus the sensible heat at 1000°F. in 1.25 times the air theoretically required for combustion, plus the sensible heat in the preheated producer gas. The heat input, then, using the above heating value (128 B.t.u.) and reading sensible heats from Fig. 63, is:

Potential Heat = 128 B.t.u. per cu. ft. at 60° F., 30 in. Hg, H_2O saturated.

$$\text{Sensible Heat in Air} = \frac{\text{Air Used}}{\text{Air Required}} \times \text{Cubic Feet Theoretical}$$

Air per Cu. Ft. Producer Gas × Sensible Heat at 1000°F.
$= 1.25 \times 1.08 \times 17.2 = 23.2$ B.t.u.

Sensible Heat in Producer Gas:

(*a*) In CO, H_2, and N_2 =
(Cu. ft. CO + Cu. ft. H_2 + Cu. ft. N_2) per Cu. ft. Producer Gas × Sensible Heat at 1000°F.
$= (0.22 + 0.105 + 0.588)\ (17.2)$
$= 15.7$ B.t.u.

(*b*) In* CO_2, CH_4, and C_2H_4 =
(Cu. ft. CO_2 + Cu. ft. CH_4 + Cu. ft. C_2H_4) per Cu. ft. Producer Gas × Sensible Heat at 1000°F.
$= (0.057 + 0.026 + 0.004)\ (25.2) = 2.2$ B.t.u.

Total Heat Input $= 128 + 23.2 + 15.7 + 2.2$
$= 169.1$ B.t.u. per cu. ft. of producer gas at 60°F., 30 in. Hg, and H_2O sat.

The next step is to assume a temperature, say 3600°F., and to determine if the sensible heat in the products of combustion at that temperature is equal to the heat input. The products of combustion (from Table LXVI) per cu. ft. producer gas are found to be: 0.311 cu. ft. CO_2, 0.165 cu. ft. H_2O, 1.44 cu. ft. N_2 from the theoretical air, and $0.25 \times 1.08 = 0.27$ cu. ft. of excess air. The sensible heat at 3600°F. (reading from Fig 63) is found to be:

$$\begin{array}{lll} CO_2: & 0.311 \times 114 & = 35.5 \text{ B.t.u.} \\ H_2O: & 0.165 \times 100 & = 16.5 \text{ B.t.u.} \\ N_2 + \text{air}: & (1.44 + 0.27) \times 72 & = \underline{123.0} \text{ B.t.u.} \\ & & \quad 175.0 \text{ B.t.u.} \end{array}$$

which is greater than the heat input, and therefore 3600°F. is above the theoretical flame temperature under these conditions.

Repeating the above process for 3500°F., the total heat is found to be:

$$\begin{array}{lll} CO_2: & 0.311 \times 110.5 & = 34.4 \text{ B.t.u.} \\ H_2O: & 0.165 \times 95.7 & = 15.8 \text{ B.t.u.} \\ N_2 + \text{air}: & 1.71 \times 69.7 & = \underline{119.0} \text{ B.t.u.} \\ & & \quad 169.2 \text{ B.t.u.} \end{array}$$

---

* Assuming heat capacity of CH_4 and C_2H_4 the same as CO_2, which is sufficiently accurate for this calculation and for the temperature range in question.

which agrees well enough with the heat input, and the theoretical flame temperature under the specified conditions therefore is 3500°F.

Where it is desired to know the theoretical flame temperature of a gas under a number of different conditions it is convenient to have a chart such as Fig. 64, which shows the total heat in

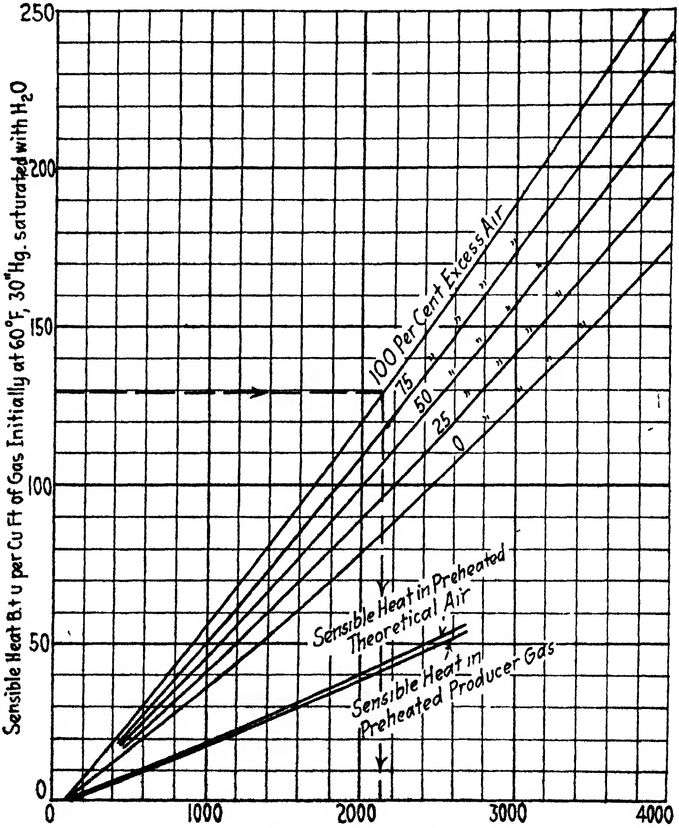

FIG. 64.—Sensible heat in products of combustion of producer gas above 60°F.

the products of combustion of the typical producer gas given in Table LXVI, when the gas is burned with theoretical air, and with 25, 50, 75, and 100 per cent excess air. With the aid of figures of this kind it is possible (knowing the heat input) to read the flame temperature directly. For example, with producer gas and air initially at 60°F., the heat input equals the low heating value of the gas, and, reading across from 128 B.t.u. (the low heating value of the gas), the theoretical flame temperature, when 100 per cent excess air is used, is seen to be about 2130°F.

Theoretical flame temperatures determined in this way are plotted as a function of the amount of air in Fig. 65. The top curve shows the theoretical flame temperature when both gas and air are preheated to 1000°F. (538°C.). It will be seen that the flame temperature is decreased nearly 900°F. (500°C.) by increasing the excess air to 100 per cent. The curves in the lower part of Fig. 65 show the much lower flame temperatures obtained when the degree of preheating the gas and air is decreased.

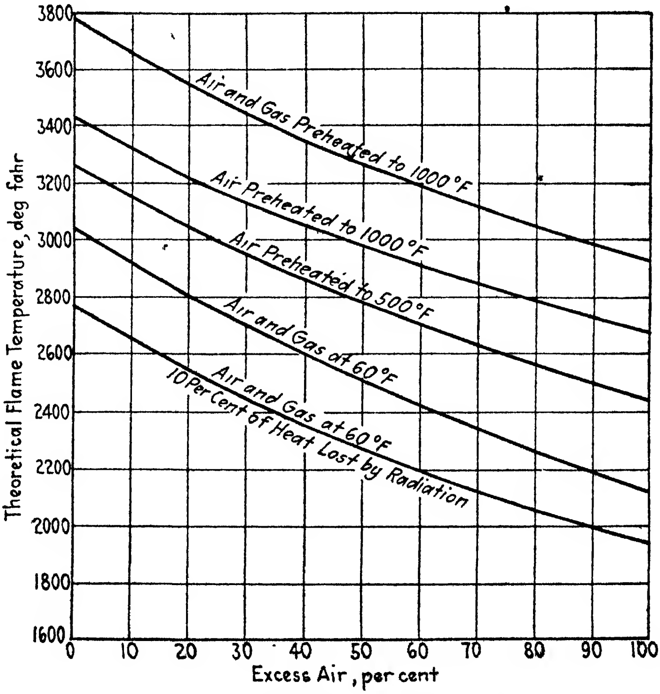

FIG. 65.—Theoretical flame temperatures (neglecting dissociation) of producer gas burned with varying amounts of excess air.

The theoretical flame temperatures in Table LXVI and the calculations just made are based on the assumption that all the heat of combustion of the gas is available for heating the products of combustion and will appear as sensible heat in them. This is not actually the case, and, although very valuable for comparing different fuels for specific purposes, the calculated theoretical flame temperature is always higher than is obtained in practice. This discrepancy between the calculated and the actual flame temperatures is the result of the following factors: (1) CO_2 and water vapor, at temperatures above 3000°F. (1650°C.), are slightly dissociated according to the reactions

$$2CO_2 \rightarrow 2CO + O_2 - 244{,}800 \text{ B.t.u.}$$
$$2H_2O \rightarrow 2H_2 + O_2 - 246{,}200 \text{ B.t.u.}$$

and, as indicated by the equations, the dissociation absorbs heat; (2) the specific heats of gases are not known with accuracy at temperatures above 3800°F. (2093°C.); (3) from 5 to 20 per cent of the heat of combustion may be radiated directly from the flame; (4) in the combustion of gases, the heat is not generated instantaneously, and during the heating period a certain amount of heat is lost by conduction or convection. Our present knowledge of radiation from flames is still too meager to permit accurate prediction of the effect of this factor on flame temperature, and the amount of heat lost by conduction during combustion obviously will vary greatly with local conditions. It is possible, however, to estimate closely the amount of heat consumed by dissociation of CO_2 and H_2O, and by taking this factor into account it is possible to calculate the flame temperature more accurately than by the preceding method. This calculation is explained in the following paragraphs.

The degree to which CO_2 and water vapor will dissociate under given conditions can be calculated from the equilibrium constants (K) for the reactions involved. Equations expressing the variation with temperature in equilibrium constants for the reactions

$$CO + \tfrac{1}{2}O_2 = CO_2 + 122{,}400 \text{ B.t.u.}$$
$$H_2 + \tfrac{1}{2}O_2 = H_2O + 123{,}100 \text{ B.t.u.}$$

have been given in Eqs. (46) and (47) of Chap. VIII. The per cent dissociation of CO_2 and H_2O according to the reverse of these two reactions is shown[16] in Fig. 66. The dissociation is given for 1 atmosphere partial pressure and for 0.1 atmosphere partial pressure. The latter values have been included since the partial pressures of CO_2 and H_2O in combustion products are about 0.1 atmosphere. For almost all combustion calculations involving the dissociation of these two gases it is sufficiently accurate to consider the partial pressures as 0.1 atmosphere.

The calculation of flame temperature, correcting for dissociation, then resolves itself into determining the temperature at which the sum of the sensible heats in the entering gas and air plus the potential heat in the fuel gas *minus* the heat which is absorbed by the dissociation (and therefore is not available for heating the flue gases) just equals the total sensible heat in the products of combustion. This temperature is easily determined with the aid of Fig. 67, which gives: (1) the heat absorbed by the dissociation of CO_2 and H_2O at 0.1 atmosphere partial pressure at temperatures from 3000 to 4600°F., (2) the sensible heat in the $CO_2 + CO + O_2$ resulting from the dissociation at each temperature, (3) the sensible heat in the $H_2O + H_2 + O_2$ resulting from the dissociation, and (4) the sensible heat in the nitrogen—all expressed as B.t.u. per cu. ft. of gas (*i.e.*, CO_2, H_2O, or N_2) initially measured at 60°F., 30 in. Hg pressure and saturated with water vapor. As an illustration of the use of Fig. 67, suppose it is desired to determine the flame temperature of CO, correcting for dissociation. It was shown in Table LII that the heating value of CO is 317.1 B.t.u. per cu. ft., that the products

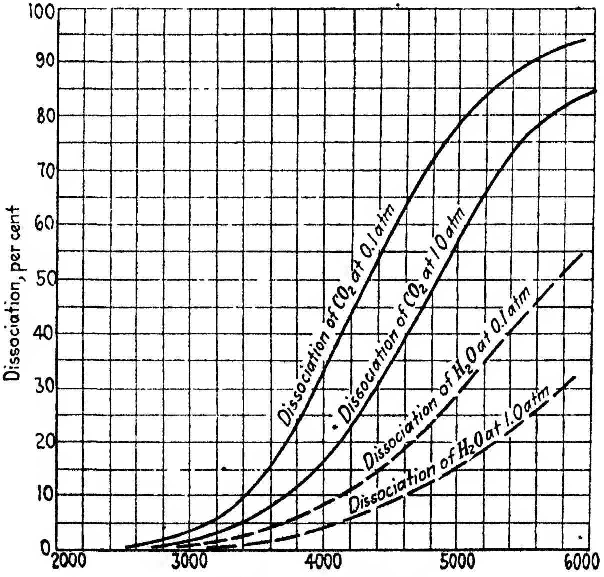

FIG. 66.—Dissociation of CO_2 and H_2O at high temperatures. (*Adapted from "Combustion," Am. Gas. Assn.* (1924).)

of combustion (per cu. ft. of CO) are 1.0 cu. ft. of CO_2 and 1.88 cu. ft. of N_2, and that the theoretical flame temperature, uncorrected for dissociation, was 4475°F.

As a first trial in determining the corrected flame temperature, assume 4000°F. At this temperature the heat input is 317.1 B.t.u. potential heat in the CO *minus* 91 B.t.u. absorbed in the dissociation of CO_2 (read from Fig. 67), or a total of 226.1 available B.t.u. The heat required for a flame temperature of 4000°F is 127 × 1 = 127 B.t.u. sensible heat in the $CO_2 + CO + O_2$ from the dissociation of CO_2 and 1.88 × 81 = 152 B.t.u. sensible heat in the nitrogen. The total heat required is 127 + 152 = 279 B.t.u., or more than the heat input; therefore the flame-temperature corrected for dissociation is below 4000°F.

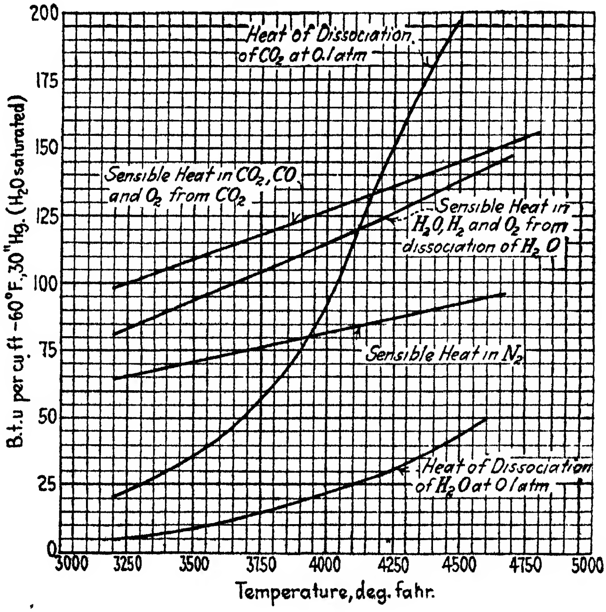

FIG. 67.—Heat of dissociation of CO_2 and H_2O, and sensible heat in $CO_2 + CO + O_2$ and $H_2O + H_2 + O_2$. (*Adapted from "Combustion," Am. Gas Assn.* (1924).)

Taking 3750°F. for the second trial, the heat input is:

Potential heat in CO	317.1 B.t.u.
Heat lost through dissociation	57.0 B.t.u.
Total	260.1 B.t.u.

The heat required is:

Sensible heat in CO_2 (+ CO + O_2 from dissociation)	118 B.t.u.
Sensible heat in N_2 = 1.88 × 76	143 B.t.u.
Total	261 B.t.u.

which agrees quite closely with the heat input, and the theoretical flame temperature of CO, *corrected for dissociation*, is, therefore, about 3750°F., as compared with an uncorrected temperature of 4500°F.

Figure 67 has been drawn up on the assumption that no excess air is present, since the presence of oxygen would decrease the dissociation. For strictly accurate calculations, when excess air is present it is necessary to recalculate the degree of dissociation of the CO_2 and H_2, taking into account the amount of oxygen in the gas. Since an excess of oxygen hinders the dissociation of CO_2 and H_2O, it is clear that higher actual flame temperatures are reached with a slight amount of excess air than with the theoretical amount of air.

There are three principal ways in which the maximum possible flame temperatures may be attained:

1. By liberating the heat of combustion as fast as possible (*i.e.*, by surface catalysis) in order to minimize conduction losses from the flame.

2. By using only a small amount of excess air, at the same time speeding up combustion by intimately mixing the combustible gas and air, and by surface catalysis. For example, in Fig. 65 it was shown that by increasing the excess air from 0 to 100 per cent the theoretical flame temperature* of a producer gas was lowered from about 3800°F. (2093°C.) to nearly 2900°F. (1593°C.), a decrease of about 25 per cent.

3. The flame temperature may also be increased by increasing the sensible heat in the gas-air mixture. For example, Fig. 65 showed that the theoretical flame temperature* of a producer gas, burned with 25 per cent excess air, was increased from 2570°F. (1510°C.) to about 3500°F. (1927°C.) by preheating the gas and air to 1000°F. (538°C.).

In practice, the most common (and usually unintentional) way in which flame temperatures are lowered is by the use of amounts of air largely in excess of the amount required for com-

* Neglecting dissociation.

plete combustion. A reduction of flame temperature by this means results in: (1) a loss of sensible heat out the stack, due to the greater volume of outgoing gases, and (2) a further loss in heat, since these stack gases usually leave the furnace at a higher temperature (which is due to the less efficient absorption of heat within the furnace when a large amount of excess air is used). Such dilution of the burning gases usually represents an unnecessary loss.

There are a few cases, however, where dilution with an inert gas may be sound economically. Fuel oil, for example, is often atomized and sprayed into a furnace by means of a jet of steam (see p. 477). Although this steam results in the dilution of the furnace gases, it is often the most satisfactory and economical method of atomization. There are also cases where dilution is necessary in order to protect the product from being overheated, but such dilution should be accomplished as outlined later (see Chap. XVI, on the recirculation of flue gases).

In many cases the flame temperature does not give a good visualization of the difference between combustible gases or show clearly the tremendous effect of dilution. This is particularly true in processes which require the absorption of large quantities of heat at a high temperature. The melting of metals, for example, and the decomposition of limestone both require heat at high temperatures. Consider the process of manufacture of lime from limestone, in which calcium carbonate ($CaCO_3$) is decomposed according to the reaction

$$CaCO_3 = CaO + CO_2 - 76{,}500 \text{ B.t.u.}$$

This decomposition occurs only (at ordinary atmospheric pressure) at temperatures above 1630°F. (888°C.).* If, for the sake of illustration, we consider this limestone being burned with methane (CH_4), the effect of excess air on the amount of heat available for the decomposition of the limestone and on the fuel consumption will be as shown in Table LXVII.

It will be noticed that increasing the air used for burning the methane from that theoretically required to 100 per cent excess air cuts the possible production from 2.83 to 1.42 mols of lime per mol of methane. Since the effect of dilution is not always obvious from an inspection of flame temperatures, particularly where heat must be absorbed at a high temperature, it is, therefore, often advisable to calculate the heat available per mol or

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* In actual lime burning decomposition begins below the temperature.

TABLE LXVII

EFFECT OF EXCESS AIR IN BURNING LIMESTONE WITH METHANE

Excess air, per cent	Mols of flue gas per mol of methane	B.t.u. required to heat products of combustion of 1 mol of methane to 1630°F.	Total heat liberated per mol of methane, B.t u	Heat available (per mol of CH_4) for decomposition of the limestone	Mols of CaO (56 lb. per mol) that can be decomposed per mol of methane (359 cu. ft. t S. C.)
0	10.52	129,400	346,000	216,600	2.83
50	15.28	183,200	346,000	162,800	2.13
100	20.04	237,200	346,000	108,800	1 42

per cu. ft. of flue gas at the temperature at which the heat is needed. The fifth column in Table LXVII shows the result of such a calculation, and the decreasing effectiveness of the products of combustion is at once apparent.

Regeneration and Recuperation.*—Many industrial processes require high temperatures for successful operation, these temperatures being necessary not only to make the process possible, but also to enable its completion within a reasonable length of time. Preheating the air (and in many cases the gas as well) offers the only means of attaining the necessary temperatures, except with certain rich gases which cannot be used on account of their high cost. Preheating is generally accomplished by means of regenerators or recuperators which add to the entering air a part of the sensible heat in the outgoing stack gases. Thus, in addition to raising the flame temperature, regeneration and recuperation also recover a certain amount of heat, an extremely important function in many industrial furnaces, where without this recovery, the heat actually applied usefully in the process would often be as low as 5 to 15 per cent of the total heat input.

In regenerative furnaces one or more chambers containing checkerbrick are provided at each end of the furnace. The flow of fuel gas and air, and the direction of flame travel, are reversed periodically. In one period the hot stack gases give up part of

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* For a more comprehensive treatment of the subject of regeneration and recuperation, the reader is referred to GROUME-GRJIMAILO,[17] DAMOUR,[18] "A Study of the Open Hearth,"[19] "A Study of the Blast Furnace,"[20] "Calco Handbook of Recuperation,"[21] and TRINKS "Industrial Furnaces."[22] GROUME-GRJIMAILO and TRINKS give a number of examples of recuperator and regenerator calculations.

their heat to the checkerbrick in the chambers at one end of the furnace, while the entering air, or air and gas, are preheated by passing over the checkerbrick at the opposite end, heated previously by the stack gases. Thus regeneration is a discontinuous process, in which brickwork serves alternately to pick up, and to act as a reservoir for heat from the flue gases, and then to pass this heat on to the entering air or gas.

Recuperation, on the other hand, is a continuous process, the heat from the hot flue gases being transmitted to the cold entering air through a partition which separates flues or ducts carrying the two gases (stack gas and air).

The use of preheated air, whether obtained by regeneration or by recuperation, in addition to making possible the attainment of high flame temperatures and adding materially to furnace efficiency by the return of heat to the cycle, possesses another decided advantage. All combustion reactions are greatly speeded up by increasing the temperature, and therefore with preheated air it is possible to burn fuel completely with much less excess air than would be required if cold air were used. This fact has much to do with the success of air preheaters in boiler furnaces (see Chap. XIV) and assumes an added significance in many metallurgical furnaces. Here oxidation of the metal at high temperatures (as evidenced by scaling) must be prevented, usually by burning the fuel with zero excess or a slight deficiency of air. In such cases preheated air permits efficient combustion without the necessity for using excess air in such quantities that excessive spoilage of the material being heated would result.

Regenerators are simple in construction and are almost always used to give the high degrees of preheat necessary, for example, in open-hearth furnaces. Moreover, there can be no leakage of stack gas into the preheated air ducts, as is possible in some recuperators. To offset the advantages of simplicity of construction and ability to deliver highly preheated air uncontaminated with stack gas, regenerators have the following disadvantages: (1) The process is discontinuous, and therefore the temperature of the preheated air varies; (2) relatively complicated valves are required for reversing the flow of flue gases and air; and (3) there is a labor cost involved in operating the reversing mechanisms.

Recuperators, being continuous in operation, give uniform degrees of preheat, require but half the number of chambers used in regenerators, require no reversing mechanism, and call for no extra labor for operation, thus overcoming many of the disadvantages inherent in an intermittent process like regeneration. Recuperators have certain disadvantages, however, among which are the following: (1) Special refractory forms are used in recuperators to deliver air preheated above 1400°F. (760°C.); these forms are costly and leaks are liable to develop; (2) only a small amount of surface (in comparison with regenerators) is available for heat transfer and large equipment is necessary; (3) smaller degrees of preheat are economically possible than with regenerators; and (4) they do not have the ability to carry overloads as well as regenerators. The disadvantage of leakage has been practically eliminated in heat-resisting metallic recuperators which can be used to deliver air preheated up to about 1200°F. (649°C.).

As previously stated, regenerators are mainly used for obtaining high degrees of preheat (usually above 1400°F., or 760°C.), while recuperators find their widest application in operations which require air preheated to lower temperatures.

Types of Regenerators.—The two largest applications of regenerators are in open-hearth furnaces and in blast-furnace stoves. For the open-hearth furnace, one or two regenerator chambers (one for air, and one for gas when lean gases, such as producer gas, are used) are provided at each end of the furnace, the regenerators, in general, being placed independent of the furnace body, so that molten metal leaking through cracks in the furnace bottom cannot fill the checkerbrick in the chambers.

When producer gas is used the air regenerator is made larger than the gas chamber on account of the large volumes of air required; the ratio of these two volumes is roughly 1.3:1.

The passage of hot stack gases is *down* through the checkerbrick, and the passage of cold air and fuel gas is *up* through the brickwork, in accordance with the principles of gas flow discussed in Appendix I.

The brickwork in the regenerator chambers generally is laid to form a number of vertical chimneys or passages, the exact arrangement and size of brick employed varying in different installations. The brick chosen should have a high heat capacity, and the amount of brickwork required to store the proper amount of heat under given conditions may be calculated empirically or rationally by the methods given by Groume-Grjimailo[17] and Trinks.[22] Ordinarily, in open-hearth furnace design the volume of checkerbrick in all four chambers will range from 130 to 150 cu. ft. per ton of furnace capacity. The general considerations involved in questions of heat transfer are discussed in Appendix II.

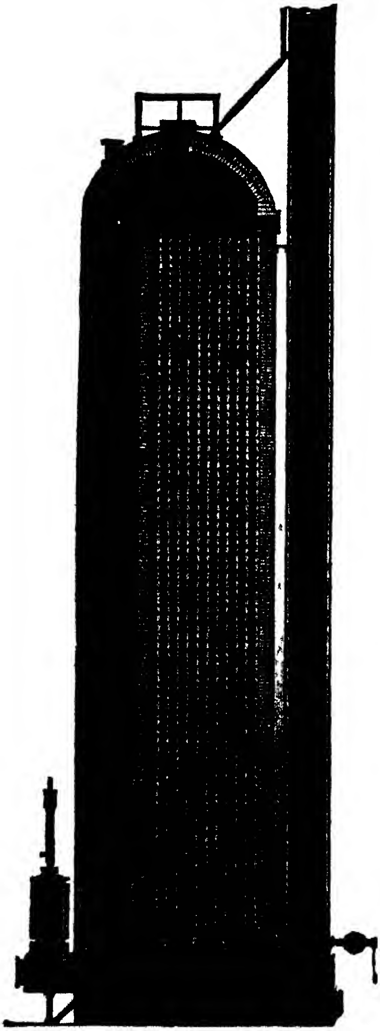

FIG. 68.—Foote patent blast-furnace stove. (*Harbison-Walker Refractories Company.*)

The second major use of regenerative-type furnaces is for blast-furnace stoves. These stoves (Fig. 68), are tall, fire-brick-lined chambers filled with checkerbrick. They are heated by burning in them a part of the furnace gases. When the checkerbrick has been heated sufficiently by this combustion, a blast of cold air is turned on (in the opposite direction to the flow of the hot flue gas), the heat previously absorbed by the brickwork serving to preheat the air. The air temperature gradually falls during the air blow, and, since a relatively constant temperature air blast is necessary in blast-furnace work, the hottest air at the start of the air run is cooled by admitting the proper amount of cold air. The cooling air is gradually reduced in amount until, at the end of the air run on a given stove, the blast furnace is using only the heated air, no cooling being necessary. Since the hot-air blast to the furnace must be continuous, several stoves are necessary, four generally being used.

The efficiency of the heat transmission from and to the checkerbrick is greatly influenced by the cleanliness. Thus, as pointed out by Walker, Lewis, and McAdams,[23] a number of tests have shown that the coefficient of heat transfer with clean brick surfaces is nearly twice as great as when the stoves are dirty. Much better heat transmission is possible when the gases are given a staggered path around the bricks.

Types of Recuperators.—In recuperators, two passages, divided by a wall, carry the countercurrently flowing streams of hot flue gas and the air which is being preheated; part of the sensible heat in the flue gases flows through the dividing wall and serves to preheat the entering air prior to combustion. For preheated air temperatures in excess of about 1400°F. (760°C.) it is necessary to construct the air and flue-gas ducts of refractory materials. Recuperators made of ordinary iron and steel give satisfactory service for flue-gas temperatures up to about 800 to 1000°F. (427 to 538°C.); for preheated-air temperatures[21] between 800 and 1200°F. (427 to 649°C.), however, it is necessary to use heat-resisting metals, since the ordinary irons and steels are subject to excessive oxidation and scaling over this range. Even the heat-resisting metals (such as calorized metal, *i.e.*, steel with an alloyed surface coating of aluminum) are not recommended for use with flue-gas temperatures in excess of 2200°F. (1204°C.), which would correspond to a preheated-air temperature of about 1200°F. (649°C.).

An example of the application of a modern tile recuperator is shown in Fig. 69, and a section of the recuperator tile work, clearly illustrating the flow of gas and air, is given in Fig 70. The flow of air is straight upward through small ducts, and the hot gases pass downward and across through horizontal channels in the tile. The gas flow is in accordance with the principles treated in Appendix I. and the process is practically counter-current. The tile are held in place laterally by slabs that direct the flow of the hot gases, and overlapping joints are used in order to diminish the tendency for leakage between the gas and air ducts. Even with these precautions, however, leakage will result if the recuperator is cooled frequently, and this type of heat-recovery equipment should not be used in intermittent operations where periodic cooling is necessary.

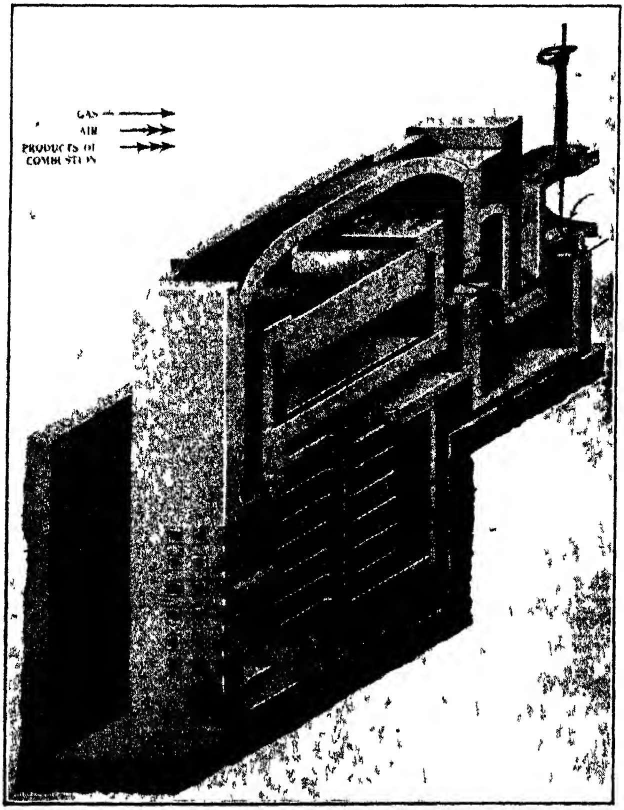

FIG. 69.—Section of typical Chapman-Stein forge furnace for small material.

The heat-absorbing capacity necessary for successful regenerators is not called for in recuperators, since the heat is transferred through the wall dividing the flue-gas and air ducts. These walls should, therefore, be made relatively thin. The main resistance to heat flow in recuperators, however, is in the relatively quiet film of flue gas on one side and in the film of air on the other side of the dividing partition (see Appendix II).

Fig. 70.—Chapman-Stein leakproof recuperator (patented).

From this it follows that considerable variations in the heat conductivity of the brick will have but little effect on the rate of heat transfer. In connection with the subject of heat transfer it is interesting to note that, by corrugating the walls of the flue-gas and air ducts on the side *away* from the partition through which heat is transmitted, it is possible to obtain from 30 to 50 per cent better heat transfer than if the corrugations are not used. The corrugations in the gas duct pick up heat from the gases and radiate it to the dividing wall between the two ducts, which, in turn, loses heat by radiation to the corrugated walls in the air passage. These corrugated walls can then give up more heat to the air.

Several types of metallic recuperators used for moderate degrees of preheat (up to about 700 or 800°F.), are described in Chap. XIV. The metallic recuperators used for more highly preheated air (made of heat-resisting metal) are practically always of the tubular type, on account of the lesser construction difficulties. These recuperators occupy less space than the tile type, and if properly constructed are practically free from leakage. Manufacturers' data[21] show that the conductivity of the thin *metal* walls is so great in comparison with that of the thick tile walls that more efficient heat transfer is possible with the metallic construction.

Heating Value of Gases.—Industrial gases are usually mixtures of CO and H_2, together with smaller quantities of methane, ethylene, etc. with varying amounts of inerts (N_2, CO_2, H_2O), as shown in the following diagram:

Commercial Gases

Base	Enriching Gases	Inert Material
H_2	CH_4	N_2
CO	C_2H_4	CO_2
	etc.	H_2O

The heating values of these elementary fuels under various conditions have been given in Table LII, and the distinction between gross (high) and net (low) heating value discussed on page 206. The composition and the heating values for representative industrial gases are shown in Table LXVI.

When a gas is richer than 300 to 325 B.t.u. (high value) per cu. ft. the gas contains appreciable quantities of hydrocarbons in amounts depending on the heating value of the gas. If the heating value of the gas is below 300 to 325 B.t.u. per cu. ft., the gas contains inerts, such as nitrogen; the lower the B.t.u. value the greater the quantity of inert gas.

Amount of Air Required for Combustion.—If the number of cubic feet of air required for the combustion* of 1 cu. ft. of gas is

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* See Chap. X, p. 233 for method of calculation.

plotted against the heat content of the various combustible gases in B.t.u. per cu. ft., it is found that the result is a fairly straight line (see Fig. 71). If the plot were constructed for the saturated hydrocarbons (methane, ethane, etc.) alone, the line would be almost perfectly straight, because the amount of heat developed when a unit volume of oxygen (or air) burns with the theoretical amount of any saturated hydrocarbon is approximately constant.* Thus, if the required amount of methane is burned with 1 cu. ft. of air, the heat evolved will be the same as when the proper amount of propane or any other saturated hydrocarbon is burned with 1 cu. ft. of air.

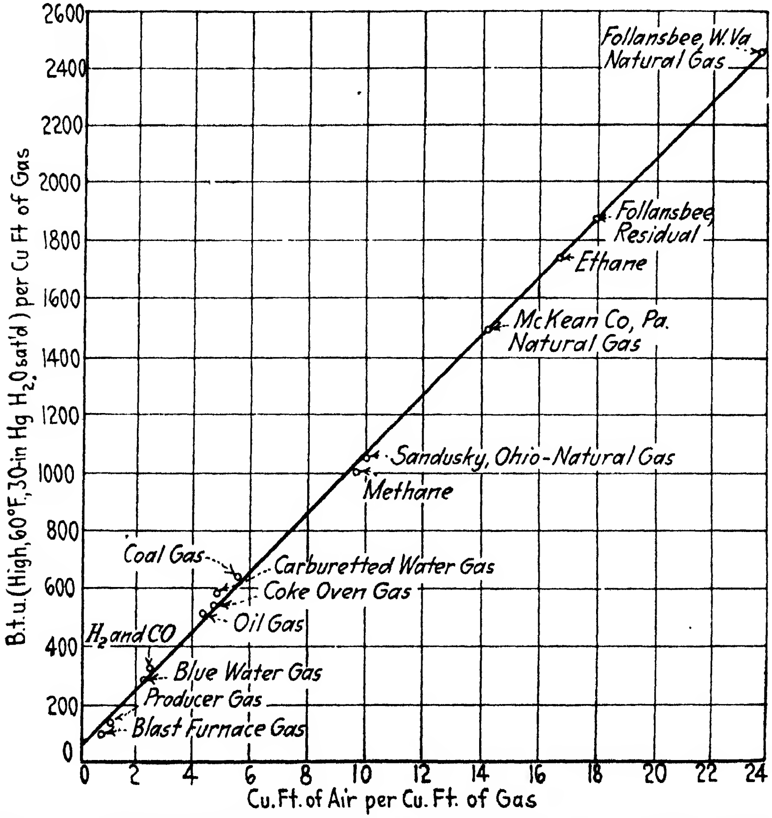

FIG. 71.—Cubic feet of air theoretically required to burn 1 cu. ft. of various gases.

Fortunately, the heating value for CO and H_2 and the amount of air required for their combustion are such that the points for

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* This may be deduced from Thornton's rule.

these gases also lie close to the extrapolated line for the hydrocarbons. Although not strictly correct in the case of the unsaturated hydrocarbons such as acetylene, ethylene, and propylene, these compounds are usually present in such small amounts in most commercial gases that the above relationship affords a simple and fairly accurate method for determining the amount of air required for the combustion of various gases from a knowledge of the B.t.u. content of the combustible gas.

The equation of the line in Fig. 71 is:*

$$\text{Cubic feet of air required to burn 1 cu. ft. of gas} = \frac{(\text{B.t.u.})^{1.06} - 52}{100}.$$

If a combustible gas and the theoretical amount of air for its combustion are mixed together, the B.t.u. per cu. ft. of this mixture can be calculated easily by using the above formula, namely:

$$\text{B.t.u. of 1 cu. ft. of mixture (air + gas)} = \frac{\text{B.t.u. of 1 cu. ft. of gas}}{x + 1} \text{ where } x = \text{cubic feet air}$$

required per cu. ft of gas.

$$\text{Then the B.t.u. per cu. ft. mixture (air + gas) is } \frac{\text{B t u. gas}}{\frac{(\text{B t u})^{1.06} - 52}{100} + 1} = \frac{100\ (\text{B.t.u.})}{(\text{B.t.u.})^{1.06} + 48}.$$

For most calculations it is sufficiently accurate to use this formula in a slightly modified form:

$$\text{B.t.u. per cu. ft. of mixture (air + gas)} = \frac{100\ (\text{B.t.u.})}{\text{B.t.u.} + 50}.$$

This formula indicates that the B.t.u. per cu. ft. of gas-air mixture is approximately constant over a wide variation of B.t.u. of the gas when the B.t.u. per cu. ft. of gas is high. Under these circumstances (using the approximate formula) the importance of the constant 50 drops out and the heating value of the gas-air mixture is approximately $\frac{100\ (\text{B.t.u}).}{\text{B.t.u.}} = 100$ B.t.u. per cu. ft. of mixture.

These conclusions are borne out by Fig. 72, in which the B.t.u. per cu. ft. of gas is plotted against the B.t.u. per cu. ft. of mixture (gas plus required air) for various commercial and natural gases. The figure shows that the B.t.u. of the mixture

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* A handy approximate formula is

$$\text{cu. ft. air per cu. ft. gas} \doteq \frac{\text{B.t.u. of gas}}{100} - 0.5$$

This is fairly accurate for high B.t.u. gases (such as manufactured or city gas) but is subject to error in the case of the leaner gases (such as producer gas and blast furnace gas).

remains approximately constant at a value of 92 to 100 B.t.u. per cu. ft. even though the heating value of the gas changes from about 300 to over 2000 B.t.u. per cu. ft.

The presence of inerts in a rich gas does not disturb this relationship. For example, a cubic foot of methane has a high heating value (at 60°F., 30 in. Hg and saturated with H_2O vapor) of 995 B.t.u. and, from Fig. 71, requires 9.52 cu. ft. of air for combustion. The B.t.u. per cu. ft. of the mixture therefore = $\frac{995}{9.52 + 1} = 94.5$. If the methane now is diluted with nitrogen until the gas is 50 per cent CH_4 and 50 per cent N_2 the heating value will be 497.5 B.t.u. and the gas will require 4.76 cu. ft. of air for combustion. The heating value of a cu. ft. of the final mixture of diluted gas and required air is $\frac{497.5}{4.76 + 1} = 86.5$ B.t.u., or a 50 per cent dilution of methane by nitrogen only caused a decrease of 9 per cent in heating value of the gas-air mixture.

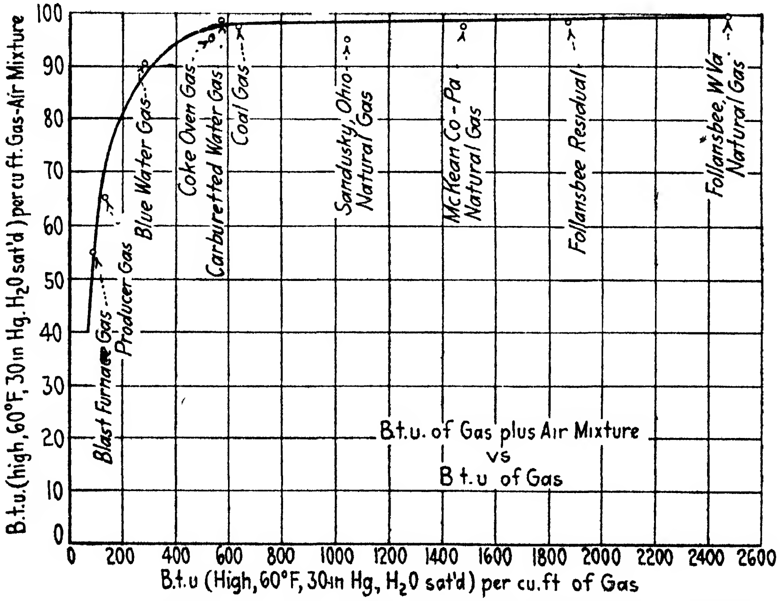

FIG. 72.—Heating value per cu. ft. of gas—theoretical air mixture for various gases.

In gases of lower heating value, dilution causes a greater decrease in the B.t.u. of the combustible mixture. For example.

1 cu. ft. of hydrogen has a high heating value (at 60°F., 30 in. Hg and saturated with H_2O) of 318.8 B.t.u. and requires 2.38 cu. ft. of air. The B.t.u. per cu. ft. of mixture (gas and required air) is $\frac{318.8}{2.38 + 1}$, or 94.5. If the hydrogen now is diluted with 50 per cent N_2, a cu. ft. of a mixture of diluted hydrogen and the air required for combustion will have a heating value of $\frac{159.4}{1.19 + 1}$, or 72.8 B.t.u. per cu. ft. (a decrease of about 23 per cent).

In general, a variation of 300 to 3000 (tenfold) in the B.t.u. per cu. ft. of combustible gas produces less than a 10 per cent change in the B.t.u. per cu. ft. of a mixture of combustible gas and the air required for combustion.

Surface Catalysis.*—In 1817, Sir Humphrey Davy, in his researches on flames, placed a heated platinum wire in a jar of gas-air mixture which contained sufficient excess gas to prevent explosion. The wire became red hot and continued so until the oxygen was practically all consumed. From the discovery of Davy and subsequent investigations by Graham, Dulong, Thenard, and others it was concluded: (1) that hot solids, particularly of the platinum group, speed up gaseous combustion at relatively low temperatures, and (2) that hydrogen is more susceptible to this action than other combustible gases.

Interest in the subject gradually died out, however, and no major researches were carried out in this field until 1902, when Bone[25,26] and his collaborators started their systematic investigation of the general phenomenon of surface combustion. They found:

1. The power of accelerating gaseous combustion is possessed by all surfaces (in differing degree) at temperatures below the ignition point.

2. This acceleration is dependent on the adsorption of the combustible gas, and probably the adsorption of the oxygen, by the surface. The surface becomes "activated" (and probably ionized) by this process. For example, photomicrographs of silver gauze taken before and after acting as a catalyst for the combustion of hydrogen at 752°F. (400°C.) showed that after acting as a catalyst for combustion the gauze was full of small, deep pits.

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* For a complete exposition of the subject of surface combustion and its technical applications, see Bibliography, references 24, 25, 26, and 27.

3. The surface itself becomes electrically charged during the process. Thus, the silver gauze just mentioned became negatively charged during the experiment.

4. At low temperatures, platinum, ferric oxide, and nickel oxide are the best catalysts for surface combustion.

Similar results were obtained at high temperatures, although it was found that above a red heat practically all surfaces have an equal catalyzing power. Thus, although platinum, ferric oxide, and nickel oxide are the best catalysts at low temperatures, at higher temperatures incandescent coal, fire brick, and other refractory materials are equally good and work out very satisfactorily.

Mechanism of Surface Combustion.—Catalytic combustion on surfaces depends primarily on the condensation or adsorption of one (or both) of the reacting gases by the incandescent surfaces. The condensed layers have the same effect as increasing the concentrations of the reacting gases and thus, according to the Law of Mass Action, increase the rate of combustion. This increase in the rate of combustion at the surface also raises the temperature of the surface, which again tends to increase further the rate of combustion at the surface. Through this pyramiding mechanism the rate of combustion at the surface increases until all of the gas burns there. It is known that hot incandescent surfaces emit large numbers of electrons and, according to Sir J. J. Thomson, the energizing of the field brought about in this way is also a cause for the catalytic effect.

The rate at which the gases combine is proportional to the pressure of the gas and air, provided the gases are in their combining volumes. When either gas or air is present in excess, the rate of combustion is, in nearly all cases, proportional to the partial pressure of the combustible gas.

Industrial Development of Surface Combustion.—In the commercial applications of surface combustion an explosive gas-air mixture in the theoretical proportions for complete combustion (or with slight excess air) is burned *without flame* in contact with an incandescent solid. The advantages claimed for this system are: (1) combustion is greatly accelerated and may be concentrated where heat is desired, (2) combustion is perfect with a minimum of excess air, (3) very high temperatures can be attained without the aid of elaborate regenerative devices, and

(4) owing to the large amount of radiant energy developed, heat transmission is very rapid.

An application of the surface combustion principle to a semi-muffle-type furnace is shown[28] in Fig. 73. Air and gas, mixed in the correct proportions, are introduced at both sides of the furnace, and the mixture plays directly on piles of refractory material, which become incandescent and on which surface combustion takes place. Heat is radiated from the incandescent material up to the furnace roof or arch and then back to the material which is to be heated. The burner shown is constricted at the tip in order to maintain sufficient gas velocity into the furnace to prevent back-firing. The metal burner shell extends back out of the furnace, where cooling fins are provided to dissipate the heat conducted back from the burner nozzle. If this heat (radiated to the nozzle from the refractory bed) were not dissipated, the temperature might rise sufficiently to cause pre-ignition of the gas-air mixture or to burn out the burner shell.

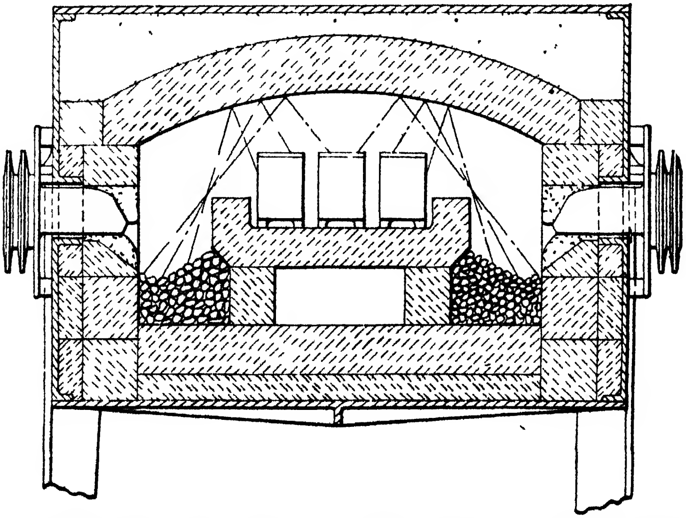

FIG. 73.—Semi-muffle type surface combustion furnace. (*Surface Combustion Company*)

Another method of application of this type of burner is shown in Fig. 74. In this case, instead of the gas-air mixture impinging on refractory material on the bottom of the furnace, surface combustion is obtained by causing part of the combustion to take place in a refractory-lined tunnel in the furnace wall, and then

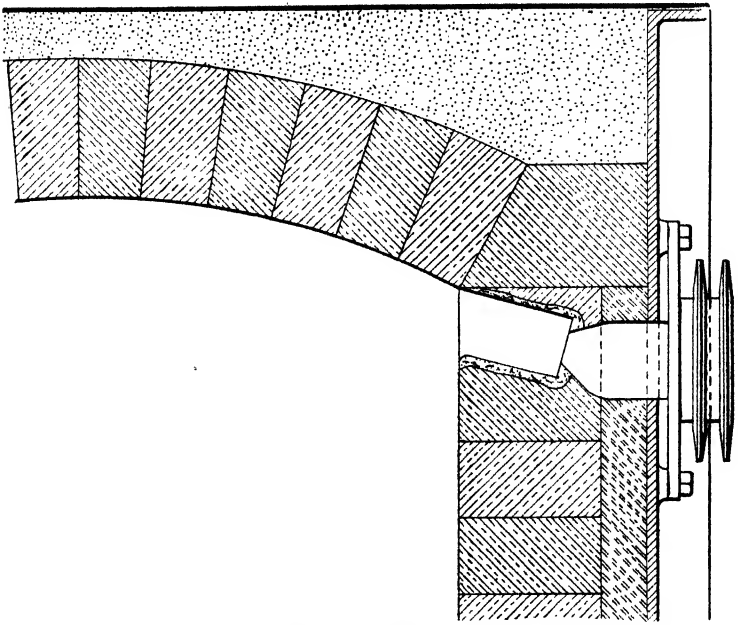

FIG. 74.—Tunnel burner of the arch type. (*Surface Combustion Company.*)

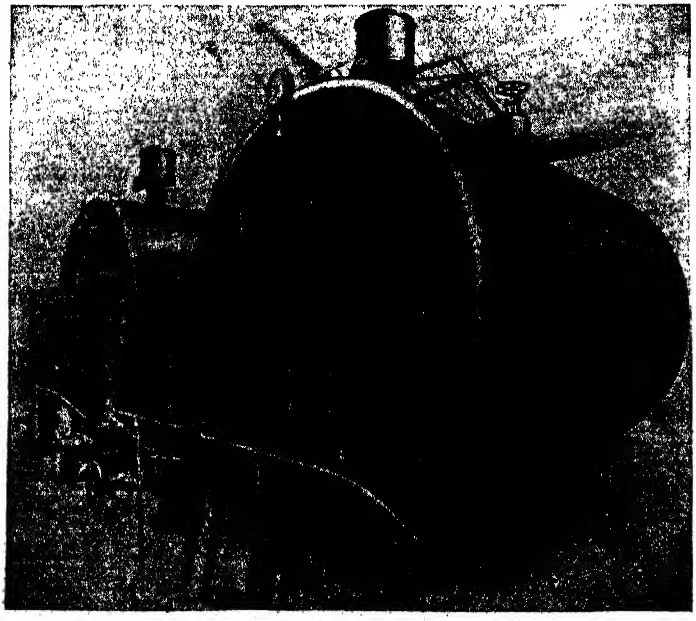

FIG. 75.—Front view of Bonecourt boiler. (*From W. A. Bone: "Coal and Its Scientific Uses, p. 469 (1922), by permission of Longmans Green Co.*)

combustion is completed as the hot gases wipe the furnace roof or arch. The roof becomes incandescent and radiates its heat to the material in the furnace. This same effect is obtained in upright furnaces by bringing the gas-air mixture in tangential to the side walls of the furnace, and allowing the combustion to go to completion as the hot gases swirl around in contact with the walls.

Surface combustion principles have also been successfully applied in the *Bonecourt boiler* Figures 75 and 76 show the front and back views of two surface combustion boilers (Bonecourt) each 10 ft. in diameter by 15 ft. long, fitted with 38 6-in.-diameter tubes. The tubes are packed with a rigid system of refractory blocks, on which the surface combustion takes place. Each section of refractory block fits into the next until the boiler tube is filled. Each tube has its own gas supply, and the gas-air mixture is drawn through the tubes by the suction of the fan shown at the rear of the furnace. The evaporation per linear ft. of each tube is shown diagrammatically in Fig. 77. The diagram shows the great speed of combustion possible even in a water-jacketed furnace. In this boiler, heat can be developed at the

FIG. 76.—Rear view of Bonecourt boiler (*From W A Bone "Coal and Its Scientific Uses," p 469 (1922), by permission of Longmans Green Co*)

rate of 700,000 B.t.u. per cu. ft. of combustion space per hr., or over 30 times as much as is developed per cu. ft. in an ordinary coal-fired boiler furnace.

In a single tube of this type 88 per cent of the net heat developed by the combustion was transmitted to the water, and in trials on a 10-tube experimental furnace efficiencies of 94.3 and 93.3 per cent were obtained. In the test giving 94.3 per cent efficiency the steam was at 100-lb. gage (338°F., or 170°C.), the flue gas leaving the boiler was at 446°F. (230°C.), and the temperature of the gases as they left the feed-water heater or economizer was 203°F. (95°C.). The rates of evaporation in the two tests were 21.6 and 33.9 lb. per sq. ft. of heating surface per hr., or at seven to eight times the rate for ordinary boilers.

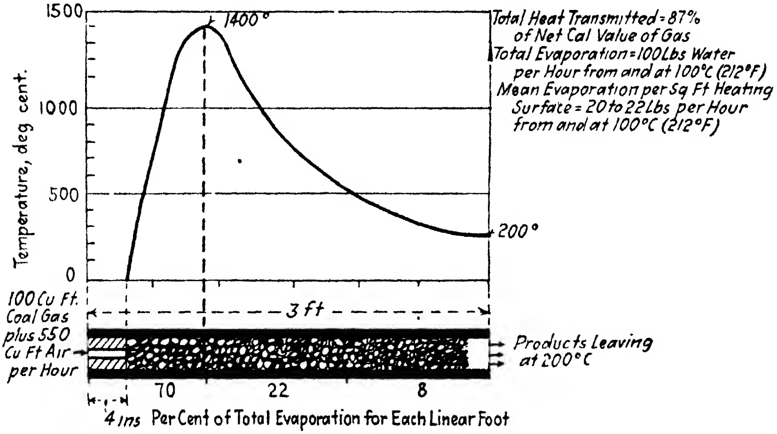

FIG. 77.—Evaporation per lin. ft. of Bonecourt boiler tube. (*From W. A. Bone: "Coal and Its Scientific Uses," p. 459 (1922), by permission of Longmans Green Co.*)

A 110-tube boiler, after five months' actual running, indicated the heat balance shown in Table LXVIII.

TABLE LXVIII

HEAT BALANCE ON BONECOURT BOILER

	PER CENT
Heat utilized	92.7
Heat lost:	
As sensible heat in flue gases	3.0
In unburned gas	0.0
By radiation, etc	4.3
	100.0
Rate of evaporation in pounds per sq. ft. of heating surface per hr	14.1

The Bonecourt boiler has the following advantages: (1) an exceptionally high thermal efficiency (the thermal efficiency of an ordinary gas-fired boiler ranges from 70 to 84 per cent), (2) a very high evaporative efficiency (an average of 20 lb. per sq. ft. of heating surface per hr. as compared with 10 lb. in a locomotive boiler and from 3 to 5 lb. in the average steam boiler), and (3) no formation of scale, due to the high rate of evaporation.

There is a heat loss, however, if coal must be gasified, and there is a question as to the strength of boiler tubes after this severe treatment, although apparently they stand up in a satisfactory manner.

GAS BURNERS

The main functions of any gas burner are: first, to proportion the air and the gas properly, and, second, to mix them and to deliver the mixture to the proper point. The way in which various domestic and industrial burners meet these requirements, and certain features of burner design, are discussed in the remainder of this chapter.

Domestic Gas Burners.*—Domestic gas burners are mainly of the atmospheric or Bunsen type, in which the momentum of the gas stream is utilized to inject part of the air necessary for combustion. This primary air and the gas are mixed together in passing through the burner tube, and the requisite secondary air is added as the gas burns at the outlet of the burner tube.

There are two parts to the flame obtained with atmospheric burners: (1) *the cone,* where the gas burns with a portion (primary) of the total air required for combustion, and (2) *the upper, or main, part of the flame,* where the balance of the gas burns with additional secondary air supplied from the outside by diffusion. In general, the gas burner must inject as much primary air as possible, be of such construction that the flame will neither blow away from the ports nor snap back, and give steady operation over as wide a variation in operating conditions as possible, *i.e.,* over a wide variation in gas pressure, gas density, and in capacity. To accomplish these purposes, and particularly to entrain the maximum amount of air, the energy of the gas stream must be conserved and transferred over to the stream of mixed primary air and gas. The maximum entrainment of air is possible only

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* This section on the design and efficiency of domestic gas burners is based largely on the results of investigations carried out at the U. S. Bureau of Standards.[29,30]

when the resistance to the flow of air into the burner and the back pressure within the burner itself are small. To accomplish this, the burner requires four main parts (shown for a typical burner in Fig. 78), the functions of which are as follows:

1. The *nozzle* or *orifice*, through which a fine, high-velocity stream of gas flows, should be of such design that, for a given volume, the gas has the maximum velocity possible.
2. The *injection tube*, which carries primary air and gas, should be of such design that eddy currents are prevented and the resistance to the flow of air and gas is a minimum.
3. The *burner ports*, through which the gas and primary air flow and at which combustion starts, should be of sufficient area to give minimum resistance to the flow of gas.

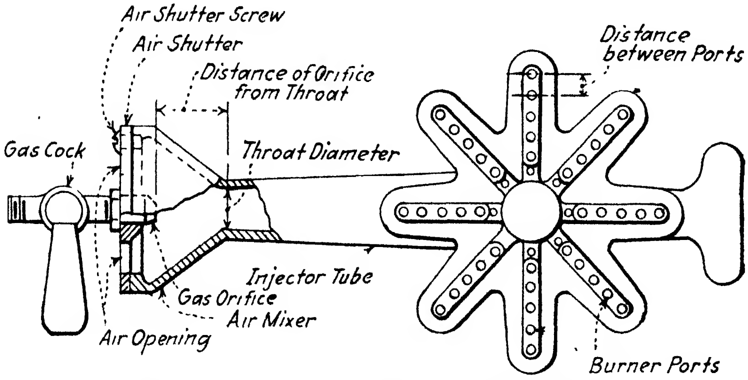

FIG. 78. Atmospheric-type domestic gas burner.

4. The *air shutter*, which controls the amount of air, should interpose only a slight resistance to the flow of primary air.

Orifice.—Two types of orifices are used, the sharp-edged and the channel type of orifice, shown in Fig. 79. The flow of gas through the sharp-edged orifices may be calculated from the formula $V \doteq 40.4\sqrt{\frac{H}{D}}$, where V = velocity of the gas in feet per sec., H = pressure of gas on the orifice in inches of water, and D = specific gravity of the gas (air = 1). This formula is a modification of the general formula given in Eq. 18 of Appendix I, using an orifice coefficient of contraction of 0.61. Channel types of orifices have a higher coefficient of contraction, namely 0.89, and consequently the equation for the flow of gas from a channel-type orifice is $V = 59\sqrt{\frac{H}{D}}$. The two following equations (which show the product of the velocity times the area of the orifice) give the rate of flow through sharp-edged and channel orifices where A = area of the orifice in square inches and Q = the cubic feet of gas flow per hr.

$$Q \text{ (sharp-edged orifice)} = 1010\,A\sqrt{\frac{H}{D}}$$

$$Q \text{ (channel orifice)} = 1474\,A\sqrt{\frac{H}{D}}$$

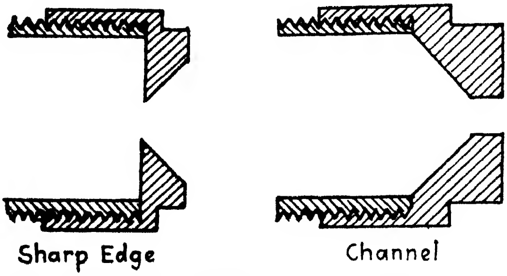

FIG. 79.—Sharp edge and channel type of orifice.

Since the sharp-edged type of orifice has the greater contraction (smaller coefficient of contraction), under conditions of domestic burner operation the velocity through a sharp-edged orifice will be greater than that through the same diameter of channel-type orifice for a given rate of flow. Consequently, the momentum of a gas stream is the greater when coming from a sharp-edged orifice, and therefore a sharp-edged orifice will inject more primary air into a gas burner than will a channel-type orifice.

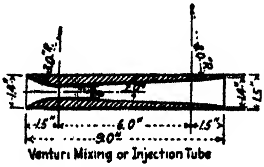

FIG. 80.—Most efficient type of atmospheric burner injecting tube.

Mixing or Injection Tube.—Since primary air is injected due to the transfer of momentum from the gas to the stream of gas plus primary-air mixture, the quantity of air injected into a burner will depend upon the velocity, the quantity, and the density of the gas. The venturi type of injection tube, in which the gas is gradually contracted and then allowed to expand naturally with a small amount of resistance, is best. This type is shown in Fig. 80. Even with this type of injection tube, due to friction, the entire energy of the gas stream is not conserved. The efficiency with which the air is injected may be measured by noting the ratio of the momentum of the gas stream to the momentum of the stream of gas and primary-air mixture. This ratio, R, is given by the following equation:

$$R = \frac{1658.5XK\sqrt{Hd}}{Q(r + d)\,(r + 1)},$$

where X = the cross-sectional area of the injection tube in square inches.

K = the contraction coefficient of the orifice (0.61 for sharp-edged and 0.89 for channel type).

H = the pressure of the gas on the orifice in inches of water.

d = specific gravity of the gas with reference to air.

Q = the rate of flow of gas in cubic feet. per hr.

r = the ratio of primary air to gas in the mixture passing through injection tube.

Each setting of any particular burner has a constant ratio R, which is characteristic of the burner itself. The injection efficiency of the burner is equal to $\frac{1}{R} \times 100$. The use of this formula is shown by the following example. In a pipe burner injecting gas at the rate of 50 cu. ft. per hr. the air-gas ratio was 5:1. This gas was coming in through a sharp-edged orifice and into an injection tube 1¼ in. in diameter, the specific gravity of the gas being 1.0, the gas pressure being 4.0 in. Under these circumstances, the ratio R is:

$$R = \frac{(1658.5)1.23 \times 0.61\sqrt{4 \times 1}}{50(5 + 1)\,(5 + 1)}$$

$$R = 1.38.$$

In other words, the injection efficiency of this burner is $\frac{100}{1.38}$ = 72.5 per cent. In order to get the maximum efficiency of air injection it has been found (see Fig. 80) that the mixing or injection tube should meet the following requirements:

1. As the gas flows through the injection tube the change in the inner contour of the tube should be gradual.
2. The approach should follow approximately a curvature which should have not less than a 3-in. radius for a ⅝-in. throat. Other sizes should be proportioned approximately the same.

3. The gas outlet angle should be about 2 deg.

Such an injection tube is somewhat hard to construct and consequently, while not quite so efficient, the burner that meets the following specifications (see Fig. 81) will serve quite well:

1. The approach should be on a 10.5-deg. angle.
2. The outlet angle should be approximately 3.5 deg.

In the injection tube the angle of approach is not so important as the outlet angle, but it is important that the gas stream be directed straight along the axis of the injection tube.

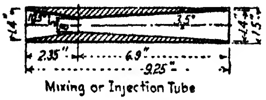

FIG. 81.—Satisfactory design of atmospheric burner injecting tube.

Ports.—The ports of the burner should be drilled so that the total area is equal to about 75 per cent of the cross-sectional area of the injection tube. If the rate of flame propagation of the gas is apt to change to any great extent, it is advisable to counterbore the burner ports in order to make them flare at the top.

The curves in Fig. 82 show for water gas the velocity of the air-gas mixture through ports of burners when the air-gas ratio is sufficient to cause the flames to blow from the ports. For example, with a 3:1 air-gas ratio the flame will blow out when the air-gas mixture has a velocity through the burner ports greater than 15 ft. per sec. If the velocity drops down below 3.5 ft. per sec. the mixture flashes back through the ports into the injection tube. With increasing air-gas ratio the limits at which the flame will blow from the ports or flash back into the injection tube come closer together.

Shutter.—In order that the resistance of the shutter which regulates the amount of primary air be sufficiently small, it is necessary that the velocity through the shutter should not be over 4 to 5 ft. per sec.

Efficiency of Domestic Gas Burners.—The efficiency of a domestic gas burner as measured by the ratio of the heat absorbed to the potential heat of the gas burned is very nearly constant with any given burner, even though the B.t.u. of the gas itself varies over a wide range. For example, the B.t.u. of a gas may vary from 300 to 600 B.t.u. per cu. ft., yet the quantity of the gas required for a given performance varies inversely as the heating value of the gas. This, of course, assumes that the gas burner is suitably adjusted for this change in heating value. In general, domestic gas burners operate with an efficiency of around 37 per cent.

Factors Affecting the Type of Flame.—The character of the flame depends greatly on the amount of primary air with which the gas is mixed before combustion starts. Combustion is extremely rapid when the gas is thoroughly mixed with the total amount of air required for combustion. For example, when the amount of air theoretically required is thoroughly mixed with city gas, combustion is so rapid that only 0.6 per cent CO is left unburned after $\frac{1}{70}$ sec.[31] If 8 per cent of excess air is thoroughly mixed with manufactured gas, combustion is complete within $\frac{1}{70}$ sec. and no unburned CO is present in the gas. Under these conditions (*i.e.*, when the air and gas are thoroughly premixed) the temperature of the furnace wall has no effect on the rate of combustion or, in other words, on the length of the flame. Figure 83 shows[31] that a change in wall temperature from 230 to 2012°F. (110 to 1100°C.) had no appreciable effect on the rate with which the gas burned. This is important in connection with the use of

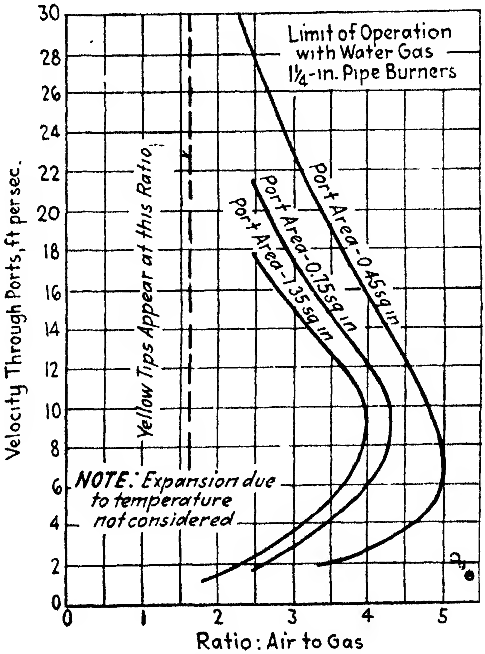

FIG. 82.—Blow out (upper limit) and snap back (lower limit) limits of velocity for water gas.

gas under widely different conditions, such as in a high-temperature heat-treating furnace or in a domestic boiler. In the first case the walls are very hot, and in the second case, of course, the walls are water cooled. If the gas and the air are premixed, the rate of combustion will be approximately the same in each case.

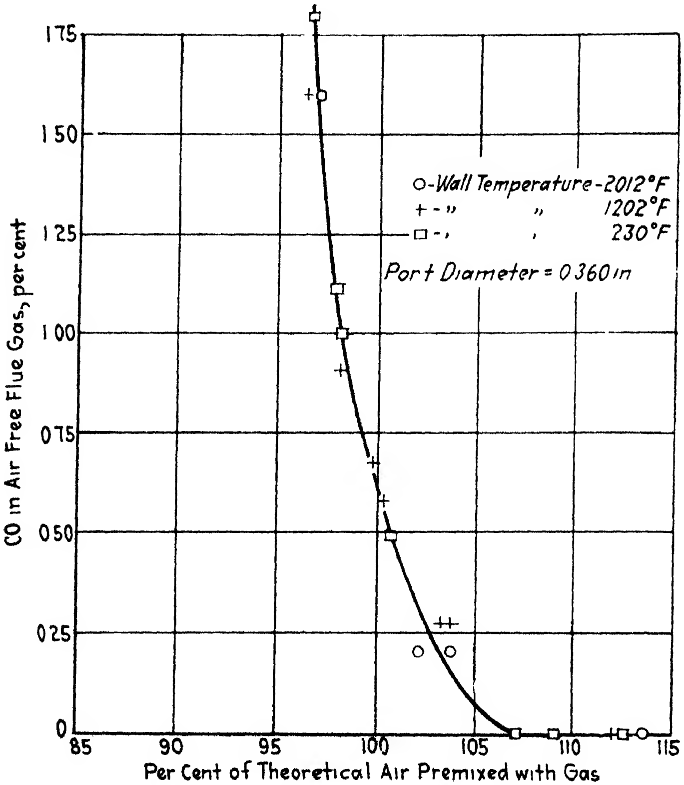

FIG 83 —Effect of primary aeration on the per cent of CO in air-free flue gas, showing negligible effect of wall temperature on speed of combustion

Effect of Primary Air on the Height of Cone.—It is important to note the effect of the amount of primary air injected into the burner upon the height of the cone and upon the height of the flame itself. The height of the cone goes down very rapidly as the amount of primary air increases up to 75 per cent of the theoretical air. This fact is illustrated[31,32] by Fig. 84, which shows that when 100 per cent of the theoretical air required for combustion (4.7 cu. ft. of air per cu. ft. for the particular gas used in these experiments, or, in other words, an air-gas ratio of 4.7

to 1) was premixed with the gas, the cone height is only about 1.5 in. under the conditions specified. With *half* the air premixed (air gas ratio = 2.35), on the other hand, the cone was 10 in. high. When only a small amount of primary air is injected into the injection tube of the burner (2 cu. ft. primary air to 1 cu. ft. of gas) the height of the cone is almost directly proportional to the velocity of the air-gas mixture. With a high degree of aeration (air-gas ratio 4:1), however, the height of the cone is small and is almost independent of the air-gas velocity. With low aeration (air-gas ratio 2:1) and constant gas velocity, the height of the cone is proportional to the diameter of the portholes. With greater aeration the port diameter does not seem to have so much effect.

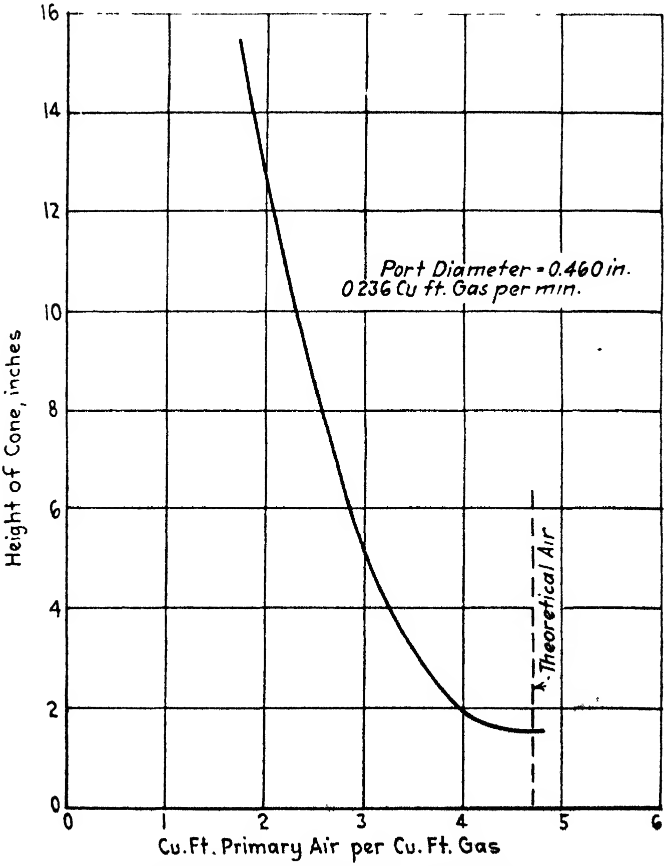

FIG. 84.—Effect of primary aeration on height of flame.

Effect of Primary Air on the Flame.—Increasing the velocity of the gas-air mixture through the ports or keeping the velocity constant and increasing the diameter of the port both tend to lengthen the total height of the *flame itself*. Also, the more primary air injected into the flame the shorter the flame. These points are shown[31,32] in Table LXIX.

TABLE LXIX

EFFECT OF PRIMARY AIR ON TOTAL HEIGHT OF FLAME

Amount of primary air, cubic feet air per cu. ft. gas	Height of flame when velocity of air and gas mixture is	
	3 ft. per sec., inches	10 ft. per sec., inches
0	17	28
1	6	19
2	2	14
3	1	10

Industrial Gas Burners.—The choice and the design of gas burners for industrial furnaces to a large extent depend on the specific purpose for which the heat from the combustion of the gas is to be employed and also on the B.t.u. of the gas used. The degree of mixing of gas and air is one of the major factors controlling the rapidity and the completeness of combustion—the greater the mixing the more rapid the burning of the gas. Thus for furnaces in which all the heat must be liberated quickly in a confined space, intimacy of mixing of gas and primary air is essential. Furthermore, as just shown for domestic gas burners, the larger the ratio of primary air to gas the shorter the flame. For other purposes, where a long, soft, lazy flame is desired and where the process of combustion must extend over considerable distances (as, for example, in reheating furnaces), mixing of gas and air must be retarded, and the type of burner selected with this end in view.

Industrial gas burners may be grouped arbitrarily as follows: (1) The port type, in which separate streams of gas and air are passed into and mix within the furnace; this type of burner is used mainly with highly preheated gas and air which, if mixed together outside the combustion chamber, would preignite—or with gas containing large quantities of dirt or tarry matter which would interfere with the operation of ordinary burners. (2) The aspirator type of burners, in which either gas or air (or both) is under pressure, and the momentum of one stream (gas or air) is used to induct the other. In this type of burner enough primary air or gas may be induced to make the theoretical mixture, or the momentum of an incomplete mixture may be used to induce the remainder of the necessary air, or one of several other combinations may be used. (3) In the third type, "proportioning machines," the gas and all the air necessary for combustion are mixed in a fan or chamber and then delivered as a blast to the various furnaces. This classification is decidedly arbitrary, and, as will be seen in the subsequent discussion of specific burners, a burner may embody features from more than one of the above classes.

Port-type Burners.—In burning producer gas in open-hearth furnaces, on account of the high flame temperatures required, (3000 to 3300°F.) it is necessary to preheat both air and gas to very high temperatures. The ports (openings through which gas and air are admitted to the furnace and by means of which the direction of the flame is controlled) are very important in good open-hearth design and are a frequent source of trouble in furnace operation. One or the other end of the furnace is always being subjected to an extremely rapid flow of hot gases containing bits of limestone, slag, etc., and this tends to wear away and destroy the alignment of the ports. In many furnaces there are two air ports and one gas port at each end of the furnace, although sometimes two gas and three air, two gas and two air, or one gas and one air port are used.

The number of ports used, the angles at which they enter the furnace, the point at which the air and gas streams are designed to converge, and the relative velocities of the two streams depend on the type of furnace. Where all the heat must be developed in a relatively restricted space, the streams of gas and air must be mixed as quickly as possible. Such mixing can be brought about: (1) by providing a relatively large number of small ports, and (2) by causing the gas and air ports to enter the furnace at such angles that the two streams converge soon after they leave the ports. In other types of furnaces (such as the open-hearth and certain kinds of heat-treating furnaces) the flame must extend over considerable distances, the temperature must be reasonably uniform throughout the furnace, and therefore the combustion must be gradual. This delayed or progressive combustion can be obtained by using a relatively small number of large ports, and by permitting only a gradual convergence of the gas and air as the two streams pass through the combustion space. In certain cases it may be advantageous to make the gas and air ports nearly parallel in order to delay mixing.

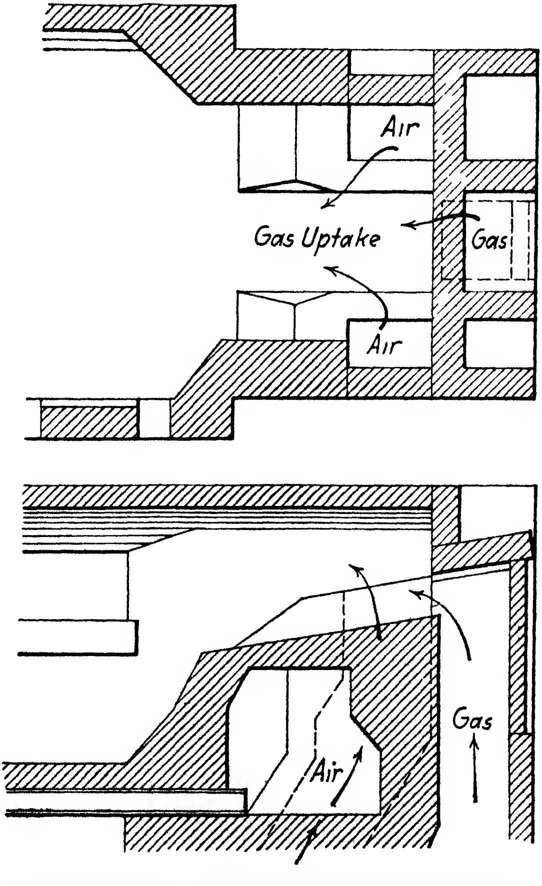

Fig 85.—Typical port construction of an open-hearth furnace.

In open-hearth furnaces the ports are inclined downward to make the flame come into contact with the molten bath; otherwise poor heat transfer would result and the roof would be burned out rapidly. The air usually enters above the gas and the air stream is deflected downward more than the gas stream.

The air thus is kept above the gas, preventing too severe heating of the roof and also preventing oxidation of the metal. The streams of gas and air meet from 3 to 5 ft. above the surface of the bath; if this point is below 3 ft., the gases are somewhat chilled by the bath, thus preventing maximum flame temperature, and if above 5 ft. the roof is rapidly burned out. In addition to the inclination of the ports, the flame is also directed by the use of high gas and air velocities. In European practice this varies from 40 to 60 ft. per sec. and in American furnaces the velocity through the ports is usually much higher—up to 160 ft. per sec. Details of the port construction of typical open-hearth furnaces are shown in Figs. 85 and 86.

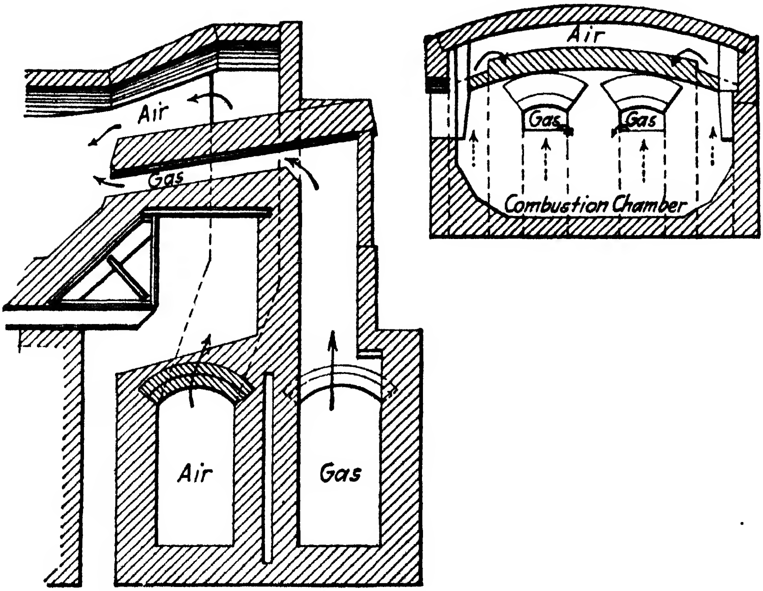

Fig. 86.—Typical port construction of an open-hearth furnace.

Ordinarily, ports are made up of silica brick, but on account of the rapidity with which this brickwork is burnt out a number of water-cooled ports have been installed. Details of a water-cooled port that has given good service at the Lackawanna Steel Company's plant are shown[33] in Fig. 87.

A number of modifications in open-hearth design have been tried out in an attempt to lengthen the life of a furnace and to shorten the time necessary for melting and refining a charge. One of the more recent developments is the Loftus system[34] shown in Fig. 88, which injects regenerated, hot air under pressure into the producer-gas line. The gas-air mixture thus produced passes through a nozzle into a mixing chamber, thereby inducing the secondary supply of air necessary for complete combustion. A supply of air (see figure) is withdrawn from the regenerator and passed through a special blower driven by a variable-speed motor. This air is then injected into the center of the gas stream. The chief advantages claimed for this type of port are: (1) By regulating the volume and the pressure of the injected air it is possible to obtain either a long or a short flame, (2) the travel of the gas may be altered to suit the conditions by rotating the port, (3) a port can be replaced easily and quickly.

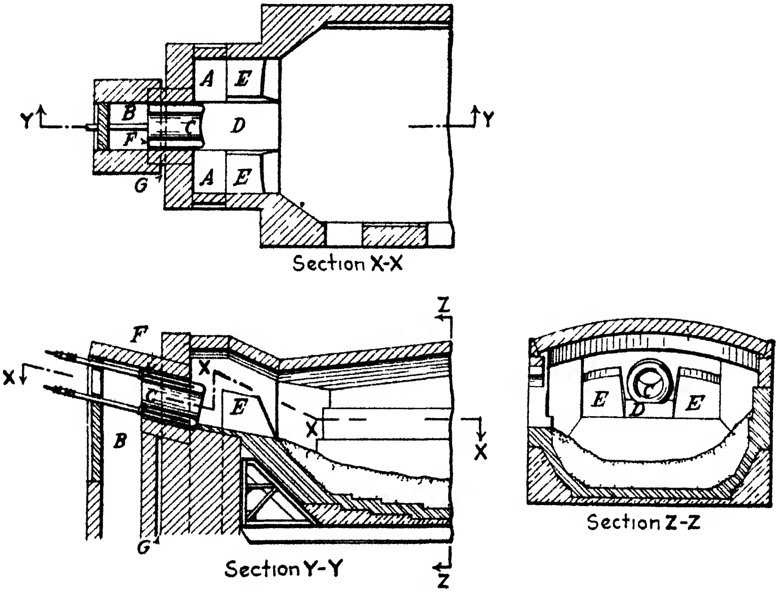

Fig 87 —Circular water-cooled port in use at plant of Lackawanna Steel Company *A*, air uptakes, *B*, gas uptake, *C*, port cooler, *D*, combustion chamber, *E*, blocks, *F*, ground silica brick packing, *G*, air space between gas and air uptakes

Aspirator-type Burners.—As previously stated, in aspirator-type burners either the gas or the air (or both) may be under pressure, and the momentum of one stream (usually the gas) is used to induce the desired flow of the other stream. There are a great many variations of this principle, as is shown below in the description of individual burners.

In burning blast-furnace gas, on account of its leanness (low B.t.u.), it is important to provide thorough mixing of the gas and air, and to provide long flame travel to allow complete combustion of the gas. Furthermore, gas coming directly from a blast furnace contains suspended ore, coke, and dust, which collect on heating surfaces and lower heat transmission. The gas also contains relatively high percentages of moisture, which increases the heat loss as sensible heat of the gases. The dust and moisture contents may be reduced by washing the gas in

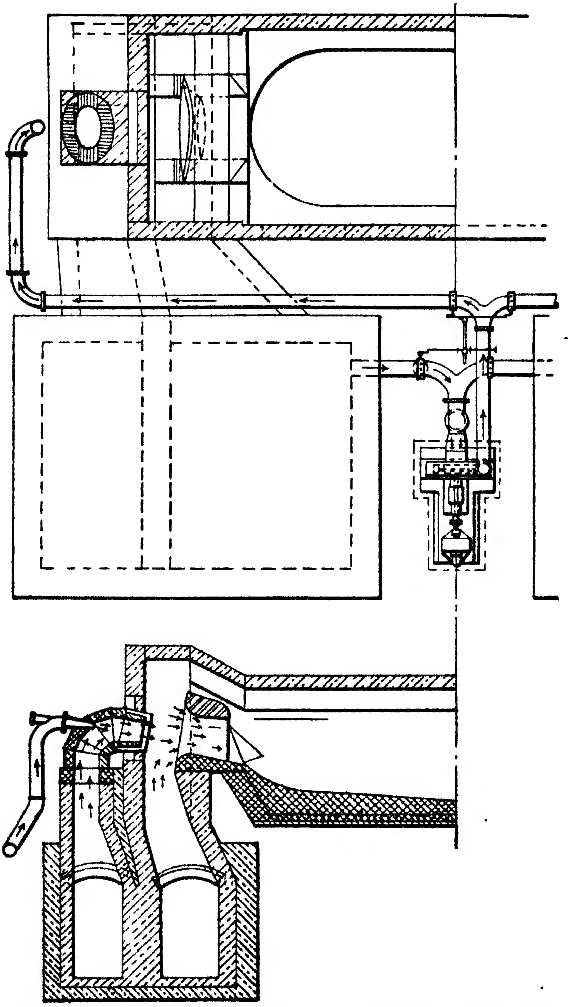

Fig. 88.—The Loftus port for open-hearth furnaces.[34]

a spray chamber and, in spite of the sensible heat lost (spraying cools the gases from about 300 to 80°F., 150 to 27°C.), the increased efficiency which may result from the use of a cleaner gas often offsets this loss.

Several types of burners used for burning blast-furnace gas are shown* in Figs. 89, 90, 91, and 92. The "common" type of burner illustrated in Fig. 89 has been the one most generally used. The gas is blown in through holes in the furnace front, and by its momentum induces a flow of air around the burner body. Mixing is very poor, however. Furthermore, the gas velocities are usually too high.

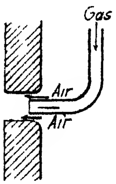

Fig. 89 — Common type blast-furnace gas burner

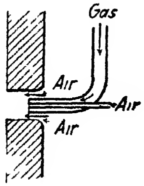

Fig. 90 — Bunsen-type burner

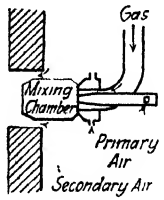

Fig 91 — Birkholtz-Terbeck burner for blast-furnace gas

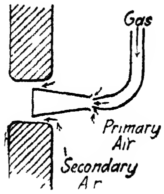

Fig 92 — Bradshaw burner for blast-furnace gas

The burner shown in Fig. 90 is of the Bunsen type, some of the air necessary for combustion entering through the small central tube shown, and the necessary secondary air being induced at the furnace front by the velocity of the gas-primary-air mixture. This burner is an improvement over the "common" type and gives better mixing.

The Birkholtz Terbeck burner (Fig. 91) has two places for the admission of primary air, and a supply of secondary air enters at the furnace front. As indicated by the diagram, part of the primary air is drawn in through a small pipe running through the central gas duct, and an additional supply of primary air is drawn in around the gas duct just at the entrance to the mixing chamber. Thus an annular ring of gas, with a layer of air both inside and outside, passes into the chamber, where the three streams (one of gas and two of primary air) are mixed. The

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* Figures 89 to 92 are adapted from Diehl,[35] who describes these and numerous other burners used for blast-furnace gas. Diehl also gives the results of an extensive series of tests on boilers and blast-furnace stoves fired with this gas.

momentum of this mixture then induces the necessary additional supply of secondary air.

The Bradshaw burner, shown diagrammatically in Fig. 92, utilizes the venturi principle to inject primary air. The gas duct is constricted at the venturi throat and the reduced pressure causes a certain amount of primary air to be drawn in. The gas and the air are then mixed in passing through the expansion chamber of the venturi and secondary air is drawn in as the mixture passes into the furnace. The expansion tube should be so designed that the velocity of the gas-air mixture in it is sufficiently greater than the rate of flame propagation to prevent back-firing.

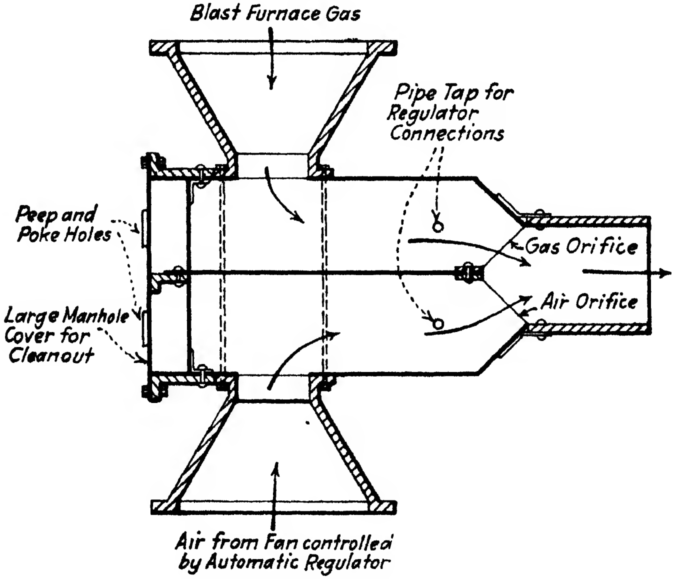

FIG. 93.—Steinbart automatic blast-furnace gas burner.

All of the above burners rely on manual operation to adjust the air and gas supply. Thus, when, for any reason, the gas pressure changes, the ratio of air to gas also changes to a certain extent. This, obviously, leads to changed (and usually lowered) efficiencies. The Steinbart burner shown in Fig. 93 has been used in an attempt to overcome this difficulty. The gas and air (both under pressure) pass through separate orifices at the front of the furnace, the orifices being so designed that if the pressures in the gas and air ducts are equal the proper amount of air will be admitted. A pressure regulator interconnected between the two ducts controls the speed of a forced-draft air fan, thus maintaining substantially equal pressures of gas and air. This method of regulation operates satisfactorily unless the orifices are eroded or become blocked up by dirt in the gas.

It should be emphasized again that the function of a gas burner is to admit the proper amount of gas and air (in the correct proportions) and to mix them so that the desired control of combustion can be maintained. The performance of all burners must be viewed in the light of these two requirements.

The burners of the aspirating type so far described have all been used primarily for blast-furnace gas. The requirements for a burner to be used with producer gas are somewhat different and will be discussed next. Producer gas (unless carefully cleaned) usually contains considerable quantities of tar and soot, and, as pointed out on page 318, port-type burners are ordinarily used. For many industrial purposes, however, particularly in small furnaces, the many advantages of raw producer gas make it highly desirable to have a burner of the aspirating type suitable for this gas, and several satisfactory burners have been developed.

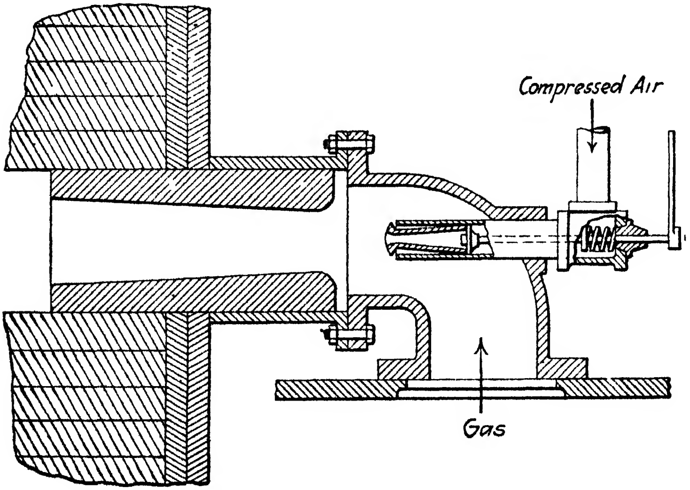

Fig. 94.—Flinn and Dreffein producer-gas burner.

In burning raw, cooled producer gas the flow of gas cannot be efficiently regulated by means of valves, since the tar in the gas will accumulate wherever there is a valve or gate. For this reason the necessary control of air and gas flow in raw, cooled producer-gas burners is secured by regulation of the air either through pressure or by the amount sent directly through an injecting nozzle (to induce the proper gas flow). In the Flinn-Dreffein burner (Fig. 94) the air supply is controlled by an air nozzle, which not only admits all the air necessary for combustion but also induces the flow of producer gas from the gas flue. This nozzle is so arranged that a jet of air (tending to induce gas flow from the gas main) can be sent through the small central tube, or a cone of air (which tends to retard gas flow) can be sent

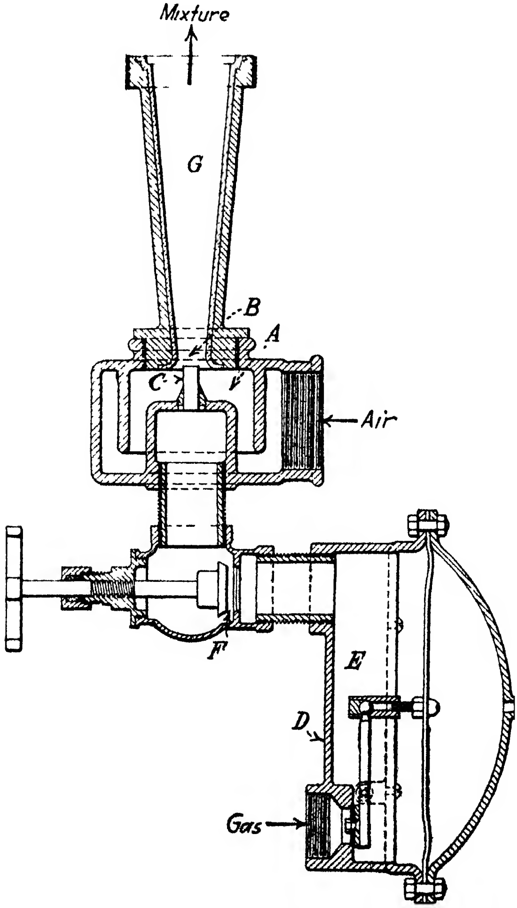

Fig. 95.—Low-pressure system of Surface Combustion Company.

around the outside of the central nozzle tube. By adjusting the nozzle to obtain a desired balance between these two effects, it is possible to keep the air-to-gas ratio constant and to vary the amount of gas fed to the furnace between quite wide limits. After leaving the nozzle, the gas-air mixture passes into a refractory-lined tube in which a small amount of initial combustion takes place, thus raising the temperature to a point where entrained tar is atomized and can be burned efficiently. This burner is also equipped with a vertical lift valve (not shown in the diagram) which is left wide open while the burner is in operation and is closed only for complete shutdown. Tar solidifies

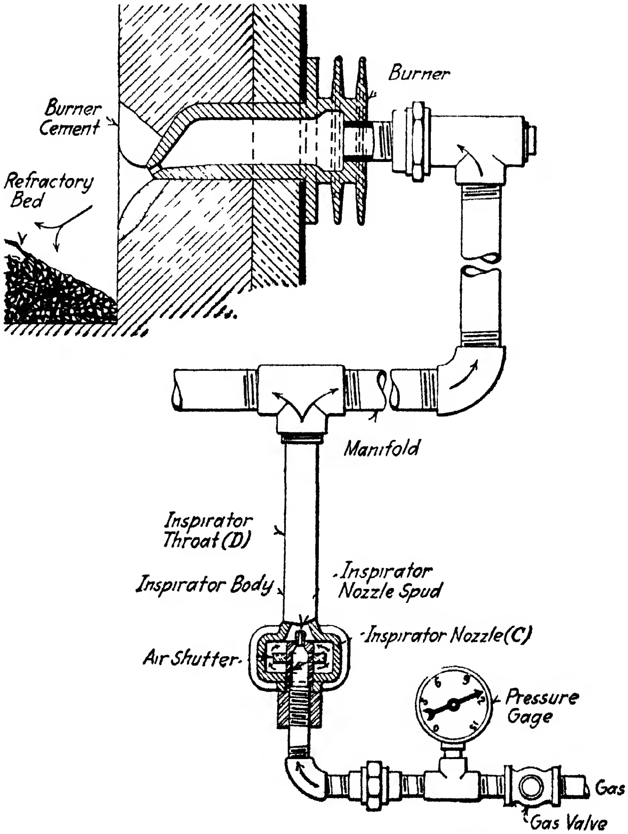

FIG. 96.—High-pressure system of Surface Combustion Company.

around the valve seat while the burner is shut off, and on starting up again it is necessary to shatter the tar seal either by pressing downwards or by striking the valve stem.

Two burners (really proportioning devices) which have been used mainly with rich, high B.t.u. gases are shown[28] diagrammatically in Figs. 95 and 96. In the so-called "low-pressure system" (Fig. 95), the gas is used at a pressure of from 2 to 6 in. water, and the air (all that is required for combustion) is kept at a pressure of from 0.2 to about 4 lb. per sq. in. The air enters the chamber *A* and in passing through the throat *B* induces a flow of gas through the nozzle *C*. The gas is regulated to furnace or atmospheric pressure by means of the governor *D*, and passes through the constricted opening *F* on its way to the mixing nozzle. The gas and the air are mixed in the venturi-like throat and from the expansion tube *G* pass through pipes (as a combustible mixture) to burners, usually similar to those illustrated in Figs. 73 and 74, at the furnaces. The amount of combustible mixture passing to the burners is controlled by varying the air supply, and the gas-air ratio adjusted by changes in the setting of the valve at *F*. This system requires double pipes and is mainly used in small furnaces.

In the "high-pressure system" shown in Fig. 96 the gas is compressed up to about 10 to 25 lb. per sq. in. pressure and, in flowing through the nozzle *C*, induces a flow of air through the air inlet and into the throat *D*. Here the gas and the air are mixed and the combustible mixture led to individual burners (usually of the type shown in Figs. 73 and 74). An adjustable shutter over the air inlet can be used to fix the proportions of gas and air, and the quantity of gas-air mixture supplied then can be controlled by regulation of the gas pressure. This system is well adapted for use with rich gases which require large volumes of air for combustion, since with commercial appliances it is practically impossible to induce this air unless the gas pressures are substantially as high as those maintained with this type of equipment. In this system, as well as in the type shown in Fig. 95, an explosive combustible mixture is being handled at all times.

Proportioning Machines.—In proportioning machines, the gas and all or part of the air required for combustion are mixed together in a fan or chamber, and then delivered as a blast around the plant.

The Reeves burner (or inspirator) system, although compressing a mixture of gas and air and delivering this mixture as a blast around the plant, does not have the disadvantage of carrying an explosive mixture, since the blast is composed of only 1 part gas and 2 of air (*i.e.*, outside the limits of inflammability). This non-explosive mixture, under a pressure of from 0.5 to 3 lb. per sq. in. pressure, passes through the orifice shown in Fig. 97 and induces a stream of secondary air which mixes with the gas-primary-air mixture in the burner tube. The amount of the final complete mixture of gas and air supplied to the furnace is controlled by the pressure on the gas-primary-air mixture and the final gas-air ratio is adjusted by setting the needle valve to give the proper orifice opening.

The Maxon premix burner, in which gas and air are mixed by a fan, is shown in Fig. 98. The admission of air and gas at the blower intake is controlled by a single valve, which, when opened or closed to change the amount of gas supplied, simultaneously alters the setting of the air shutter, thereby maintaining the gas-air ratio essentially constant. The gas and the air are mixed by the fan and delivered under slight pressures direct to the furnace. In most cases no secondary air is used, since all the air necessary for combustion is admitted with the gas at the suction side of the blower. In certain cases, where the flame is to be used in the open, and where no means are provided for slowing down the velocity of the gas-air mixture, considerable difficulty may be experienced in preventing the flame from blowing away from the burner tip. This trouble is remedied (as shown in Fig. 98) by supplying a casing, through which several small holes are drilled, around the burner tip; a small amount of the gas-air mixture passes through these holes, is ignited, and burns in the casing, and serves as a pilot to keep the main body of the mixture burning.

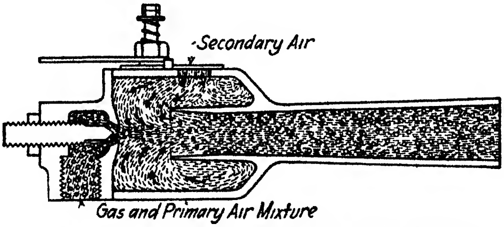

FIG. 97.—Reeves gas burner. (*Industrial Gas Equipment Co.*)

With this system, and with other systems in which the gas-air mixture (requiring no additional secondary air) reaches the furnace under pressure, the furnace draft need not be relied upon to bring in secondary air. The combustion chamber, therefore, may be run under a slight (but appreciable) positive pressure and combustion conditions will not be disturbed by the opening and closing of furnace doors Moreover, with positive pressure in the furnace, air leakage is, of course, prevented—an important consideration wherever it is necessary to keep a reducing atmosphere in order to prevent any oxidation of the material being heated.

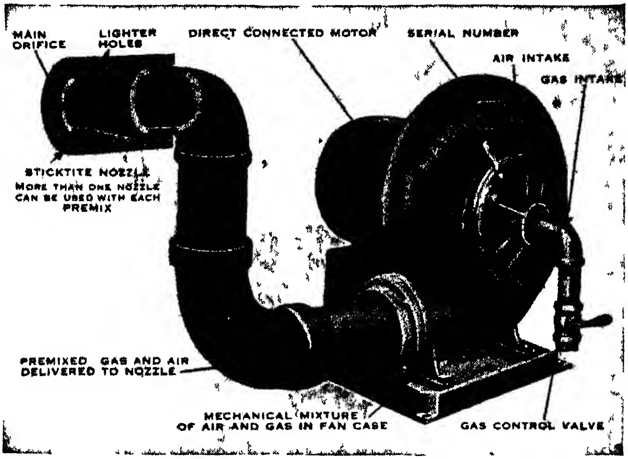

Fig 98 Maxon premix burner (*Maxon Premix Burner Company*)

BIBLIOGRAPHY

1 Tausz and Schulte. "Uber Zündpunkte u Verbrennungs-vorgange im Dieselmotor," *Mitt. Chem -techn Inst Technischen Hochschule Karlsruhe*, **2** (1924).
2. Dixon and Coward: *J* Chem Soc, **95**, 514 (1909).
3. Greenwood: "Industrial Gases," p. 296, D. Van Nostrand Co. (1919).
4. Dodge: *Chem. Met. Eng.*, **26**, 416.
5. Berl and Fischer: *Z. f. Elektrochem.*, **30**, 29 (1924).
6. Mason and Wheeler: *J.* Chem. Soc., **113**, 45T (1918).
7. Payman and Wheeler: *ibid.*, **123**, 426T (1923).
8. Haward and Otagawa: *ibid.*, **109**, 83T (1916).
9. Payman and Wheeler: *Fuel in Science and Practice*, **1**, 185 (1922). *Cf.* Also Payman: *J.* Chem. Soc., **115**, 1436T, 1446T, 1454T (1919);

ibid., **117,** 48T (1920); *ibid.*, **121,** 363T (1922). PAYMAN and WHEELER: *ibid.*, **115,** 36 (1919).
10. MALLARD and LE CHATELIER: *Ann. mines* (VIII), **4,** 274 (1883).
11. DIXON: *Phil. Trans.*, **184,** 97 (1893).
12. VON HELMHOLTZ: "Die Licht und Warmestrahlung verbrannter Gase," Berlin (1890).
13. CALLENDAR, J.: *Gas Lighting,* **111,** 644 (1910). *Cf.* also *Brit. Assoc. Repts.* (1910), 469.
14. HASLAM, LOWELL, and HUNNEMAN: *Ind. Eng. Chem.*, **17,** 272 (1925).
15. LENT and THOMAS: *Z. Ver deut. Ing.*, **58,** 1017 (1924).
16. "Combustion," Am. Gas Assoc. (1924).
17. GROUME-GRJIMAILO: "Flow of Gases in Furnaces" (translated by Williams), John Wiley & Sons (1923).
18. DAMOUR and QUENEAU: "Industrial Furnaces," *Eng. Mining J* (1906).
19. "A Study of the Open Hearth," Harbison-Walker Refractories Co. (1911).
20. "A Study of the Blast Furnace," Harbison-Walker Refractories Co. (1911).
21. "Calco Handbook of Recuperation," The Calorizing Co. (1924)
22. TRINKS: "Industrial Furnaces," John Wiley & Sons (1923).
23. WALKER, LEWIS, and MCADAMS: "Principles of Chemical Engineering," 236–241, McGraw-Hill Book Co., Inc. (1923).
24. BONE: "Coal and Its Scientific Uses," chap. XXII, Longmans, Green & Co. (1922).
25. BONE: Howard Lectures on Surface Combustion, *Roy. Soc. Arts* (1914).
26. *Proc.* Am. Gas. Inst., **6,** 565; and *Ber.*, **46,** 5 (1913).
27. BLAKE: "Surface Combustion," *Eng. Soc. West. Pa*, **36,** 145-204 (1920). *Cf.* Also chapter on Surface Combustion by BLAKE in BACON and HAMOR: "American Fuels," McGraw-Hill Book Co., Inc. (1920).
28. RICHARDSON: *Proc. Pa. Gas Assoc.* (April, 1916).
29. BERRY, BRUMBAUGH, MOULTON, and SHAWN: "Design of Atmospheric Gas Burners," U. S. Bur. Standards, *Tech. Paper* 193 (1921).
30. BERRY, BRUMBAUGH, EISEMAN, MOULTON, and SHAWN: "Relative Usefulness of Gases of Different Heating Value and Adjustments of Burners for Changes in Heating Value and Specific Gravity," U. S. Bur. Standards, *Tech. Paper* 222 (1922).
31. REMBERT: Mass. Inst. Tech. *D. Sc. Thesis* (1925).
32. REMBERT and HASLAM, *Ind. Eng. Chem.*, December, (1925).
33. MILLER: *Year Book* Am. Iron Steel Inst., **11,** 214 (1921).
34. *Blast Furnace and Steel Plant,* **11,** 61 (1923).
35. DIEHL: *Am. Iron Steel Inst.,* **5,** 314–425 (1915).

CHAPTER XII

THE COMBUSTION OF COAL ON GRATES

In many furnaces important special conditions are encountered which involve exceptional problems of design. Furthermore, these special factors vary widely in different industries and operations, and to solve them satisfactorily requires a thorough understanding of the fundamental facts concerning the way in which coal burns on a furnace grate.

When green coal is put into a furnace the volatile matter distills off, leaving coke on the grate. The volatile matter consists essentially of the elementary fuels, gaseous hydrocarbons, and CO plus hydrogen. The final products of the combustion of these elementary fuels are water and CO_2 (see Chap. IX). The coke left on the grate is partly burned to a mixture of CO_2 and CO, the CO being burned later to CO_2. On the fuel bed, therefore, distillation occurs, followed by partial combustion of the coke. *Over* the fuel bed, combustion of the volatile matter (methane, hydrogen, CO, tar, etc). takes place, as well as the combustion of the CO resulting from the partial combustion of the coke.

The distillation of volatile matter (see Chap. II) begins at about 660°F. (350°C.) and is quite rapid at 1300°F. (700°C.), the hydrocarbons appearing first, followed by hydrogen and CO at the higher temperatures. The approximate heat and material distribution resulting from the distillation of the coal are shown in Table LXX.

That is, the volatile matter, tar, and gas which are distilled off carry about 30 per cent of the heat of the coal. Furthermore, the coke left behind on the grate is incompletely burned and, as a result, from 40 to 60 per cent of the heat in the coal usually is developed in the combustion space over the fuel bed. The air comes in under the grate and passes up through and partially burns the coke to a mixture of CO and CO_2. This portion of the total air required for combustion is called *primary air*. The extremely important and necessary additional air which must be

TABLE LXX

HEAT AND MATERIAL DISTRIBUTION ON COAL DISTILLATION

Substance	Per cent by weight	Per cent of total heating value of coal
Water vapor	6.6	0
Gas	14.6	13
Tar	8.5	17
Coke	70.3	70
	100.0	100

admitted *over* the fuel bed in order to burn the volatile matter and the CO is called *secondary air.*

There are four distinct zones in the fire bed, as indicated in Fig. 99. In the *distillation zone* at the top of the fuel bed the coal is separated by destructive distillation into volatile matter and coke. Part of the coke thus formed is burned to CO_2 in the *oxidation zone* by the primary air entering at the bottom of the fuel bed. The CO_2 produced by this oxidation travels up through the bed and is partly reduced to CO by hot coke in the *reduction*

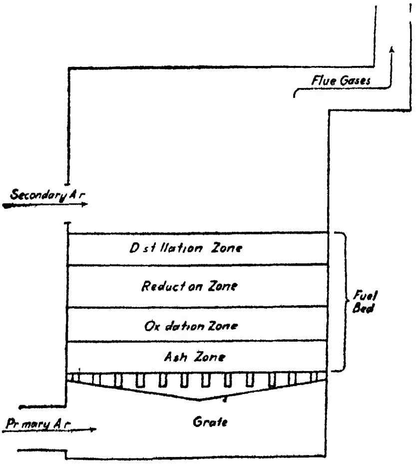

FIG. 99.—Zones in the fuel bed of a furnace. Primary air enters *under* the grate, and secondary air is admitted *over* the fuel bed in order to complete combustion.

zone. An *ash* zone below the oxidation zone protects the grate bars from excessive heat.

The combustion reactions in coal- and coke-fired furnaces have been studied by the U. S. Bureau of Mines[1,2,3] and by Audibert[4] in France. The results obtained clearly indicate the importance of, and distinction between (1) combustion of coal on the grate and (2) combustion in the combustion space.

COMBUSTION ON THE GRATE

In the Bureau of Mines investigation,[1] the apparatus consisted of a cylindrical furnace placed on an air-tight ash box, to which air was supplied under pressure through an orifice. Several water-cooled gas samplers were inserted radially into the furnace at different distances above the grate. The grate was in two sections and about 1 sq. ft. in area. The fuel was fed through the open top of the furnace, and the flue gases allowed to escape out a stack.

Before starting a test the fuel bed was built to the desired thickness and the furnace was heated for 20 to 30 min. The grate was then shaken and the test started. During a test the rate of firing, the thickness of fuel bed, and the rate of air supply were kept constant. Fuel was charged in equal quantities at intervals of from 1 to 5 min. With high rates of combustion charging was frequent and with low rates the intervals were longer. The thickness of the fuel bed was measured with a specially constructed hook. Four sets of temperatures were taken on each test, the measurements being made with an optical pyrometer sighted through the openings provided for the gas-sampling tubes. Three types of fuels were used in the tests, namely, Pittsburgh bituminous coal, anthracite coal, and coke. The rate of combustion was varied from 3 to 125 lb. of fuel per sq. ft. of grate area per hr., and two thicknesses of fuel bed (6 and 12 in.) were used.

Composition of Gases in Fuel Bed.—By analyzing the gases within a fuel bed at varying distances from the grate, and plotting the per cent of the various gases against the distance from the grate in inches, curves similar to Fig. 100 were obtained. The actual data vary somewhat from the lines as shown, on account of the serious difficulties in obtaining representative gas samples. The following conclusions may be drawn:

1. *Primary Air.*—The oxygen present in the primary air (21 per cent) is consumed rapidly, until at a distance of 3 to 4.5 in. above the grate it is practically zero.* The rate of combustion of the fuel (*i.e.*, as measured by the disappearance of oxygen) is proportional to the amount of oxygen present, that is, very rapid at the start, slowing up gradually as the oxygen is consumed. CO_2 only is formed as long as there is much oxygen present, since any monoxide which might be produced would immediately be burned to the dioxide.

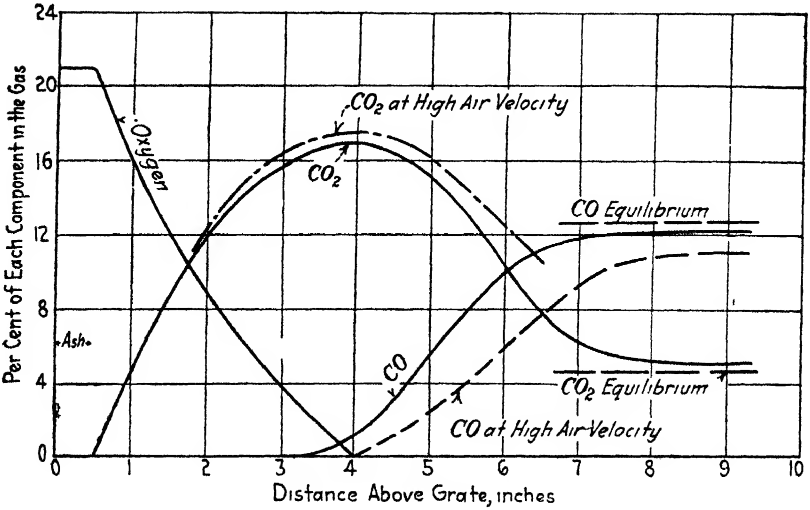

FIG. 100.—Composition of the gases within the fuel bed of a coal-fired furnace. All the oxygen is consumed at about four inches above the grate

2. *Formation of CO_2.*—The oxygen goes to form CO_2, since in the initial stages the curve of CO_2 formation is the exact invert of the oxygen disappearance curve. These two curves cross at approximately 10.5 per cent O_2 and 10.5 per cent CO_2, since in the early stages little or no CO is formed. The oxidation of carbon to the dioxide takes place mainly in the first 3 to 4½ in. from the grate.† As soon as much CO_2 is formed, reduction to the monoxide begins, and for this reason the CO_2 never reaches

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* Except in the case of lignite. With this fuel the oxygen is all consumed within 1½ to 2 in. from the grate.

† In this discussion, it is considered that the fuel bed has been cleaned of ash. Otherwise the distances from the grate refer to the distance from the top of the ash zone.

21 per cent (the maximum is usually about 16.5 per cent). When the CO_2 content of the gases is high, the reduction of the dioxide by hot carbon is rapid and the per cent CO_2 falls until, with a very thick fuel bed, it approaches the value dictated by the equilibrium of the reaction $C + CO_2 = 2CO$ at the temperature of the top of the fuel bed. The first 3 to 4.5 in. of a fuel bed, therefore, make up the *oxidation zone,* in which coke is oxidized largely to CO_2.

3. *Formation of CO.*—Only a very small amount of CO is formed until the oxygen is practically all consumed. The per cent of CO_2 is then at its maximum and CO is, therefore, formed at a rapid rate. At distances further up from the grate the rate of CO formation drops, due to the decreased CO_2 content of the gases and to the decrease in temperature. The CO content of the gases continues to increase as the distance from the grate increases, but never exceeds the composition corresponding to the equilibrium of the reaction $C + CO_2 = 2CO$ for the temperature at the top of the fuel bed. With high-capacity, modern furnaces the temperature at the top of the fuel bed is so high that, as far as equilibrium of this reaction is concerned, all the CO_2 could be reduced to CO. Under these circumstances, however, the time of contact between the gases and the hot coke is insufficient for complete reduction. That portion of the fuel bed above the oxidation zone is, therefore, a *reducing zone* where the CO_2 formed in the oxidation zone is partially reduced by the coke to CO. The thicker the .bed the more CO is formed. Since the object sought in the combustion of coal on a grate is to burn it with oxygen from the air, any reduction of CO_2 to CO is a step backwards. Consequently, the fuel bed should be no thicker than is necessary to prevent blowholes. Blowholes are caused by unevenness in the distribution of coal on the grate or by anything, such as clinkers, that produces uneven resistance to the passage of the primary air up through the fuel bed.

The Effect of Rate of Firing on Gas Composition.—Inasmuch as all the oxygen in the air that comes in under the grate and passes up through the fuel bed (primary air) is consumed within the fuel bed (except, of course, in thin fuel beds with clinkers or blowholes), increased *rate* of firing is brought about by increasing the amount and the velocity of the primary air. A change in the rate of firing (*i.e.*, a change in the rate at which the gases pass through the fuel bed) influences the composition of the gases at varying distances from the grate as follows:

1. *Oxygen.*—It is a very interesting and important fact that all the oxygen is *always** consumed by the fuel within 3 to 4.5 in. from the grate, even if the rate of firing is changed from 3 to 125 lb. of fuel per sq. ft. of grate area per hr. In other words, a change in rate of firing (or in primary air velocity) does not change the position or shape of the oxygen curve in Fig. 100. This is because the rate of reaction

$$C + O_2 = CO_2 \qquad (1)$$

is limited only by the rate at which oxygen can diffuse through the stationary gas film around each carbon particle. The chemical rate of reaction between the carbon and the oxygen is so much faster than the rate of oxygen diffusion that the latter controls the speed of combustion. The rate of oxygen diffusion is approximately proportional to the air velocity, since increased velocity thins down the stationary surface gas film and, within a given time, permits a larger amount of oxygen to diffuse through to the carbon surface. Hence, if the velocity is doubled, the amount of oxygen diffusing through the stationary film is also about doubled and twice the amount of CO_2 is formed. However, since twice the amount of air is present to dilute this CO_2, the composition of the gas in the lower part of the bed remains substantially unchanged.

2. *Formation of CO_2.*—Increasing primary air velocity or rate of firing brings about no change in the proportion of CO_2 in the gases in the lower part of the fuel bed, since there is no change in the rate of oxygen consumption. After the maximum CO_2 content is reached, however, increased velocity slows down the "amount per inch" reduction of CO_2 to CO, since this increased velocity allows a shorter time of contact between the CO_2 and incandescent carbon. Reaction (2),

$$C + CO_2 = 2CO, \qquad (2)$$

is a "time reaction" and the amount of CO formed depends on the time of contact (temperature and other factors being kept

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* This has also been verified by GLADDING and ENTWISTLE,[5] who found that with anthracite coal burning at the high rate of 75 lb. per sq. ft. per hr., the oxygen was completely consumed in passing through 3 in. of live coals. The U. S. Bureau of Mines have shown that the oxygen is completely consumed in 1½ to 2 in. when the fuel is lignite. This is probably due to the great increase in actual surface per unit volume of fuel.

constant). Hence, with increasing rate of firing (*i.e.*, increasing velocity of gases through the fuel bed), after the maximum CO_2 content has been reached, more CO_2 is present in the upper part of the bed, since less time has been allowed for reduction to CO.

3. *Formation of CO.*—With increasing rate of firing, less CO is formed, or, in other words, less CO_2 is reduced, since the time allowed for reduction of CO_2 to CO is lessened. There is practically no CO in the first 1 or 2 in. of the fuel bed when the primary air velocity is higher than that corresponding to 6 lb. of fuel being burnt per sq. ft. grate area per hr., although at a distance of 1 or 2 in. above the grate CO is present to the extent of about 2 per cent when the firing rate is as low as 3 lb. of fuel per sq. ft. of grate area per hr.

Effect of Rate of Firing on Temperature of the Fuel Bed.—The effect of increased rate of firing on the temperature of the fuel bed (as determined by the Bureau of Mines[1, 3]) in the case of a 6-in. bed of coke is shown by Table LXXI.

TABLE LXXI

EFFECT OF RATE OF FIRING ON TEMPERATURE OF FUEL BED

POUNDS OF COKE FIRED PER SQ. FT PER HR	TEMPERATURE 6 IN FROM GRATE
20	2250°F. (1232°C.)
51	2523°F. (1383°C.)
71	2628°F. (1442°C.)
106	2748°F. (1509°C.)

A fivefold increase in the rate of firing raises the temperature of the top of the fuel bed only 500°F. (275°C.). A part of the increased heat developed by combustion of the fuel on the grate is carried away as sensible heat in the gases and the remainder is lost through radiation to colder bodies, as, for example, a steam boiler. Radiation increases as the fourth power of the temperature difference *i.e.*, the heat radiated will be determined by [(temperature of the fuel bed)^4 − (temperature of the boiler)^4], so only a relatively small temperature rise is necessary in order to increase greatly the heat transferred from the fuel bed by radiation.

The fuel-bed temperatures reached with rates of firing below 20 lb. of coke per sq. ft. of grate area per hr. correspond to the ash-fusing point of class 3* coals (ash-fusing point below 2280°F.,

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* See Chap. II, page 26.

or 1249°C.), whereas rates of firing above 50 lb. per sq. ft. per hr. require class 1 coals (ash-fusing point above 2600°F., or 1427°C.). The extent to which the ash in a coal fuses or clinkers depends to a great extent on actual furnace and firing conditions, such, for example, as the disturbance of the fuel bed, etc. It should be kept in mind that the temperature within the fuel bed is greater, as a rule, than the temperature of the surface. With thin fuel beds of anthracite coal protected by arches, however, the surface may be at the highest temperature.

Effect of Air Velocity on Rate of Combustion.—It has already been pointed out that carbon is consumed by the two reactions:

$$C + O_2 = CO_2 \tag{1}$$

and

$$C + CO_2 = 2CO. \tag{2}$$

If only the first reaction took place, then the rate of feeding primary air and the rate of combustion would be exactly the same, since the extent to which reaction (1) progresses is dependent on diffusion of oxygen to the fuel. However, the extent to which reaction (2) proceeds depends on the time of contact between the carbon dioxide and the fuel. Consequently, this second reaction takes place to a lesser extent when the air passes through the fuel bed at higher velocities. Table LXXII gives the pounds of primary air per lb. of coal consumed at different rates of firing.

TABLE LXXII

EFFECT OF AIR VELOCITY ON RATE OF COMBUSTION

Rate of Firing, Pounds Coal per Hr.	Pounds Air through Fuel Bed per Pound Coal Fired
20.5	6.5
47.5	6 0
79.0	6.1
103.0	5.9
124.0	6.2

The above data, which were obtained by the U. S. Bureau of Mines, using Pittsburgh bituminous coal, show conclusively that the amount of coal burned or the capacity of a furnace depends wholly on the amount of primary air coming into the furnace under the grate.

Audibert's Experiments.—It is well known that the gas velocities around the edges of a circular grate are much higher than the velocities at the center, and that the reactions are not so far advanced at the edges as they are at the center (where the Bureau of Mines[1] gas samples were taken). In an attempt to determine the *average* composition of the gases rising from the top of a fuel bed (taking edge leakage into account), Audibert[4] carried out a number of experiments, using coke as a fuel, with the results outlined below.

The apparatus used was rectangular grate 3.3 by 1.65 ft., built at one end of a horizontal cylindrical combustion chamber about 30 in. in diameter and about 30 ft. long with a stack at the opposite end. The stack served merely to carry away the gases, since the air necessary for combustion was forced in under the grate by a fan. *Therefore in these experiments no secondary air was admitted.*

It was found that, with constant air supply, the pressure drop between the ash pit and the fire box, the gas temperature, and the gas composition varied as combustion proceeded. The variation depended on the time between charges and the method of cleaning fires. The gas temperature rose regularly to a maximum (corresponding to heating the brickwork), remained constant for a few minutes, and then began to fall as blowholes in the fuel bed allowed the admission of greater amounts of excess air.

The method of operation was to hold the air supply constant, bring the fuel bed to a given thickness, and measure the temperature carefully. When the temperature began to drop, the door was opened, the grate cleaned, and fresh fuel charged. Under these conditions the interval between firings remained constant for given values of amount of air supplied, coke size, and thickness of fuel bed.

Gas samples taken at intervals during a firing period almost always contained CO_2, O_2, and CO in addition to N_2. The coexistence of CO and O_2 shows that under these conditions, even with coke, the combustion is not necessarily completed on the grate, since unconsumed oxygen (excess air) can leak around the edges of the grate or through holes in the fuel bed.

The gas at any instant during the interval between firings may be represented by a point having the time as abscissa and, as an ordinate, the per cent O_2 if oxidizing and half the per cent CO (negative) if reducing. Thus the distance *above* the zero line represents the number of volumes of O_2 which must be subtracted per 100 volumes of flue gas to get complete combustion without excess air. The distance *below* the zero line represents the number of volumes of oxygen that must be added per 100 volumes of flue gas to get complete combustion without excess air. Figure 101 shows how the gases vary in composition during a firing period.

The gas composition is represented by a series of U-shaped curves, which either are entirely above the zero-oxygen line or cut it in two points. The decrease in excess oxygen represented by the sloping curves to the left of the minimum point is due to the more rapid combustion resulting from the temperature increase during the first part of the interval between firings. As combustion proceeds, however, blowholes form in the fuel bed, more air leaks through, and the oxygen content of the flue gases increases, as shown by the portion of the curves to the right of the minimum point. Three types of operation may be distinguished: (*a*) the oxidizing type, in which there is always an excess of oxygen (*i.e.*, the entire curve is above the zero-oxygen line as in curve 1, Fig. 101); (*b*), the "reducing type," in which the curve, although above the zero-oxygen line at the start, later becomes reducing (*i.e.*, the gas contains CO); and (*c*) a type in which the

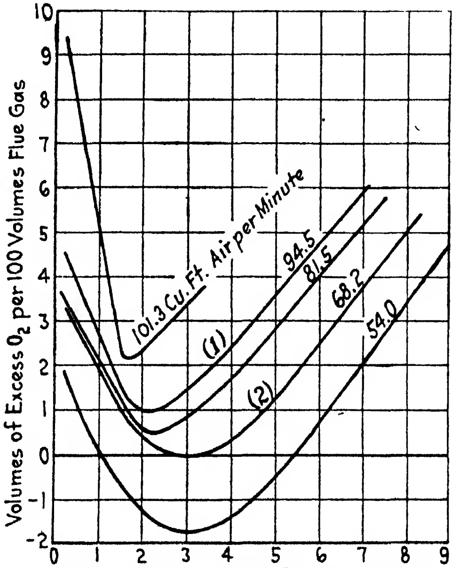

FIG. 101.—Change in the composition of gases arising from a 6 in. fuel bed of 1 in. diameter coke.

curve at its minimum is tangent to the zero-oxygen line, as in curve 2, Fig. 101. In a firing period in which the gas composition may be represented by a curve of the third type there is less *potential* heat loss (CO) than in the reducing type (if no secondary air is admitted), and there is less *sensible* heat loss (excess air) than in the oxidizing type. However, since in all industrial furnaces secondary air is admitted, there is a given curve of the reducing type (with proper regulation of the secondary air supply), which represents the best operating conditions.

From Audibert's work it is possible to draw several conclusions to supplement those resulting from the U. S. Bureau of Mines investigation. These supplementary conclusions, dealing with the operation of a coke-charged fixed grate supplied with a constant amount of air, may be summarized as follows:

1. Air leakage around the edges of the grate and through holes in the fuel bed is of great importance. Thus, although all the oxygen may be removed from the primary air passing up through the center of the grate, air leakage may cause the gases at the top of the fuel bed to contain considerable oxygen.

2. In any event, the gases rising from the top of the fuel bed are rich in CO, and the total heat of the coke can be recovered only by subsequent combustion of CO over the grate. This secondary combustion may be effected either by excess oxygen in the gases coming up through blowholes or around the edges of the fuel bed, or, more usually, by the use of secondary air admitted over the fuel bed.

3. The quantity of air required to give continuously a gas containing oxygen over the fuel bed (with no secondary air) varies directly with the thickness of the fuel bed and inversely with the size of coke.

4. To maintain efficient combustion and keep the gases at their highest possible temperature, the coke must be fed at short intervals. These intervals must be shorter when the combustion rate is high and when the coke is small

THE COMBUSTION OF GASES OVER THE FUEL BED

As pointed out in the early part of this chapter, from 40 to 60 per cent of the total heat in the coal remains as unburned gases rising from the top of the fuel bed. This unconsumed fuel is partly the volatile constituents of the coal (hydrocarbons, CO, H_2, soot) and is partly CO formed within the fuel bed itself.

This is well illustrated[1] by Fig. 102, which shows the composition of gases in the fuel bed and in the combustion space. At the top of the fuel bed it will be seen that the gases are very rich in combustible matter. Then, as secondary air is admitted, the combustible gases burn as they traverse the combustion space. When the end of the combustion space has been reached most of the combustible matter has been consumed.

From the discussion of combustion on the grate it is to be noted that all the oxygen in the primary air is completely consumed (except for blowholes and edge leakage) in passing through a fuel bed of 5 in. or over. Therefore, the *capacity* of a coal-fired furnace or the amount of fuel burned per hr. depends *wholly on the amount of primary air supplied.* From this it follows that if there is any excess air in the exit gases from a furnace this excess air usually comes in as *secondary* air through openings *over* the fuel bed. These openings are usually in the fire door or along the sides of the fuel bed and just over it. Oftentimes, however, the excess air gets in through thin cracks in the furnace walls and excess air from this source is of no benefit; in fact, such leakage greatly lowers the furnace efficiency. Therefore, the *amount, distribution, and turbulence of the secondary air determines the efficiency of the furnace.* These facts may be summarized as follows:

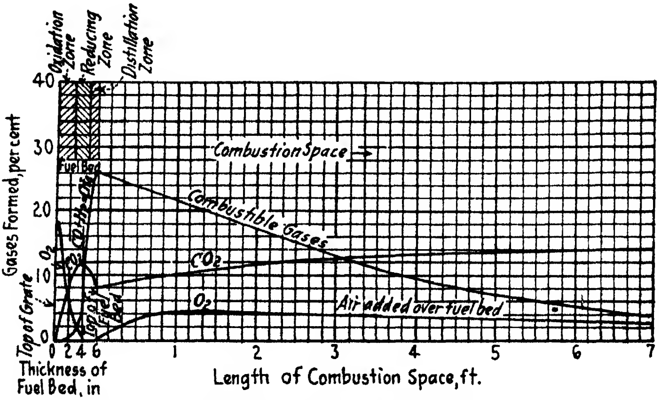

FIG. 102.—Composition of gases in the fuel-bed and combustion space of a coal-fired furnace.

KIND OF AIR	SOURCE	FACTOR AFFECTED
1. Primary	Under the fuel bed	Capacity of the furnace, pounds of coal burned per hr.
2. Secondary	Over the fuel bed	Efficiency with which the fuel is burned

To supplement our general knowledge of gaseous combustion (see Chap. XI) we have considerable quantitative data concerning the combustion of gaseous fuels in the space above the fuel bed of a furnace, the results being mainly from experimental work conducted by the U. S. Bureau of Mines.[3]* The tests were carried on in a furnace having a combustion chamber about 40 ft. long and of nearly constant cross-section. The furnace was fed by a standard Murphy stoker and the hot flue gases were passed from the further end of the long combustion chamber through a Heine boiler. Air was supplied under pressure, both under the grate and over the fuel bed, an exhaust fan in the stack being used merely to give a sufficient reduction of pressure at the discharge end of the furnace. In each experiment gas samples and temperatures were taken at intervals of about 5 ft. along the combustion space. The coal was carefully weighed.

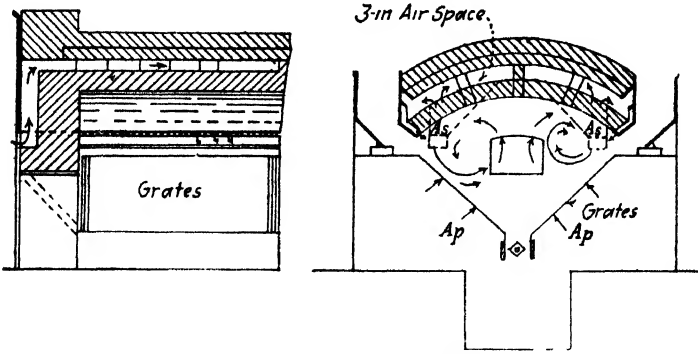

FIG. 103.—Section through Murphy stoker used in the Bureau of Mines investigation, showing construction of arch and method of admitting primary and secondary air.

The combustion space was essentially a long, straight, brick tunnel 3 by 3 ft. in cross-section, having at one end the Murphy stoker with 25 sq. ft. of projected grate area. Two sections through the stoker are shown in Fig. 103. The primary air A_p was admitted under the grate and the secondary air A_s was admitted through a series of tuyéres over the fuel bed. The hot gases rising from the fuel bed met the incoming secondary air and quite rapid mixing resulted, as indicated by the arrows.

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* The subsequent discussion is based entirely on the findings of the Bureau of Mines and to a large extent is modeled on the matter contained in *Bull.* 135. Certain changes (as indicated in the text) have been made to permit the application of these data to ordinary industrial practice.

Following this mixing, the gas passed into the long combustion space.

During each test the thickness of fuel bed, the rate of firing, and the air supply were kept as nearly constant as possible. Three coals were used in the tests, namely, Pocahontas, Pittsburgh, and Illinois coal. The composition and the heating value of each coal are given in the Table LXXIII.

TABLE LXXIII

ANALYSES OF COALS USED IN BUREAU OF MINES TESTS

	Per cent		
	Pocahontas	Pittsburgh	Illinois
(*a*) Proximate analysis (as received)			
Moisture	2.21	2.51	16.16
Volatile matter	15.78	30.28	34.09
Fixed carbon	71.65	56.82	39.19
Ash	10.36	10.39	10.56
	100.00	100.00	100.00
(*b*) Ultimate analysis (dry basis),			
Hydrogen	3.92	4.82	4.66
Carbon	80.90	76.57	69.63
Nitrogen	1.06	1.55	1.49
Oxygen	2.97	4.99	9.55
Sulfur	0.56	1.41	2.08
Ash	10.59	10.66	12.59
	100.00	100.00	100.00
(*c*) Heating value as received (B.t.u. per lb.)	13,762	13,365	10,433

Variations Affecting Data.—In spite of the precautions taken to insure uniformity, considerable variations occurred during the tests. The greatest difficulty was encountered in obtaining representative gas samples. There is considerable tendency for stratification of gases through any combustion space, in spite of the whirling of the gases indicated in Fig. 103. This stratification continues for some distance beyond the fuel bed and it was found very difficult to obtain an average gas sample.

Other variables which affected the results were the completeness and the rapidity with which combustible gas and air mixed. With good mixing combustion is rapid and with poor mixing the process may be very slow. This is one of the major factors in determining the completeness of combustion in a given combustion space. Mixing of gas and air in the Bureau of Mines furnace was more rapid than is ordinarily obtained in practice. This must be borne in mind in adapting their results to problems of design.

The ratio of length of flame travel to volume of combustion space is also an important variable, and in the Bureau of Mines investigation this ratio was much greater than is customary in the average furnace. That is, the length of gas travel, and the consequent time for mixing and interaction of a given volume of gas and air, was greater than is usually possible. Furthermore, the most active portion of the fuel bed was kept at a high temperature because so little of it was "in sight" of the distant boiler. This has a decided tendency to speed up combustion.

In spite of these variations from practical conditions, the data are the only quantitative measure we have concerning the relative importance of the various factors that affect combustion of the unburned fuel over the fuel bed of a furnace, and, as will be shown later, the data are of considerable value in problems of furnace design.

General Conditions.—The progress of combustion over the furnace grate is shown in Fig. 104, in which the composition of the gases is plotted at various distances from the grate. The ordinates are the percentages of the various constituents and the abscissae show both the volume of combustion space up to the point at which the gas is sampled and the length of path of gas travel.

The curves show that the gases leaving the surface of the fuel bed contain over 25 per cent (ordinarily between 25 and 40 per cent) of combustible gases, which represents from 40 to 60 per cent of the total heat of the coal. No oxygen should come through the fuel bed except around the edges or through blowholes, and, due to the difficulty of getting air to flow down against the rising stream of hot gases, very little air comes to the top of the bed. Also at this point the gases are very hot and any oxygen present is rapidly converted into CO_2. Thus, in Fig. 104, although the secondary air is admitted just over the fuel bed, the per cent of oxygen in the gases does not increase very rapidly. The CO curve falls very rapidly, however, and the CO_2 curve rises very rapidly, showing that the combustion is rapid even though the concentration of oxygen is low. As the gases pass away from the fuel bed the combustion slows down, and, in the particular test shown in Fig. 104, the combustion was practically complete at a distance of 13½ ft. from the grate.

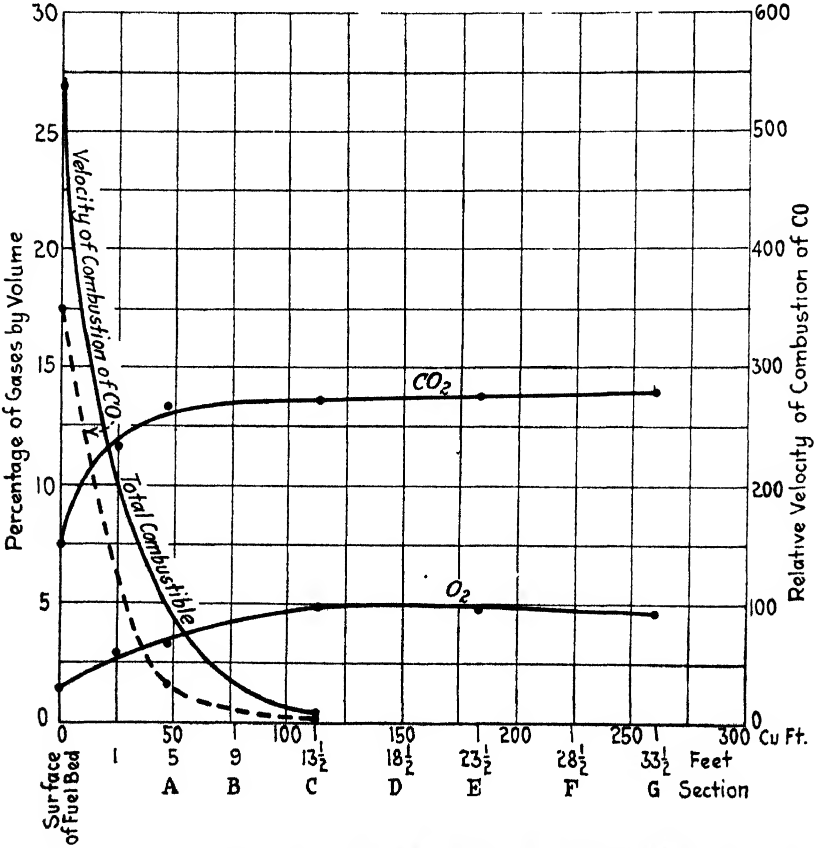

FIG. 104.—Composition of furnace gases at various distances above the fuel bed. Coal used, Pittsburgh screenings; rate of combustion, 35.6 lb. per sq. ft. per hr.; sampling period, 20 minutes.

Size of Combustion Space Required.—The combustion of gases over the fuel bed of a furnace can best be taken up by studying the size of combustion space required under different conditions. It has already been shown that secondary air comes in over the grate, and burns the combustible gases, that primary air controls only the *rate* of burning, and that *secondary air controls the efficiency of combustion.*

The length or volume of the combustion space required depends on four main factors:

1. The amount of excess air used.
2. The rate of combustion, *i.e.*, the pounds of fuel burned per sq. ft. of grate area per hr.
3. The completeness of combustion desired, *i.e.*, the amount of undeveloped heat,* as CO, that goes out the stack.
4. The kind of coal being fired.

There are, of course, factors of construction, such as the efficiency of mixing the unburned gases and air (an important factor in maintaining efficient combustion), the relation of length of combustion space to its volume, and the like, but for a given style of furnace the four factors outlined above are of the most importance.

A study of the way in which each of these four factors affects the size of combustion space affords a good quantitative method of studying the combustion of the unburned gases over a fuel bed.

SIZE OF COMBUSTION SPACE REQUIRED FOR ANY DESIRED COMPLETENESS OF COMBUSTION, RATE OF FIRING, AND EXCESS OF AIR

The relation between the required size of combustion space and any given rate of firing, per cent excess air, and completeness of combustion is given in Figs. 105 and 106. Each of these figures gives the relationship for two of the coals tested by the U. S. Bureau of Mines. The curves to the left of the heavy vertical line dividing Fig. 105 into two equal parts represent Illinois coal, and those to the right represent Pocahontas coal. Each half of the figure contains seven groups of curves, each group representing one degree of completeness of combustion, and each curve in any one group represents one percentage of excess air. Each group is a complete figure in itself, having for ordinates rates of firing ranging from 20 to 60 lb., and as abscissae the size of combustion space, which is given in three scales at the bottom. Reading from top to bottom, the first of these scales gives the volume of the combustion space in cubic feet; the second gives

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* Undeveloped heat signifies the heat lost due to unburned CO in the flue gases. The amount of undeveloped heat is usually expressed as a percentage of the heating value of the coal.

the ratio of the volume of combustion space in cubic feet to the area of the grate in square feet (hereafter, this will be referred to as the ratio of volume to area, or $\frac{V}{A}$); the third scale gives the

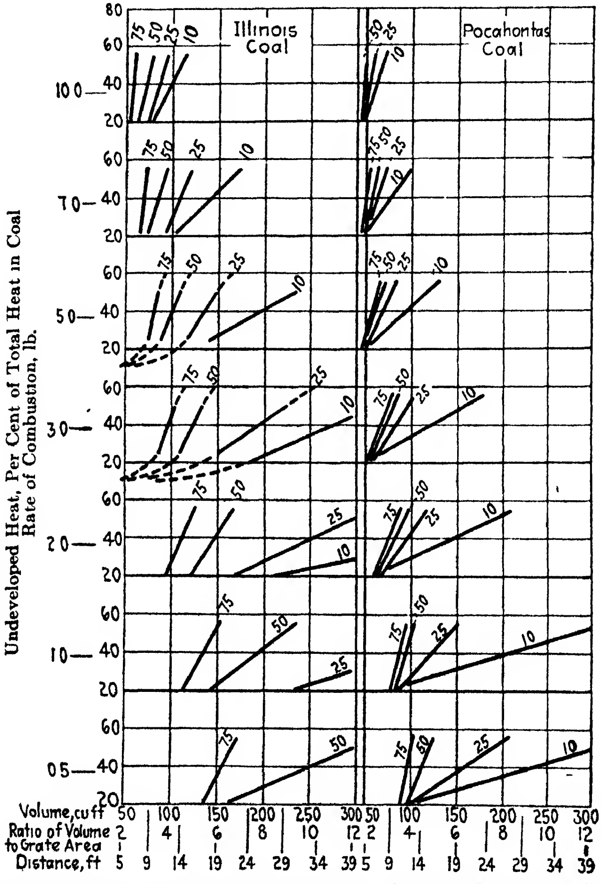

Fig. 105.—Results of tests with Illinois and Pocahontas coal showing relation between the required size of combustion space, a given completeness of combustion, the rate of firing, and the percentage of excess air. Curves on the left are for Illinois coal; those on the right for Pocahontas coal. Figure on each curve indicates percentage of excess air; figures at left margin opposite each group of curves indicates percentage of undeveloped heat and rate of combustion. Figures at the bottom indicate volume of combustion space, ratio of volume of combustion space to grate area and length of path of gas travel.

average length of the path of gas travel from the top of the fuel bed in feet.

The amount of undeveloped heat is expressed as a percentage of the heating value of the coal; that is, the heat in the unburned, combustible gas is so many per cent of the total heating value of the coal. In calculating this value, each of the combustible constituents of the gases was weighted on the basis of its heating

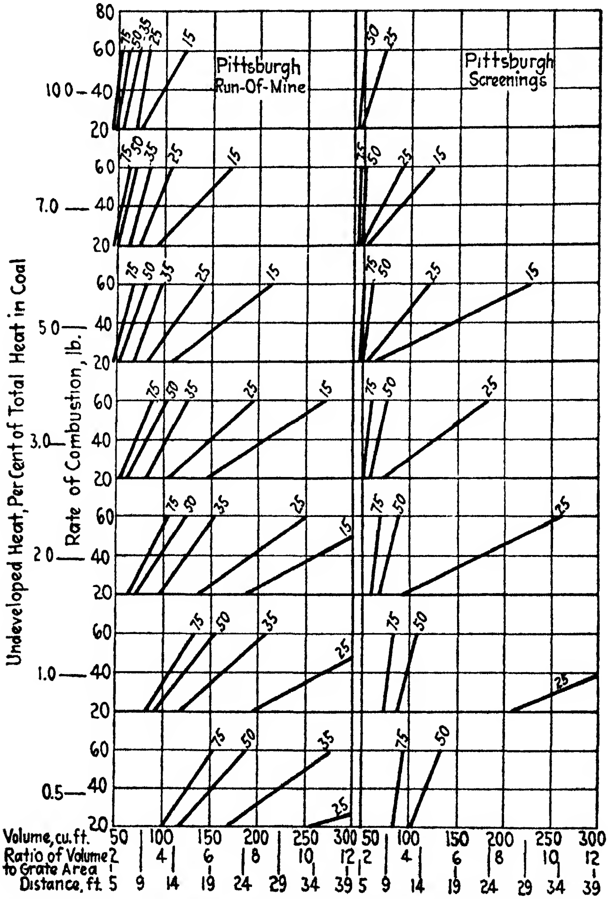

FIG. 106.—Results of tests with Pittsburgh run-of-mine and Pittsburgh screenings showing relation between the required size of combustion space, a given completeness of combustion, the rate of firing, and the percentage of excess air. Curves on the left are for run-of-mine coal; those on the right, for screenings. Figure on each curve indicates percentage of excess air. Figures at left margin opposite each group of curves indicate percentage of undeveloped heat and rate of combustion. Figures at the bottom indicate volume of combustion space, ratio of volume of combustion space to grate area and length of path of gas travel.

value. The unsaturated hydrocarbons were considered as ethylene (C_2H_4). The general formula used[3] was as follows:

$$H = \left(\frac{CO \times 12 \times 10{,}000 + CH_4 \times 16 \times 23{,}500 + H_2 \times 62{,}000 + C_2H_4 \times 28 \times 21{,}300}{12(CO_2 + CO + CH_4 + 2C_2H_4) \times \text{Heating Value of Moisture- and Ash-free Coal}}\right)C.$$

The term H is the undeveloped heat of the combustible gases in the furnace as a per cent of the total heating value of the coal (on a moisture- and ash-free basis) in B.t.u. per lb. C equals the per cent of carbon in the moisture- and ash-free coal. The terms CO_2, CO, CH_4, H_2, and C_2H_4 are the percentages of the respective gases as determined by volumetric analysis. The value of H as calculated in this way is slightly in error (probably not over 5 per cent), since heat in soot and tar rising from the fuel bed is not taken into account. The correct value is approached as tar and soot disappear from the furnace gases.

The Bureau of Mines experiments were run in a furnace which gave exceptionally rapid mixing of the combustible gases and the secondary air admitted over the fuel bed. Furthermore, a long flue (40 ft.) permitted the process of mixing and combustion to extend over a considerable distance. Thus the furnace was hot and the gases were not cooled until practically complete combustion had taken place. Therefore, the percentages of excess air found for a given completeness of combustion were lower than those usually required in the ordinary furnace.

For these reasons, after using the data in Figs. 105 and 106 in problems of furnace design, the calculated per cent of excess air to meet a given set of conditions should be increased by a suitable factor of safety. The actual factor of safety employed depends on the extent to which the furnace combustion space under consideration deviates from the construction of the Bureau of Mines furnace. Usually a factor of safety of 1.5 to 2.2 must be used. For example, if the Bureau of Mines data indicate that 20 per cent excess air is sufficient to give a certain per cent of undeveloped heat under definite conditions, it will usually be found necessary to multiply this by a factor of safety varying from 1.5 to 2.2. In other words, in practice, about 40 per cent excess air would be required to give the specified undeveloped heat loss. All the data in the tables of this chapter (except those in the example at the end of the chapter) are based on the direct data of the U. S. Bureau of Mines without using any factor of safety.

The curves are drawn as straight lines since only two points were determined for each condition, the points being partly obtained by averaging and partly by interpolation. The curves are approximate within perhaps 10 to 20 per cent. Since the range takes in the rates of combustion used in both large and small power stations, they are of considerable practical importance in furnace design for the determination of size of combustion space when other conditions are given. The figures given for the first three coals are more accurate than those obtained with Pittsburgh screenings, since not enough runs were made with the latter to insure reliable points.

Size of Combustion Space as Affected by Excess Air.—The way in which the size of combustion space is affected by the amount of excess air is well shown by Figs. 105 and 106. From the plots it can be seen that, for any given degree of completeness of combustion, the size of combustion space decreases with increase in excess air. *With very low excess air, complete combustion* in direct coal-fired furnaces can be obtained only with great difficulty. This is well illustrated by taking a specific example. Suppose a furnace is firing an Illinois coal at a constant rate of 40 lb. per sq. ft. of grate area per hr. and the amount of undeveloped heat going out the stack is 3 per cent, then, reading from the left-hand side of Fig. 105, the results shown in Table LXXIV are obtained.

TABLE LXXIV

EFFECT OF EXCESS AIR ON SIZE OF COMBUSTION SPACE

WITH THE EXCESS AIR EQUAL TO, PER CENT	THE FOLLOWING VOLUME OF COMBUSTION SPACE PER SQ. FT. OF GRATE AREA $\left(\frac{V}{A}\right)$ IS REQUIRED
10	11.0
25	7.8
50	5.0
75	3.8

This clearly shows that with low excess air a large ratio of volume of combustion space to grate area (*i.e.*, large combustion space) is required, even for a relatively large loss of undeveloped heat (unburned fuel).

Size of Combustion Space as Affected by Rate of Combustion.—The way in which the size of combustion space is affected by the rate of combustion may be shown by further reference to Fig. 105. Suppose it is desired to fire a furnace with Illinois coal, using 50 per cent excess air and allowing only 0.5 per cent of undeveloped heat to escape from the stack. The results are shown in Table LXXV.

TABLE LXXV

Effect of Rate of Combustion on Size of Combustion Space

With the Rate of Firing Equal to (in Pounds Coal per Sq. Ft. per Hr.)	The Following Volume of Combustion Space per Sq. Ft. of Grate Area $\left(\frac{V}{A}\right)$ Is Required
20	6.5
30	8.0
40	10.0
50	12.0

Table LXXV shows that increasing the rate of firing two and one half times has less than doubled the ratio of volume to area. Thus with harder forcing, each cubic foot of combustion space is relatively more effective. This would be expected from the discussion in the first part of the chapter, in which it was pointed out that, with increased rate of firing (and consequent increased primary air velocity), less CO is formed in the fuel bed. Furthermore, the higher temperature and the greater gas velocity and turbulence due to increased rates of firing all tend to promote better combustion.

Size of Combustion Space as Affected by Completeness of Combustion.—To illustrate the way in which the size of combustion space is affected by the degree of completeness of combustion desired, suppose an Illinois coal is being fired at the rate of 40 lb. per sq. ft. of grate area per hr., and the excess air is being kept constant at 50 per cent. Then, reading from Fig. 105, the values shown in Table LXXVI are obtained.

TABLE LXXVI

Effect of Completeness of Combustion on Size of Combustion Space

With the Per Cent Undeveloped Heat Equal to	The Following Volume of Combustion Space per Sq. Ft. of Grate Area $\left(\frac{V}{A}\right)$ Is Required
10	2.8
5	4.0
3	5.0
2	5.9
1	7.8
0.5	10.0

Thus, for a given rate of firing and excess air, the loss of undeveloped heat must be relatively large in a furnace having a small combustion space

Size of Combustion Space as Affected by Quality of Coal.—It is known that coals of different compositions burn with different efficiencies in a combustion space of a given size and shape. The characteristics of a coal that are most likely to affect the size

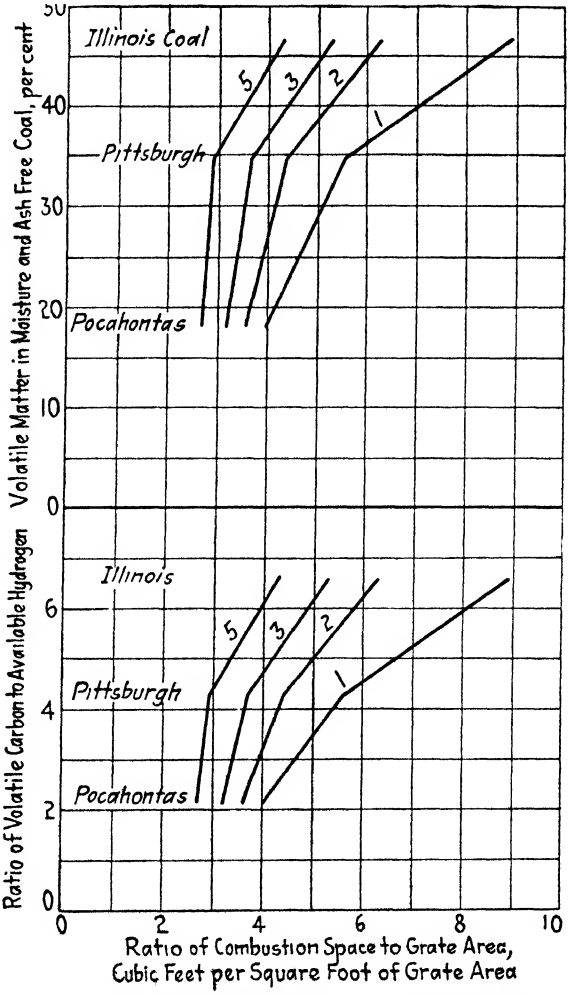

FIG. 107.—Relation between volatile matter of coal and size of combustion space required. Upper group shows relation between the size of required combustion space and the percentage of volatile matter; lower group shows relation between the size of required combustion space and the quality of the volatile matter as shown by the ratio of volatile carbon to available hydrogen. Figure near each curve indicates percentage of heat not developed, owing to incomplete combustion. In all tests the rate of combustion is 50 pounds and the excess of air 50 per cent.

of the combustion space are the quantity and the quality of the volatile matter.

1. *Quantity of Volatile Matter.*—Coals with high percentages of volatile matter require more combustion space than coals containing a low percentage of volatile matter. The relation between the per cent of volatile matter and the size of combustion space for the three coals studied by the Bureau of Mines is shown in the upper part of Fig. 107. The small figures on each curve, 1 to 5 inclusive, show the per cent of undeveloped heat (CO). The volume of combustion space required is *not* proportional to the volatile matter. For example, the Pittsburgh coal (35 per cent *V. M.*) and the Pocahontas coal (18 per cent *V. M.*) differ greatly in per cent volatile matter, but the change in volume of combustion space per sq. ft. of grate area $\frac{V}{A}$ is comparatively small. This is also shown by the change in combustion space required when the rate of combustion is changed. Doubling the rate of firing would cause double the amount of volatile matter to be produced. Doubling the rate of firing, however, does not double the size of the combustion space required to give the same completeness of combustion with the same excess air. This is shown by Table LXXVII.

TABLE LXXVII

EFFECT OF QUANTITY OF VOLATILE MATTER ON SIZE OF COMBUSTION SPACE

Per cent undeveloped heat	Rate of combustion, pounds per sq. ft. per hr.	Ratio of volume of combustion space (cubic feet) to grate area (square feet) with 50 per cent excess air		
		Pocahontas coal	Pittsburgh coal	Illinois coal
5	50	2.7	2.9	4.3
5	25	2.0	2.2	3.5
3	50	3.2	3.7	5.3
3	25	2.3	2.7	4.4

Thus, for a given coal to burn the double quantity of volatile matter which results from a doubled combustion rate with the same excess air and completeness of combustion requires an increase in the size of combustion space in the above case, of less than 30 per cent.

2. *Effect of Quality of the Volatile Matter.*—The quality of the volatile matter of a coal may be expressed as the ratio of volatile carbon to available hydrogen,* *i.e.*,

$$\frac{\text{Per Cent Total Carbon} - \text{Per Cent Fixed Carbon}}{\text{Per Cent } H_2 \text{ Content of Moisture- and Ash-free Coal} - \frac{1}{8} \text{ Per Cent Oxygen Content}}.$$

The ratio of required size of combustion space to grate area is plotted against this factor in the lower part of Fig. 107. The figure shows that the Pittsburgh coal contains about twice as much volatile carbon and the Illinois coal nearly three times as much volatile carbon as the Pocahontas coal. In burning Pocahontas coal (which has a low ratio) the volatile matter is probably distilled off as light, fixed gases which are readily burned in the combustion space. The proportionality between this factor and the size of combustion space required is not good, and should not be used in attempting to determine the size of combustion space with a coal other than those tested by the Bureau of Mines. In Fig. 107 it will be seen that there is a marked change in slope as the completeness of combustion is changed, and that it is practically impossible to burn Illinois coal with small percentages of heat lost from the stack as undeveloped heat unless the combustion space is very large.

3. *Combined Effect of Quality and Quantity of Volatile Matter.*—Figure 107 shows that both the quantity and the quality of the volatile matter influence the burning of a coal in a similar manner. Therefore it appears probable that the quality and quantity each affect the volume of combustion space required. The product of these factors $\left(\text{Volatile Matter} \times \frac{\text{Volatile Carbon}}{\text{Available Hydrogen}}\right)$ is plotted against the ratio of a required size of combustion space to grate area in the top of Fig. 108. Each curve shows the relation between the size of combustion space and the above factor for the degrees of undeveloped heat indicated on the curves.

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* If the value of this ratio is low, the volatile matter contains considerable quantities of hydrocarbons which are saturated (such as the paraffins, C_nH_{2n+2}) or which are only slightly unsaturated (such as the olefines, C_nH_{2n}). When the ratio is high, on the other hand, the volatile matter is rich in unsaturated hydrocarbons (*i.e.*, less hydrogen is present in proportion to the carbon than with the paraffins). Unsaturated hydrocarbons usually decompose thermally or deposit soot more easily than the saturated hydrocarbons.

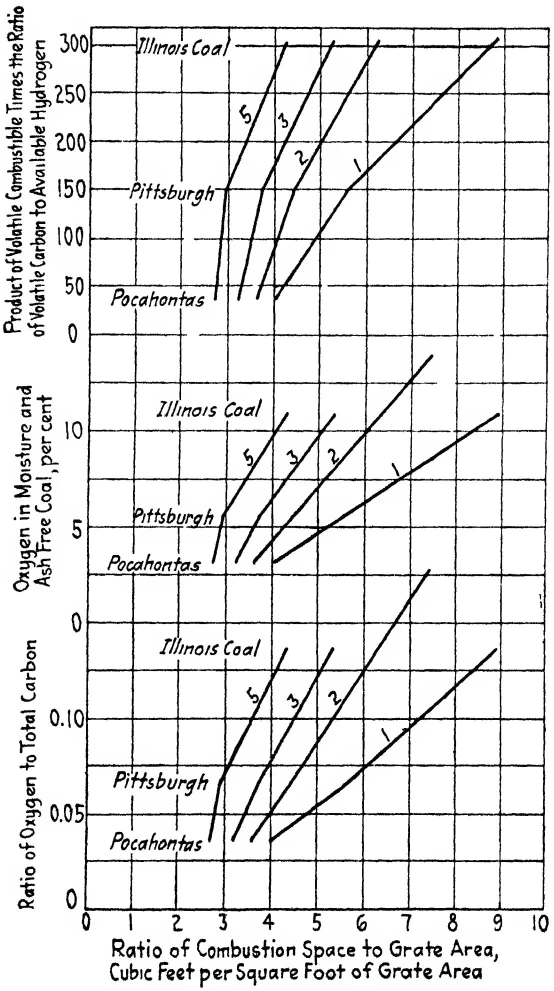

FIG. 108.—Relation between size of combustion space required and the chemical properties of three coals. The upper group of curves has for its ordinates the product of the percentage of volatile matter times the ratio of volatile carbon to available hydrogen; the middle group, the percentages of oxygen in "moisture and ash free" coal; and the lower group, the ratio of total carbon to oxygen in "moisture and ash free" coal. All three groups have a common abscissa, the ratio of the combustion space to grate area. Figure near each curve indicates percentage of heat not developed, owing to incomplete combustion. In all tests the rate of combustion is 50 pounds and the excess of air 50 per cent.

The size of combustion space per sq. ft. of grate area $\left(\frac{V}{A}\right)$ is very nearly proportional to this factor, particularly under circumstances where combustion is practically complete, as indicated by the fact that the curves in Fig. 108 are nearly straight lines.

4. *Oxygen Content of Coal as an Indication of Burning Quality.*—The effect of oxygen on the heating value and coking properties of coal has been pointed out in Chap. II, and the oxygen content in moisture- and ash-free coal has been considered as a possible indication of the burning quality. This factor is plotted against the ratio of required size of combustion space to grate area in the curves in the center of Fig. 108. Curves are drawn for 1, 2, 3, and 5 per cent undeveloped heat. From the figure it can be seen that the proportionality between the per cent of oxygen in the coal and the required-size combustion space $\left(\frac{V}{A}\right)$ is much better than in the preceding plots. The oxygen content of coal is, therefore, useful as an indication of the burning quality of a coal; coals with low oxygen contents do not require as large a combustion space as the higher-oxygen coals.

5. *Ratio of Oxygen to Total Carbon.*—Another factor which indicates the burning quality is shown in the lower set of curves in Fig. 108. The ratio of oxygen content to the total carbon (both in moisture- and ash-free coal) is plotted against the ratio of required size of combustion space to grate area required to give a definite loss of undeveloped heat, other factors being kept constant. Thus, if from the tests of the three coals used by the Bureau of Mines, carried out with the same rate of firing and the same completeness of combustion, the ratios of required combustion space to grate area are plotted as ordinates against the ratios of oxygen to total carbon as abscissae, it is found that the proportionality between this factor and the size of the combustion space is good and the lines connecting the points for the three coals will be nearly straight. This relationship is also useful in judging a coal for furnace design.

Summary of Coal Quality Factors.—In designing a furnace for burning a given coal, equal weight should be given to the three following factors: (1) Volatile Matter $\times \left(\frac{\text{Volatile Carbon}}{\text{Available Hydrogen}}\right)$,

(2) oxygen content of moisture- and ash-free coal, and (3) the ratio of oxygen to total carbon. For example, suppose it is desired to design a furnace for burning each of the coals given in Table LXXVIII, with 2 per cent undeveloped heat, when the rate of firing is 50 lb. per sq. ft. grate area per hr. and the excess air is maintained at 50 per cent.

TABLE LXXVIII

Analyses of Three Given Coals and Factors Influencing Their Burning Qualities

	New Mexico coal, per cent	Virginia coal, per cent	New River (W. Va.) coal, per cent
1. Moisture (as received)	11.90	3.37	1.85
2. Volatile matter (as received)	37.85	33.75	20.31
3. Fixed carbon (as received)	41.57	56.41	73.87
4. Ash (as received)	8.68	6.47	3.95
5. Carbon (moisture- and ash-free basis)	78.48	84.70	88.50
6. Hydrogen (moisture- and ash-free basis)	5.52	5.32	5.02
7. Oxygen (moisture- and ash-free basis)	14.02	7.53	4.28
8. Nitrogen (moisture- and ash-free basis)	1.28	1.55	1.55
9. Sulfur (moisture- and ash-free basis)	0.70	0.90	0.65
10. Volatile matter (moisture- and ash-free basis)	47.6	37.5	21.55
11. Fixed carbon (moisture- and ash-free basis)	52.4	62.5	78.45
12. Volatile carbon ÷ available H_2	6.9	5.1	2.24
13. Line 10 × line 12	327	191	48.3
14. Line 7 ÷ line 5	0.178	0.089	0.049
15. Heating value (as received), B.t.u.	13,762	13,365	10,433

By the use of the three factors in lines 7, 13, and 14, and Fig. 108, it is possible to approximate the size of combustion space required. Thus, with the New Mexico coal the per cent oxygen is 14.02 and the point on the prolonged curve for 2 per cent undeveloped heat in the center of Fig. 108 shows the size of combustion space required is 7.4 cu. ft. for each square foot of grate area. Similarly, for the factor, Volatile Matter × $\frac{\text{Volatile Carbon}}{\text{Available Hydrogen}}$, opposite 327 in the upper part of Fig. 108, we read for 2 per cent undeveloped heat that 6.7 cu. ft. of combustion space are required for each square foot of grate. The ratio of oxygen to total carbon (line 14 in the table) is 0.178, and, reading from the lower set of curves in Fig. 108, it is found that 7.4 cu. ft. of combustion space are required per sq. ft. of grate.

Thus, from the three factors, it has been found that the required cubic feet of combustion space per sq. ft. of grate $\left(\frac{V}{A}\right)$ varies from 6.7 to 7.4, and the value to be used is the average, or $\frac{6.7 + 7.4 + 7.4}{3} = 7.1$.

In the same way the ratio of $\left(\frac{V}{A}\right)$ for the Virginia coal is found to vary between 4.9 and 5.1, and for the New River coal from 3.7 to 3.9.

It should be noted that the curves given in Fig. 108 are drawn for only one set of conditions, namely, 50 lb. of coal fired per sq. ft. of grate per hr., with 50 per cent excess air. To design a furnace for a different rate of firing and different excess air it is necessary to construct curves similar to those in Fig. 108, but for the new conditions. Data for such curves can be readily obtained from the curves in Figs. 105 and 106.

Effect of Moisture in Coal on Combustion.—The amount of moisture present in a coal may also affect its combustion. In small amounts, up to 5 per cent, as shown in Chap. IX, moisture speeds up the combustion of CO, but beyond this point increase in water content retards the reaction.

Aside from its influence on the speed of combustion of CO, moisture also functions as a diluent of the reacting gases, and thus makes combustion more difficult. For example, in burning Illinois coal (16 per cent moisture) with 50 per cent excess air with 3 per cent of undeveloped heat at a rate of 50 lb. per sq. ft. of grate per hr., the analysis of the flue gases leaving the furnace is as shown in Table LXXIX.

TABLE LXXIX

ANALYSIS OF FLUE GASES WHEN BURNING ILLINOIS COAL

(50 per cent excess air; 50 lb. per sq. ft. per hr.; 3 per cent undeveloped heat)

CONSTITUENT	PER CENT
Nitrogen	75.4
Carbon dioxide	13.0
Water	7.0
Combustibles:	
Oxygen	4.0
Carbon monoxide	0.6

Thus, although the water vapor is only 7 per cent of the furnace gas, it is greater in volume than the combined volume of the two reacting gases, CO and O_2. In such cases the moisture probably retards the process of combustion, since it reduces the concentration of the reacting gases and cuts down the temperature in the combustion chamber. Moisture, of course, increases the stack loss. Coal is sometimes "tempered" with water before firing. This retards the rate at which the volatile matter is distilled and cuts down the smoke.

THE MOST ECONOMICAL AIR SUPPLY FOR A GIVEN FURNACE AND A GIVEN COAL

For a given furnace and a given coal there is a definite air supply which gives the most economical operation, *i.e.*, which allows the greatest overall efficiency. In the design of a new furnace there are two other variables: (1) size and kind of furnace and grate, and (2) kind of coal. With a furnace already built, however, there are only two variables, *i.e.*, the kind of coal and the amount of excess air. Let us consider an installed furnace with the kind of coal to be used as having been determined.

Under these conditions, the percentage of excess air which gives the greatest overall efficiency depends on two factors: (1) the amount of heat lost up the stack as sensible heat in the flue gases, and (2) the amount of heat lost up the stack as undeveloped heat (unburned CO). With the correct amount of excess air, the sum of these two losses is at a minimum. If the excess air is increased over this definite optimum amount, the sensible heat loss will amount to more than the gain from greater completeness of combustion. Also if the amount of excess air is decreased, the per cent of CO goes up and the heat lost in this way more than counterbalances the saving in sensible heat. With a change in quality of coal the excess air necessary to obtain these optimum conditions changes, low-quality coals requiring high excess air, but for a given coal the most efficient per cent of excess air in any given furnace is fixed.

Thus, for the Illinois coal tested by the Bureau of Mines, fired at the rate of 25 lb. per sq. ft. of grate per hr., the optimum amounts of excess air are shown in Table LXXX. The most desirable per cent excess air decreases as the size of combustion space increases.

TABLE LXXX

MOST EFFICIENT PER CENT EXCESS AIR FOR BURNING ILLINOIS COAL
(Rate of combustion = 25 lb. per sq. ft. per hr.)

Ratio of volume of combustion space (cubic feet) to grate area (square feet)	Optimum per cent excess air	Most desirable per cent CO_2
3	100	9 0
4	68	10 8
5	54	11 8
6	46	12 5
7	40	13 0

This situation is shown graphically for a typical boiler and coal in Fig. 109. The stack loss due to the volume of hot flue gases is shown as dotted curve 1, while the loss of heat due to unburned CO is shown as dotted curve 2. The sum of these two losses is the determining factor and is shown as full curve 3. It is seen that as the amount of air is increased over that theoretically required, the total heat loss (curve 3) drops on account of the rapid decrease in the loss due to unburned CO. As the amount of air is further increased, however, the rate at which heat is gained through a decrease in the unburned CO gradually slows up until, at about 25 per cent excess air (in the case given), this gain is just counterbalanced by the increased loss as sensible heat in the stack gases. Figure 109 is drawn on the assumption that increasing the per cent excess air does not change the stack temperature. Actually, the stack temperature usually *increases* with increase in excess air (see Fig. 202, Chap. XVI) and this calls for a still lower excess air for maximum efficiency. The use of a very large amount of excess air would, of course, again lower the stack temperature, owing to the great dilution of hot furnace gases with cold air.

The temperature of flue gas going out a boiler stack increases as the excess air is increased over that required to give maximum efficiency. As the excess air increases, the temperature of combustion and hence the temperature of the furnace walls decreases and a smaller amount of heat is transmitted directly to the boiler tubes by radiation. This leaves more heat to be taken out of the hot flue gases by conduction and, obviously, due to the increased load on the boiler tubes, the temperature of stack gases goes up. With a water-tube boiler the rate of transfer of heat from the hot flue gases to the cooler water tubes is proportional to about the six-tenths power of the velocity. Therefore, when the flue gases contain larger amounts of excess air the rate at which heat will be taken out of the diluted gases does not go up so fast as the increased volume. Consequently, the temperature of the flue gas will go up until the effect of dilution is so great that the temperature finally falls.

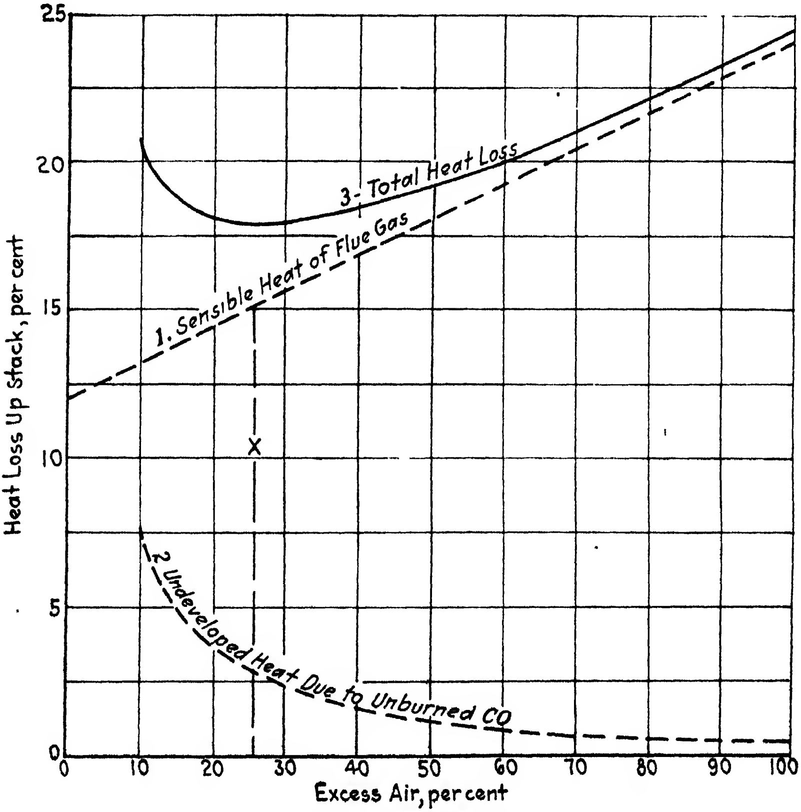

FIG. 109.—Variation in stack loss with excess air, and the most desirable per cent CO_2 for a given boiler furnace and given coal.

To summarize the best conditions for furnace operation: first, with a given furnace and coal, there is a certain per cent of excess air which must be used for most economical operation; second, with a furnace having a high stack temperature the best percentage of excess air is lower than with a furnace having a low stack temperature; and, third, for the majority of furnaces the most desirable per cent of excess air is such that there is a very slight amount of CO present in the stack gas. If there is absolutely no CO or soot in the flue gas, it is certain that the furnace is not being operated under the most economical conditions, since too great an excess of air is being used. It should be borne in mind that relatively small amounts of CO cause a large loss of heat and, as shown in Fig. 101, the composition of the gases coming from a hand-fired fuel bed is not constant; at one moment there may be no CO and at another a large amount. Therefore, the extent to which in practice one can approach the optimum percentage of excess air depends on conditions specific to each particular case. For example, with automatic stokers or with pulverized coal the fluctuation can be largely eliminated, and consequently it is possible to operate constantly at the most efficient excess air. Soot causes a double loss, of potential or undeveloped heat, and, in addition, of sensible heat due to coating the heat-absorbing surfaces and preventing heat transfer. It is obvious that, with a very smoky coal and a hand-fired furnace, the proper per cent of excess air to be used will be higher than with a non-smoking coal.

To determine the per cent of excess air and the amount of heat lost up the stack as sensible heat in the flue gas and as unburned CO, it is only necessary (for reasonably accurate results) to obtain the temperature and analysis of the stack gases. If the gases are analyzed carefully, it is not necessary to determine either the heating value of the coal or its complete chemical analysis. The charts given in Chap. X obviate these determinations.

The following illustrative problem is included merely to show the *manner* in which the Bureau of Mines data may be used to solve problems of design and the influence of the various factors.

An industrial furnace is to be heated by burning an Illinois coal having a heating value of 12,000 B.t.u. per lb. and costing $7.50 per ton. The coal is to be fired at the rate of 25 lb. of coal per sq. ft. of grate area per hr. The net useful heat required by the furnace is 22,000,000 B.t.u. per hr. Tests on similar furnaces indicate a stack temperature of 680°F. above atmospheric temperature and losses by radiation, through the ash pit, etc., amounting to 15 per cent of the gross heat.* The furnace is to be operated 24 hr. a day for 300 days in the year.

The company purchasing this equipment expects to depreciate this furnace in five years and also expects the cost of maintenance during this period to be 10 per cent per year. Furthermore, the company can earn, in its regular line of business, 25 per cent on the money it invests. The cost of similar furnaces has been found to be equal to (\$7500 + \$20 X), where X is the total combustion space in cubic feet. Under these conditions what will be:

1. The best ratio of combustion space to grate area?
2. The volume of the combustion space?
3. The most desirable percentage of excess air for the proposed furnace?
4. The heat loss as unburned CO figured as per cent of the heat in the coal?

The solution of a problem of this type consists first in determining the minimum heat loss for each ratio of combustion space to grate area (V/A), and then, using these values, determining the minimum cost for coal and fixed charges. An assumption of the coal analysis must be made, but even a relatively large error in this item will make little difference in the result. The sensible heat loss up the stack for various percentages of excess air can then be calculated as follows:

From the table of coal analyses in Chap. III, a coal from the Orient mine in Franklin County, Illinois, is found to have the following analysis as received:

	Per Cent
Proximate analysis:	
Moisture	7.4
Volatile matter	36.7
Fixed carbon	47.9
Ash	8.0
	100.0
Ultimate analysis:	
Carbon	69.0
Hydrogen	5.4
Sulfur	1.0
Nitrogen	1.6
Oxygen	15.0
Heating value (as received)	12,164 B.t.u. per lb.

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* For simplicity's sake no account will be taken of the variation of stack temperature with excess air. The percentage of heat lost by radiation also is considered constant.

From these data it is possible to calculate the sensible heat loss in the stack gases, assuming complete combustion.

Basis: 100 lb. of coal

This coal contains 69 lb. of carbon, or $^{69}\!/_{12} = 5.75$ mols carbon. The 15 lb. of oxygen in the coal will combine with $^{15}\!/_{8} = 1.875$ lb. H_2 (since O_2 combines with H_2 in the ratio of 8:1) to form $\frac{1.875}{2} = 0.937$ lb.-mols of water (*i.e.*, the amount of moisture and combined water in the coal). The total hydrogen content of coal minus the hydrogen equivalent of the oxygen content is $5.4 - 1.875 = 3.53$ lb. of net hydrogen, or $\frac{3.53}{2} = 1.77$ lb.-mols of hydrogen to be burned by the oxygen of the air.

The oxygen required for combustion is:

For carbon	5.75 mols O_2
For net hydrogen $\left(\frac{1.77}{2}\right)$	0.88 mols O_2
Total O_2 required	6.63 mols O_2

Thus 6.63 mols of oxygen are equivalent* to $\frac{6.63}{0.21} = 31.5$ mols of air or $31.5 \times 0.79 = 24.9$ mols of nitrogen. The flue gas coming from 100 lb. of coal therefore contains:

CO_2		5.75 mols
H_2O	$1.77 + 0.94 =$	2.71 mols
N_2		24.9 mols

Now, assuming the average outside air temperature to be 60°F., the average temperature of the stack gases will be $680 + 60 = 740$°F. The sensible heat per mol of CO_2, H_2O, and N_2 can be read from Fig. 49, Chap. X, as follows:

Gas	Sensible Heat at 740°F., B.t.u. per Mol (Above 60°F.)
CO_2	6760
H_2O	5800
N_2	4760

The sensible heat in the flue gas from 100 lb. of coal is then found to be:

For CO_2	$5.75 \times 6760 =$	38,800 B.t.u.
For H_2O	$2.71 \times 5800 =$	15,700 B.t.u.
For N_2	$24.9 \times 4760 =$	118,400 B.t.u.
Total sensible heat in flue gas (between 740 and 60°F.) per 100 lb. coal	=	172,900 B.t.u.

The per cent of the heat input that is lost as sensible heat is, then, the sensible heat in the flue gas divided by the heating value of the coal, or $\frac{172{,}900 \times 100}{12{,}164 \times 100} = 14.2$ per cent. Similarly, when 100 per cent excess air

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* Air contains 21 per cent oxygen and 79 per cent nitrogen.

(theoretical air previously shown to be 31.5 mols) is used, in addition to the sensible heat of the gases as calculated above, there is a further loss due to the 31.5 mols of air:*

$$31.5 \times 4760 = 150{,}000 \text{ B t u.}$$

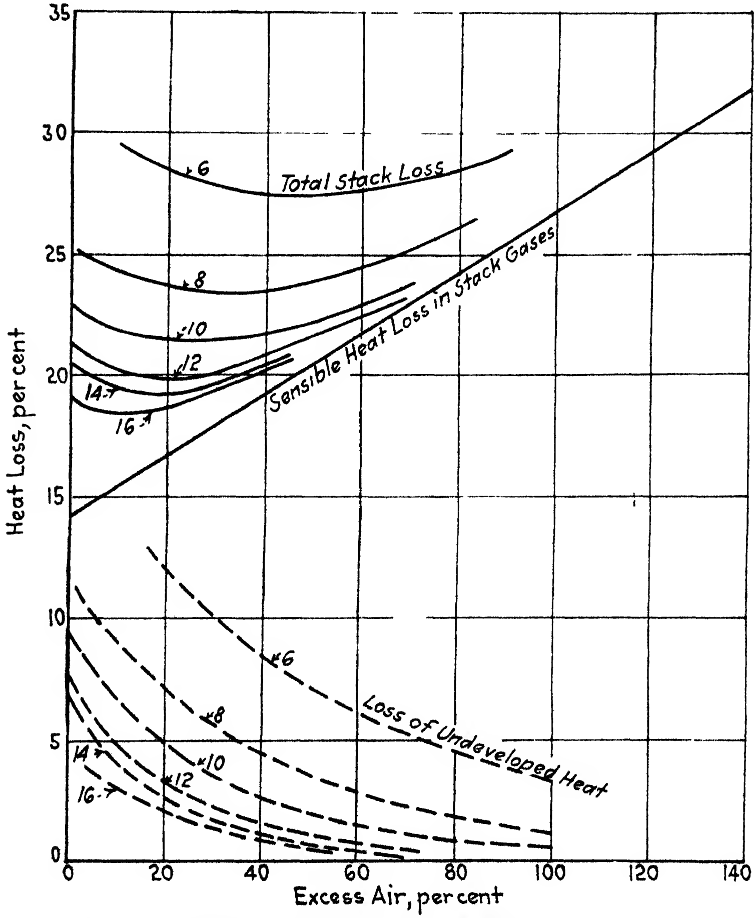

FIG. 110.—The variation in sensible-heat loss, undeveloped-heat loss, and total stack loss with excess air. The lower series of curves show the loss of undeveloped heat in furnaces having the ratio of volume of combustion space to grate area specified on the curve. The *total* stack loss for furnaces having these ratios is shown in the upper series of curves, which represent the sum of the sensible-heat and the undeveloped-heat losses.

Therefore, the total sensible heat loss with 100 per cent excess air is

$$\frac{150{,}000 + 172{,}900}{12{,}164 \times 100} = 26.6 \text{ per cent.}$$

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* Air composed of O_2 and N_2 will have the same sensible heat as the N_2 in the preceding calculation.

Then, plotting the per cent excess air against the per cent heat loss and drawing a straight line between the two points just calculated as in Fig. 110, it is possible to read off the sensible heat loss for any desired per cent excess air.*

The coal analysis used in making the above calculations is near enough to that of the Illinois coal of the Bureau of Mines investigation for the data in Fig 105 to be used directly. Therefore, from Fig 105, the per cent excess air and the per cent of undeveloped heat can be read off for each value of the ratio of volume of combustion space (cubic feet) to grate area in square feet (V/A) at 25 lb. of coal per sq. ft. of grate area per hr. Thus, with a ratio of $V/A = 6$, reading from the plot, the values of excess air for each degree of undeveloped heat are found to be as given in Table LXXXI. Using the values of excess air thus obtained, the corresponding sensible heat loss can now be read from the Sensible Heat Loss curve in Fig. 110. The *total* stack loss is the sum of the undeveloped heat loss and the sensible heat loss, as indicated by Table LXXXI.

TABLE LXXXI

HEAT LOSSES WITH VARYING EXCESS AIR AND COMPLETENESS OF COMBUSTION

(Illinois coal; 25 lb. per sq. ft. per hr.; ratio of combustion space (cubic feet) to grate area (square feet) = 6)

Per cent undeveloped heat	Per cent excess air	Sensible heat loss, per cent	Total stack loss, per cent
0.5	68	22.4	22.9
1	52	20.5	21.5
2	35	18.4	20.4
3	21	16.7	19.7
5	10	15.4	20.4
7	1	14.4	21.4

A series of values of undeveloped heat and of total heat loss for given percentages of excess air obtained in this way are plotted in Fig. 110. The figures included on each line represent the value of the ratio of combustion space to grate area. *These ratios, however, have been multiplied by a factor of* 2 *in order to*

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* It should be noted that the curves in Fig. 110 apply *only* to the coal under consideration and to the specified conditions of stack temperature.

make the Bureau of Mines data more nearly conform to average plant conditions.[*] The variation in undeveloped heat with varying excess air for the different values of the corrected ratio of volume to area is represented by the lower series of curves (dotted lines). The *total* stack loss (*i.e.*, the sum of the undeveloped heat and the sensible heat loss) is represented by the curves in the upper part of the figure (full lines).

The minimum values of total stack loss (not including the constant loss due to heat used in evaporating the water in the coal) for each value of the corrected ratio of volume to area may then be read from Fig. 110, as shown in Table LXXXII.

TABLE LXXXII

MINIMUM HEAT LOSSES WITH VARYING RATIOS OF COMBUSTION SPACE TO GRATE AREA

Corrected ratio of volume to area ..	6	8	10	12	14	16
Minimum stack loss	27.4%	23.4%	21.4%	20.6%	19.2%	18.4%
Total heat loss = stack loss + 4.1 per cent loss‡ due to latent heat in water vapor + 15 per cent radiation, etc .	46.5%	42.5%	40.5%	39.7%	38.3%	37.5%

‡ A certain amount of the heating value of the coal must be expended in vaporizing the 2.71 mols or 2.71 × 18 = 48.8 lb. of water that are present in the stack gases from each 100 lb. of coal. The heat of vaporization of water is 1000 B.t.u. per lb., so this loss can be calculated as $\frac{48.8 \times 1000}{12{,}164 \times 100} \times 100 = 4.1$ per cent.

The furnace efficiencies[†] for different values of the corrected ratio of combustion space (cubic feet) to grate area (square feet) are 100 minus the total heat loss as shown in Table LXXXIII.

TABLE LXXXIII

MAXIMUM EFFICIENCIES WITH VARYING RATIOS OF COMBUSTION SPACE TO GRATE AREA

Corrected ratio of volume to area.................... ...	6	8	10	12	14	16
Efficiency..................	53.5%	57.5%	59.5%	60.3%	61.7%	62.5%

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[*] The factor of safety to be used depends on conditions, as previously mentioned.

[†] These efficiencies are not to be taken as the efficiencies to be obtained with Illinois coal in *boiler* practice.

The amount of coal which must be burned per hr. to meet the heat demands of the furnace is $\frac{22{,}000{,}000 \times 100}{12{,}164 \times \text{efficiency}}$. The weight of coal per hr., the grate area (which equals the pounds of coal per hr. divided by the rate, 25 lb. per sq. ft. per hr.), and the volume of combustion space to give maximum efficiency at various corrected ratios of combustion space to grate area are shown in Table LXXXIV.

TABLE LXXXIV

POUNDS COAL PER HR., GRATE AREA, AND VOLUME OF COMBUSTION SPACE TO GIVE MAXIMUM EFFICIENCY

Corrected ratio of combustion space to grate area	6	8	10	12	14	16
Pounds coal per hr	3390	3150	3045	3000	2935	2895
$\frac{\text{Pounds per hr.}}{25}$ = grate area, square feet	135	126	122	120	117	116
Volume of combustion space, cubic feet	810	1010	1220	1440	1640	1856

The coal cost per day is $\frac{\text{Pounds per hr.} \times 24 \times \$7.50.}{2000}$.

This cost calculated for the various corrected ratios of combustion space to grate area is shown in Table LXXXV.

TABLE LXXXV

COAL COST FOR VARYING RATIOS OF COMBUSTION SPACE TO GRATE AREA

Corrected ratio of combustion space to grate area	6	8	10	12	14	16
Cost of coal per day	$305.00	$283.50	$274.00	$270.00	$264.00	$260.50

The fixed charges, including the 25 per cent that must be charged off on account of the interest the company can earn on money invested in its regular line of business, including also the 20 per cent per year for depreciation (*i.e.*, 100 per cent in five years) and 10 per cent for maintenance,* are:

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* The furnace is to be operated 300 days a year.

$$\frac{(7500 + 20X)(0.25 + 0.20 + 0.10)}{300} = (7500 + 20X)(0.00183).$$

Fixed charges calculated in this way, and the total cost (*i.e.*, coal cost plus fixed charges) are shown in Table LXXXVI.

TABLE LXXXVI

FIXED CHARGES AND TOTAL COST WITH VARIOUS RATIOS OF COMBUSTION SPACE TO GRATE AREA

Corrected ratio of combustion space to grate area	6	8	10	12	14	16
Fixed charges, dollars per day	$ 43.50	$ 50.80	$ 58.50	$ 66.50	$ 74.00	$ 82.00
Total cost,* dollars per day (fixed charges plus coal)	348.50	334.50	332.50	336.50	338.50	342.50

* To simplify, the effect of changing grate size on cost of furnace is omitted. Furthermore, since labor and other charges are constant and do not affect this problem they have been omitted. Attention should be called to the small change in cost of operation with a change in furnace size *when each size furnace is operated at its particular optimum efficiency*. This is due to the compensating character of the two stack losses.

Table LXXXVI shows that the minimum cost is, therefore, with a furnace having a ratio of volume of combustion space to grate area equal to 10. From Fig. 110, with a value of V/A of 10, the minimum stack loss is found to be 21.4 per cent, using about 30 per cent of excess air. Under these conditions there is a loss of about 3.6 per cent due to undeveloped heat. The grate area from the above calculations is 122 sq. ft. and the volume of the combustion space is 1220 cu. ft. The first cost of the furnace would be $7500 + 20 × 1220 = $31,900.

BIBLIOGRAPHY

1. KREISINGER, OVITZ, and AUGUSTINE: U. S. Bur. Mines, *Bull.* 137 (1917).
2. KREISINGER, AUGUSTINE, and KATZ: U. S. Bur. Mines, *Tech. Paper* 139 (1918).
3. KREISINGER, AUGUSTINE, and OVITZ: U. S. Bur. Mines, *Bull.* 135 (1917).
4. M. E. AUDIBERT: *Rev. ind. minèrale*, **37**, 351 (1922).
5. GLADDING and ENTWISTLE: Mass. Inst. Tech. *Thesis* (1924).

CHAPTER XIII

THE OPERATION OF HAND-FIRED FURNACES

The desirability of thin fuel beds, from 4 to 5 in. deep, has been pointed out in the previous chapter. When firing coal by hand, however, due to the unavoidable formation of blowholes in the fuel bed, and the unevenness of ash distribution, it is necessary to carry a somewhat thicker fuel bed. In hand-fired furnaces, therefore, the thickness of fuel bed must be increased from 4 to 5 on up to 6 or 8 in. Usually fuel beds thicker than 8 in. are not only unnecessary but they also cause low efficiency, due to the increased formation of CO, and increase the tendency for the formation of clinkers. Moreover, the high drafts which must be used with thick fuel beds cause air leakage through the furnace setting, thus increasing the sensible-heat stack losses.

The previous chapter also showed the large amount of heat lost if the volatile matter of the coal is not burned. Moreover there is an additional, indirect loss, since resulting soot depositing over boiler surfaces decreases the heat-absorbing capacity and the temperature of the stack gases thus is increased. Therefore, it is necessary to know the correct firing methods, in order to obtain the full heating value from the volatile matter of the coal and to prevent the formation of smoke.

The best methods of firing coal by hand and means of reducing losses in hand-fired furnaces have been ably discussed in a pamphlet issued by the U. S. Bureau of Mines.[1] The more important features developed in this pamphlet are summarized in this chapter under the headings: (1) Placing Coal on the Fire, (2) Cleaning Fires, (3) Frequency of Firing, (4) Clinker Formation, (5) Draft Regulation, and (6) Thickness of Fuel Bed.

Placing Coal on the Fire.—The two most commonly used methods of firing coal by hand are: (1) distributing the green coal over the entire fire bed, taking particular care to cover thin spots, and (2) coking the coal on a coking plate at the front of the furnace and then distributing the coke over the fire. The first method, covering the thin spots in the fuel bed, is the easier of the two and is the one most generally used. It has the disadvantage, however, that with badly caking coals the raw fuel may fuse, making it practically impossible for sufficient primary air to come up through the spot where the coal was fired. Moreover, the accurate placing of coal on all the thin spots in the fuel bed requires considerable skill. In the second method, coking the coal on a coking plate at the front of the furnace, the volatile matter is driven off in the path of the secondary air coming in through the fire doors and the combustion of the gaseous hydrocarbons is further aided by passing the rich volatile gases across the highly heated coals, thus preventing chilling before combustion is complete. Then, if the coal cakes badly, the coke may be broken up and distributed over the thin spots in the fuel bed. This method has the disadvantage, however, that excess air is drawn into the furnace while the coke is being broken up, and in distributing the coke it is difficult to prevent stirring up the fuel bed, thus causing clinker formation by bringing the ash into a zone of incandescent fuel. In general due to the fewer operations involved, the method of spreading raw coal over thin spots in the fuel bed is the one most widely adopted, and good efficiencies can be obtained with careful firing.

In firing by either method, one of the most important considerations is to cover the thin spots in the fuel bed. Coal on the grate never burns uniformly over the whole surface, and thick and thin spots are bound to occur. Where clinker forms, or the coal cakes, less air can get through the fuel bed and thus combustion is less rapid, causing a thick spot to form. The coal should always be thrown on the thin spots, which usually can be recognized by their brilliancy. If the thin spots are not filled up, air leakage results. In other words, the proper method of firing is *not* to spread the coal evenly over the whole fire bed at each firing, but rather to concentrate it where the bed has become thin.

Fresh coal always should be fired in thin layers, because if the fireman attempts to fill up the deep holes in the fuel bed at one firing, a crust will form and a new high spot will result. If no coal is placed on the high spot for one or two firings, the crust will crack, and air can get through to allow a return to the normal condition. Of course, if the thick spot is due to an accumulation of ash or clinker, the flow of air will not return to normal until the clinker is taken out with a fire tool.

If the thin spots in the fuel bed are not properly covered, they may burn through entirely, allowing an inordinate amount of excess air to come up through the grate. Furthermore, when fresh coal is placed on these spots, some of it will fall through the grate, thus causing an additional loss. Therefore, in order to prevent the formation of holes, the interval between firings must be short. If bad thick and thin spots have been allowed to form, however, the quickest way of leveling the fuel bed is to break up the caked coal forming the thick spots and spread it over the depressions. This method has three disadvantages: (1) It causes extra work for the fireman, (2) while the door is open for cleaning a large and unnecessary excess of air is admitted, and (3) the leveling disturbs the ashes and is liable to start clinker formation. By placing the coal where it is needed, this extra work may be avoided, the fuel bed may be kept level for hours, and clinker may be greatly lessened.

Cleaning Fires.—In hand-fired furnaces the coarse pieces of ash and the clinker must be removed at intervals to prevent obstruction of the air spaces. Two general methods of cleaning fires are the "side method" and the "front-to-rear method." In the side method one side of the fire is cleaned at a time, the good coal being pushed away from the clinker and ash. The clinker and the ash are then removed and the process repeated with the other side of the furnace. When the burning coal from the second side is pushed back onto the clean portion of the grate, fresh coal must be added in order to have sufficient fuel to cover the whole grate after the cleaning.

In the front-to-rear method the burning coal is pushed against the bridge wall, instead of to one side. After the clinker is removed, the burning fuel is pulled back on the grate and fresh coal charged. This method is somewhat quicker than the side method, but is not quite so thorough, since some clinker is always left at the bridge wall.

The larger the air spaces in the grate the greater the amount of ash that falls through into the ash pit, and, consequently, the longer are the intervals between cleanings. For most bituminous coals the air spaces should be at least ½ in. wide and for some coals they should be ¾ in. Since almost all bituminous coals cake on the grate, no small pieces remain loose in the furnace, and thus, even with large air spaces, only a very small amount of combustible falls through into the ash pit.

Frequency of Firing.—Soft coal should be fired at short intervals, in small amounts, depending on the size of grate and the amount of draft. When the draft in the uptake is 1 in. of water, 2 to 2½ lb. of coal should be fired (per firing) for each square foot of grate. Thus from 100 to 125 lb. of coal, or from six to nine shovelfuls, should be charged on a 48-sq. ft. grate. If the draft is high, coal should be fired about once in 3 min. and with a slow fire about once every 8 min. On the average, the intervals between firings should be about 5 min. and never longer than 10. Small, frequent firings make the ratio of coal to air supplied more nearly constant; this is important, since in hand-fired furnaces the air supply is approximately constant. Furthermore, coal does not cake as badly when fired frequently, since with small firings the thin layer of fresh coal is burned through before a hard crust can form.

One of the principal reasons for frequent firing is to maintain a gradual distillation of volatile matter. In the preceding chapter, it was pointed out that when green coal is charged on a furnace grate the coal is rapidly heated and from 40 to 60 per cent of the combustible is given off as volatile matter and CO. This process of distillation is usually nearly complete in from 2 to 5 min. after firing. It is evident, then, that, if the air supply is correct when the volatile combustible matter is being given off, it must be *incorrect* during the period after the volatile has been consumed. Thus theoretically a large amount of air should be admitted while the distillation is taking place, whereas during the latter part of the interval between firings only a small amount of air need be added. Such constant variation of the air supply, though possible with the aid of automatic devices (such as mechanically controlled steam jets), is difficult to maintain, and the best results (even with these automatic devices) are obtained by small, frequent firings to make the distillation of volatile matter more nearly uniform and proportional to the amount of air supplied.

This is shown graphically in Fig. 111. In case *A* (upper part of figure) the firings are 5 min. apart, while in case *B* (lower part of figure) firings are 15 min. apart. The same amount of coal is fired, and therefore the same total amount of volatile matter is distilled in both cases. The amount of this volatile matter is represented by the shaded areas under each part of the saw-tooth curve. The horizontal dotted lines represent the constant air supply sufficient to burn the volatile matter under each line. Thus, for a given air supply, the shaded area above it represents a loss due to unburned volatile matter. The unshaded area under the line represents a loss due to excess air. As was shown in the preceding chapter, as the air supply is increased, the loss due to unburned volatile matter decreases, but at the same time the loss due to excess air is increased. The optimum

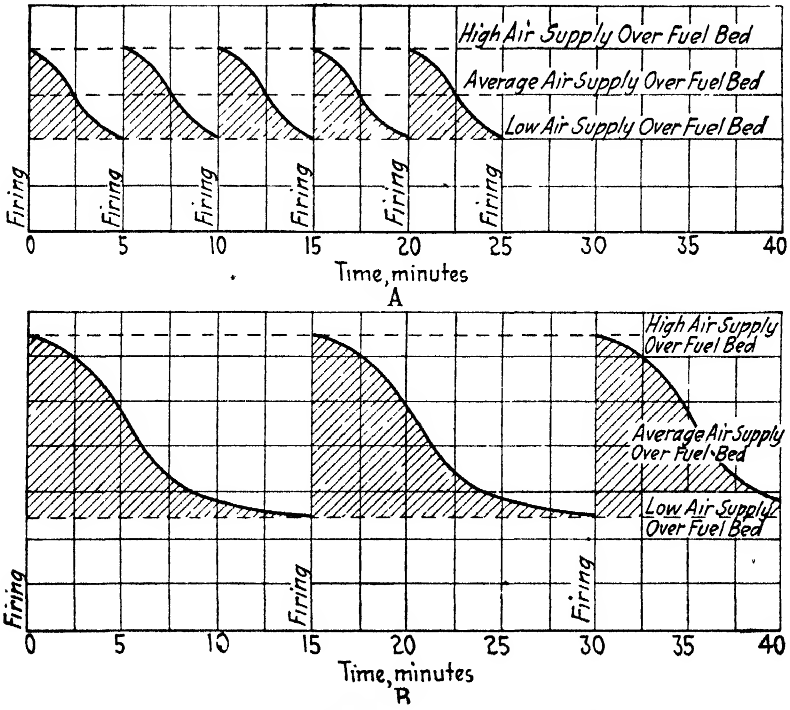

FIG. 111.—Relation of rate of distillation of volatile matter and necessary air supply. A shows the air supply and the distillation of volatile combustible when the firings are 5 minutes apart, B when the firings are 15 minutes apart. In both cases the amount of coal fired per hour and the weight of volatile combustible distilled from the coal are the same. This weight of volatile combustible is represented by the shaded area under the saw-tooth curve.

operating condition (*i.e.*, where the sum of these two losses is a minimum) is with the air supply shown by the middle dotted line marked "Average air supply over fuel bed." The total net loss is much less in *A* than in *B*, that is, small frequent firings are more economical than large firings at long intervals. It is also possible with many coals to slow up the rate of volatilization by moistening or "tempering" and thus again more nearly approximate a constant evolution of volatile matter.

Clinker Formation.—Clinker is fused ash and entrapped combustible material, and any ash will fuse if heated to a high enough temperature. It has previously been pointed out (Chap. II) that the fusing point of the ash depends on its chemical composition, although the exact influence of each constituent is unknown. It is known, however, that ash in an oxidizing atmosphere fuses at a higher temperature than in a reducing atmosphere, since some constituents of the ash, such as oxide of iron, are reduced to oxides with lower melting point. The melting points of three ashes are given in Table LXXXVII.

TABLE LXXXVII

EFFECT OF ATMOSPHERE ON FUSION POINT OF ASH

Ash from	Fusing temperature in reducing atmosphere	Fusing temperature in oxidizing atmosphere
Pocahontas	2509°F (1376°C.)	2642°F (1450°C.)
Mingo bed, Tennessee	2606°F (1430°C.)	2642°F (1450°C.)
Coal Creek bed, Tennessee	2395°F (1313°C.)	2489°F (1365°C.)

Clinkering is commonly caused by (1) thick fires, (2) excessive stirring of fires, (3) burning coal in the ash pit, (4) much slack in the coal, and (5) closing the ash-pit doors. In other words, any factor which tends to markedly increase ash temperature tends to promote the formation of clinkers.

The use of a thick fuel bed is the most common cause of clinker. With a thick fire less heat escapes from the fuel bed by radiation and consequently the ash is subjected to higher temperatures, and clinkering may result. The gases passing through the fuel bed contain more CO, *i.e.*, there is a reducing atmosphere and, as shown by Table LXXXVII, the fusion of ash can take place at a lower temperature.

If the fire is stirred frequently, as, for example, to break up the crust, the underlying layer of ashes is lifted onto the burning coal and clinker results. Similarly, burning coal in the ash pit, which may either come from holes in the fuel bed or from leveling of fires, heats and fuses the ash on the grate. Closing the ash-pit door cuts down the amount of air through the grate and thus permits the heating and melting of the ash.

Prevention of Clinker.—Clinker formation may generally be prevented by observing the following precautions, all of which tend to reduce the temperature of the ash:

1. Use thin fires and keep the fuel bed level by placing fresh coal on the thin spots. Disturb the fuel bed as little as possible.
2. Fire the coal in small charges, to minimize crust formation.
3. Do not allow burning coal in the ash pit. It is a good plan to keep water in the pit, or, if it is not water-tight, to blow waste steam through the grate if the coal clinkers *badly*.
4. Keep the ash-pit door open, and regulate the draft by means of the damper.

Draft Regulation.—In most plants the load fluctuates during the day, and this varying demand is met by burning more or less coal. It has already been shown that the amount of coal that can be burned depends on the amount of primary air forced through the fuel bed. The proper air supply for the combustion of a given amount of coal is controlled by regulation of the draft; that is, a high draft is required for a rapid rate of combustion, and a low draft for a slow sluggish fire. It is, of course, possible to meet a low demand by keeping up a high rate of combustion with a high draft, provided a large excess of air is drawn in over the fuel bed, but this is extremely wasteful and allows a high stack loss.

The best method of draft regulation is to use a damper in the uptake or breeching and there should be one for each furnace. The connections should be so arranged that the fireman can regulate the damper from the floor in front of the boiler; otherwise the draft will not be properly regulated. The draft should never be regulated by closing the ash-pit doors, as this causes an increase in secondary air at a time when a decrease is desired. The ash-pit door and the grating over the firing door should be adjusted so that each admits half the air. Closing the ash-pit door prevents air from coming up through the grate and allows more air to be sucked in over the fuel bed. This is harmful in two ways: (1) when the air supply through the fuel is cut down, the ash on the grate and the grate itself become heated, which results in troublesome clinker, and repeated heating of the grate may cause it to warp; and (2) when the amount of air coming through the grate is decreased, more air is drawn into the furnace over the fuel bed and the sensible heat lost up the stack is increased. Figure 112 shows diagrammatically the correct and the incorrect methods of air admission. In the correct method, shown at the left of the figure, the ash-pit door is left open, the air is controlled by means of the stack damper, enough primary air being admitted through the ash-pit door to cool down the ashes on the grate, and enough secondary air being admitted through the fire door to burn efficiently the gases rising from the top of the fuel bed. In the incorrect method, shown at the right of the diagram, the air is controlled *not* by the damper in the uptake, but by closing the ash-pit door. This results in heating and fusion of the ash on the grate, and since an extra amount of air is drawn in through the fire door (more than is required to burn the volatile matter) there is a large and unnecessary waste of sensible heat up the stack.

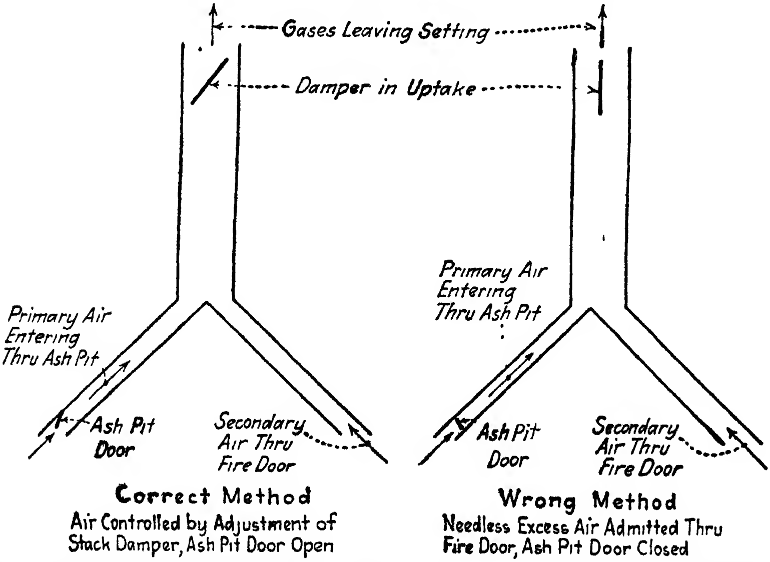

FIG. 112.—Correct and incorrect methods of draft regulation in a hand-fired boiler furnace.

The amount of the draft cannot be determined from the position of the damper, and the draft should always be regulated with the aid of a draft gage. If the resistance of the fuel bed is kept constant, the amount of air supplied, and consequently the rate of combustion, will be roughly proportional to the square root of the draft over the fire. Thus, if the draft over the fire is $\frac{1}{2}$ in. of water at 25 lb. of coal per sq. ft. of grate area per hr., then, with the same resistance of fuel bed, with a draft 1 in. of water, the rate of combustion will be $25 \times \sqrt{\frac{1}{\frac{1}{2}}} = 25 \times \sqrt{2} = 35$ lb. per sq. ft. of grate per hr. If the draft is low when the damper is wide open, there may be holes in the fire bed, and if the draft is very high there usually is an accumulation of clinker on the grate. Thus, the use of a draft gage not only allows the proper draft regulation to maintain a desired rate of combustion and air supply, but also serves to indicate the condition of the fire.

Thickness of Fuel Bed.—The proper thickness of fuel bed varies somewhat with the draft available and the kind of coal used. In furnaces having a draft of 1 in. of water, where run-of-mine coal is used, the fuel bed should be between 4 and 8 in. thick. In locomotives, on the other hand, the draft in the fuel box is high (frequently up to 10 in. of water) and a 6- to 10-in. fuel bed gives good results. Under ordinary furnace conditions, however, there is no reason for carrying a fuel bed thicker than 8 in. If the coal is fine, only a 4-in. bed is required, but with a coarse coal which contains little or no fines the bed may be 8 in. thick. The proper maintenance of a thin fire, however, requires considerable skill, the main difficulty being to prevent blowholes in the bed.

If the fuel bed is thicker than necessary, the resistance to air flow is too large and consequently the rate of combustion and the furnace capacity are decreased. If coal is properly fired in a thin bed it is often possible to keep up steam pressure when it would be impossible to do so using a thick fire. Furthermore, a thick fuel bed is the most common cause of clinkering troubles.

Summary.—Methods of smoke prevention, methods of preventing clinker formation, and means of obtaining maximum efficiencies in hand-fired furnaces may be summarized as follows:

Methods of Smoke Prevention in Hand-fired Furnaces.—In order to prevent smoke formation, the rich, volatile gases rising from the top of the fuel bed should be mixed with air and then passed over the incandescent coals in order to promote combustion before the gases are chilled. The coal should be fired in small amounts at short intervals, since in this way only small amounts of volatile matter are given off at any one time. The use of automatic devices which cause the admission of an extra quantity of air just after firing (when volatile matter is being given off and more air is needed) will also aid in smoke prevention. Moistening or "tempering" the coal may slow down the rate of volatilization and thus make the evolution of volatile matter more nearly uniform.

Method of Preventing Clinker Formation in Hand-fired Furnaces. Clinker is fused ash and every effort should be made to keep the ash cool. A thin fuel bed should be carried in order to permit cooling of the ash by primary air coming up through the grates. It is important not to disturb the fuel bed, thus bringing the ash up into a hot reducing zone where clinker will form at a lower temperature. Burning coal must be kept out of the ash pit in order to allow the ash to cool. The draft should be regulated by a stack damper and the ash-pit door should be kept open in order to admit air under the grate. The practice of bleeding in steam under the grates, although necessary with some badly clinkering coals, lowers the efficiency and should only be used in case of necessity.

Methods for Obtaining Maximum Efficiency in Hand-fired Furnaces.—In order to obtain the maximum efficiency, the fuel bed should be kept thin and uniform, *i.e.*, the same depth all the time. The bed must be kept free from blowholes or thin spots, particularly around the edge of the grate, where air leakage may cause clinker formation. The coal should be fired frequently in small amounts in order to obtain the maximum heat from the volatile matter. The ash-pit door should be kept open, the opening in the fire door fixed, and the draft regulated by the stack damper. The excess of secondary air should be kept low and air infiltration through cracks in the furnace setting should be kept at a minimum. The breeching must be tight. The surfaces of the boiler, both inside and out, should be kept clean, as this increases heat absorption and lowers the temperature of the waste stack gases. Draft gages should be used, one connected just over the bed and another in the breeching; the first shows the pull through or resistance of the fuel bed and the second shows the total pull. This arrangement, therefore, tells the condition of the fire.

BIBLIOGRAPHY

1. H. Kreisinger: "Hand Firing Soft Coal under Power Plant Boilers," U. S. Bur. Mines, *Tech. Paper* 80 (1916).

CHAPTER XIV

MECHANICAL STOKERS AND FURNACES

The preceding chapters (XII and XIII) have shown the benefits derived from firing coal frequently in small amounts, and from distilling the volatile matter of the coal in an oxidizing atmosphere. Both of these conditions are made possible by mechanical stokers, and therefore they give less smoke and permit greater efficiencies than can be maintained with hand-firing. Making any process continuous usually increases the efficiency. Mechanical stokers were developed because of these advantages, because of the saving in labor, and the ability of stoker fired furnaces to carry high overloads, thus greatly reducing the cost of equipment necessary for handling peak loads. The question of ability to carry peak loads has become more and more important with the development of the present-day, high-capacity, large-size boiler furnaces.

When coal is fired intermittently (see Chap. XII) there is a loss due to unburned volatile matter, immediately after firing, and during the remainder of the interval there is an inordinate loss due to excess air. With mechanical stokers, on the other hand, the coal is fired uniformly in small increments, the fuel is heated slowly, and the volatile matter is driven off continuously. This uniform evolution of volatile matter permits the use of a steady draft, and therefore the constant admission of the proper amount of air for efficient combustion. Moreover, a thinner fuel bed can be carried with mechanical stokers, therefore giving a higher capacity per cu. ft. of combustion space. This more efficient utilization of the combustion space is possible because the gases coming off the top of a thin fuel bed are higher in CO_2, *i.e.*, a greater proportion of the combustion has taken place in the fuel bed, and therefore less CO and gaseous hydrocarbons remain to be burned in the space over the fire.

The principal methods of feeding coal and air in industrial furnaces are shown[1] diagramatically in Fig. 113. In hand-fired furnaces (*A*), primary air passes through a bed of ashes and comes in contact with partly burned fuel. All the oxygen is used up before the air reaches the green coal. Thus the volatile hydrocarbons distilling from the top of the fuel bed come into contact with gases depleted of oxygen, and therefore "hydroxylation" is made difficult and the deposition of soot results easily (see Chap. IX). In the overfeed type of mechanical stokers (*B* in Fig. 113) coal from a hopper is fed onto the sloping grate and is coked and distilled in the upper part of the furnace by the heat reflected from an arch. The resulting coke burns as it travels down the grate to the ash pit. With chain-grate or traveling-grate stokers (*C* in Fig. 113) the green coal is fed onto a moving horizontal grate. The coal is coked and distilled by heat reflected from an arch just as the grate enters the furnace; the resulting coke is burned during the travel of the grate through the furnace; the ash is dumped by the grate into the ash pit. In the underfeed type of mechanical stoker (*D* in Fig. 113) the green coal is fed from the underneath part of the fuel and distilled volatile matter passes up through the bed of incandescent coke. Thus, in overfeed stokers and in chain-grate stokers part of the primary air, and in underfeed stokers prac-

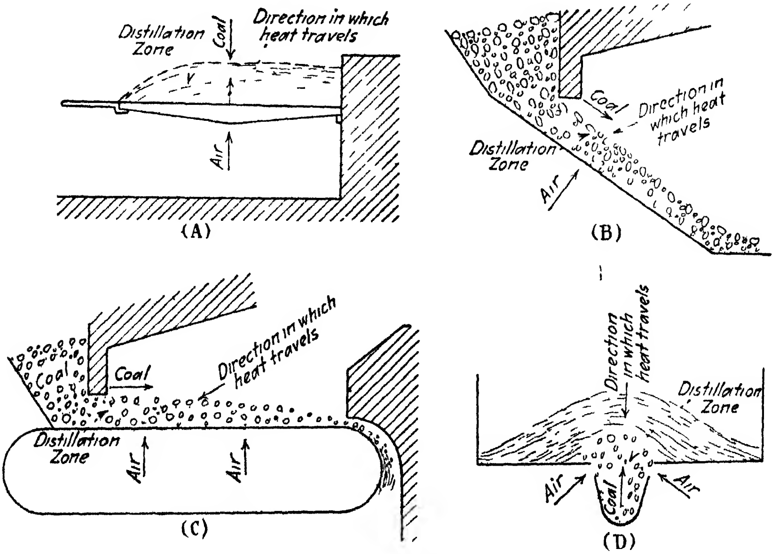

FIG. 113.—Principal methods of feeding coal in industrial furnaces. (A) Hand fired furnace. (B) Overfeed stoker. (C) Chaingrate stoker. (D) Underfeed stoker.

tically all the entering primary air, passes through the green coal as it is being distilled. The distillation of volatile matter therefore takes place in an oxidizing atmosphere, the entrance of oxygen into the hydrocarbon molecule (hydroxylation) is made easier and the tendency for "cracking" or soot deposition is decreased. Moreover, with mechanical stokers the fresh coal is heated gradually, and the volatile matter is driven off relatively slowly, again tending to promote hydroxylation. The amount of coal fired and the amount of air supplied the furnace may be regulated automatically with practically all types of stokers; therefore it is possible to feed into the furnace the coal and air properly proportioned, and in amounts depending on the load.

Historical Outline of Stoker Development.*—The early development of mechanical stokers was brought about almost entirely by the attempts to secure smokeless combustion. Although the idea of a continuous feed of fuel dates back to the days of alchemy, the first adaptation of the stoker principle to the steam boiler must be credited to James Watt, the inventor of the steam engine. In 1785 Watt patented a stoker-fed furnace, and in modified form applied the idea of progressively distilling the coal so that the gases could be burned in passing through the fuel bed. Probably the first truly mechanical stoker was brought out in England by William Brunton in 1819. Brunton patented and successfully applied a traveling-grate stoker consisting of a rotating circular plate. The original chain-grate stoker was invented by Juckes in 1841, and embodied most of the principles involved in present-day chain-grate stokers. The development of mechanical stokers operating on the overfeed principle dates from 1845, when Hall in England patented a stoker in which the fuel was pushed onto inclined reciprocating grate bars. Juckes, the inventor of the traveling-grate stoker, in 1838 brought out a stoker of the underfeed type. In 1889 Jones patented an underfeed stoker which, with certain changes, is still being operated successfully.

The adoption of these mechanical stokers in American plants was very slow, due for the most part to lack of knowledge of stoker operation and to the faulty design of many early stoker types. By about 1900, however, quite a large number of stokers had been installed and were operating in a fairly satisfactory

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* For complete and interesting details of the development of mechanical stokers, the reader is referred to P. K. CLARK[2] and ROBERT JUNE.[3]

manner at the low capacities demanded at that time. Gradually, however, the demands on furnaces increased until at the present time many boilers are operating for short periods at as high as 500 per cent of normal rating. These increasing demands have had a far-reaching effect, not only on stoker design, but also on the design of the furnaces themselves.

CLASSIFICATION AND DESCRIPTION OF MECHANICAL STOKERS

Stokers now used commercially fall into three main groups: (1) the chain or traveling grates, (2) stokers of the overfeed type, and (3) stokers of the underfeed type. In the subsequent discussion the characteristics of the three types of stokers* are described, representative makes of stokers are discussed, and the kind of coal used, the rate of combustion, and the draft requirements are given for each type. In considering individual makes of stokers, some of the main points to be considered are: dependability, freedom from manual labor, protection through *easily accessible* shear parts, etc.

Chain- or Traveling-grate Stokers.—Chain-grate or traveling-grate stokers consist of endless jointed webs which pass over sprockets at the front and rear of the furnace. Coal is fed onto the grate from hoppers, the grate (before entering the furnace) passing under an adjustable gate which regulates the thickness of the fuel bed. As the grate with its layer of coal enters the furnace, the coal is distilled and ignited by heat reflected from the hot portion of the fire to the front of the grate by an ignition arch. Combustion of the coal then proceeds gradually as the grate passes to the rear of the furnace. When burning soft coal care must be taken to prevent large lumps from rolling to the side of the grate and then, by burning out quickly, causing blowholes. The cinders are dumped automatically as the grate passes over

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* For a complete description of the important mechanical stokers manufactured in the United States the reader is referred to "Mechanical Stokers," by Worker and Peebles[4] and to the "Condensed Catalog of Mechanical Stokers."[5] These references give the dimensions of the various stokers, detailed explanations of the actuating mechanisms, etc. British stoker practice is discussed by David Brownlie,[6] and recent advances in American stoker design are discussed in reports of the Prime Movers Committee of the National Electric Light Association.[7] Much valuable information regarding mechanical stokers, and the design, operation, and characteristics of stoker-fired furnaces, is given by Marsh.[8]

the rear sprocket which is above the ash pit and the grate is cooled as it returns to the feed hoppers.

Tuyères under the grates permit the admission of the proper amount of air at any desired point, and air leakage through the ash pit or around the edges of the grate usually is prevented by suitable seals and ledge plates. It is necessary to prevent air leakage between compartments and to distribute the air evenly across the width of the grate. Usually only a little air is distributed to the front compartment, more to the next, and the maximum amount to the third. Sometimes secondary air is taken in through the ignition arch and throat to avoid stratification of gases. Chain-grate stokers are either natural draft or forced draft. The natural-draft type were the first developed and are now most widely used where the load is fairly steady and where only medium capacities are required. The forced-draft stokers of this type have higher capacities and can successfully burn a wider range of fuels than the natural-draft type. Descriptions of representative makes of chain-grate stokers are given herewith.

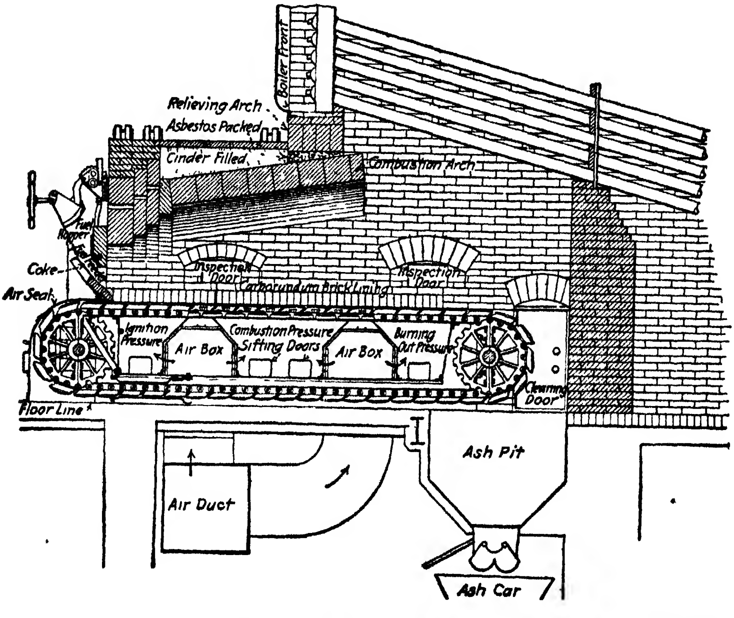

Fig. 114.—Sectional drawing of Coxe traveling grate. (*Combustion Engineering Corporation.*)

The Coxe stoker, shown in Fig. 114, is of the forced-draft traveling-grate type and is quite widely used for firing coke breeze and the smaller sizes of anthracite. It is also used in connection with soft coals from Ohio, Indiana, and Illinois The stoker consists of two side frames provided at the rear with bearings to hold the driving shaft and at the front end with idler-shaft bearings. Forced-draft connections are made through openings in the side frames. A sheet-steel box under the active portion of grate is divided into compartments by tuyères or air boxes as shown in Fig. 114. Air from the tuyères or air boxes is distributed through air compartments extending the whole

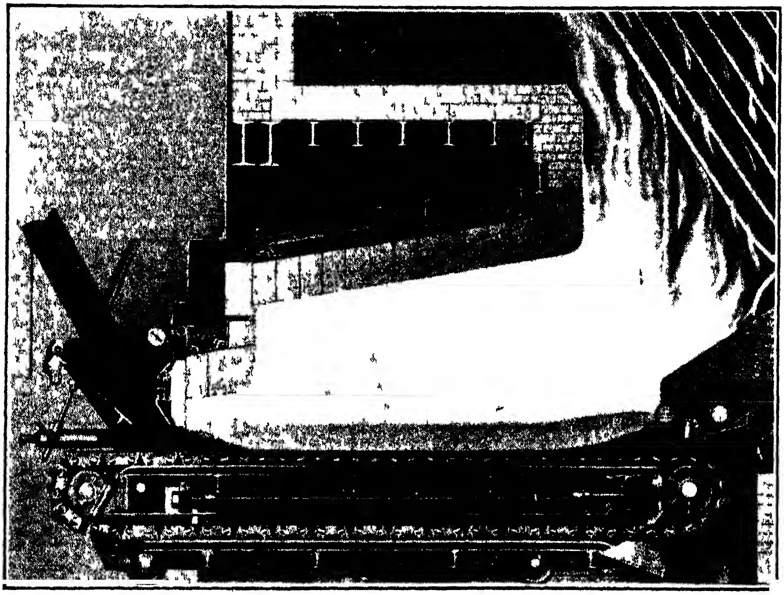

Fig. 115.—Green natural-draft chain-grate stoker (*Combustion Engineering Corporation*.)

width of the grate, the amount of air sent to each compartment being controlled by the adjustment of a slide damper in the tuyère sides. The number of tuyères and air compartments used depends on the length of fire desired.

The front part of the grate is covered by an ignition arch (see Fig. 114), which varies from one-half to three-fourths the length of the fire, depending on the size of stoker and the ease with which the fuel can be ignited. The height of the arch above the grate[5] is at least 3 to 5 ft., depending on and increasing with the length of the stoker. In burning steam sizes of anthracite a rear arch is used and the operation is such that hot, live coals from the rear are thrown forward onto the green coal, thus aiding ignition.

One type of arch suspension is indicated in Fig. 115, which shows the Green natural-draft chain-grate stoker. This figure also shows the use of a "fuel retarder," which is essentially a water-cooled member extending across the rear of the grate. This device is lowered to prevent any unburned fuel passing over into the ash pit and, conversely, is raised to permit all the ash formed to be dumped without building up any undue accumulation of ash on the grate.

Coals Suitable for Use on Chain-grate Stokers.—Chain grate stokers, in general, can be used to burn either high- or low-volatile, non-caking coals, steam-size anthracite, and coke breeze. They should not be used with caking coals, because, after the coal fuses and cokes, further heating causes this coke to shrink and crack. Much of the primary air then goes up through these cracks, resulting in too much air at the voids and not enough elsewhere. Thus with caking coals, chain-grate stokers give high stack losses, low capacities, and high ash-pit losses, since large lumps of coke are still unburned by the time the coal has traveled through the furnace.

The time the fuel is in the furnace depends on the length and rate of travel of the grate, because when the coal or ash reaches the end it is automatically dumped off. For this reason mixed fuels, such as a high percentage of bituminous coal with either coke breeze or anthracite, cannot be burned on a chain grate *if the coals are all mixed together.** The rates of combustion and ignition of the two fuels are different, and if the grate is set for the easy burning fuel the slow-burning coke breeze or anthracite will pass into the ash pit unburned. This, of course, results in a high ash-pit loss. On the other hand, if the rate of grate travel is set to permit complete combustion of the fuel of low combustibility, the capacity is greatly decreased and the quick-burning fuel is burned out long before the grate enters the ash pit. This results in blowholes in the fuel bed and consequently large quantities of excess air leak through, causing a high stack loss. The way in which such mixed fuels can be burned most

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* Small percentages of soft coal are mixed with anthracite coal and coke breeze to aid ignition.

successfully on a chain grate is in layers, with the fuel of low combustibility on the bottom, so that it will come into contact with the maximum amount of oxygen and thus burn first. A hopper for feeding separate layers of bituminous coal and coke is shown diagrammatically in Fig. 116. An adjustable gate is provided to regulate the amounts and relative thickness of the coke layer and the coal layer.

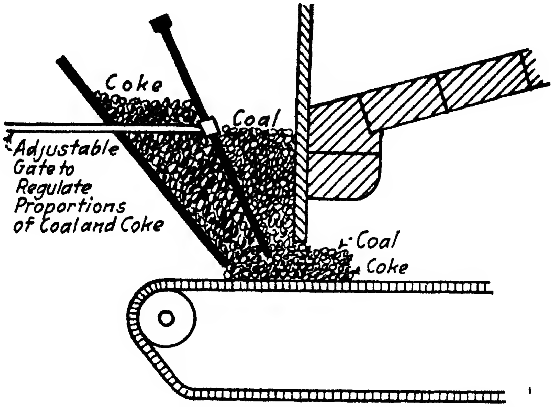

FIG. 116.—Hopper for feeding layers of coal and coke on a chain grate stoker. Diagrammatic

Chain grates can be used to burn coals with ash of low fusion point, because the fuel bed is not disturbed and the ash is kept in an oxidizing atmosphere, thus decreasing the tendency for clinker formation (see Chap. XIII). Therefore non-caking coals of any kind or amount of ash may be used. Traveling-grate stokers are to be preferred for burning coke breeze, anthracite, semi-anthracite, and high ash (above 10 per cent), clinkering, non-coking bituminous coals. Traveling grates for burning lignites and subbituminous coals should be equipped with liberal ignition arches to permit ready ignition of the coal.

Draft Requirements of Chain-grate Stokers.—Chain-grate stokers may use either natural or forced draft, the latter being necessary when the resistance of the fuel bed is high on account of the fineness of the coal or the small size of the grate openings. The draft for natural-draft stokers varies between 0.25 and 0.60 in. of water, depending on the load. In general, for natural-draft furnaces about 10 lb. of coal can be burned per sq. ft. of grate per hr. for each 0.1 in. of water draft through the fuel bed.

In forced-draft stokers, the pressure varies between 1 and 4 in. of water, depending on the load.

Rates of Combustion with Chain- or Traveling-grate Stokers.—The rate of combustion with chain-grate stokers averages about 30 to 35 lb. of coal per sq. ft. per hr. For carrying peak loads it is possible to maintain rates from 45 to 60 lb. per sq. ft. per hr. over short periods.

Regulation of Combustion Rate with Chain-grate Stokers.—The rate of combustion on chain grates can be regulated (1) by changing the height of the feed plate or the thickness of the fuel bed, (2) by changing the speed of the grate (20 to 120 ft. per hr. is the average speed), and (3) temporarily by changing the blast, either as to amount or distribution under the front or the rear of the grates. Thus for a light load the grate speed would be lowered and the air to the rear of the grate would be checked.

Overfeed Stokers.—Overfeed stokers are said to be "single-inclined, front-feed" stokers if the coal is fed from the front of the furnace only, or "double-inclined, side-feed stokers" if the coal is fired from both sides of the furnace. In most overfeed stokers the grate extends into the furnace at a rather steep incline, although the coal will not flow down this incline by gravity alone. A portion of the grate is fixed, but reciprocating sections are provided to help move the coal through the furnace. Coal is fed onto the grate from hoppers and is distilled and coked by heat radiated from arches above. Secondary air to burn the volatile matter is usually supplied through openings in the coking arch or through the lower part of the grate. In the front-inclined feed type the reciprocating motion of the grate bars carries the fuel to the rear, where it is dumped as cinder after combustion is completed. In the double-inclined, side-feed type of stoker the movement of fuel is toward the middle of the furnace and a central clinker grinder discharges the cinder into the ash pit.

Overfeed stokers will handle all types of coal from lignite to anthracite, though they are mainly used with midwestern, high-volatile, high-ash coals. They may also be used with refuse fuels, such as sawdust, tanbark, etc.

The Detroit V-type stoker, shown in Fig. 117, is of the double-inclined, side-feed type of overfeed stoker. Coal is fed into hoppers extending the whole length of each side of the furnace. From the hoppers the coal is moved to a coking plate, where it is distilled and the volatile matter is mixed with air which has been preheated by passing over the top of the radial arch shown in Fig. 117. From the coking plate the fuel passes onto the inclined grates, alternate bars of which are movable Motion imparted to these bars gradually advances the fuel down the grate surface, and, when combustion is complete, forces the cinder into the cinder grinder (shown in Fig. 117) The motion imparted by the rocking bars is greatest at the bottom to agitate the partially burned coal. The cinder grinder consists of a series of interlocking, toothed iron segments which crush any clinker and deliver it to the ash pit. This type of stoker is adapted for use in small- or medium-sized plants which have a relatively steady load.

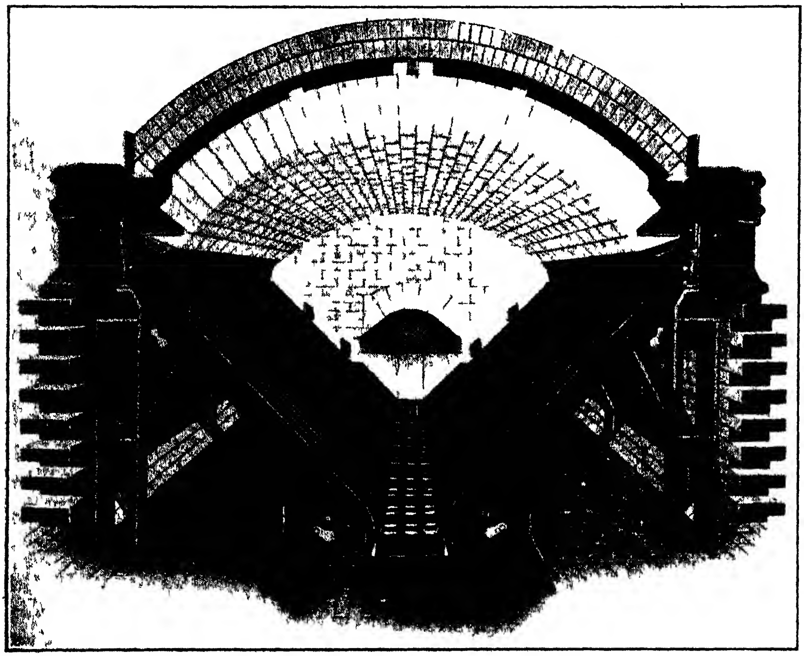

FIG. 117.—Detroit V-type overfeed stoker with radial sprung arch. (*Detroit Stoker Company.*)

The Westinghouse-Roney stoker shown in Fig. 118 is of the single-inclined, front-feed type. Coal from a hopper extending along the front of the furnace is forced mechanically at the desired rate onto a dead plate, where the coal is coked by heat radiated from the overhanging arch. From the coking plate the fuel is forced onto the grate bars, which by their rocking motion gradually cause the coal to move down toward the ash pit. A wide, partly balanced dump grate is built at the lower end of the stoker and means are provided for raising the grate (to prevent unburned coal from falling into the ash pit) when the dump grate is open Air leakage through the ash pit and around the dumping grate is kept as low as possible. Like the Murphy or Detroit stoker, the chief field of application of the Westinghouse-Roney stoker is in plants where the load is fairly uniform.

Draft Requirements with Overfeed Stokers.—About 50 lb. of coal per sq. ft. of grate area per hr. is the maximum rate of firing even with forced pressure and with caking coals the maximum rate is about 35 lb. per sq. ft of grate per hr. The average combustion rate is from 20 to 35 lb. per sq. ft. per hr. with free-burning coals.

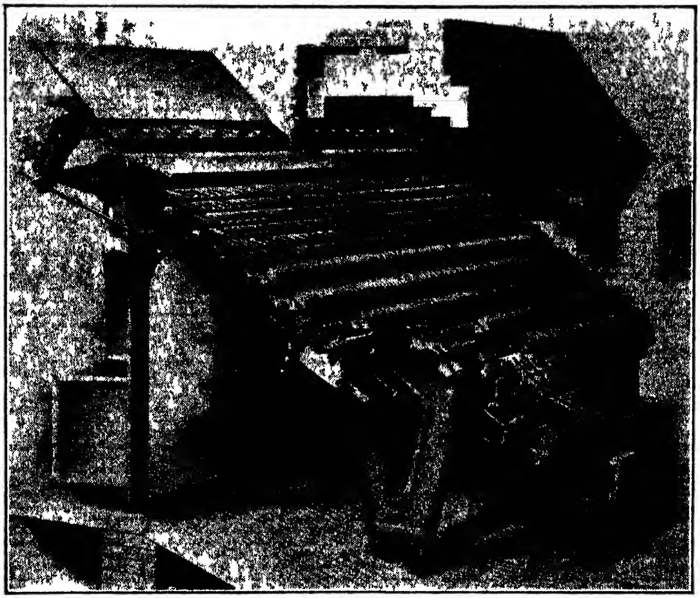

Fig. 118—Westinghouse-Roney overfeed stoker. (*Westinghouse Electric & Manufacturing Co*)

Regulation of Combustion Rate with Overfeed Stokers.—Overfeed stokers are regulated by changing (1) the rate at which coal is fed through the hoppers, (2) the rate at which ash is removed by the ash dumper, or (3) by changing the amount or distribution of the blast. Boilers may be operated continuously up to 200 per cent of rating without overheating the grates.

With this type of stoker more labor is required than on chain grates to keep the fire bed open and to make the ash pass through the ash grinder. The fires sometimes have to be cleaned of clinkers.

Hand-operated Stokers.—Hand-operated stokers are sometimes used in small plants where the installation of complete mechanical stokers may not be advisable, due to the high initial investment, or where the possible labor saving may be small. Stokers of this kind are mainly of the overfeed type.

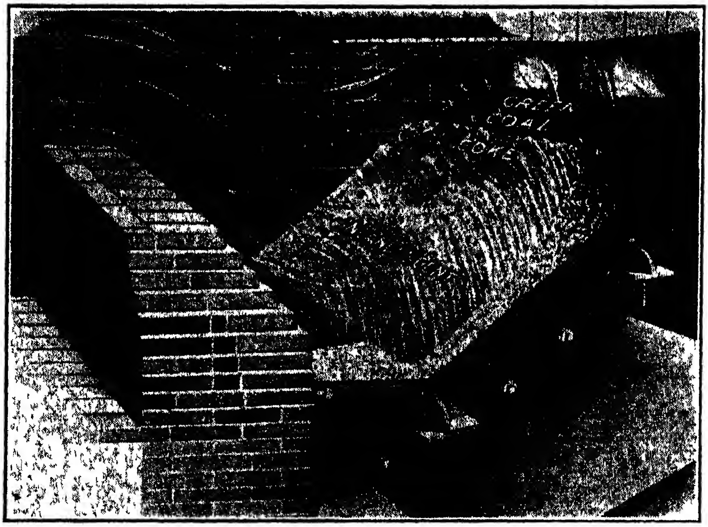

Fig. 119.—National hand-operated stoker. (*The M. A. Hofft Company.*)

In the stoker shown in Fig. 119 coal is fired by hand onto a coking plate just in front of the furnace doors. Here the coal is coked and then gradually moved down the sloping grate surface by the motion of hand-operated grate bars. This kind of stoker possesses many of the advantages of the complete mechanical stoker, such as the distillation of the coal in an oxygen-rich atmosphere and the relatively easy removal of cinder, but has the disadvantages that the doors must be opened to fire fresh coal, thereby chilling the furnace and increasing the stack loss while the doors are open.

The semi-mechanical stoker shown in Fig. 120 uses mechanical methods of feeding the coal but uses hand operation for the remainder of the process. Coal from the hopper *A* passes over a coking shelf *B* which extends the full width of the furnace and is pushed directly under the coking arch *C*. After being coked, the coal falls onto the grates, where the hand-operated stoker bars *G* distribute the fuel over the bed, and, as combustion progresses, gradually force the cinder on the drop bars *H* Volatile matter, driven off on the coking plate, is mixed with secondary air coming in through the ducts *D*, while a further supply of secondary air is admitted through the dead-plate shutters *E* and the dead-plate bars *F* to mix with the rich gases rising from the top section of the fuel bed

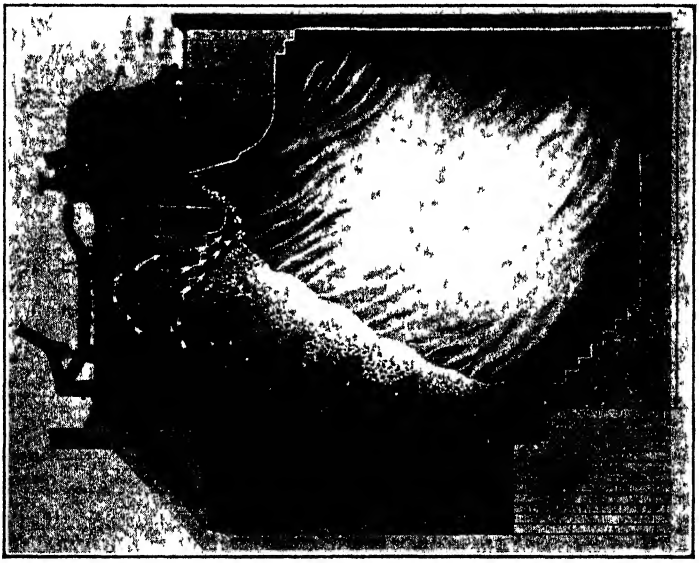

FIG. 120—Huber semi-mechanical stoker. (*Flynn & Emrich Co.*)

Underfeed Stokers.—Underfeed stokers are either horizontal or of the "gravity" type, in which the stoker is inclined toward the rear of the furnace. In these stokers green coal is fed to the *underneath* portion of the fuel bed, either forced in small increments by means of a power-driven ram, or continuously by means of a screw conveyor. In the lower part of the fuel bed the coal is distilled and coked in an oxidizing atmosphere. The hydrocarbons burn in an atmosphere rich in oxygen and the gases then work their way up through the hot incandescent fuel bed, thus lessening the tendency for the formation of soot. The coked coal enters the combustion zone from beneath, is burned, and when combustion is complete the remaining cinder is forced out onto side dead plates, side dump plates, or rear dump plates, depending on the kind of stoker. Underfeed stokers are of the single- or multiple-retort type, depending on the number of retorts to which the coal is fed.

In these stokers, the hottest part of the fuel bed is at the top, and, as pointed out above, there is practically no chance for soot formation. Thus the boiler tubes are kept clean, good heat transfer results, and high efficiencies are possible. Also, no arch is used, and a large area of boiler tubes is "in sight of" the red-hot top of the fuel bed. Therefore, a large amount of heat is transferred by radiation and the stack temperature remains low even with high overloads. This feature is highly important, since arches are the most expensive part of a furnace to keep in repair. Moreover, there is practically no smoke with high

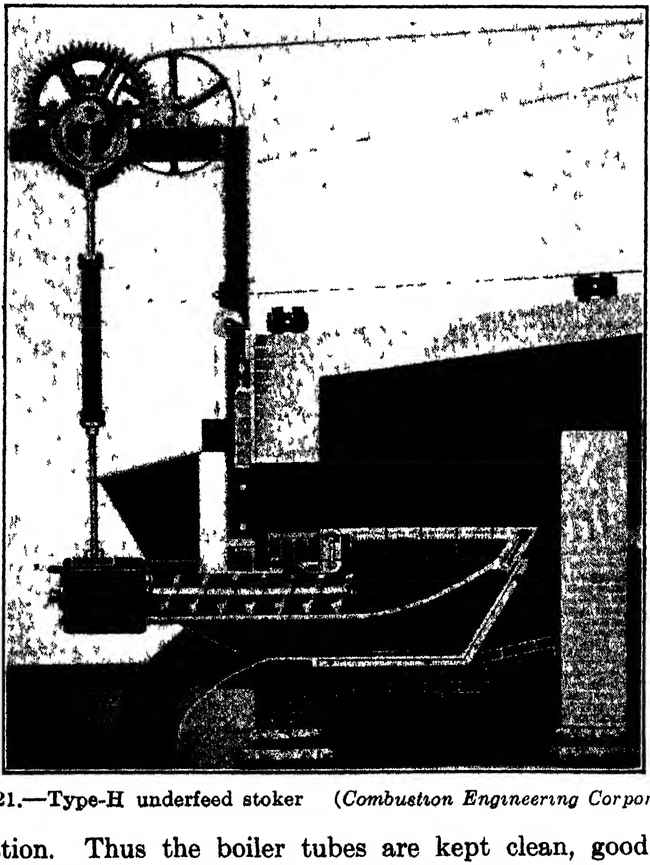

Fig 121.—Type-H underfeed stoker (*Combustion Engineering Corporation.*)

loads and this may be an important consideration in districts having strict smoke ordinances. For these reasons underfeed stokers are used mainly for high-duty loads and have been very successful in large plants for as high as 350 to 400 per cent of normal rating. However, due to the high overloads carried by these stokers and to the consequent high velocity through the fuel bed, a relatively large amount of cinder is blown out the stack. The development of large-size multiple-retort stokers has been an important factor in determining the trend of central-station design. In addition to the high capacities possible, stand-by and banking losses are low, and the efficiencies which can be maintained over a wide range of operating conditions are high.

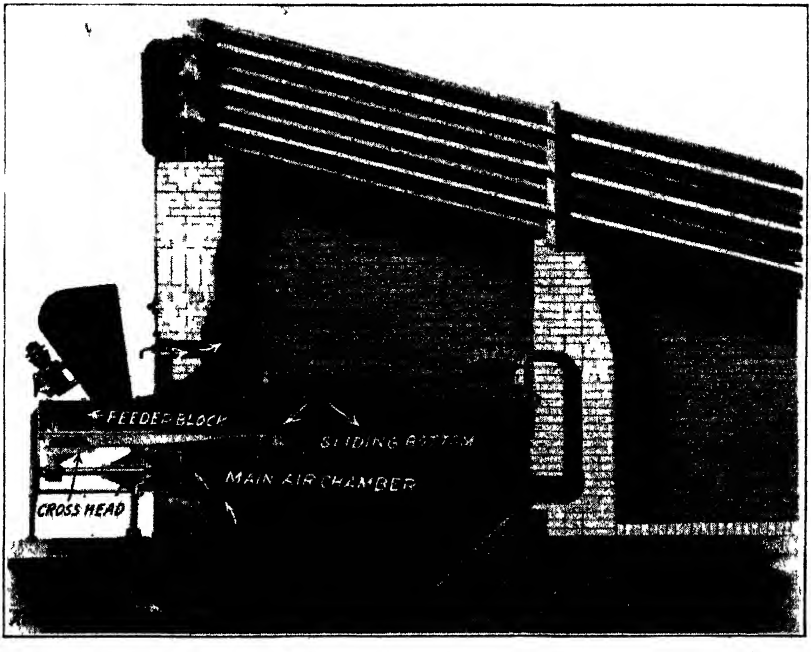

FIG. 122.—Type-E underfeed stoker. (*Combustion Engineering Corporation.*)

The type-H stoker shown in Fig. 121 is of the underfeed single-retort type. Coal from the hopper is fed in the stoker retort by means of a screw conveyor. The burning fuel is forced upward and spreads over side grates. Air from a wind box below the retort enters through tuyères, mixes with and burns the volatile matter from the coal and consumes the coke on the side grates. This type of stoker is mainly used on industrial and heat-treatment furnaces.

The type-E stoker shown in Fig. 122 is also of the underfeed retort type and is adapted for boilers from 150 to 600 hp. Coal from the hopper is fed into the retort by a pusher plate which forms the bottom of the retort. Grate bars on each side of the retort are alternately moving and fixed. The moving bars aid in forcing coal from the retort out onto dump trays. Air from a central duct beneath the retort is admitted through tuyères and through the grate bars themselves. It is also possible to admit pressure air under the dump trays in order to increase the effective grate surface or to burn out combustible remaining in the cinders. The dump trays are operated from the front of the furnace and deliver the cinders to the ash pit, which with this type of stoker need not be below the firing-floor level.

Fig. 123.—Jones AC underfeed stoker. (*Sanford Riley Stoker Company.*)

The Jones "A-C" stoker shown in Fig. 123 is of the inclined, multiple-retort type of underfeed stoker. Coal is fed into the retorts by the action of a steam-driven ram. Air from the windbox is admitted to the fuel bed through a system of tuyères between the retorts (see the diagram). As combustion progresses the fuel is gradually worked back to a dead plate at the rear of the furnace. Cinder from the dead plate is gradually forced onto balanced dump plates which deliver into an ash hopper below

The Taylor stoker shown in Fig. 124 is of the inclined multiple-retort type, such as are used with large furnaces. The stoker consists of inclined retorts divided by raised tuyères through which air is admitted. The fuel is fed into the furnace retorts and up over the tuyères by a main pushing ram. Subsequent motion of the coal down the sloping grate and onto dead plates is brought about by the motion of secondary rams which form the bottom of each retort. Air under pressure is admitted through the tuyères and through perforated blocks at the side of the stokers. These air-cooled blocks tend to prevent the adhesion of clinker to the side walls. The stoker shown in Fig. 124 is equipped with a rotary clinker grinder, which crushes the cinders and delivers them to the ash pit below. This makes operation more nearly continuous.

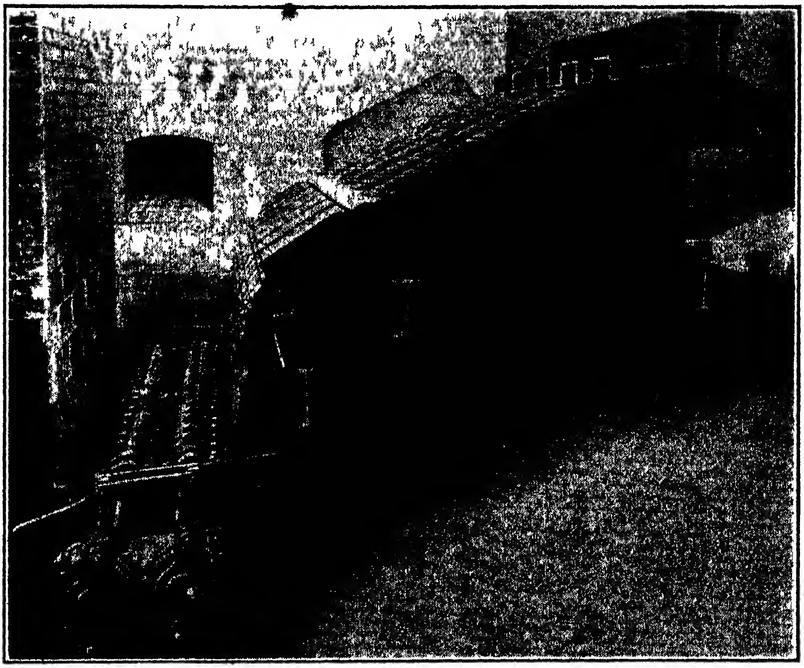

FIG. 124.—Taylor underfeed stoker with rotary ash discharge. (*American Engineering Company*)

The Riley stoker shown in Fig. 125, is of the inclined multiple-retort type. Coal from the hopper is forced into the inclined retorts by means of a plunger and is distributed in the furnace by a reciprocating motion imparted to the retort sides. The coal is distilled in the presence of air, which is admitted through a series of perforated grates. The coked fuel from the retorts gradually works its way down to "overfeed grates," where an additional supply of air is admitted to complete the combustion. Cinder and refuse are then discharged and crushed by rocker plates, which deliver into the ash pit.

A view, from the rear, looking forward, of the Frederick stoker is shown in Fig. 126. This stoker, like the Taylor and the Riley, is of the inclined multiple-retort type. The fuel-feeding and fuel-agitating parts of the stoker consist of banks of tuyères, which are cooled by the air which is admitted through them. As indicated in the diagram, there is a row of perforated blocks along the side of the stoker; these blocks are also cooled by the air passed through them and help to prevent the growth of clinker along the side walls. After passing out of the retorts the fuel is forced over a short section of overfeed grate bars and then onto the dump plate.

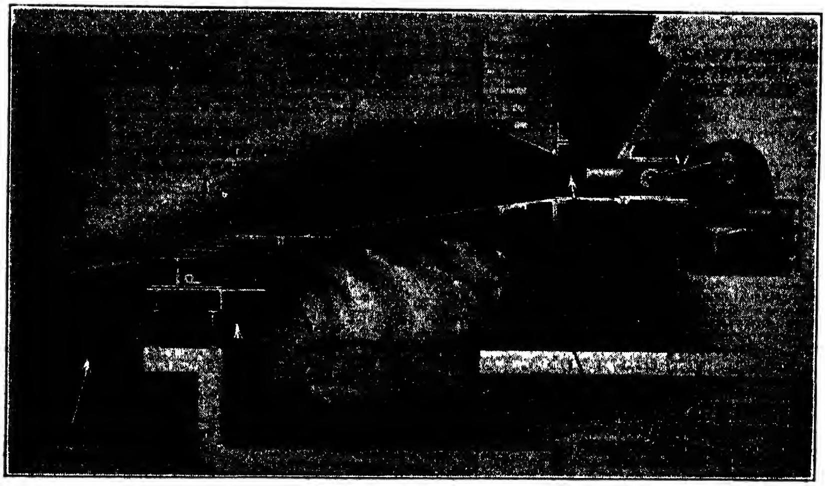

Fig. 125.—Riley underfeed stoker. (*Sanford Riley Stoker Company*)

Coals Used with Underfeed Stokers.—Underfeed stokers operate on the gas-producer principle and run with thick fuel beds. This thickness depends on the stoker and varies from 12 to 24 in. Low-grade, high-volatile, or coking coals are often used and slack coals or fines are common. This type of stoker is not satisfactory with coals of low-fusion-point ash, as the ash comes up through the highly incandescent fuel bed and is fused into clinkers.

Draft Requirements with Underfeed Stokers.—All underfeed stokers use forced draft, the normal pressure in the windbox being from 1 to 3 in. of water, though for high combustion rates the pressure may be increased to from 5 to 7 in. of water. In general, 1 in. of pressure is required for each 10 lb. of coal burned per sq. ft. of grate per hr., although this figure varies with the size and the caking properties of the coal. The maximum rate of firing is from 60 to 80 and the normal rate from 30 to 40 lb. per sq. ft. of grate per hr.

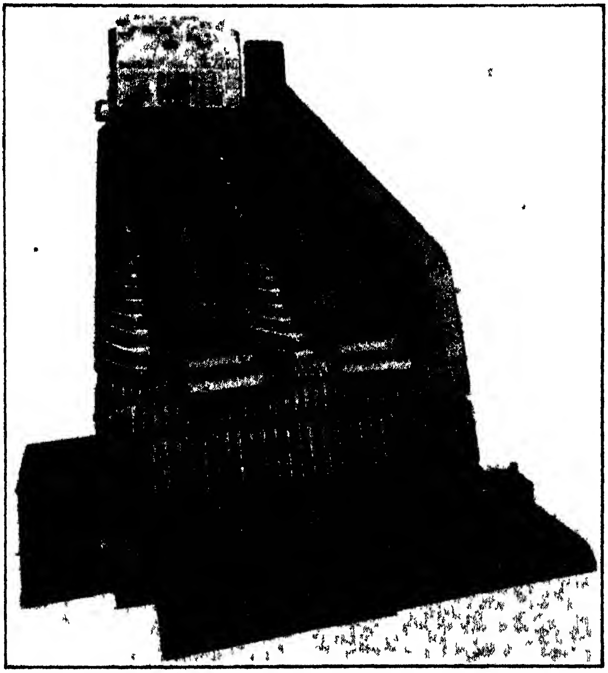

FIG. 126.—Rear view of two-retort section of the Frederick stoker. (*Combustion Engineering Corporation.*)

Regulation of Combustion Rate with Underfeed Stokers.—The combustion rate with underfeed stokers is varied (1) by changing the rate of feed, and (2) by changing the amount and the distribution of the air.

Summary on Stokers.—Chain-grate and underfeed stokers are used more widely than the overfeed types. In choosing the fuel for use with chain-grate stokers, the combustible matter of the coal must be watched—the coal substance must not cake. With underfeed stokers, the type of ash is important; it should be small in amount and not clinker badly. Overfeed stokers while handling any coal or refuse fuel have to be watched and controlled by the operators more than the chain-grate or underfeed type. Pressure inside the furnace must be avoided, particularly with soft coal, if maintenance of furnace walls is to be kept down. The salient points concerning the three types of mechanical stokers are further summarized in Table LXXXVIII, and the combustion rates used in a number of American installations are given[7] in Table LXXXIX.

BOILER SETTINGS WITH MECHANICAL STOKERS

Three main points must be considered in the design and construction of stoker-fired furnaces: (1) the raw coal must be heated properly to its ignition temperature and the temperature must be kept high until the gases are burned, (2) the volatile hydrocarbons and the combustible gases rising from the fuel bed must be adequately mixed with sufficient secondary air, and (3) sufficient time of contact must be provided for the combustion of gases in the combustion space. In the early stages of stoker development these factors were not so clearly recognized as at present and many installations gave low efficiencies, due to lack of appreciation of the necessity for ample combustion space, mixing of combustible gas and air, etc. In modern stoker installations the three important features are taken care of by the proper design of the combustion space, furnace throat, and with chain grate or overfeed stokers by means of ignition arches, combustion arches, and baffles, as illustrated by the representative furnace constructions shown diagrammatically in Figs. 127, 128, 129, and 130. Figure 127 and 128 illustrate the type of furnace and arch construction employed with chain grate and overfeed stokers respectively. With these two types a front arch over the grate must be provided to ignite the entering coal; the furnace throat (*i.e.*, the distance between the front ignition arch and the rear wall) must be designed to shield the fuel bed from any excessive cooling that may result from too great exposure to the relatively cold boiler tubes, and must also promote adequate mixing of the combustible gases and secondary air. Figures 129 and 130 show the application of single- and multiple-retort underfeed

TABLE LXXXVIII

Summary of Information on Stokers

	Chain grates	Overfeed stokers	Underfeed stokers
Description	Fuel from hopper is carried horizontally into furnace on a continuous web. The web is cooled coming out through the ash pit and returns to coal hopper.	Grate extends from hopper into furnace at a rather steep incline, although coal only flows down if a rocking or plunger movement is imparted to grate bars	Raw coal is pushed up through the fuel bed and cinder falls off onto cinder plate. Gases are distilled in an oxidizing atmosphere. Capable of high overloads.
Fuel used. .	1. Coke breeze, steam sizes of anthracite, or 2. High-volatile midwestern coals. In general, any noncoking, clinkering coal may be used, but preferably no mixtures	All coals can be used. Good all-round stoker. Is mainly used on midwest fuels. Coking coals can be used. Will burn refuse fuels.	High-volatile coals, coking coals, and slack or fines may be burned. Ash must not be easily fusible.
Draft . . .	Natural: 0.25 to 0.60 in. water. Forced: 1- to 2-in. water pressure with coke. 1- to 4-in. water pressure with Illinois and similar coals	Natural: 0.25 to 0.6 in. water Forced: 1- to 3-in water pressure	All forced draft. Normal: 2- to 4-in. water pressure in wind box. Maximum: 5- to 7-in. water pressure in wind box.
Rate of combustion, pounds per sq ft per hr.	Average: 30–35 Maximum: 45–60. 10 lb. per 0.1 in water, for natural drafts.	Average: 25–35. Maximum: 40–50.	Average: 30–40 Maximum: 60–80. 10 lb per 1-in water pressure.
Means of regulation.	1. Height of coal gate 2. Speed of grate. 3. Amount and distribution of air.	1. Rate of plunger feeder. 2. Rate of ash removal. 3 Amount and distribution of air.	1. Rate of feed. 2. Amount and distribution of blast.
Miscellaneous . .	Watch for live coals going over end of grate.	Fireman needed for ash removal at times and fire must be cleaned.	Air is admitted over clinker plate to burn out cinder. Fuel bed from 12 to 24 in. deep.

TABLE

RATE OF COAL BURNING

Plant	Stoker	Pressure in air duct	Draft at damper	Class of coal
Brunots Island, Pittsburgh, Pa	Murphy	0 80 in.	0.80 in.	Pitts. bit.
Marion Station, Jersey City	Roney	1 08 in.	0.20 in.	Lancashire
Ashley St. Station, St. Louis	Westinghouse	3 0 in.	0 85 in.	So. Ill. bit.
Ashley St. Station, St. Louis	Sanford Riley	3 0 in.	0.88 in.	So. Ill. bit.
Brunot Island, Pittsburgh, Pa	Westinghouse	3 5 in.	0.90 in.	Pitts. bit.
Brunot Island, Pittsburgh, Pa	Westinghouse	4.25 in.	0 70 in.	Pitts. bit.
Cannon St., New Bedford	Frederick	4 0 in.	0.90 in.	New River
Essex Station, Newark, N. J.	Riley	7.72 in.	0.88 in.	Cent. Penn. bit.
Essex Station, Newark, N. J	Riley	8.0 in.	0.96 in.	Cent. Penn. bit.
L St. Station, Boston	Riley	2.3 in.	0.24 in.	New River
L St. Station, Boston	Westinghouse	6.0 in.	2.4 in.	New River
Marion Station, Jersey City	Westinghouse	5.65 in.	3.43 in.	Lancashire
Schuylkill #2, Philadelphia, Pa	Taylor	0 to 4 in.	0.9 in.	Clearfield bit.
Sherman Creek, New York City.	Taylor	2 to 6 in.	0.1 to 1.0 in.	Pocahontas
Waterside #2, New York City	Taylor	3.5 to 5.5 in.	0.98 in.	East. Penn.
West End Station, Cincinnati, O.	Westinghouse	3.8 in.	0.89 in.	W. Va. bit.
West End Station, Cincinnati, O.	Westinghouse	3.8 in.	0.63 in.	W. Va. bit.
West End Station, Cincinnati, O	Sanford Riley	4.5 in.	0.64 in.	W. Va. bit.
Lacombe Station, Denver	Westinghouse	2 in.	0.5 in.	No. Colo. lignite
South St. Station, Providence.	Taylor	6 in.	1 in.	Penn. semi-bit.
6th St. Station, Cedar Rapids, Iowa	Taylor	. . . . . .	. . . .	Waukee Iowa.
Barbadoes Island, Pa.	Westinghouse	4.5 in.	0.65 in.	W. Penn. bit.
Chester Station, Chester, Pa	Taylor	0 to 5.2 in.	1.7 ft.	Clearfield bit.
Colfax Station, Cheswick, Pa	Westinghouse	6.5 in.	1.64 in.	Bit.-Harwick
Delaware Station, Philadelphia, Pa	Taylor	0 to 7 in.	1.7 in.	Clearfield bit.
Essex Station, Newark, N. J.	Taylor	6.75 in.	1.55 in.	Lancashire
Essex Station, Newark, N. J.	Taylor	5.13 in.	1.47 in.	C. Penn. bit.
Hell Gate, New York City	Taylor	1 to 6 in.	0.05 to 1.6 in.	Pocahontas
Windsor-Beach Bottom, W. Va.	....	5.7, 3.48, 3.48 in.	−1.49 and −1.22	Pitts. #8
Windsor-Beach Bottom, W. Va.	Westinghouse	5.7, 5.6, 4.36 in.	4.4, 4, 1.35 in.	Pitts. #8
66th St. Brooklyn, N. Y.	Westinghouse	5.48 in.	1.034 in.	Pocahontas

LXXXIX

ON VARIOUS TYPES OF STOKERS

Size	Proximate analysis of coal							Coal per sq. ft. stoker per hr.			Remarks
	Moisture	Volatile matter	Fixed carbon	Ash	Sulfur	B.t.u. (as fired)	Ash fusion temperature	1 hr. peak	2 hr. peak	Contin.	
Overfeed stokers											
R.O.M.*	1 5	34.0	54 5	10.0	2.0	13,200	2500°	54	50	38	*Run of mine
R.O.M.	1 94	14.77	64 04	11.19	1.67	13,818	. . .	..	...	33 18	
Underfeed Stokers with Ash Dump											
Crushed nut,	10 9	33 76	43 5	11 84	.. . .	10,876	..	55	48	41	
pea & slack	10 9	33 76	43 5	11.84	....	10,876		70	60	52	
R.O.M.	3.5	32 0	53 0	11.5	1 4	13,300	2600°	52	49	40	
R.O.M.	3 5	32 0	54 0	10 5	1.5	13,500	2600°	60	56	40	
R.O.M.	2 12	17.85	72 58	7 45	....	14,210	. ..	50	44	32.9	
R.O.M.	1.92	26 45	64 73	8.82	2.12	13.970	.....	...	80 1	60.1	
R.O.M.	1.48	22.98	61.59	15.43	2.35	12,786	. .	.	. . .	38 99	
R.O.M.	3.11	22.31	69 17	6.08	0.95	14,603	2500°	51 81	32 4	22.29	
R.O.M.	3.11	22 31	69.17	6 08	0 95	14,603	2500°	67 0	61 0	49.0	
R.O.M.	4 42	18 65	74 32	7 03	0 80	13,940	....	.	. . .	39.7	
R.O.M.	2 9	20 17	69 15	7.98	1.40	13,788	2400°	41 0	36.5	28.0	
Slack	2.9	19.3	71 7	6.1	0.6	14,250	2400°	52	50	38	
R.O.M.	2.8	16.8	72.9	7.5	1.47	14,077	2750°	68	58	40	
Nut & slack	4.5	35.95	53.55	10.5	1.00	12,850	2500°	73.3	64 8	44.8	
Nut & slack	4 5	35.95	53 55	10 5	1.00	12,850	2500°	68 5	58 7	46.7	
Nut & slack	4.5	35.95	53 55	10 5	1 00	12,850	2500°	74.3	71.0	45.8	
Slack	24	28	44 5	3.5	0.35	9,300	2100°	72.5	72.5	49 0	
Crushed	3.22	19 26	66.63	10.89	3.25	13,745	....	45.3	45.3	39.5	
1¼ in. screen	18.4	.....	....	27	....	8,100	.....	52	.. ....	46.6	
Underfeed Stokers with Clinker Grinders											
R.O.M.	1.22	21.64	69.63	7 51	1.81	14,274	2300°	67.0	60.0	50.5	
R.O.M.	2.83	20.21	65.03	8.93	1.088	13,736	2500°	48	43	32.5	
R.O.M.	3.0	32.5	54.5	10.0	1.3	12,800	2500°	102	....	86	
R.O.M.	2.74	23.94	64.96	8.36	1.29	13,775	2500°	54	48	36.5	
R.O.M.	2.55	24.7	65.87	9.43	2.00	13,872	.....	..... .	*67	43.4	*3 hr. test
R.O.M.	3.11	26.28	63.71	10.01	2.4	13,600	......	....	*62.2	43.6	*3 hr. test
Slack	2.87	19.82	70.30	7.01	0.64	14,122	2400°	....	37	26	
Crushed R.O.M.	4	35.82	48.86	11.32	2	12,422	2100°	*65.2	**59.7	49.3	*4 hr.– **9 hr.
	4.34	36.15	48.05	11.46	2	12,134	2100°	*61	**58.3	50.9	*4 hr.– **8 hr.
	3.7	21.5	71.3	7.2	0.8	13,980	Over 2500°	....	.....	68	

TABLE LXXXIX —

Plant	Stoker	Pressure in air duct	Draft at damper	Class of coal
Ashley St, St Louis, Mo	Green		0 83 in	So Ill screen
Ashley St, St Louis, Mo	Green		0 78 in	So Ill screen
Brunot Island, Pittsburgh, Pa	Green		0 75 in	Pitts bit
6th St Station, Cedar Rapids, Iowa	Green			Harrisburg, Ill
6th St Station, Cedar Rapids, Iowa	Green			Montgomery Co
Northeast Station, Kansas City	B and W		0 4 to 2 2 in	Springfield
Northeast Station, Kansas City	B and W		1 2 to 1 8 in	Missouri bit
Northeast Station, Kansas City	B and W		0 4 to 1 3 in	Kansas bit
Northeast Station, Kansas City	B and W			Ok semi anth
Nebraska Power, Omaha	Coxe E '	2 5 in	1 3 in	Kansas bit
Nebraska Power, Omaha	Coxe E '	2 7 in	1 5 in	Cent Ill bit
Nebraska Power, Omaha	Coxe E '	2 8 in	1 5 in	So Ill bit
Nebraska Power, Omaha	Coxe E	2 0 in	0 9 in	Missouri bit
Northeast Station, Kansas City	Illinois	1 to 2 in	0 6 to 0 9 in	Ill bit
Northeast Station, Kansas City	Coxe	1 to 2 in	0 6 to 1 5 in	Missouri bit
Northeast Station, Kansas City	Coxe	1 to 2 in	0 4 to 1 5 in	Ill bit
Northeast Station, Kansas City	Coxe	1 to 2 in	0 8 to 2 in	Kansas bit
Northeast Station, Kansas City	Illinois	1 to 2 in	0 6 to 1 9 in	Kansas bit
Hauto, Pa	Coxe	5 1 in	0 3 in	Panther anth
Hauto, Pa	Coxe	4 12 in	—0 76 in	Panther anth
Milton, Milton, Pa	Coxe	4 71 in		Shamokin anth
Northeast Station, Kansas City	Coxe	1 5 to 3 in	0 65 to 1 5 in	Ark. anth
Northeast Station, Kansas City	Illinois	1 to 2 in	0 6 to 2 in	Ark anth
Northeast Station Kansas City	Coxe	1 7 to 2 7 in.	0 7 to 1 5 in	Okla semi-anth
Amsterdam, Amsterdam, N. Y	Coxe	3 in	2 in. max *	Scranton anth
Harwood Mines, Harwood	Coxe	4 in	0 5 to 1 47 in	Hazleton anth

(*Continued*).

Size	Proximate analysis of coal							Coal per sq ft. stoker per hr			Remarks
	Moisture	Volatile matter	Fixed carbon	Ash	Sulfur	B.t.u (as fired)	Ash fusion temperature	1 hr. peak	2 hr peak	Contin.	
Chain Grates (Natural Draft)											
Nut, pea	12.5	31.06	42.44	14.0		10,422		43	43	43	
& slack	12.2	30.29	33.8	23.71		8,979		40	40	40	
Slack	5.0	31.5	52.3	11.2	1.2	12,600	2800°	32	30.5	27	
1¼ in screen	12.8			13.5		10,770				30.7	
	18.4			17		9,150		45		37.7	
R.O.M	15.81	28.96	39.32	15.91	4 to 5	9,592	2050°	40 to 50	40 to 50	38	
R.O.M	17.6	21.68	34.58	26.14	4 to 5	8,009	1950°	40 to 45	40 to 45	35	
Slack	7.3	29.33	48.44	14.93	3 to 4	11,612	2050°	35 to 40	35 to 40	35	
Slack	8.66	12.17	65.2	13.97	1 to 2	11,888	2000°	35 to 40	35 to 40	35	
Chain Grates (Forced Draft) Bituminous Coals											
Mill	11.0	26.3	44.9	17.8	3.27	10,507	2170°	71	65	56	
#5 screen	11.5	31.5	41.8	15.2	3.08	10,293	2160°	64	59	53	
¾ in screen	9.1	28	50.7	12.2	2.78	11,320	2175°	74	65	59	
Mill	13.4	30.0	38.8	17.8	5.79	10,149	1990°	59	57	55	
Slack	15.81	28.96	39.32	15.91	4 to 5	9,592	2000°	55	55	40	
Slack	17.6	21.68	34.58	26.14	4 to 5	8,009	1950°	55	55	40	
Slack	15.81	28.96	39.32	15.91	4 to 5	9,592	2000°	60	60	40	
Slack	7.3	29.33	48.44	14.93	3 to 4	11,672	2000°	65 to 70	65 to 70	40	
Slack	7.3	29.33	48.44	14.93	3 to 4	11,672	2000°	60 to 65	60 to 65	40	
Chain Grates (Forced Draft) Anthracite Coals											
Culm	10.0	*6.1	*68.6	*25.3	0.45	9,730	2550°			34.4	*Dry
#3 buck	9.5	*5.8	*72.2	*22.0	0.27	10,100	2550°			54.6	*Dry
R.O.M	8.6	*7.4	*72.0	*20.6	0.48	10,754	2550°			58.3	*Dry
Slack	9.5	10.83	64.24	15.43	1.6	11,244	2150°	60	60	40	
Slack	9.5	10.83	64.24	15.43	1.6	11,244	2150°	60 to 65	60 to 65	40	
Slack	8.66	12.17	65.20	13.97	1 to 2	11,888	2000°	60	60	40	
#3 buck.	8 to 9	6 to 7	65 to 70	16 to 18	0.5 to 0.75	11,300		55	55	51	*Induced
		*5.1	*71.9	*23	0.56	10,077	2650°			56.6	Dry

stokers to boiler furnaces, and illustrate the relatively simple construction (freedom from arches) possible with this type.

Certain features of furnace design, such as average size of combustion space, arch construction, the application of baffles for controlling the path of gases within the furnace, and various means of cooling furnace walls in order to prevent excessive depreciation, are taken up below.

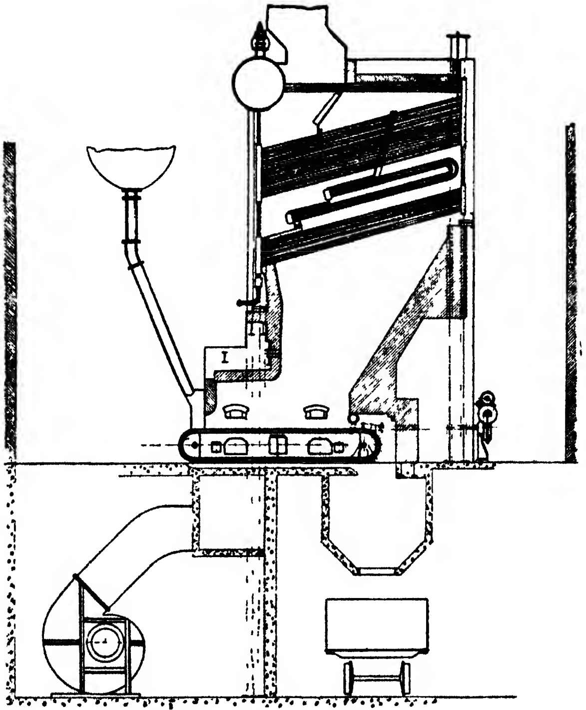

Fig. 127.—Coxe chain-grate stoker installed under a cross-drum boiler. Setting designed for burning midwest semi-bituminous coal. (*Combustion Engineering Corporation.*)

Average Size of Combustion Space.—In several large central stations using forced-draft chain-grate stokers, about 38,000 B.t.u. per hr. are liberated in each cu. ft. of combustion space. The average furnace volume is from 2 to 4½ cu. ft. per 10 sq. ft. of boiler heating surface.[10]

With underfeed stokers the size of combustion space is such that from about 30,000 to 46,000 B.t.u. are liberated per cu. ft.

of combustion space per hr., depending on the coal used. The average for seven of the large American central stations is about 38,000 B.t.u. per cu. ft. of combustion space per hr. From 3 to 4½ cu. ft. of combustion space are provided per 10 sq. ft. of boiler heating surface.[10]

These data for the amount of heat liberated per cu. ft. of combustion space with mechanical stokers are to be compared with central-station practice in powdered-coal and oil-fired furnaces. The average for six of the largest powdered-coal installations is about 18,500 B.t.u. liberated per cu. ft. of combustion space per hr., and for three oil-burning stations is about 35,000 B.t.u. liberated per cu. ft. per hr.

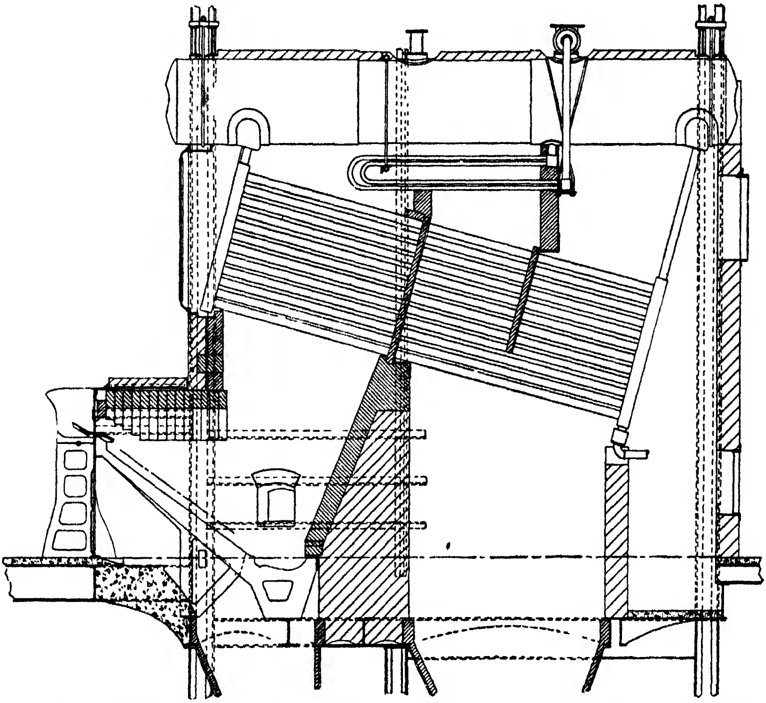

FIG. 128.—Westinghouse Roney stoker applied to a B. & W. boiler. (*Westinghouse Electric & Manufacturing Co.*)

Arches.—The ignition arch over the front portion of a chain grate or overfeed stoker has three main functions: (1) to heat up, distill, and ignite the incoming coal; (2) to mix the air and combustible gases; and (3) by regulating the path of the gases, to promote mixing, prevent stratification, and direct the gases toward the heat-absorbing surfaces of the furnace. The rate of heat flow from the combustion space toward the front of the grate must be exactly the same as the rate of grate or fuel travel into the furnace. If the rate of heat flow is *less*, the combustion zone will move further and further into the furnace, and finally the fire will go out. If the rate of heat flow is *greater* than the rate of grate or fuel travel, the combustion zone will work gradually back toward the feed hoppers, the coal will all be burned before the grate enters the ash pit, and a large amount of excess air can enter on account of the decreased resistance of the fuel bed.

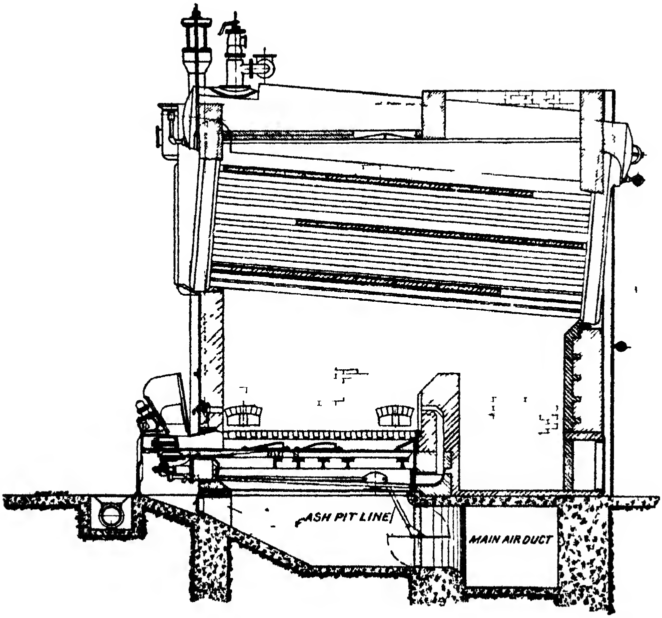

Fig. 129.—Type-E underfeed stoker applied to a Heine boiler. (*Combustion Engineering Corporation.*)

The main variables affecting the length of arch to be employed in a given case are: (1) the heating value of the coal, and (2) the rate at which the coal is to be ignited. For this second factor Marsh[8] has used the term "ignition rate," *i.e.*, the pounds of coal ignited per ft. of stoker width per hr. For example, if the firing rate is 30 lb. per sq. ft. of grate per hr. and the grate travel is 10 ft., the "ignition rate" is 30 × 10, or 300 lb. of coal ignited per ft. of stoker width per hr. The approximate length of arch required to ignite a coal containing about 25 per cent volatile matter can be determined (for natural-draft conditions) with the aid of the formula:*

$$L = \left[1.0 + 0.16\left(\frac{15{,}000 - \text{B.t.u.}}{1000}\right)\right]\frac{W}{100},$$

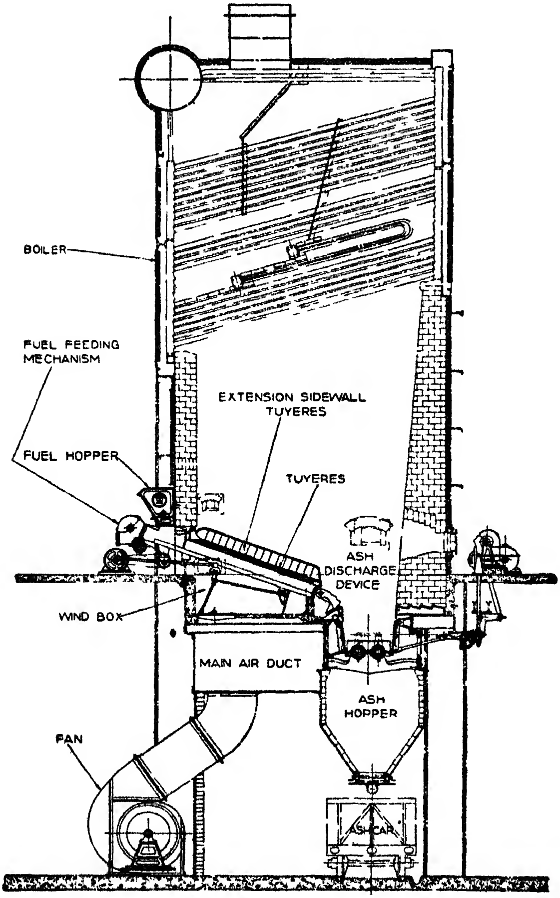

FIG. 130.—Installation of Westinghouse underfeed stoker. (*Westinghouse Electric & Manufacturing Co.*)

where L equals the length of arch in feet, B.t.u. signifies the heating value of the coal in B.t.u., and W is the ignition rate as previously defined. For example, the length of arch to be used

---

* Adapted from data given by MARSH.[8]

with a stoker 15 ft. in length, burning 25 lb. of a 12,000 B.t.u. coal per sq. ft. of grate per hr., is determined as follows:

$$L = \left[1.0 + 0.16\left(\frac{15{,}000 - 12{,}000}{1000}\right)\right]\left(\frac{25 \times 15}{100}\right)$$
$$= [1.0 + 0.48]3.75 = 5.5 \text{ ft. length of arch.}$$

These figures apply to natural-draft conditions and for the combustion of a bituminous coal containing 25 per cent or more of volatile matter. With coals of lower volatile matter or with coals of lower B.t.u. the arch length must be increased. Present practice as to the height of arch[8] is to set the arches, in natural-draft installations (in which the combustion rates are necessarily low), from 2 to 3 ft. above the grate with a slope of 2 to 3 in. per ft. toward the rear; forced-draft arches are higher and flatter than natural-draft arches, on account of the higher combustion rates, and commonly are set at least 4 ft. above the grate with pitch of 1 in. per ft. toward the rear of the furnace.

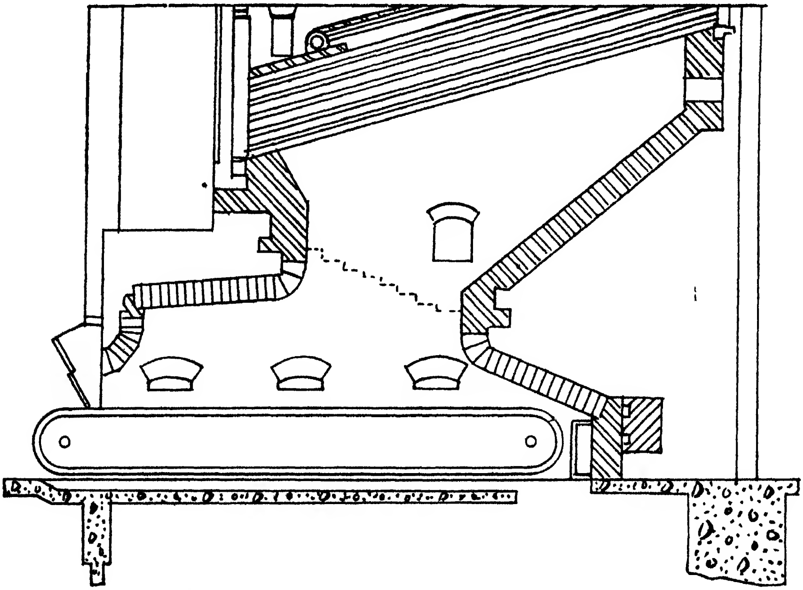

FIG. 131.—Reverse arch construction employed at the Adirondack Power and Light Co.

A second important function of arches is to promote mixing of the combustible gases and air, thus utilizing the available combustion space most efficiently. The effect of arch construction and of the use of auxiliary mixing arches is convincingly shown by a series of experiments recently carried out[9] by the Adirondack Power & Light Co. A series of tests were made on three boilers using Coxe stokers for burning small sizes of anthracite. One boiler was of the customary design for use with the Coxe stoker, such as has already been shown in Fig. 127. The second boiler, as indicated in Fig. 131, was equipped with a reverse arch, designed to bring forward any air coming up through the ash-pit end of the grate and mix it with the combustible gases. The third boiler tested was equipped not only with front and rear arches, but also with a third mixing arch above the two lower arches as indicated in Fig. 132. The superiority of the setting illustrated in Fig. 132 over the other two settings as regards overall efficiency is shown by Fig. 133, in which the results of the tests are given graphically. This chart clearly shows that the boiler with the three arches (*i.e.*, Fig. 132, called boiler No. 1, and results shown with full line in Fig. 133) gave a much higher and more uniform percentage of CO_2 in the stack gases, had less loss as carbon in the cinder, and showed considerably higher combined efficiency than either of the other two settings.

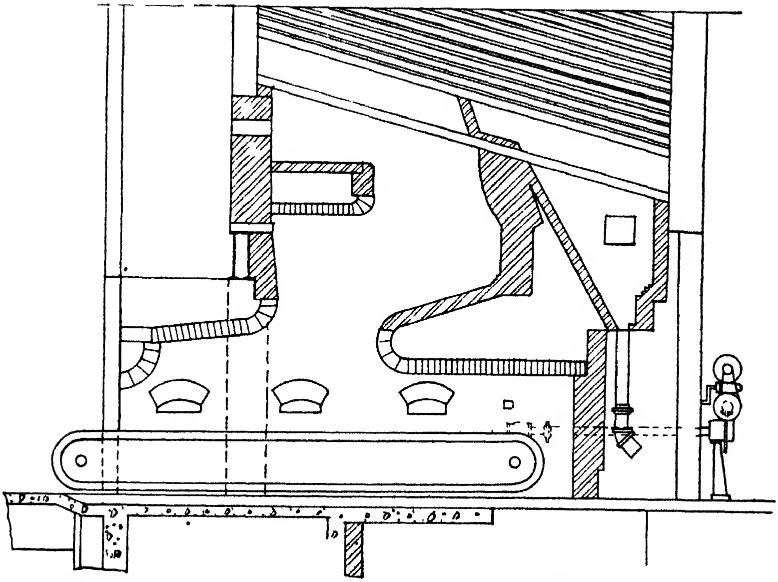

FIG. 132.—Boiler setting showing reverse arch and mixing arch construction.

Details of several types of arch construction and of the method of hanging tile are shown in Fig. 134. Arch maintenance is apt to be high on account of the exposure to intense heat, and the construction always should be such that replacements are easy.

Furnace Throat.—The furnace throat (that is, the narrowest width between the front and the rear arches), if too large, causes a cold fuel bed on account of too complete exposure to the cool boiler tubes. A small throat, on the other hand, promotes good mixing of the combustible gases and air, gives good efficiencies (little or no CO is left unburned in the stack gases), and maintains the proper hearth temperature. If the throat is too small, a high pressure drop is created (if not venturi shaped), the

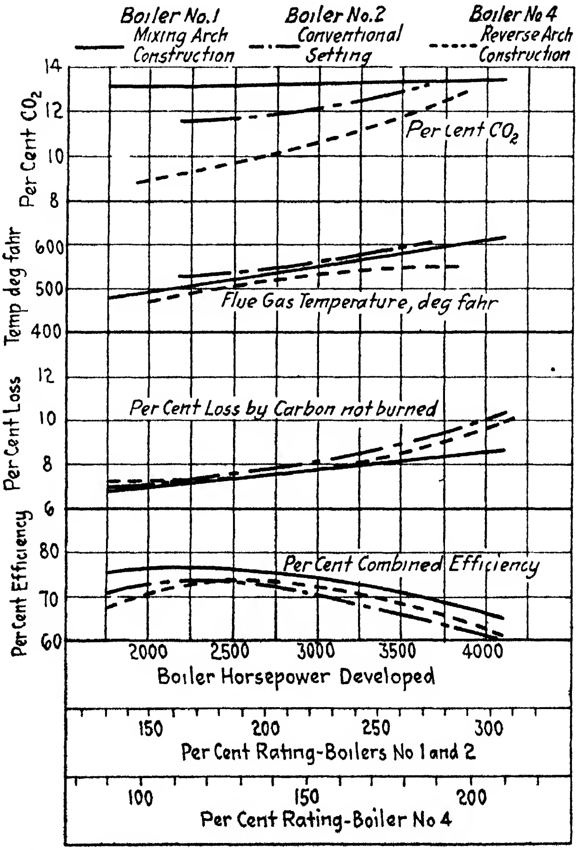

Fig. 133.—Results of tests showing effect of arch construction on boiler operation.

capacity is lowered, and the efficiency drops because too little heat is transmitted by radiation. In general, the area of the furnace throat should be about one-third of the grate area.

Baffles.—Baffles are placed in furnaces or among boiler tubes to direct the flow of the hot gases and to eliminate "dead spaces" where the stagnant gas reduces the rate of heat transfer. In boilers these baffles may be either vertical or horizontal. Vertical baffles in a horizontal-tube boiler (as illustrated in Fig. 128) force the gases up and down across the boiler tubes. They require a higher stack draft to pull the gases through the furnace but give better transfer of heat from the gases to the tubes and consequently give a lower stack temperature. With horizontal baffles (see Fig. 129) the path of the gases is largely parallel to the tubes and in consequence the draft required is less, but the stack temperature is higher than with vertical baffles. Horizontal baffles, however, are easier to clean and to keep tight, and tube replacement is, in general, simpler than with the vertical type.

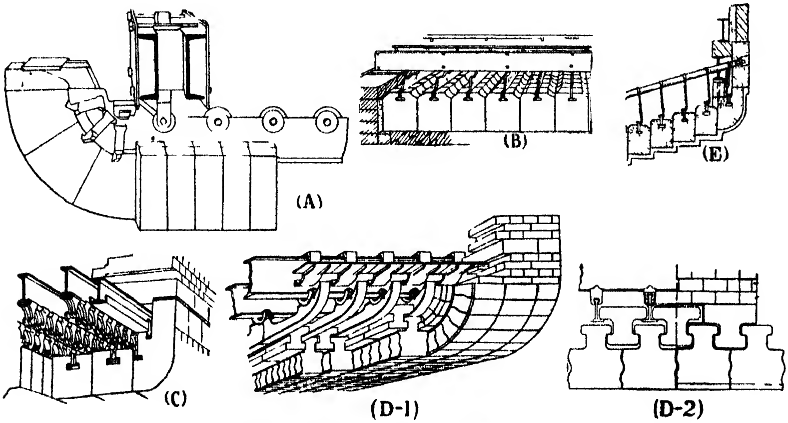

Fig. 134.—Types of arch construction. (A) Method of hanging tile, The M. H. Detrick Co., (B) method of suspending Detrick flat arch, (C) McLeod and Henry Co. suspended arch, (D-1) (D-2) Liptak double suspended arch construction, (E) method of suspending sloping arch, Bigelow Arch Company.

The flue gases are, of course, cooled in passing over the heat-absorbing surfaces of the boiler, and the baffled passages usually are constricted in the direction of cooling to decrease the cross-section and thus maintain an essentially uniform gas velocity over the tubes. The gas velocities employed range from about 25 to 40 ft. per sec. It is important to avoid such marked constriction as will greatly lower the draft and therefore seriously limit the capacity.

Water-cooled and Air-cooled Furnace Walls.—The maintenance of furnace linings has become a serious problem with the growing use of large furnaces and the demands for high capacities. The melting and the erosion of refractories, and the adhesion of clinker to the walls, have been two of the major problems in the development of powdered-coal firing (see Chap. XV) and recently have caused a great deal of trouble in stoker-fired furnaces. In powdered-coal installations the trouble has been overcome to a great extent by water cooling and air cooling the furnace walls. These two methods have been applied recently to stoker-fired furnaces, and although they are still relatively new developments,

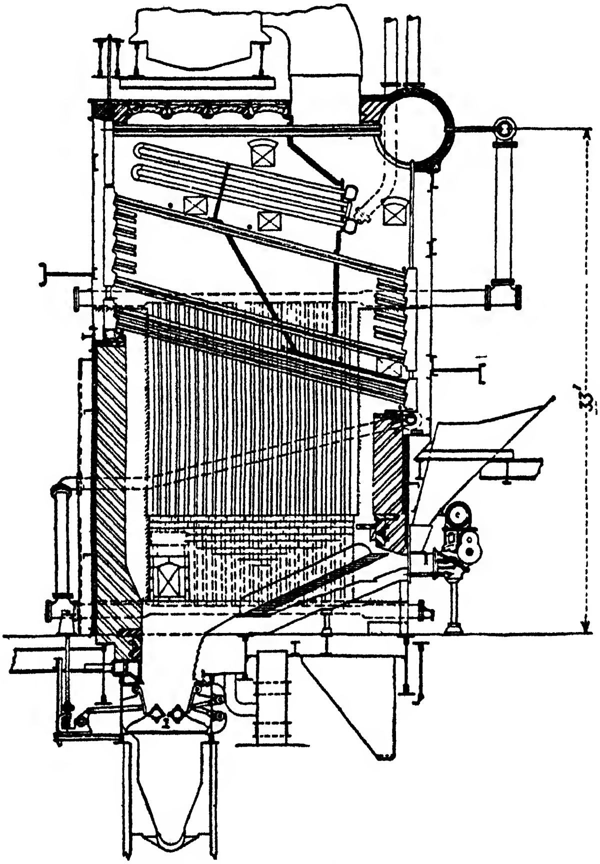

Fig. 135.—Water-cooled (fin-tube) furnace for underfeed stoker setting.

water cooling seems to offer greater possibilities than air cooling, particularly with large-size multiple-retort underfeed stokers. Some operators have felt that trouble would be experienced in attempting to apply hollow air-cooled walls to stoker furnaces on account of the high air pressures ordinarily used.

Water-cooled furnace walls* usually are composed of water-filled upright tubes or flat elements which are connected with the circulating system of the boiler. These water-cooled elements absorb a large amount of heat from the fuel bed by radiation, thus adding considerable extra heat-absorbing surface to the boiler. They eliminate erosion of the refractory walls, prevent clinker adhesion, and markedly lower radiation losses. The furnace efficiency is therefore increased, unless so much cooling surface is exposed that combustion is retarded and there is a loss of unburned fuel up the stack and in the cinder.

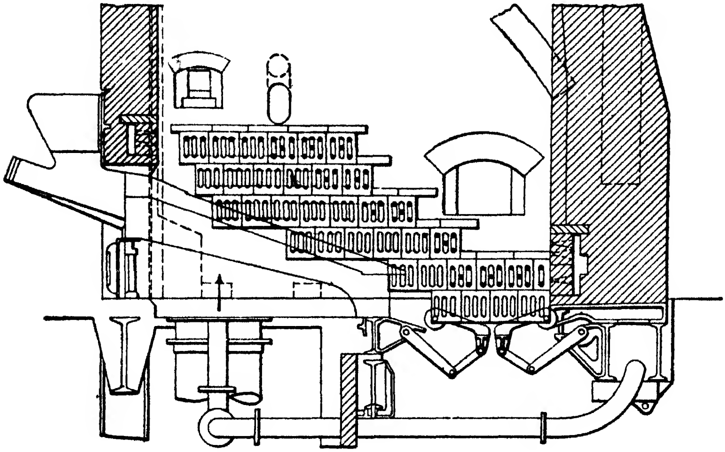

FIG. 136.– Air-cooled furnace blocks for preventing clinker adhesion to side walls.

One type of water-cooled furnace (the "fin furnace") is shown in Fig. 135. The front and rear walls of this furnace are simple vertical walls. The side walls, however, are almost entirely composed of 4-in. water tubes, about 20 ft. long, set on 7 ⅝-in. centers. These tubes are connected with the boiler circulation, and each tube has a fin about 2½ in. wide welded on either side. These fins receive radiant heat from the furnace and transmit it by conduction to the water tubes. The tubes are backed by

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* The subject of water cooling of furnace walls is also taken up in Chap. XV, p. 458.

special tile, insulating material, and a steel casing which forms the outer shell of the furnace. A series of tests conducted on a boiler with this type of construction have shown excellent results, the combined efficiency of the boiler, furnace, grate, and economizer being 86.8 per cent at 243 per cent of rating, and 81.5 per cent at 346 per cent of boiler rating.

A second method of cooling the walls and preventing clinker adhesion is illustrated by Fig 136. A bank of perforated refractory blocks is installed in the side, front, and bridge walls of the furnace on the level with, as well as extending above, the fire line. The secondary air which must be added above the fuel bed is admitted in thin jets through the perforated blocks, cooling them and decreasing the formation of troublesome clinker along the sides. In the installation shown in Fig. 136, the line of side-wall blocks extended considerably above the fire line and the admission of too much secondary air through the unobstructed holes caused trouble. In this particular case it was found necessary to plug about one-third of the holes in the side-wall blocks.

AIR PREHEATERS AND THE USE OF PREHEATED AIR

Regeneration and recuperation (see Chap. XI) have been used for a long time in the metal industries in order to recover heat from the stack gases and to raise furnace temperature, but the active application of air preheaters to boiler furnaces is of relatively recent origin.[11 12,13] The earliest recorded air-preheating system made in the United States—installed by Hoadly in Lawrence, Mass., in 1882—clearly demonstrated the gain in efficiency made possible by preheating the entering air by means of the hot flue gases. Active interest in the United States concerning this important subject dates back only three or four years in spite of the success of Hoadly and in spite of the development of preheaters in Europe. The preheaters which have been installed in the United States in the past few years have demonstrated the value of this type of equipment and many of the large power stations now being erected include air preheaters. These heaters are as well adapted (and in some respects better adapted) for use with oil-fired and powdered-coal-fired furnaces, as they are with stoker-fired furnaces. The remainder of this chapter is devoted to a description of various types of air preheaters now in use, and to a discussion of the results which have been obtained by using preheated air in boiler furnaces.

Types of Air Preheaters.*—Flue-gas-to-air preheaters now in use are of three types: (1) the plate type, in which the flue gas and the air are passed countercurrently through alternate spaces divided by thin, flat plates; (2) the tubular type, in which air is passed countercurrently over the outside of tubular elements through which the flue gas is exhausted to the stack; and (3) the rotating type, in which hot flue gas and air pass countercurrently through a divided cylindrical passage; a rotating element in the cylindrical passage absorbs heat from the flue gas on one side of a central dividing partition, and as the heated section rotates into the air side of the passage the heat is transferred to the air.

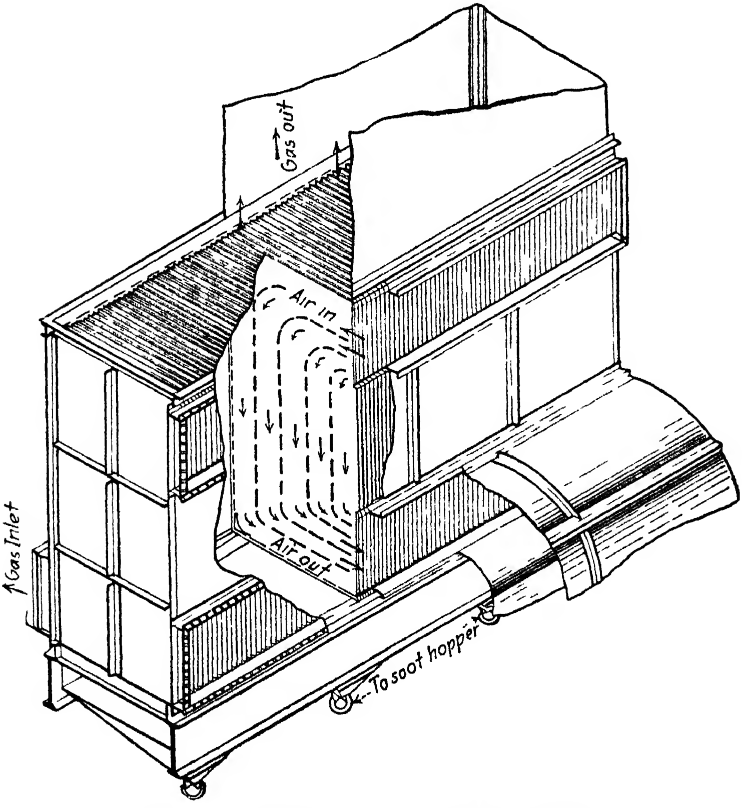

FIG. 137.—Plate type of flue-gas air preheater.

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* Descriptions of a number of preheaters not discussed in this section are given in reference 12 of the Bibliography.

In the heater of the plate type (Combustion Engineering Corporation) shown[12] in Fig. 137, the flat plates are welded in place. Angle irons between the plates are used to prevent shortcircuiting of either gas or air, and to give a uniform rate of heat transfer. The gas and the air pass countercurrent to each other, as indicated by the arrows in the diagram, and suction from the fan used to force air into the furnace is utilized to pull air through the preheater. Any particles of dirt lodging on the gas passages can be removed by brushing from the top into a soot hopper below the preheater. Stack-gas by-passes around the preheater may be used to increase the uptake openings in order to take care of the large volumes of gas leaving the furnace under peak-load conditions.

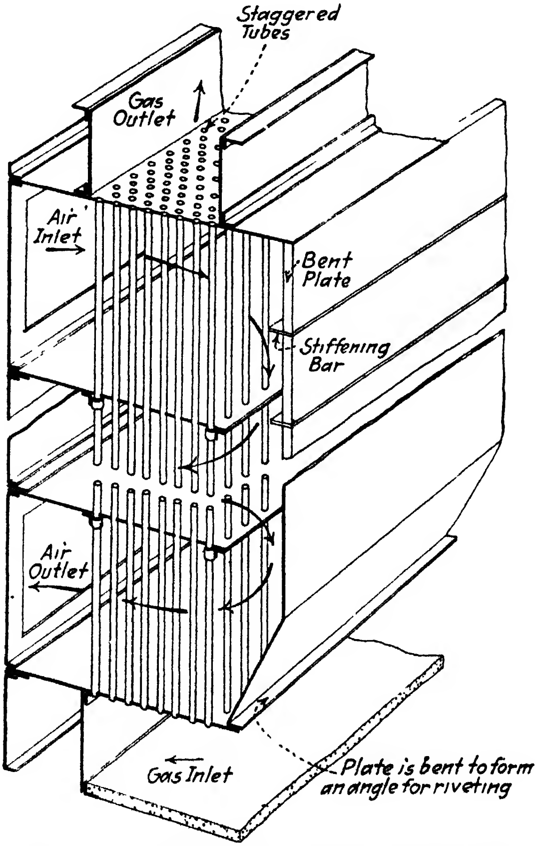

FIG. 138.—Tubular type of flue gas air preheater.

In the tubular air preheater (one type of Babcock & Wilcox Co.) shown in Fig. 138, the stack gas passes up through a number of staggered 2-in. tubes. Air is distributed over the outside surface of the tubes by a series of baffle plates as indicated. This type of preheater is cheap to construct and affords a good opportunity to make the preheater free from leaks.

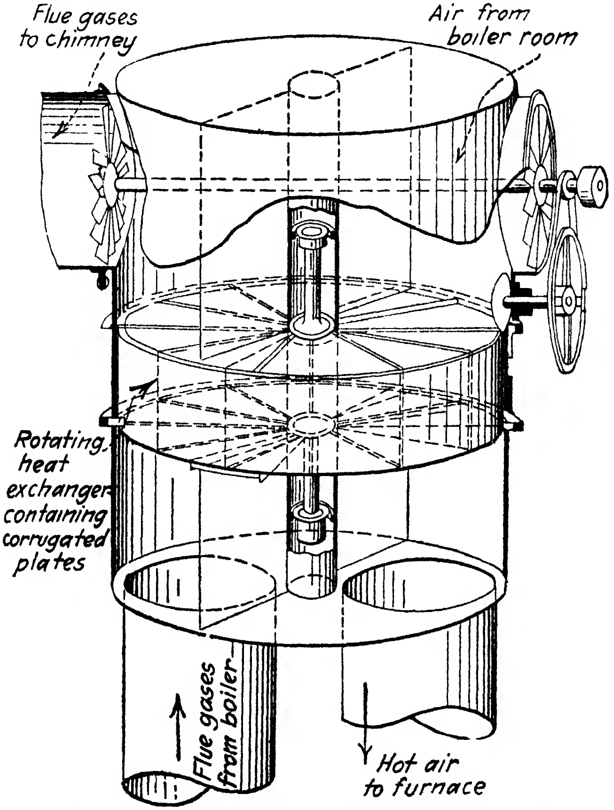

FIG. 139.—Ljungstrom type of rotating flue gas air preheater.

A rotating type of preheater developed by Ljungstrom is shown in Fig. 139. As indicated, the heater is divided into two vertical compartments; air is drawn down through one and hot flue gases are passed up through the other. A rotating heat exchanger occupies the middle of the preheater. This heat exchanger is composed of concentric, alternate elements of thin corrugated steel and of plain sheet steel. In this manner a large number of vertical passages are provided for the free up or downward movement of the air and gas. This heat exchanger rotates about six times a minute, being heated by the flue gas while on the flue-gas side of the central dividing partition, and giving up heat to the air on the other side. Packing above and below the rotating exchanger allows very little clearance and prevents excessive leakage of air into the flue-gas chamber. Leakage is further reduced by dead spaces which separate the flue-gas section from the air section. A forced-draft fan is used to drive air through the preheater, and a second fan mounted on the same shaft is used to exhaust the stack gases. The flue-gas section is larger than the air section on account of the increased flue-gas volume as compared with the volume of entering air. With this type of preheater, air may be preheated to 300°F. or better.

In several preheaters now being installed, steam bled from the main turbine is used to preheat the air. These preheaters are usually of the plate or tubular type.

Results Obtained with Preheated Air.—Although air preheaters have been adopted but recently for boiler furnaces in the United States, and test data are still incomplete, it has been demonstrated that the use of preheated air possesses several important advantages over and above the recovery of heat from the stack gases. Furnaces equipped with air preheaters and using only the moderate degrees of preheat adopted in the early American installations (200 to 300°F., or 110 to 170°C.) have shown an increase in efficiency of from 4 to 10 per cent. The total efficiency increase is due to the net effect, (1) of the actual recovery of heat and its return to the furnace, and (2) an added gain due to an increased temperature in the furnace. As pointed out in Chap. VIII, all combustion reactions are greatly speeded up by increasing the temperature, and with preheated air it is therefore possible to obtain more nearly *complete* combustion, not only on the grate, but also in the combustion space, with lower percentages of excess air. This increases the efficiency, makes possible higher combustion rates, lowers the ash-pit loss, and offers interesting possibilities with regard to the combustion of low-grade, high-moisture fuels. In chemical and metallurgical furnaces in which heat for chemical reactions or melting has to be supplied at high-temperature levels, preheating air greatly increases the fuel efficiency *over and above* that saved directly by the preheater.

Clarke,[13] on the basis of tests made on a plate type of preheater at the Colfax station of the Duquesne Light Company, has presented the results shown graphically in Fig. 140. This figure shows the comparative performance of the boiler tested, both with or without preheated air. The increase in efficiency due to preheated air, as indicated by the curves in the upper part of the diagram, ranged from a 5½ per cent increase at 114 per cent of rating, to a 7 per cent increase at 200 per cent of rating. The general effect of this increase is to "flatten" the efficiency curve, and to make higher ratings possible and still maintain high efficiency. The temperature to which the air was preheated was higher at the high ratings, in spite of the larger amount of air which passed through the preheater. This is due to the fact that the temperature of the flue gases is increased, there is a greater temperature difference between flue gas and air, and greater heat transfer is possible. Figure 140 shows the high CO_2 content of the gases (low excess air) possible with the preheater in operation, and the curves of "per cent

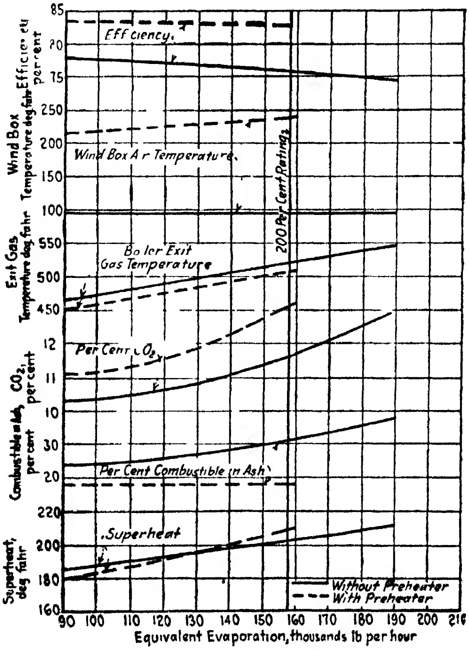

FIG. 140.—Comparative performance of boiler with and without preheated air.

combustible in ash" demonstrate the lower ash-pit losses when preheated air is used.

An analysis of the difference in efficiency with and without air preheating is given[13] in Fig. 141. The lower heavy line represents the efficiency without preheated air and the upper heavy curve shows the efficiency with the preheater in use. The heat recovered from the stack gases is seen to be the major gain, but the decreased loss due to unburned carbon in the cinder (represented by the space between the lines *AB* and *EF*) is also very important. There was a slight additional loss due to radiation and unaccounted-for losses, but the net result was, nevertheless, a highly important gain in efficiency.

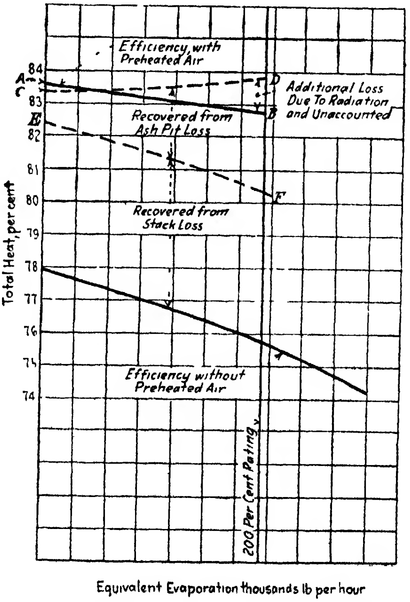

FIG. 141.—Decreased losses with preheated air.

In these tests reported by Clarke, the air was not highly preheated (below 250°F., or 121°C.) and no trouble was experienced due to clinker formation resulting from the increased temperature in the fuel bed. In one installation in this country, however, in which the air was preheated to about 500°F. (260°C.) and coal with a moderately clinkering ash was burned with an underfeed stoker, difficulties with clinker formation were encountered when attempts were made to carry heavy loads. The remedy for this situation would appear to lie in water cooling of the furnace walls, and the future trend of stoker furnaces will probably be toward the use of highly preheated air and the application of more or less complete water cooling of the furnaces in order to decrease radiation, prevent clinker troubles, and lower furnace-maintenance costs.

BIBLIOGRAPHY

1. KREISINGER, OVITZ and AUGUSTINE: "Combustion in the Fuel Bed of Hand-fired Furnaces," U. S. Bur. Mines, *Tech. Paper* 137 (1917).
2. D. K. CLARK: "The Steam Engine."
3. JUNE: "Two Hundred and Thirty-seven Years of Stoker Development," *Combustion*, **8** and **9** (1922–1923).
4. WORKER and PEEBLES: "Mechanical Stokers," McGraw-Hill Book Co., Inc. (1922).
5. "Condensed Catalog of Mechanical Stokers," compiled by the Stoker Manufacturers Assoc. (1924).
6. DAVID BROWNLIE: *Proc* Inst. Mech. Eng. (Eng.), 263–428 (1920).
7. *Rep.* of Prime Movers Committee, Nat. Elec. Light Assoc. (1922–1923 and 1923–1924).
8. T. A. MARSH: "Combustion in the Power Plant," D. Van Nostrand Co. (1924).
9. *Rep.* of Prime Movers Committee, Nat. Elec. Light Assoc., *Pub.* 2472, 7 and 8 (1924).
10. JUNKERSFIELD and ORROK: *Combustion*, **11**, 125 (1924).
11. C. W. E. CLARK: Air Preheaters, *Mech. Eng.*, **47**, 175 (1925).
12. "Air Preheaters and Their Application," *Power*, **60**, 884 (1924).
13. CLARKE: "Boiler Tests with Preheated Air," *Mech. Eng.*, **46**, 67 (1924).

CHAPTER XV

POWDERED COAL

It has already been pointed out that the combustion of coal takes place in two steps, first, the separation of the coal into the elementary fuels, gaseous hydrocarbons and coke; and, second, the combustion of these elementary fuels. In the ordinary coal-fired furnace, the volatile matter is driven off as the coal is fed onto the grate and the coke left behind combines with the primary air. With powdered coal, on the other hand, the fuel, mixed with the proper amount of air for economical combustion, is injected into the hot furnace in a finely divided state. Under these conditions, the distillation of the volatile matter takes place almost instantaneously and the small size* of the resulting coke particles makes it relatively easy for the amount of oxygen necessary for their combustion to diffuse in to the carbon surface. For this reason, although not strictly true, the combustion of powdered coal has been called "one-stage combustion."

In some ways the burning of powdered coal is like the burning of a combustible gas; both types of fuels possess certain advantages in common and, with both, one of the main problems is the thorough mixing of the fuel (powdered coal or combustible gas) with the air required for combustion. Powdered-coal burners are even modeled to some extent along the lines of burners for blast-furnace gas.

In addition to the problem of mixing, powdered coal burning has other more pressing problems. First, before burning, the coal must be properly prepared, that is, crushed, pulverized, and in many cases dried before pulverizing. This preparation requires special equipment which may not be justified if the plant is small. Furthermore, due to the high temperature of the solid coal and

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* Since the rate of the reaction between oxygen and solid carbon depends on the amount of surface exposed (see Chap. VIII), it is interesting to compare coal as ordinarily fired with powdered fuel on this basis. Thus a piece of coal 1 in. cube has 6 sq. in. of surface, but when pulverized to pass through a 200-mesh screen it has over 4400 sq. in. It is at once evident that in the powdered condition combustion must be very much more rapid.

ash particles within the furnace, a large amount of heat is radiated directly to the furnace walls and the resulting high temperatures within the furnace may cause the coal ash to fuse. This impedes ash removal, and the molten ash may combine with and seriously erode the refractory furnace lining.

Offsetting these difficulties, powdered coal possesses many advantages, the most important being: (1) great flexibility and, therefore, low stand-by and low banking losses, (2) high thermal efficiency, and (3) low cost of operation and maintenance. These and other advantages are discussed more thoroughly in the latter part of this chapter. The advisability of using coal in a pulverized form depends on the balancing of the advantages to be gained from an almost perfect method of combustion against the cost of pulverizing and the difficulties of handling the ash.

The Development of Powdered-coal Firing.—The first attempt to utilize powdered coal was made by Neipce in 1818, but this installation, as well as that of Henchal in 1831, failed, probably on account of the relatively coarse coal particles used. The first really successful application was in 1895, when the Atlas Portland Cement Company began to use powdered coal in firing cement kilns.

At the present time (1925) the total amount of pulverized coal that is being consumed by the various industries in the United States is about 20,000,000 tons per year, divided* as follows:

	Tons
Portland-cement industry	8,000,000
Generated power	5,000,000
Iron and steel industry	3,000,000
Copper industry	2,500,000
Miscellaneous	1,500,000
	20,000,000

In 1918 only about 200,000 tons of coal were pulverized for power-generation purposes, the actual amount of boiler-heating surface fired by pulverized coal at that time being less than 200,000 sq. ft. The growth of powdered-coal application in the power-generation field has been the most striking feature of the development of this type of fuel utilization, as indicated by Fig. 142. This figure shows how rapidly the amount of boiler-heating surface fired with powdered coal has increased—from less than 200,000 sq. ft. in 1918 to over 2,000,000 sq. ft. in 1925, or an increase of over 100

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* Data from Fuller-Lehigh Company.

fold in seven years. H. W. Brooks[4] of the U. S. Bureau of Mines estimates that the use of powdered coal is increasing (in 1925) at the rate of 200 per cent per year. The increase is due largely to the enormous expansion in the boiler field, particularly in large central stations, and in replacing fuel oil in the localities where the price of this latter fuel has increased greatly.

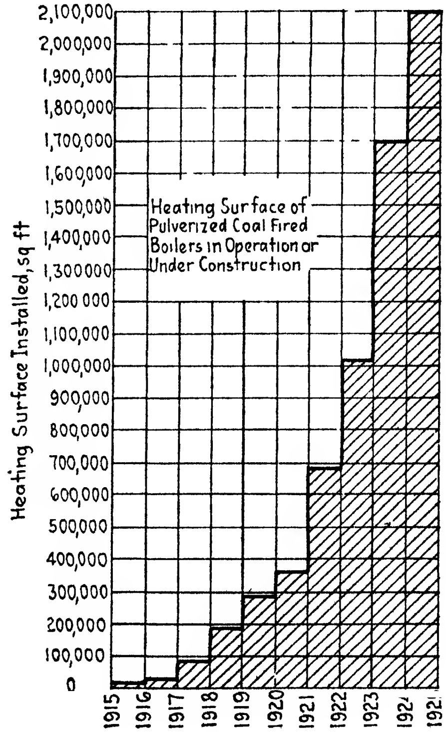

FIG. 142.—Heating surface of pulverized-coal-fired boilers in operation or under construction. (*Data from Fuller Lehigh Company.*)

POWDERED-COAL PREPARATION

Powdered coal installations may be divided into two main classes; (1) unit systems, in which the coal is pulverized and delivered (with a large proportion of the air for combustion) to the furnace by a single machine; and (2) the storage system, in which the coal is prepared centrally in a separate plant, stored, and then transported to the individual furnaces.* The operation of the unit system will be described first.

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* For a more detailed description of powdered-coal equipment, furnaces, costs, etc., the reader is referred to BLIZARD,[1] HARVEY,[2] and COLE.[3]

The Unit System.—In unit systems of powdered-coal utilization, such as shown in Fig. 143, coal, crushed so as to pass a $1\frac{1}{2}$-in ring, is fed from an overhead storage bin through a hopper to the pulverizer. The pulverizer (of which one type is shown in detail in Fig. 144) usually consists of a rotor composed of pulverizing elements and a fan mounted on a single shaft. The fan serves a dual purpose. First it carries the finely ground coal particles out of the pulverizer to the burner at the furnace, leaving behind the coarse particles for further pulverizing and, second, it supplies and mixes with the coal some of the air required for combustion. Pulverized-coal burners are described later (p. 448). A unit system in which a tube mill is employed is shown in Fig. 145. Coal from the hopper falls on to a disc feeder and is fed into the tube or ball mill. A current of primary air, preheated by passing through the hollow walls of the furnace, is drawn through the mill, where it picks up and mixes with the powdered coal and carries it to a burner at the furnace front.

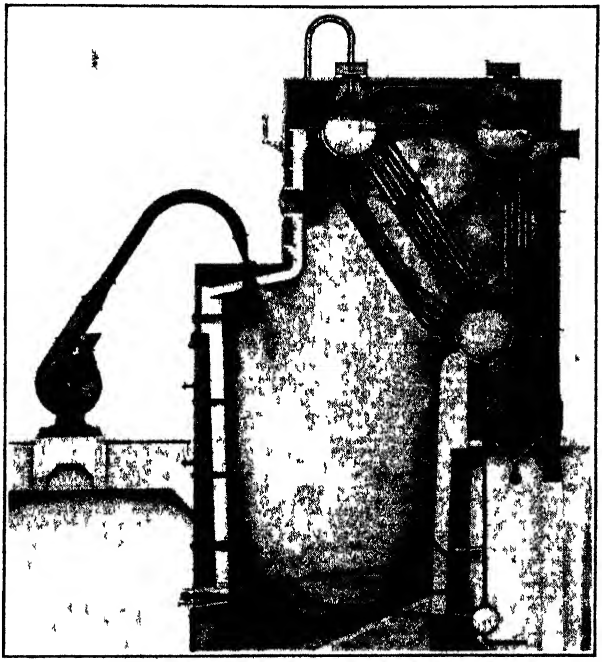

FIG 143.—Unit system of coal pulverization applied to boiler (*Erie City Iron Works*.)

Additional secondary air can be admitted through adjustable openings around the burner. The use of a tube mill does away with dangers from "tramp" iron, since any which may accidentally enter the system merely acts as an additional grinding element within the mill. Tube mills possess the advantage of delivering a uniformly fine product, which should aid in cutting down the size of furnace required. The air supply may be adjusted at will, the usual practice with unit systems being to admit the major part (75 per cent) of the air for combustion through the pulverizer.

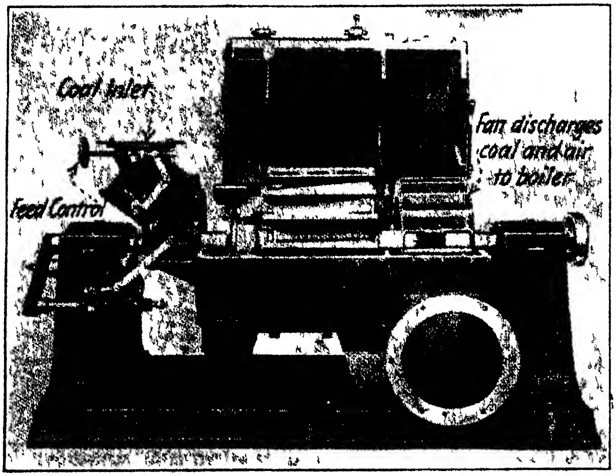

Fig 144.—Unit-system coal pulverizer. (*Erie City Iron Works.*)

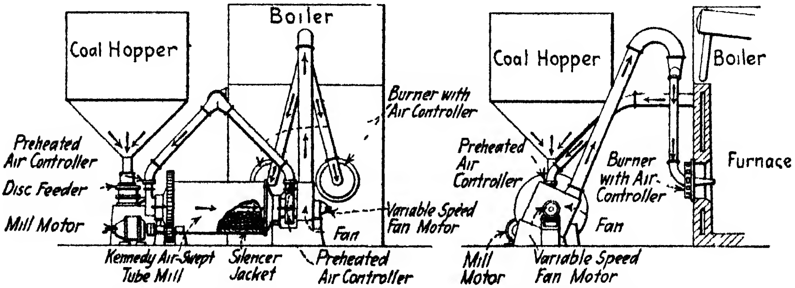

Fig. 145.—Unit powdered coal system using air-swept tube mill. (*The Kennedy-Van Saun Manufacturing & Engineering Co.*)

When this large amount of "primary" air is supplied with the coal the flame is short, and it is more difficult to ignite the coal-air stream than if the coal is premixed with a lesser amount of air (the balance being added after the burner is lighted and mixed in just after entrance into the furnace). Certain of the latest types of unit pulverizers can separate and deliver finely pulverized coal to the furnace, using only 25 per cent of the total air required for combustion. Ignition is easy with this type of pulverizer.

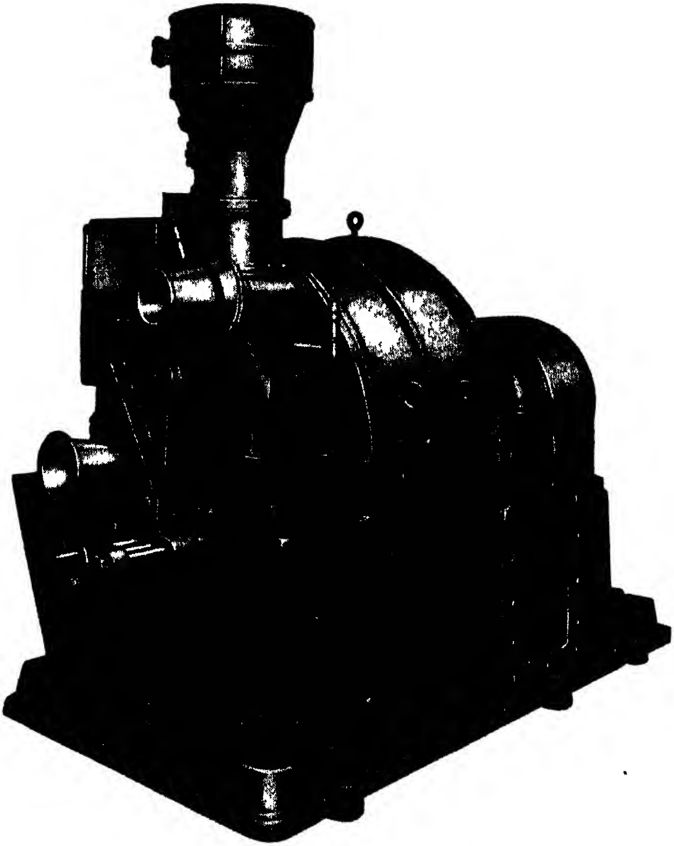

FIG. 146.—Simplex unit pulverizer with gyratory crusher. (*Furnace Engineering Company.*)

The particular advantages of the unit system are:

1. No storage of pulverized coal is required. This fact simplifies plant construction, cuts down the cost of installation, and reduces the fire hazard.

2. The air and the pulverized coal are mixed and delivered by the pulverizing unit, thus rendering unnecessary auxiliary blowers and simplifying burner design.

3. The floor space required is small. For example, the unit pulverizer shown in Fig. 146, which includes a crusher, pulverizer, and fan, requires only 10 sq. ft. of floor area per ton of coal handled per hr.

Until quite recently unit systems have been used only in the smaller power plants, consuming, say, under 50 tons of coal per day. This was due: (1) to the fact that each unit pulverizer had only a small capacity, and (2) to the high power consumption as compared with the storage system. Lately, however, they are being made in larger units (varying from ½ to 5 tons per hr.)and, as a result, the power consumption compares quite favorably with the storage system.

The power usually required to operate the various parts of a moderate-sized unit system is approximately as shown in Table XC.

TABLE XC

POWER CONSUMPTION OF UNIT POWDERED-COAL SYSTEM

	POWER PER 2000 LB COAL PER HR., KILOWATT-HOURS
Pulverizer (including fan)	20 0
Magnetic separator (for removal of tramp iron)	1 5
Coal crusher (if coal will not pass through 1½-in. ring)	0 5
Elevating coal to bunker	0 5
Total	22 5

Assuming 2 lb. of coal per kw.-hr., only 2.25 per cent of the total energy of the coal is used for pulverizing and for supplying air.

$$\frac{(22.5 \times 2)}{2000} \times 100 = 2.25 \text{ per cent}$$

With the larger, high-speed pulverizers the power consumption drops. For example, recent tests[5] show that, with a bituminous coal containing 4 per cent moisture, the total power consumption of the pulverizer shown in Fig. 146 is only 14 kw.-hr. per ton, while with a moisture content of 10 per cent the power consumption is 17 kw.-hr. per ton. A maintenance cost of 2 to 4 cts. per ton of coal ground is claimed for this type of system.

The investment required in unit systems is small, as brought out by Fig. 147, which gives the cost of motor, pulverizer, and blower per ton per hr. for units having capacities of from 0.3 to 10 tons per hr.

The average cost of preparing pulverized coal on the unit system is shown in Fig. 148. On the basis given in this figure the cost for the average plant is about 65 cts. per ton of coal ground.

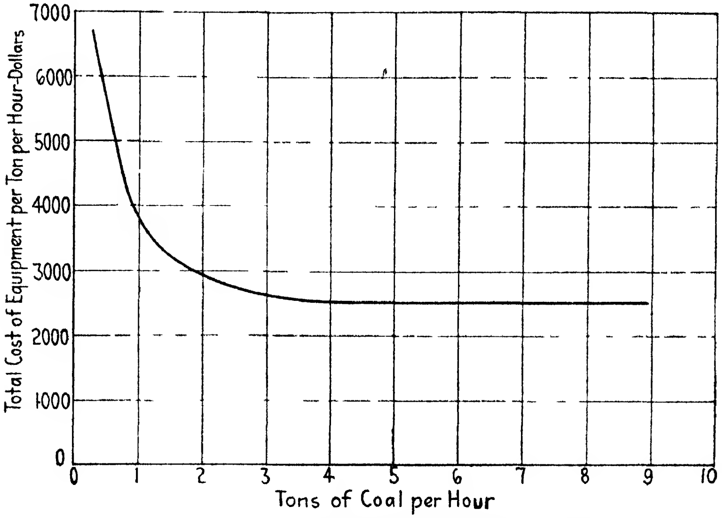

FIG. 147 Cost of unit pulverizers versus capacity.

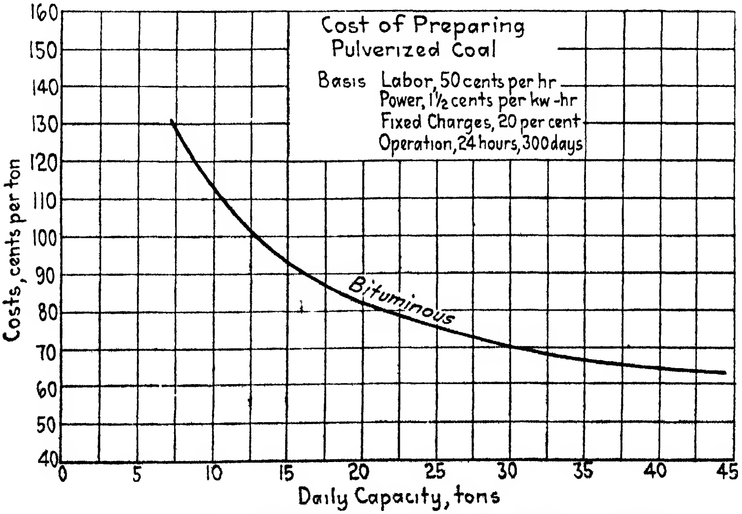

FIG. 148.—Cost of preparing powdered coal by the unit system.

As high efficiencies with this type of apparatus are possible as with the storage systems, and, on account of their low first cost, unit systems have been successfully applied in small plants where the quantity of coal used would not justify the greater expenditure necessary for the more complex storage system. Recent growth of unit systems in larger plants has been very rapid. Usually coal driers are not necessary in the unit system, since a certain amount of drying is accomplished by the air blast. However, a drier generally is necessary for a coal with a high moisture content.

The Storage System.—A powdered-coal installation of the storage type usually includes: (1) a crusher, (2) a drier, (3) pulverizers, (4) storage bins, (5) apparatus for conveying the powdered coal to the furnaces, and (6) feeders or burners. The general arrangement of a typical storage installation is shown in Fig. 149. The coal is first crushed by roll crushers into ¾ to 1 in. lumps and passed over a magnetic separator to remove any "tramp" iron, which, if not removed, would damage the pulverizer. After the magnetic separator the coal passes through the drier to the pulverizer. The pulverized fuel is then distributed to the furnaces by a system of suitable conveyors.

In many plants the raw coal is received in lumps small enough to be fed directly to the driers or pulverizer making preliminary crushing unnecessary. For this reason no description of crushing equipment has been included in the subsequent discussion of the various methods of drying, pulverizing, distributing, and burning the coal.

Drying Equipment for Powdered-coal Systems.—Since high-moisture coal lowers the capacity of the pulverizers and requires higher power consumption in preparation, practically all the coal used with earlier systems was dried before pulverization. It is conceded now, however, that it is necessary to drive off the "surface" or "free" moisture only, and that the combined and absorbed moisture (sometimes called "inherent" moisture) may be left in the coal without greatly lowering the capacity or the efficiency of the pulverizing equipment. Recent tests by Kreisinger and Blizard[6] have shown that some coals containing up to 7 or 8 per cent moisture can be burned with the same completeness of combustion as dried coal. There is, of course, a slight heat loss up the furnace stack of about 0.1 per cent for each 1 per cent of moisture in the coal. As a rule, eastern bituminous coals do not require drying, whereas western bituminous coals, on account of their high free moisture, must be

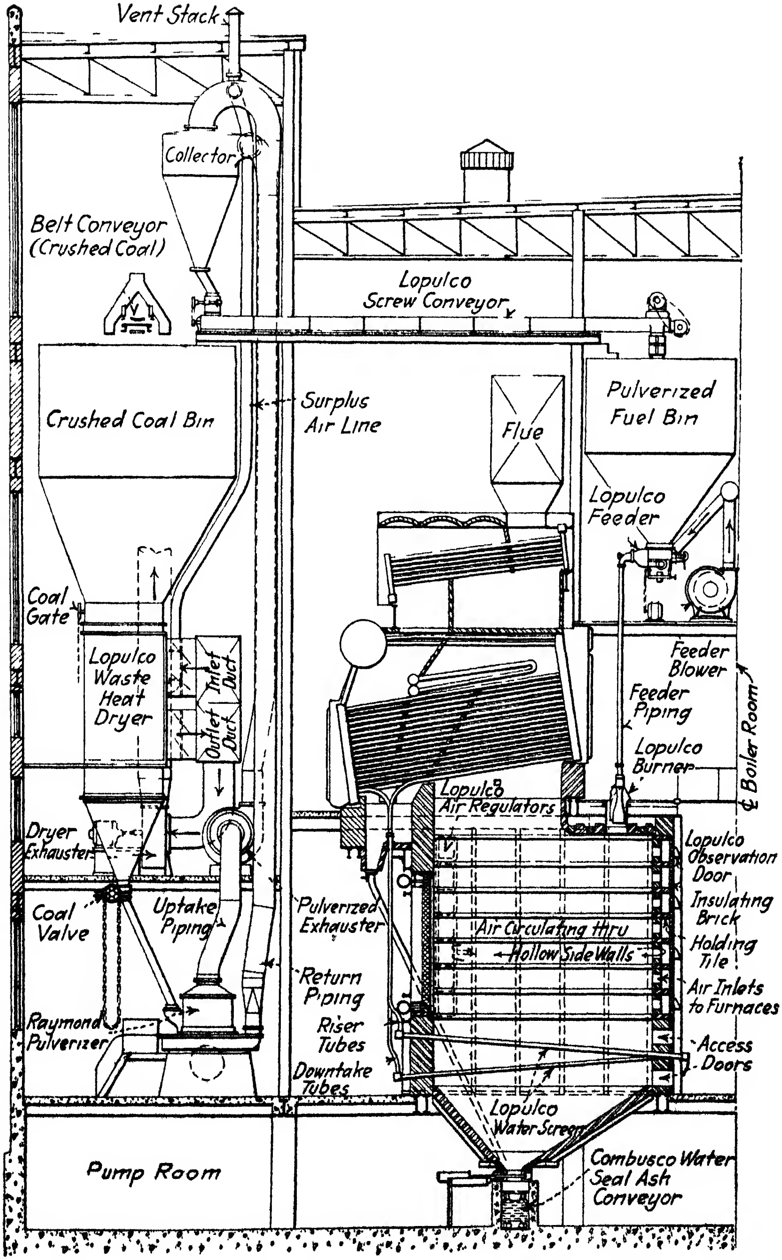

FIG. 149.—Arrangement of a typical storage system of powdered coal application. (*Combustion Engineering Corporation.*)

dried. The installation of a drier (depending on the plant location) may make a plant more nearly independent of the source of coal supply.

The advantages of drying are: (1) the capacity of the pulverizer is increased, as shown[5] by Table XCI, (2) dry coal does not pack readily or arch in bins and conveyors, (3) slightly higher flame temperatures and lower stack losses may be obtained, and (4) the possibility of spontaneous combustion is decreased. The disadvantages of drying are: (1) the increased cost of the equipment required, (2) the increased power, labor, and maintenance, and (3) the added floor space necessary.

TABLE XCI

EFFECT OF MOISTURE CONTENT ON PULVERIZER OPERATION

Moisture in coal, per cent	Capacity, pounds per hour	Approximate power consumed, kilowatt-hours per ton
4	9000	14
6	8750	15
8	8400	16 3
10	8000	17 5

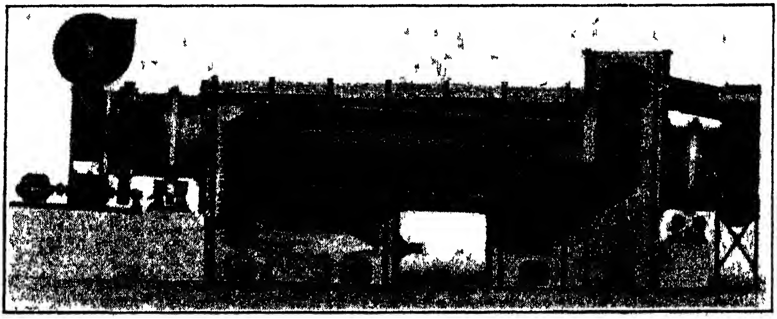

FIG. 150.—Semi-indirect-fired rotary drier. (*The Fuller Lehigh Company*)

Three types of driers are in common use: (1) rotary driers, semi-indirect fired; (2) cell-type driers, which use hot flue gas from the stack; and (3) steam-heated driers. Two driers of the semi-indirect-fired rotary type are shown in Figs. 150 and 151. In the type of drier shown in Fig. 150 the hot gases from the furnace circulate over the outside of the inclined cylindrical drier shell, and are then passed through the inside of the shell countercurrent to the fuel. The drier shown in Fig. 151 consists of a series of relatively short, standard steel pipes. The incoming coal is fed into these tubes and, as the tubes rotate, travels to the discharge end of the drier by gravity. Hot gases from the furnace pass over the outside of tubes and then up through them toward the coal feed end The chief advantage claimed for this drier is the small floor space required. In both these types of drier the gases coming in contact with the outside of the shell (at the coal feed end) can be relatively hot because the coal is kept cool by rapid evaporation of water; the gases passing directly *through* the drier, however, must be cool (about 250°F.) in order to prevent overheating of the coal. For this reason the furnaces for these driers are generally fired with sufficient excess air to cool the flue gases to the desired temperature.

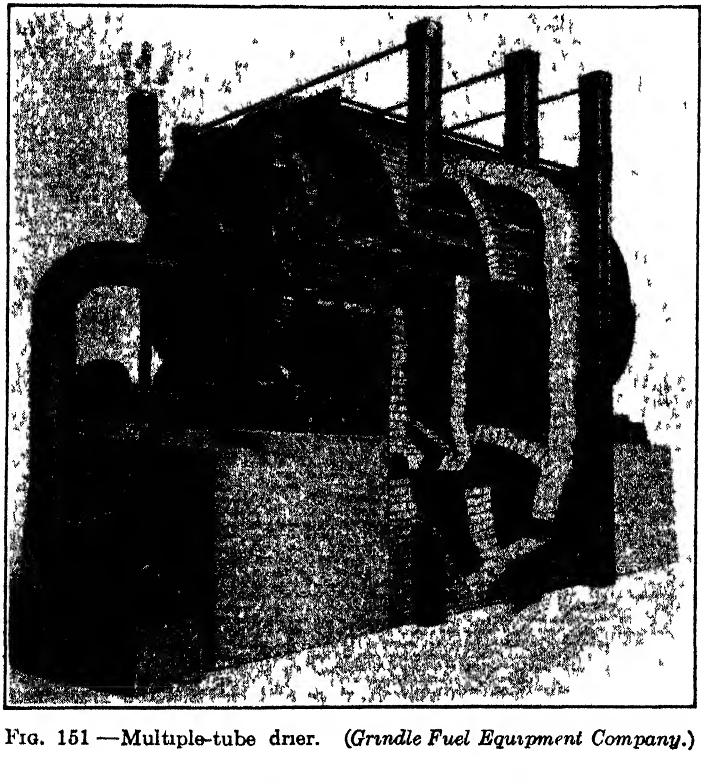

Fig. 151—Multiple-tube drier. (*Grindle Fuel Equipment Company.*)

Two types of driers which utilize hot flue gas from the boiler or furnace stack are shown in Figs. 152 and 153. In the drier shown in Fig. 152 the coal from a storage bin flows by gravity down two louvered passages and is discharged at the bottom directly to the pulverizer. Flue gas from the stack is admitted at the center (as shown), passes through the coal and is drawn out at the sides by means of an exhauster. Steam pipes are provided to extinguish any fires that might result from spontaneous combustion.

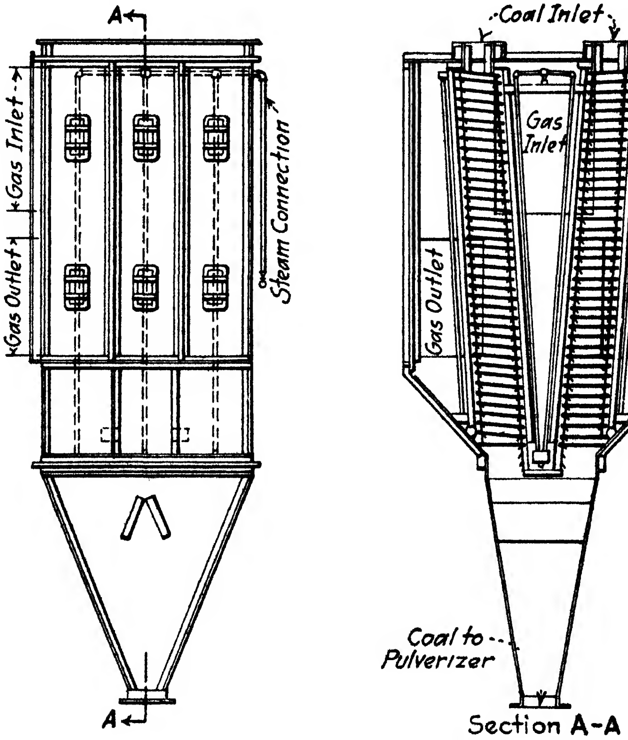

Fig. 152 —Wood type of waste heat coal drier.

The drier shown in Fig. 153 is made up of unit sections, each unit being provided with gas ports as shown in the diagram. Alternate sections have gas inlets and the intervening sections are provided with gas outlets. Stack gas is forced by a fan through the coal in the direction of the arrows, or is drawn through by an exhauster at the top of the drier. Waste-heat driers of this type effect a considerable heat saving and the power requirement is only 2 to 3½ kw.-hr. per ton of coal dried. They require about 25 sq. ft. of heating surface per ton of coal dried per hr. per per cent of moisture and take about 20 per cent of the flue gases. Driers of this type are being widely used in new powdered-coal installations. The exit gases should not be below 150°F., to insure against sweating of the drier.

A third type of drier (similar in design to the waste-heat type just described) which is being employed in some of the more newly designed installations uses steam for the heating medium. The steam is admitted to grids which heat the coal. A current of air is drawn through the coal to remove the moisture. Steam driers require less power to handle the fans, and the air-exit temperature may be run as low as 100°F. About 30 sq. ft. of heating surface are required per ton of coal per hr. per per cent of moisture and about 1 lb. of air is used per lb. of coal. This type of drier is best adapted for use in plants where coal preparation is carried out at such a distance from the stack that waste-heat driers cannot be used.

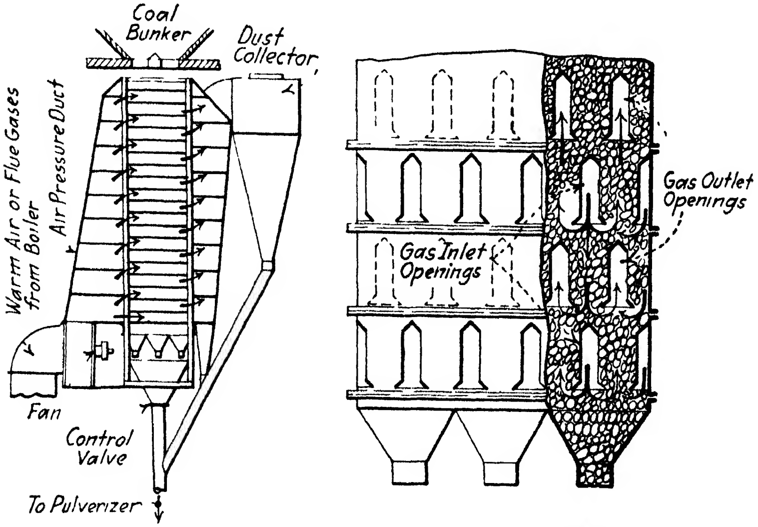

FIG 153.—Randolph type of waste heat coal drier.

Less drying is being done in modern installations than in early applications of powdered coal. Practice in this respect is changing and probably in the future most of the drying will be done in flue-gas driers.

Pulverizing Equipment.—After the coal has been dried it is pulverized to the desired degree of fineness in grinding mills.* The power used to drive these mills is the biggest power item in powdered-coal preparation, and is greatly affected by the degree of pulverization,† by the amount of moisture in the coal, and by the physical characteristics of the coal being pulverized. The cost of pulverization is dependent on the size of the plant as well as on the above factors.

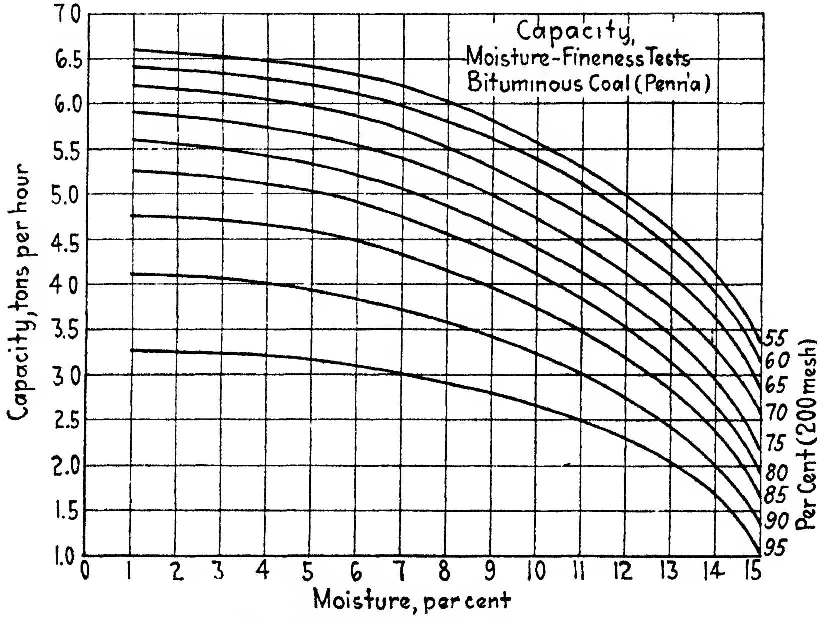

FIG. 154.—Relation between capacity, moisture, and fineness in pulverizing Pennsylvania bituminous coal.

The larger the plant the less the cost of pulverization, since the labor necessary does not increase in proportion to the capacity. Two men, for example, are required in a plant grinding 100 tons per 24 hr., while these same two men could take care of a plant grinding from 100 to 300 tons per day. Moreover, the small plant usually cannot obtain power as cheaply as the large one,

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* Where the coal does not have to be crushed before drying and no separation of iron from the coal has previously been effected a magnetic separator is installed before the pulverizer in order to remove "tramp" iron.

† Among others our thanks are due Mr. S. B. Kanowitz of the Raymond Bros. Impact Pulverizer Co. for much of the following data regarding the power consumption of pulverizing equipment and costs of pulverizing coal.

and usually is not equipped with sufficient spare equipment to make possible repairs while the plant is in operation.

The capacity of a mill is decreased with increasing fineness of grinding and with increasing moisture content of the coal, as is shown by Fig. 154. This figure gives the relationship between capacity, fineness, and moisture in grinding a Pennsylvania bituminous coal. Thus, reading from Fig. 154, if grinding specifications are 75 per cent through 200 mesh (diameter of openings 0.0029 in.) the capacity of a mill of the type on which Fig. 154 is based will be about 5.4 tons per hr. on a 4 per cent moisture coal. The same mill grinding the same coal will have a capacity of only 3.2 tons per hr. if the coal is ground so that 95 per cent will pass through 200 mesh. Similarly, a mill having a capacity of 4.7 tons per hr. with a 2 per cent moisture coal when the specifications are 85 per cent through 200 mesh, will have a capacity of only 3.7 tons per hr. if the moisture is increased to 10 per cent.

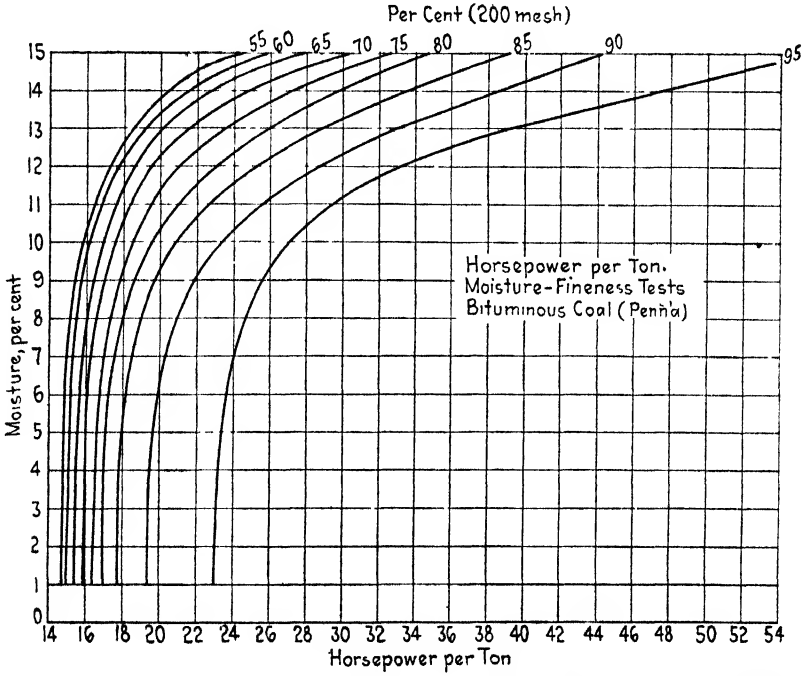

FIG. 155.—Relation between power consumption, fineness, and moisture content in pulverizing Pennsylvania bituminous coal.

The relationship between power consumption (in horsepower per ton of coal pulverized), fineness, and moisture content is shown in Fig. 155. As the figure indicates, the power consumption (and consequently the cost of pulverizing) is increased with increasing moisture content and with increasing fineness of pulverization. Thus the power consumption is about 17.8 hp. per ton when 2 per cent moisture coal is gound, so that 85 per cent will pass through a 200-mesh screen. With a 12 per cent moisture coal the power consumption is increased to 25.3 hp. per ton. Up to a certain point increasing moisture (*i.e.*, inherent or combined moisture) has but little effect, but with further increases in water content (free moisture) the capacity is greatly decreased. By reference to curves of the type shown in Fig. 155 it is often possible to determine the fineness to which a coal may be ground economically and to what extent it will pay to dry before grinding. In a number of tests the total cost of pulverization, including interest, maintenance, depreciation, power, labor, lubrication, etc., ranged from about 38 to 48 cts. per ton pulverized. The average amount chargeable to power was about 62 per cent of the total cost of pulverization (power figured at 2 cts. per kw.-hr.).

The physical characteristics of the coal also play an important part in the economy of grinding. In general, the coal becomes harder and requires more power to grind as the volatile matter decreases and the fixed carbon increases, the anthracites being harder to grind than the bituminous coals. Data on anthracite pulverization are very meager, since only a few pulverized-coal installations are using this type of fuel, and even they are engaged in trying out different kinds of furnaces and burners rather than working primarily on pulverizing problems. Furthermore, the grinding characteristics of anthracites, such as culm, river anthracite, etc., vary so widely that specific data are of little value unless the kind of anthracite handled and its source are known. For example, a roller mill with air separation having a capacity of 200 tons per 24-hr. day ground various anthracites so that 95 per cent passed a 200-mesh screen with the following costs (power 2 cts. per kw.-hr.):

	PER TON
River anthracite	$0.755
Culm	0.676
Clean anthracite	0.571

In considering the operation of pulverizing equipment, it should be borne in mind that the finer the coal the easier it is to obtain efficient combustion. It is necessary, therefore, to balance the increase in power cost against the savings made possible by finer grinding for the particular coal that is being fired. For example, coal with low-volatile matter should be ground finer than coal containing higher-volatile contents. The usual specifications for pulverizing operation are that 95 per cent of the product shall pass through a 100-mesh screen (diameter of openings 0.0059 in.) and 85 per cent shall pass through a 200-mesh screen (diameter of openings 0.0029 in.). Recent work, however, has shown that good combustion can be maintained under boilers when 90 per cent of the fuel will pass through 100 mesh and 65 per cent will pass through 200 mesh. Under these latter specifications (see Fig. 154) the capacity of the pulverizer would be increased by about 30 per cent over that possible under the first specifications, and without any marked decrease in the combustion efficiency.

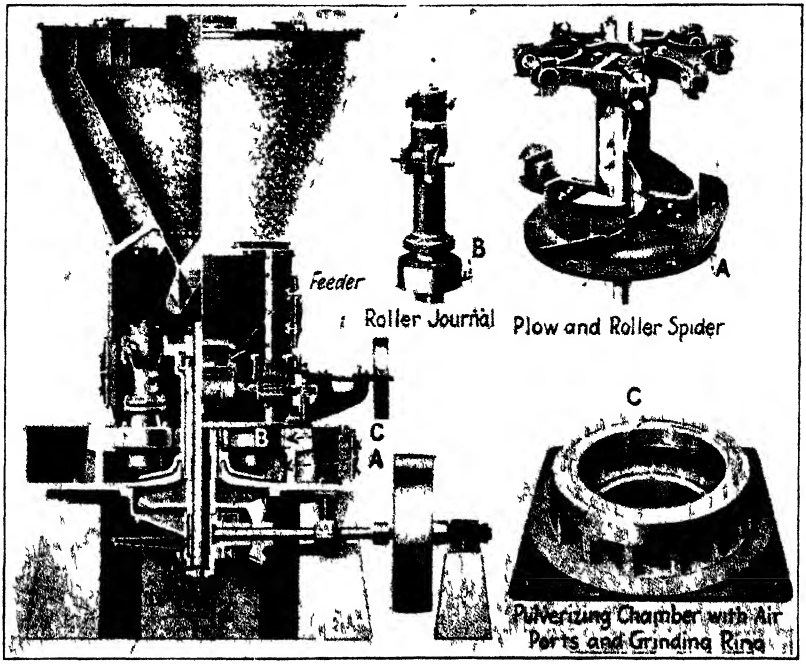

FIG. 156.—Details of roller mill and air separator. (*Raymond Brothers Impact Pulverizer Company*)

The equipment mainly used for pulverizing coal may be illustrated by the following types of pulverizing mills: (1) the Raymond, (2) the Fuller-Lehigh, (3) the Jeffrey swing-hammer, and (4) the Hardinge ball mill.

In the Raymond mill, shown in Fig. 156, the coal is fed through a spout and feeder to a grinding chamber. The plows *A* throw the coal between the rollers *B* and the pulverizing ring *C* A current of air lifts the powdered coal into a separator, situated above the mill, from which the coarse particles fall back into the mill for further grinding. The air then passes into a cyclone separator, where the finely ground material is separated and collected, the air being returned by a fan to the grinding mill. The Fuller-Lehigh mill, shown in Fig. 157, uses four steel balls as pulverizing elements. The balls are driven around the grinding ring by pushers. Two fans are used, an upper one to draw the pulverized coal into the chamber above the grinder, and a lower fan to draw the fine material through a screen. After passing through the screen the pulverized fuel is discharged through the spout shown at the side of the diagram. The same company also manufactures another mill in which the product is separated entirely by an air current rather than by screens (see Fig. 157*C*).

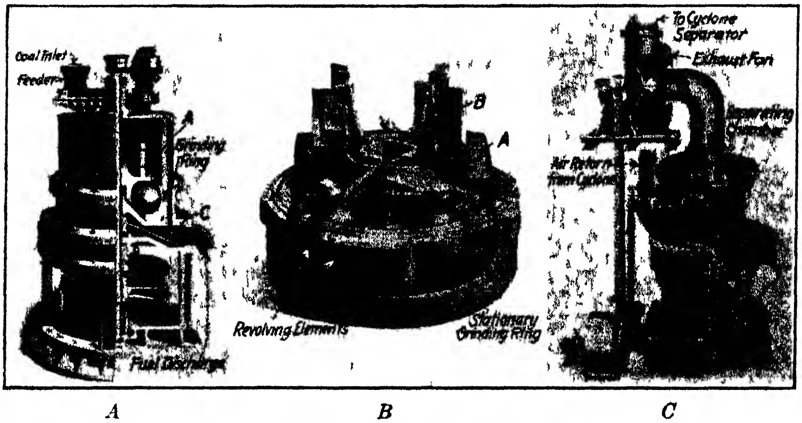

Fig 157 —Details of Fuller Lehigh roller mills.

The Jeffrey swing-hammer pulverizer breaks the coal by striking it with rapidly revolving hammers which drive the coarse pieces against breaker plates. The finely ground coal passes through a series of screen bars. and the coarse material is returned for a second grinding. This is the type of action usually used in unit-system pulverizers.

In the Hardinge mills the coal is fed into a rotating shell in which the grinding is done by the rolling or impact of balls or pebbles. Mills of this type all grind very fine, but, unless combined with air separation, usually require considerably more power for grinding coal than the types previously described.

Powdered-coal Storage.[8]—Care must be exercised in storing powdered coal, since the dust easily takes fire and smolders. Damp powdered coal is very liable to ignite spontaneously and it has been stated[9] that coal dust containing 0.75 per cent moisture and 1 per cent sulfur will invariably take fire in six days. Spontaneous combustion in the storage hoppers of a powdered-coal plant is not particularly serious, however, since the bins are always covered and there is, therefore, insufficient oxygen present to support combustion long. The most serious result of fires in storage bins is the caking of the coal and the consequent loss of its free-running properties.

Storage hoppers should have sides making 30-deg. angles with the vertical to prevent arching of the coal. The capacity of the hoppers should be from 12- to 14-hr. supply for the plant, although 96-hr. storage is said to be a safe upper limit for dry coal. Ground coal has been stored in large bins for from 10 days to two weeks, but it is best to keep it moving each day. When stored for a short time only, powdered coal is light, fluffy, and free running, but it packs into a dense mass on long standing.

Transportation of Powdered Coal to Burners.—The transportation of powdered coal is usually accomplished by one of the methods illustrated[8] in Figs. 158 and 159, that is; (1) by a combination of screw conveyors and bucket elevators, (2) as a suspension in air, (3) air transport—coal discharged through pipe lines by air pressure, or (4) forced aeration—a combination of a worm-gear pump and compressed air.

Spiral-screw conveyors in a dust-tight steel trough for moving the coal horizontally and bucket elevators for the lifts comprise the oldest system of powdered-coal transport. It has been replaced to a large extent, however, by the three systems shown in Fig. 158, although in many modern plants screw conveyors are still used to distribute the powdered coal to the bins (see *B* in Fig. 158). Screw conveyors are reliable and have a low repair cost, and, since some coal is stored in hoppers at the furnace, short delays in conveyor operation are not serious. They are heavy, however, and are not suitable for turning corners or for changing the level. On inclines, the coal is liable to flush through to the lower end and pack the conveyor. This system is dusty and requires numerous motor drives. The real field for screw conveyors is in a compact or simple layout of furnaces or boilers, adjacent to the coal-pulverizing plant (see Fig. 158*A*).

In the system shown in sketch *B*, Fig. 158, coal from the pulverizer is delivered directly to the storage bin by the exhauster. This method is in common use where the distance over which coal must be transported is small. It is simple but has the disadvantage of requiring a large amount of piping, since a return air line must be provided between the cyclone separator and the pulverizer.

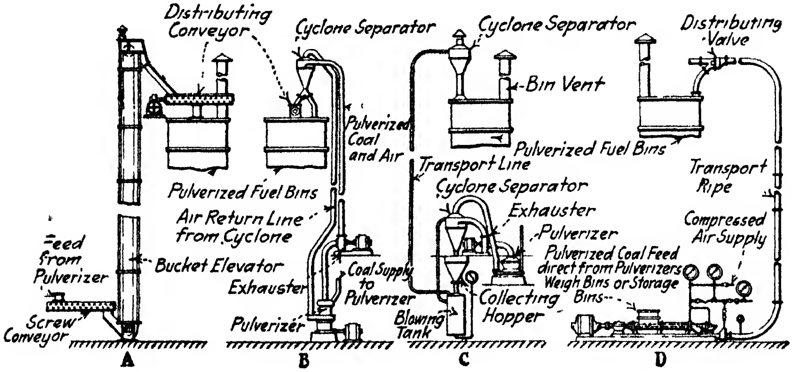

FIG. 158.—Layout of powdered coal conveying systems now in general use. Diagrammatic.

The air-transport system (*C*, in Fig. 158) has been quite widely used for powdered-coal distribution. The powdered coal from the collecting hopper flows by gravity into a blowing tank. The blowing tanks (located in the pulverizer house) are arranged on weighing platforms so that the amount of coal fed to a given furnace bin can be determined accurately. The discharge line through which the coal leaves the blowing tank reaches to within a few inches of the bottom. After the tank has been filled with the desired amount of fuel, compressed air (50 lb. per sq. in.) is admitted near the top of the tank and a stream of powdered coal forced through the transport line. The small amount of air that mixes with the coal (about 1 per cent of that necessary to form a combustible mixture) is separated from the fuel by a cyclone separator which discharges the coal by gravity into the furnace hopper. Small pipes (3- to 4-in. diameter) and normal bends are used in this system. The maintenance cost with this distributing system is very small and the power consumption is less than with most of the other methods. In the Grindle pneumatic conveying system (somewhat similar to *C*, in Fig 158) powdered coal from a pressure tank (which can be used to weigh the coal) is forced into a conveying pipe by a small amount of air (400 cu ft per ton or about 1 per cent of that necessary for complete combustion) under a pressure of 5 to 40 lb. The air is blown into the coal through an injector, aerates the coal, and forces it into furnace hoppers. The small amount of air used is separated from the coal at the furnace hopper by means of a dust arrester.

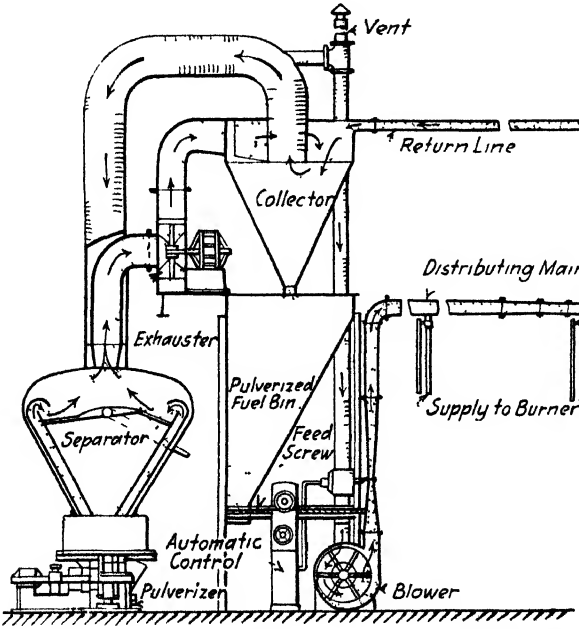

FIG 159 —General arrangement of a typical circulating system of powdered coal transport

The forced aeration system sketched in *D* of Fig. 158 uses a motor-driven worm pump, aided by a small amount of compressed air. The powdered coal falls by gravity into the pump, where it is compressed slightly. As it leaves the pump a small amount of compressed air is admitted through a number of jets.

This mixes with the coal and renders it so fluid that it may be transported readily. Little air is required (about 360 cu ft. per ton). This system uses small pipe (3 to 5 in. in diam) and is especially adapted for conveying coal long distances. The power requirement of the pump is from 1 to 2 kw.-hr. per ton of coal. In recent tests by the manufacturer, powdered coal was successfully conveyed 5400 ft. through a 5-in. pipe.

In the recirculation type of transport system (shown in Fig. 159) the air and the dust are mixed and delivered as a blast around the plant. Coal from the bottom of the bin is delivered to the suction side of a high-pressure blower by a screw conveyor. In the blower the coal is mixed with about 40 per cent of the air required for combustion and delivered to the distributing pipe. The velocity of the coal-air mixture is maintained at about 90 ft. per sec. to eliminate back-firing and to keep coarse coal particles in suspension. Any coal and air not required at the burners is returned and pumped over again. The pressure is usually maintained at 10 to 12 oz. per sq. in. and if the transport system is very long a booster fan may be installed. The disadvantages of this system are: (1) that the fan impeller and the casing are sometimes rapidly eroded by the coal-air mixture, (2) the power consumption is high, (3) coal is stored at a distance from the furnaces and any stoppage of fans causes the shut-down of the entire system. In this method of conveying powdered coal an explosive mixture is being handled at all times. There are a large number of these installations, but there have not been very many serious accidents, although explosion doors and non-return valves are recommended.

Powdered-coal Feeders or Burners.—Powdered-coal feeders or burners in common use are of two general types: (1) burners having a mechanically operated variable-speed screw or flight conveyor which brings the coal into the path of the air for combustion, and (2) burners in which the fuel is fed by compressed air, the feed being regulated by the air pressure. The air pressure usually is low (from 2 to 8 oz. per sq. in.), or a high-pressure injector (with pressures ranging from 40 to 100 lb. per sq. in.) may be used where long cutting flames are desired.

The Warford coal feeder and mixer[1] used by the Anaconda Copper Company is shown in Fig. 160. Air under a pressure of from 19 to 20 in. of water is blown in through the center pipe of the mixer. This air injects secondary air and coal from the

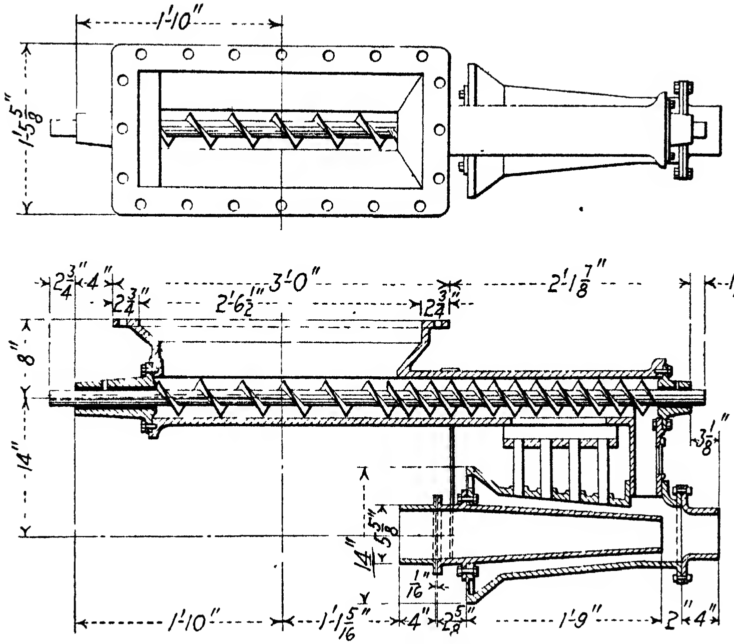

FIG. 160.—Warford pulverized coal feeder and burner

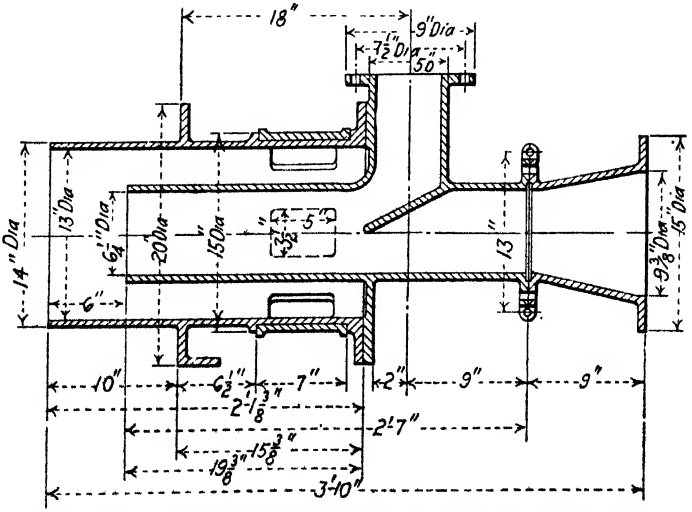

FIG. 161.—Fuller-Lehigh induction type pulverized coal burner.

screw-conveyor feeder through the annular space into a blast pipe, where additional air necessary for combustion is drawn around the blast pipe just as the combustible mixture enters the furnace

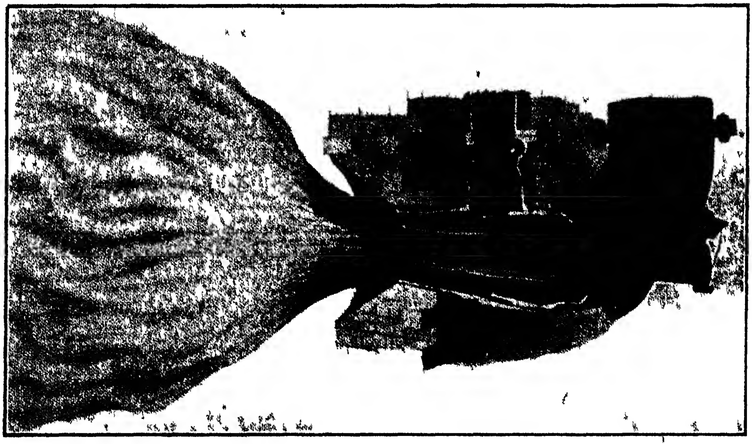

Fig 162.—Rayco powdered coal burner.

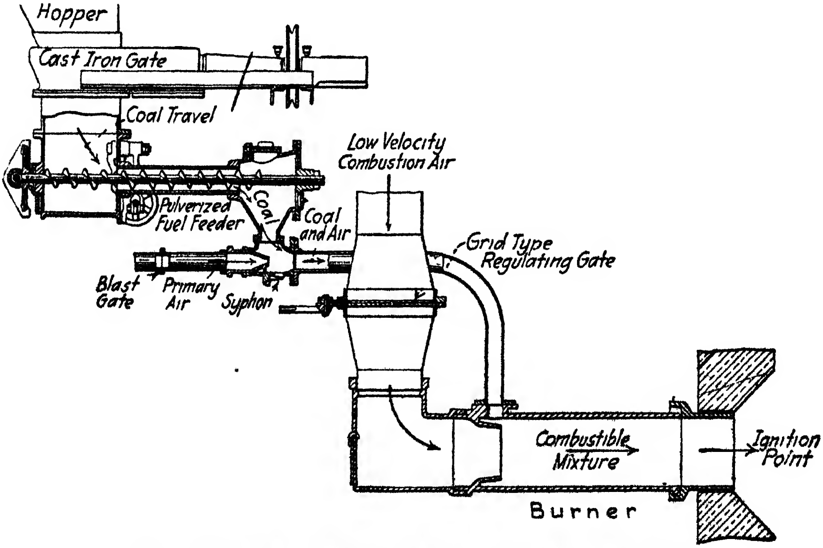

Fig. 163.—Quigley central duct type pulverized coal burner.

The Fuller-Lehigh induction-type burner[1] is shown in Fig. 161. Powdered coal from a feeder drops through the top opening and is mixed with low-pressure air coming from the right-hand opening. Secondary air is drawn in and the coal-air mixture passes into the furnace. The amount of secondary air is regulated by an adjustable register on the outer tube of the burner.

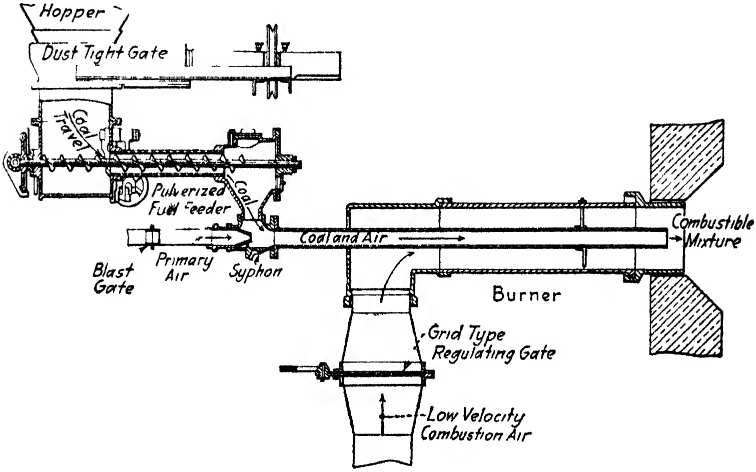

FIG. 164.—Quigley syphon type powdered coal burner.

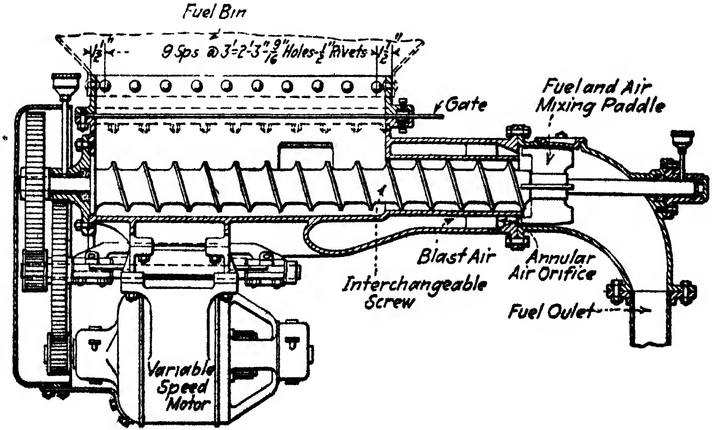

FIG. 165.—Lopulco pulverized coal feeder.

The Rayco burner of the Raymond Bros. Engineering Co. is shown in Fig. 162. The powdered coal is mixed with low-pressure air in the annular space between the two tubes of the burner and on entering the furnace meets the air coming through the center pipe of the burner.

In the Quigley system coal from the feeding mechanism falls into the path of the primary air (5-oz. pressure), as shown in Figs. 163 and 164. The coal-air mixture (about one-eighth of the air necessary for combustion) enters the burner, to which secondary air is admitted under a pressure of about $1\frac{1}{2}$ oz.

The Lopulco coal feeder and vertical burner for boilers, manufactured by the Combustion Engineering Corporation, are shown in Figs. 165 and 166. In this feeding device, air (under a pressure of 10 in. of water) and coal from the feeder screw are thoroughly mixed by the revolving paddles. The fuel and primary air then enter the burner through the inlet pipe, and induced streams of secondary air (sufficient to support ignition at the burner nozzle) are drawn in through two damper-controlled openings in the burner shell. The remainder of the air required for combustion is drawn in through inlets in the furnace front.

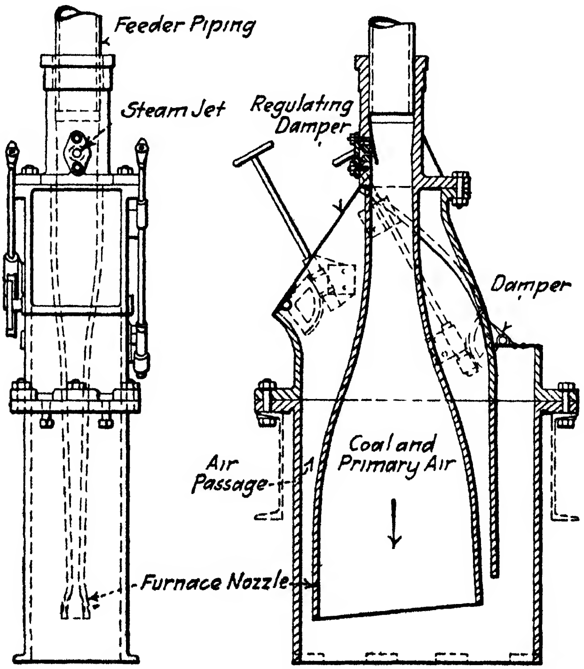

Fig 166.—Lopulco pulverized coal burner.

All the above relatively simple burners and others of similar design have been successfully applied for a wide variety of purposes. There are also several burners of more complicated construction which are designed to give better mixing of coal and air. A burner of this type, which, although used mainly in metallurgical work, has also been applied to boiler furnaces, is shown in Fig. 167. Coal from a screw feed is delivered into a drum where it is mixed with about 25 per cent of the air (low pressure) necessary for combustion. This coal-air mixture then passes into a final mixing chamber, where the remainder of the air necessary is admitted through small jets. All the air necessary for combustion, therefore, is mixed with the coal before entering the furnace.

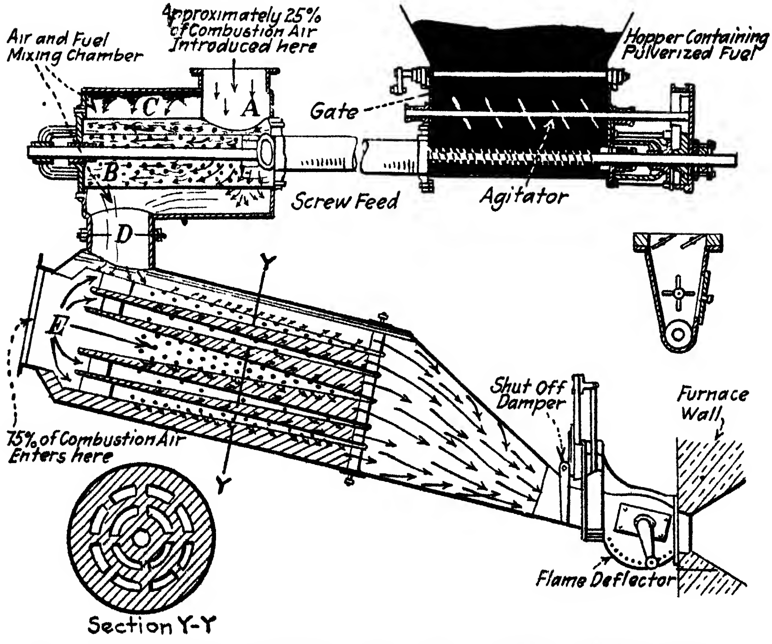

FIG. 167.—Multimix pulverized coal burner. (*Sanford Riley Stoker Co.*)

Depending on the system, the coal-air mixture may be injected into the furnace at a velocity of from 20 to 300 ft. per sec. with success. For most industrial purposes the velocity should be low, in order to give time for combustion and for the ash to fall out. For a cutting flame a high velocity is desired. In burning powdered fuel, as in other types of combustion, it is of paramount importance that the fuel and the air be intimately mixed, and burners* must be designed with this end in view. Powdered coal flames may be shortened by increasing the turbulence just as the coal-air stream enters the furnace. Streamline flow lengthens the flame. For most purposes the major part of the air for combustion should be admitted through the burner or close to it to allow ready adjustment of the flame length and quality. The fuel feed should be at a uniform rate.

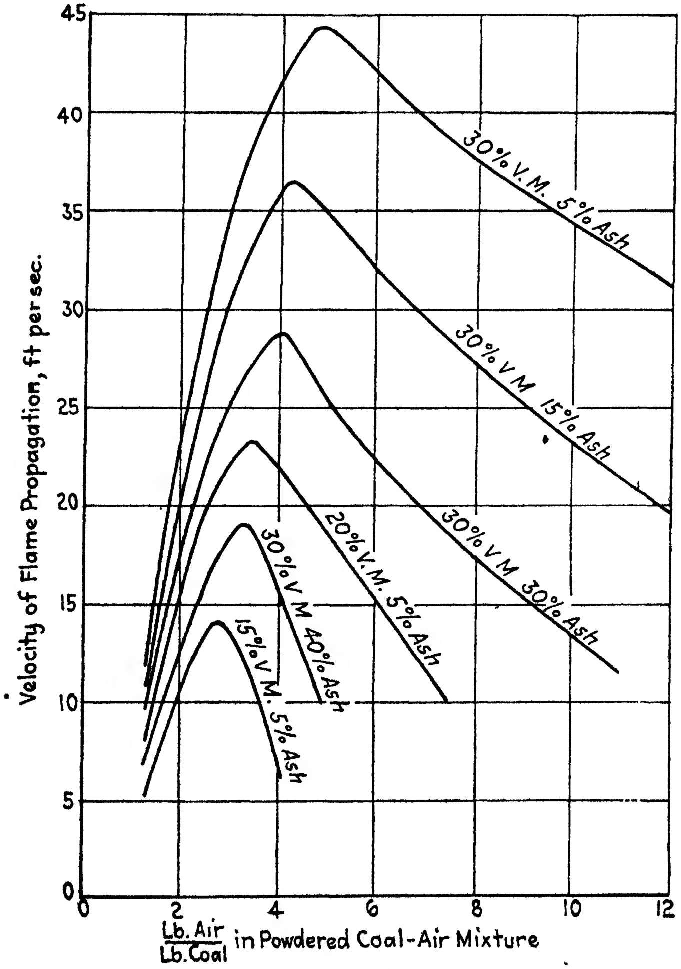

FIG. 168.—Rate of flame propagation in powdered-coal-air mixtures.

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* For certain purposes, where it is desired to have a long, lazy flame, incomplete premixing of coal and air brings about the desired result.

Fuels Suitable for Use in Powdered Form.—Practically any fuel, high or low in either ash or volatile matter, can be successfully fired in powdered form. High-volatile coals, however, on account of their greater ease of ignition and lower cost of pulverization, are better than low-volatile coals or coke. To show the wide difference in composition allowable, three coals which have been successfully fired are: (1) Rhode Island anthracite, containing 6 per cent volatile and 31 per cent ash; (2) anthracite culm, containing 8 per cent volatile and 24 per cent ash; and (3) Illinois coal screenings, containing from 35 to 40 per cent volatile matter and 16 per cent ash. The amount of moisture in the coal is of utmost importance, particularly if driers are not installed. With properly designed equipment, the quality of fuel used may be changed almost at will.

The kind of fuel used has a decided influence on the rate of flame propagation in powdered coal-air mixtures,[10] as indicated in Fig. 168. The bottom scale gives the pounds of air per lb. of coal and the scale at the left gives the speed of flame propagation in feet per sec. This figure shows that the higher the percentage of ash and the lower the percentage of volatile matter the lower will be the flame speed. With most coals the maximum-speed mixture contains from 3 to 5 lb. of air per lb. of coal. The velocity of flame propagation is important in deciding on the velocity of the coal-air mixture in conveyor pipes (to prevent back-firing) and in determining the proper speed of the mixture inside the combustion chamber. Obviously, if the coal-air mixture enters the furnace at too high a velocity the flame will be blown out, but if the proper speed is selected the flame will be steady and will burn close to the burner.

Furnaces for Burning Powdered Coal.—One of the chief objections to powdered coal in the past has been the large combustion spaces required to permit complete combustion of the fuel without excessive deterioration of the refractory furnace linings resulting from the high temperatures and the impingement of the flame on the brickwork. Good results have been obtained with combustion rates of from 0.6 to 2 lb. of coal per hr. per cu. ft. of combustion space, as shown[11] by Fig. 169. This figure shows (1) the per cent efficiency of the boiler and furnace and (2) the temperature of the flue gases leaving the boiler, both being plotted as functions of the B.t.u. liberated per hr. per cu. ft. of combustion space. In many furnaces the most efficient conditions are between about 12,000 and 18,000 B.t.u. per hr. per cu. ft. of combustion space; *i.e.*, from 1 to 1.5 lb. of coal per hr. per cu. ft. of combustion space. Above 18,000 B.t.u. per hr. per cu. ft. (see Fig. 169) the efficiency of the boiler and furnace decreases considerably. These large furnace volumes have been used to give the longest possible length of flame travel and to minimize erosion of the refractory furnace lining. With better mixing of the powdered coal and air, with preheated air, and with water cooling of furnace walls, it seems probable that the high efficiencies possible with powdered coal can be maintained with considerably smaller furnaces than are now deemed necessary.

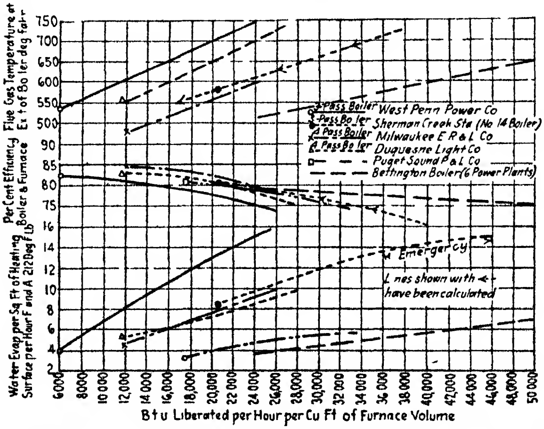

FIG. 169.—Efficiency of boiler, and temperature of flue gases versus amount of heat liberated per unit volume of combustion space

Excellent results in relatively small water-cooled combustion chambers have been obtained already. This is shown by the following table (adapted from Coutant[11]):

TABLE XCII

RATES OF COMBUSTION IN VARIOUS POWDERED-COAL-FIRED FURNACES

Type	Pounds coal burned per cu. ft. per hr.	B.t u. per cu. ft. per hr	Efficiency
Average boiler furnace	1–2	13,000–26,000	65–83
Water-wall furnace (Sherman Creek)	1.9–3 7	21,000–46,000	62–83
Water-wall furnace (Bettington boiler).	2–5.4	22,000–70,000	66–80

Burners should be kept away from the side walls of the furnace to prevent the flame from cutting the brick and allowance should be made for expansion of the gases after ignition in order to slow them up and to allow the ash to settle. Slagging of the ash depends on the ash-fusion point, and on the furnace temperature. Most slags chill solid at about 1800°F. but become soft and pasty above 2000°F. If this hot, pasty mass comes into contact with

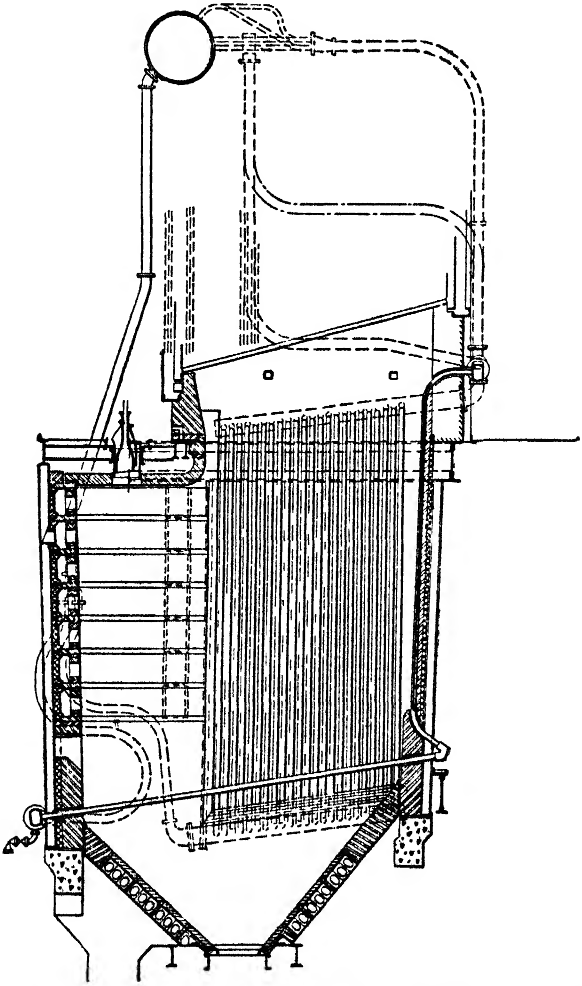

FIG. 170.—Vertical section through powdered-coal furnace. Equipped with water screen over the bottom and rear wall, and also with fin-tube side walls. (Babcock & Wilcox cross-drum boiler of 18,010 sq. ft. of heating surface, at Cahokia Station, Union Electric Light and Power Company, St. Louis, Mo. This is the latest design.)

the brickwork the refractory may be rapidly worn away and removal of the fused slag is difficult. Kreisinger[6] and his co-workers state that flat, spreading burners seem to cause less slagging and to give better results than round burners, probably due to the increased ease (in flat burners) with which heat radiates from the interior coal particles to walls, tubes, etc. They found that attempts to harden the slag and to cool the bottom of the furnace by introducing air through the slag doors caused the flames to rise and shortened their path through the furnace, thus causing a loss due to incomplete combustion. They believe that the best way to avoid slagging without increasing the excess air is to cool the bottom of the furnace by using a water screen, *i.e.*, allowing the ash to fall through water-cooled coils connected with the boiler. This practice has been quite widely adopted, since it not only solves the ash-slagging problem but also serves to increase the heat-receiving surface of the boiler.

Following the successful application of water screens to furnace bottoms, the idea was expanded, first, to the water cooling of the back walls (which, when made of brick, were often seriously eroded), and, finally, to the water cooling of the side walls as well. Figure 170 shows[12] one of the last four furnaces for the Cahokia power station. The furnace is equipped with a water screen over the bottom, with a water-cooled back wall and with fin-tube* side walls. All the water-cooled tubes are connected with the boiler circulation and serve to increase the heat absorbing surface.

In the early days of burning solid coal for steam raising, the coal commonly was fired in Scotch marine-type boilers in which the fuel bed was practically surrounded by the heat-absorbing surface of the boiler. With the methods then in vogue (hand firing on grates, the volatile matter of the coal distilling in large non-uniform masses) great quantities of smoke were produced and the efficiency was low; consequently, refractory walls of low heat conductivity were employed to hold in the heat until combustion was completed. Now, however, due to the increased temperature of the powdered-coal flame, and due to the steady input of volatile matter, the trend is back toward the marine type of boiler with its water-cooled (jacketed) walls, and, in

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* In the fin-tube furnace, metal fins are welded to the vertical tubes along the furnace side walls. The heat absorbed by these fins is transmitted to the water tubes (see Chap. XIV, p. 416).

powdered-coal furnaces, refractories are taking second place. The trend of powdered-coal furnace design will probably be toward *complete* water cooling with the use of flue-gas air preheaters to raise the flame temperature enough to prevent overcooling (and consequent incomplete combustion) of the burning powdered coal.

The use of hollow air-cooled furnace walls, as illustrated[6] by Fig. 171, has met with considerable success. From 60 to 80 per cent of the air for combustion passes through channels between the furnace lining and the outer wall. This process not only preheats the entering air, but also reduces troubles from slagging ash and furnace-wall erosion, and markedly lowers radiation losses. This method of furnace construction, although successful with coals having ash which fuses at a relatively high

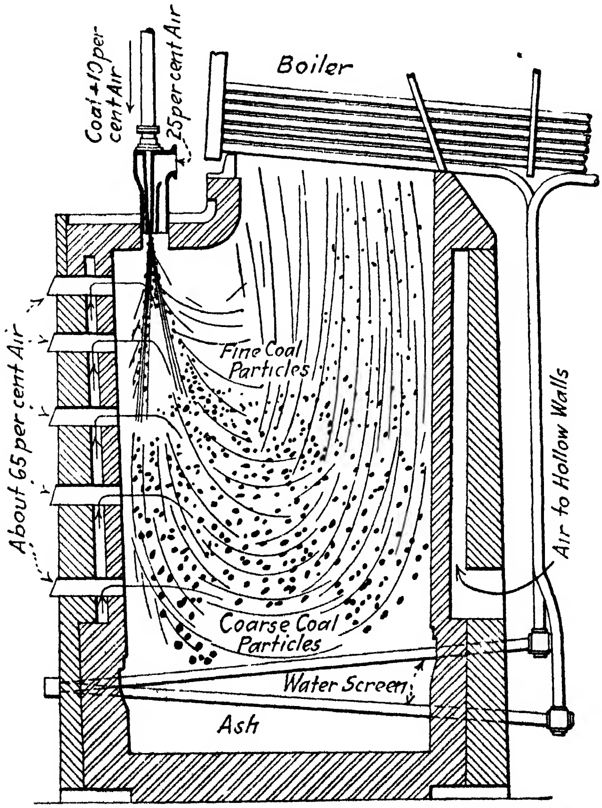

FIG. 171.—Furnace with air cooled walls and water screen.

temperature, does not seem to offer the possibilities of the water-cooled furnace.

As mentioned earlier, in powdered-coal burning the problem of mixing the finely divided coal particles with the required air for combustion is of the utmost importance, particularly if the capacity of the furnace per unit volume is to be as large as possible. It has been recognized for some time that the coal and air should be thoroughly mixed prior to being blown into the furnace. However, it has not been so well recognized that mixing, or

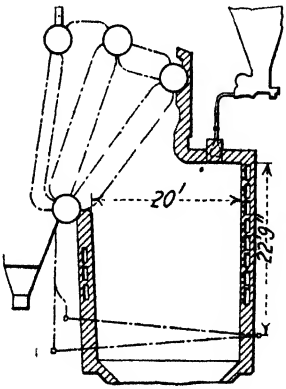

Fig. 172.

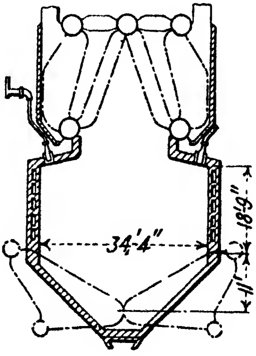

Fig. 173.

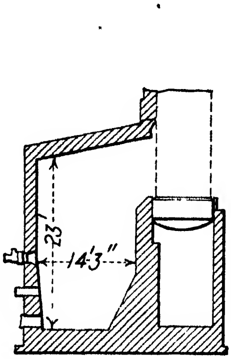

Fig. 174.

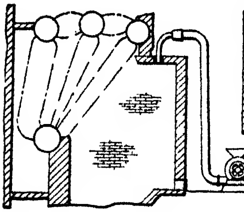

Fig. 175.

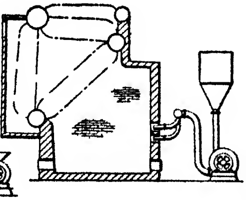

Fig. 176.

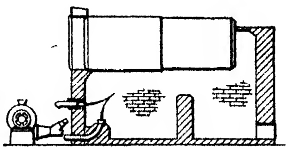

Fig. 177.

Fig. 172.—Section of inclined water-tube boiler and furnace. Middletown plant, Metropolitan Edison Company. Eight burners, 14,770 sq. ft. boiler surface, 11,000 cu. ft. furnace volume.

Fig. 173.—W-type boiler setting and furnace. Trenton-Channel plant, Detroit Edison Company. Sixteen burners, 29,087 sq. ft. boiler surface, 25,140 cu. ft. furnace volume.

Fig. 174.—Vertical water-tube boiler setting and furnace. Eline's Inc., Milwaukee. Eight burners, 5000 sq. ft. boiler surface, 3500 cu. ft. furnace volume.

Figs. 175–177.—Typical installation of unit system of powdered coal application to boiler furnace.

turbulence, within the furnace itself is also a necessity. The CO_2 from each of the burning coal particles must be removed and fresh oxygen supplied continuously if combustion is to be rapid, and turbulence within the furnace is one of the best methods for obtaining this result. Creating turbulence within the furnace is a simple matter in connection with water-jacketed furnaces as compared to refractory-lined furnaces where erosion due to molten ash must be kept in mind continually.

Diagrammatic sketches of the layouts of a number of successful installations for burning powdered coal under boilers (by both the storage and unit systems) are shown in Figs. 172 to 177. In all of these boiler furnaces the path of flame travel is long and large combustion spaces are the rule. The trend of present practice in this latter respect is toward larger combustion spaces, the Lakeside installation shown in Fig. 171 (designed in 1919) having a ratio of 1.86 sq. ft. of heat-absorbing surface per cu. ft. of combustion surface, the Cahokia setting in Fig. 170 (1922) a ratio of 1.56, the Trenton Channel installation (1923) in Fig. 173 a ratio of 1.16, while an installation now being built will have more cubic feet of combustion space than square feet of heat absorbing surface (*i.e.*, the ratio will be less than 1). As already pointed out, however, it is to be expected that, with the adoption of completely water-cooled furnaces, it may be possible to reduce considerably the size of combustion space in the future.

The results of a series of tests run on the Lakeside power station of the Milwaukee Electric Railway & Light Co. (see Fig. 171) are shown[13] in Fig. 178. The lowest curve gives the loss due to undeveloped heat, including carbon in the ash and carbon monoxide in the flue gases. This loss was about 1 per cent in most cases, but at low ratings it was somewhat higher (due to the low excess air used). This loss also increased at high ratings, probably because the velocity of the coal particles through the combustion space was increased and therefore some of the larger particles were not completely burned. The second curve shows the rate of heat evolution per cu. ft. of combustion space, the maximum economical rate for this particular furnace being 23,000 B.t.u. per cu. ft. per hr. (equivalent to about 1.8 lb. of coal per cu. ft. per hr.). The excess air curve (third from the bottom) shows that with powdered coal it is possible to maintain efficient combustion with the excess air held at 10 to 25 per cent. The thermal efficiency, shown by the two upper curves, is about 86 per cent (including boiler and superheater), and when the economizer is included the efficiency ranges between 86 and 90 per cent.

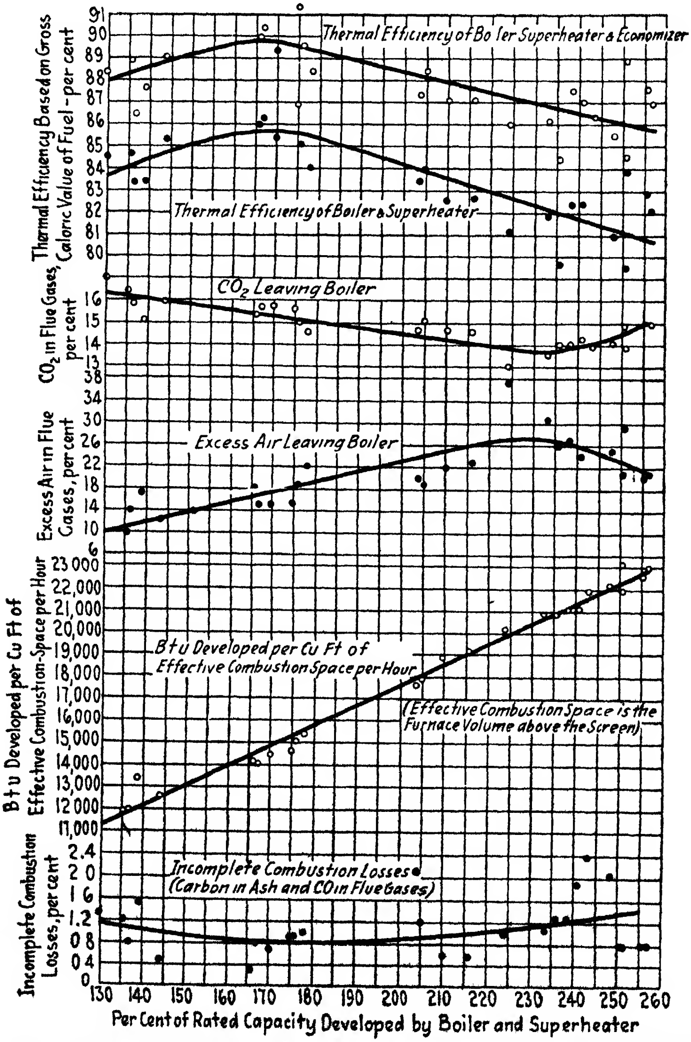

FIG. 178.—Results of tests carried out on powdered coal fired boiler furnace at Lakeside power station.

Costs of Powdered Coal.—The total costs of preparing, delivering, and firing powdered coal in a modern, well-designed plant average about 70 to 90 cts. a ton. This figure includes the cost of power, repairs, drier fuel, labor, depreciation, interest on the investment, taxes, and insurance. The total will, of course, vary in different installations, depending on the plant, the kind of coal used, the capacity, and local conditions.

Cases have been reported where the overall cost of powdered coal preparation was as low as 30 cts. a ton and as high as $1 a ton.

Although no set figures can be given on account of locally varying conditions, Fig. 179 shows the general trend in the variation in cost of preparation of powdered coal in plants of different capacities. This figure brings out the greater cost of preparing coal for small plants. It will be seen that for plants burning over 400 tons of bituminous coal per day the cost of labor, power, and other operating expense ranges from 42 to 54 cts. per ton. These data may be compared with operating figures from the Lakeside plant of the Milwaukee Electric Railway & Light Co.,[14] which show a total labor cost (from coal bin to ash hopper) of 22 cts. per ton, a cost of 9 cts. for maintenance of the pulverizing system and burners, and a cost of 3 cts. per ton for upkeep of furnace walls, making the total cost (exclusive of investment charges) 34 cts. per ton. The total power consumption at this plant is 20 kw.-hr. per ton of coal fired.

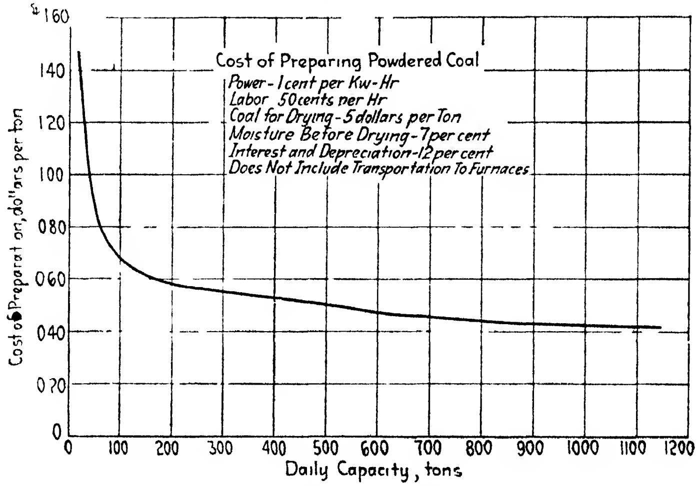

FIG. 179.—Cost of preparing powdered coal in storage systems. (*Data of Fuller Lehigh Company.*)

The large capital investment required usually makes it inadvisable to attempt the installation of storage-type powdered-coal equipment in plants using less than 80 to 100 tons of coal a day, though, with the unit-type apparatus, powdered-coal installations often may be made economically in plants burning from 15 tons a day up.

Safety in Powdered-coal Installations.*—Although in the early days of the art a number of serious explosions occurred in powdered-coal plants, the danger is not great if common sense and good housekeeping prevail. The explosions so far reported have been caused either by the ignition of an inflammable gas-air mixture (such as would result from driving off the volatile matter of a coal in an overheated drier), or by the ignition of a powdered-coal-air mixture.†

To prevent the formation of an inflammable gas-air mixture, special care must be used in order to prevent overheating of coal in the drier. The drier should never be stopped while it contains a charge of coal.

In order to prevent the formation of an inflammable coal-dust air mixture the following precautions, outlined by Tracy, should be followed: (1) The pulverizing plant and the furnace plant should be kept free from an accumulation of dust and must be adequately ventilated. (2) Dust should not be swept up without either wetting or mixing with an excess of fine non-combustible material. (3) All conveyors, elevators, and pulverized-coal bins should be tightly closed and should never be opened while running. (4) All switches should be outside the pulverizer house or else kept in dustproof casings. Guard against sparks from motors; prevent arcing from loose socket connections, etc.; ground all rapidly moving machinery to prevent trouble from static electricity. (5) Prohibit smoking, open lights, or torches in a coal-pulverizing plant.

Advantages of Powdered Coal.—The chief advantages of powdered coal lie in the great flexibility, the high efficiencies that it is possible to maintain, and the wide range of fuels which may be used successfully. Coal of all types, from the low-grade fuels, such as lignite or high-ash graphitic anthracites, to the high-rank, high-quality bituminous coals have been successfully adapted for powdered-coal firing.

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* See NEWELL and PALM[15] for a more complete discussion of powdered-coal hazards.

† The lower explosive limit of a pulverized-coal-air mixture is about 0.032 oz. of coal per cu. ft. of air.

Combustion with pulverized fuel is practically complete with only small amounts of excess air; thus, with excess air maintained at 15 to 25 per cent it is possible to have no carbon monoxide (or only a trace) in the flue gases and still leave less carbon in the ash than is ordinarily possible with mechanical stokers.[*] Due to the high temperatures of the powdered-coal flame, the low excess air possible, and the completeness of combustion, high efficiencies are obtained. Thus boiler plants can be operated continuously with an average efficiency of nearly 85 per cent, and with maximum efficiencies of about 90 per cent, including the economizer.

Powdered coal possesses many of the properties of a combustible gas and has the great flexibility and low banking losses that are possible with gaseous fuels. Thus in a boiler furnace in which *no coal* had been burned for 6 hr.[12] the boiler was brought to 300 per cent of rating in *less than* 5 *min.* This great flexibility and practically zero stand-by loss cannot be obtained with any other type of coal-firing equipment now known. Moreover, all the powdered-coal preparation and feeding equipment is *outside* the furnace, and in case of a mechanical breakdown it is not necessary to let the furnace cool down before making repairs. No arches, expensive to install and maintain, are necessary in powdered-coal furnaces. Close regulation of the desired rate of combustion is quickly and easily made.

The labor required in powdered-coal installations is small and fuel distribution is easy—similar to turning on and shutting off a gas. The energy required to crush and grind the coal is as low as that required to atomize fuel oil in oil-burning installations. High furnace capacities are often possible; thus, in Table XCIII are shown the results obtained[2] in one of the plants of the Anaconda Copper Company.

TABLE XCIII

CAPACITY OF POWDERED-COAL-FIRED VS. HAND-FIRED FURNACE

	Tons melted per day	Tons of copper melted per ton of coal
Hand-fired furnace............. ...	251	3.88
Powdered-coal-fired furnace........	476	7.08

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[*] Certainly without the use of preheated primary air.

Practically smokeless combustion can be secured, even with high-volatile fuels; this may be important in localities where smoke regulations are strict. The characteristics of the flame are under control, both as to length and as to its oxidizing or reducing character.

Disadvantages of Powdered Coal.—The chief disadvantage of powdered coal has been the cost of the equipment necessary for its preparation. This disadvantage is being overcome to a certain extent, however, although the amount of capital expenditure still precludes the use of pulverized fuel in very small plants.

There is a danger from molten ash, which may make ash removal not only difficult, but, by impinging on the brickwork, may cause rapid erosion of furnace linings. Also, some consider that the dissemination of finely powdered ash into the atmosphere is objectionable.

For a given plant it is necessary to balance the advantages of having an almost ideal method of combustion against the difficulties and cost of pulverization and ash removal. Powdered coal, on account of the savings possible through high efficiency, and on account of the many other advantages, should be carefully considered when a new installation is to be chosen. It is doubtful in many cases whether the savings possible will justify the replacement of existing efficient equipment, such as mechanical stokers, particularly if air preheaters are installed in connection with the stokers.

BIBLIOGRAPHY

1. BLIZARD: "Preparation, Transportation and Combustion of Powdered Coal," Can. Mines Branch, *Bull.* 564 (1921), and U. S. Bur. Mines, *Bull.* 217 (1923).
2. HARVEY: "Pulverized Coal Systems in America," Fuel Research Board, *Special Rept.* 1 (1919); see also HARVEY: "Pulverized Coal," Benn Brothers (1924).
3. COLE: *Power*, **59**, 900, 940, 955, 1022 (1924).
4. Review of Cleveland Powdered-coal Symposium, *Power*, **61**, 158 (1925).
5. COUTANT: *Mech. Eng.*, **47**, 183 (1925). Cf. also reference 4.
6. KREISINGER and others: "Tests of a Powdered Coal Plant," U. S. Bur. Mines, *Tech. Paper* 316 (1923), and "An Investigation of Powdered Coal as a Fuel for Power Plant Boilers," U. S. Bur. Mines, *Bull.* 223 (1923).
7. KANOWITZ: *Chem. Met. Eng.*, **37**, 199 (1925).
8. H. S. B. W. Cochrane Corporation: "Finding and Stopping Waste in Modern Boiler Rooms" (1921).

9. H. S. Taylor: "Fuel Production and Utilization," D. Van Nostrand Co. (1920).
10. Kreisinger, in discussion of Hobbs and Heller: *Proc.* Eng. Soc. Western Pa., **39**, 247 (1923).
11. Coutant: *Power*, **61** (Jan. 13, 1925). See also article in *Blast Furnace Steel Plant*, **13**, 156 (1925).
12. Kreisinger: *Mech. Eng.*, **47**, 19 (1925).
13. Kreisinger and Blizard: *Ind. Eng. Chem.*, **15**, 249 (1923).
14 Anderson: *Mech Eng.*, **47**, 228 (1925).
15. Newell and Palm: *Mech. Eng.* (Nov., 1924).

CHAPTER XVI

THE COMBUSTION OF FUEL OIL

The origin and geology of petroleum and the properties of fuel oil which are important from the viewpoint of combustion have already been discussed in Chap. VI. · It also has been pointed out that, before actual combustion, fuel oil is first changed by vaporization into the elementary fuel, "gaseous hydrocarbons." Smokeless combustion of these gaseous hydrocarbons will result if the hydrocarbons are in intimate contact with sufficient oxygen, and if the mixture is kept at a high temperature for a sufficient time to allow this oxygen to enter the hydrocarbon molecule, *i.e.*, allow hydroxylation (see Chap. IX). On the other hand, if there is a local deficiency of oxygen, the hydrocarbons will "crack" or thermally decompose to give carbon (soot) and hydrogen, and a smoky flame will result. The same result will take place if the gases are cooled too quickly. In other words, in the combustion of fuel oil there is a race between two reactions:

(1) Oil + Oxygen *by hydroxylation to* Carbon dioxide + Water Vapor
(2) Oil + Heat *by cracking to* Carbon (Soot) + Hydrogen

For the successful combustion of fuel oil, conditions must be such that the speed of the first reaction is stimulated while the second reaction is repressed.

The types and costs of fuel-oil equipment and the general considerations to be borne in mind with regard to the combustion of oil fuel are considered in this chapter. The main problems in oil burning are: (1) maintaining a uniform flow of heated oil to the burners, (2) then thoroughly atomizing the oil, and (3) mixing it with the proper amount of air. Therefore, oil-handling equipment is discussed first. The various burners for atomizing liquid fuel, air registers, and the characteristics of oil-fired furnaces are described in the latter part of the chapter.

FUEL-OIL EQUIPMENT

In handling fuel oil the following apparatus is required: (1) storage tanks, (2) strainers for removing dirt, (3) pumps, (4) oil heaters to raise the oil to the proper temperature for pumping and atomization, (5) piping, (6) meters and thermometers, and (7) burners or atomizers. The layout of a typical system is shown in Fig. 180.

Storage Tanks.—Storage facilities must be provided to maintain enough oil on hand to meet the plant requirements (when operating at its maximum load) over the longest probable period between oil deliveries. The amount of tankage necessary to insure against failure of the plant's fuel supply depends on (1) the distance from the supply, (2) the way in which it is shipped (*i.e.*, by tank car, truck, boat, etc.), and (3) the regularity with which it is shipped. For example, a plant located near the refinery supplying the oil usually does not need to provide as much storage as a plant located at a distance from the source of supply. This is particularly true in a northern climate where winter deliveries may be uncertain.

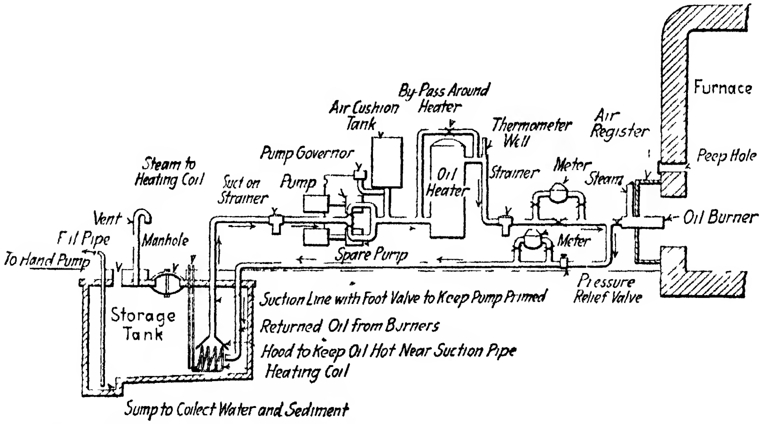

FIG. 180—Diagrammatic layout of a typical oil-burning system (steam-atomized burners), showing necessary equipment

The tanks may be located either above or underground Tanks located above ground are cheaper, but the fire hazard is greater and this type of installation is often prohibited by law within the fire limits of cities. Underground tanks, though more expensive, are subject to much less danger from fire, and in case of a breakdown the oil drains back to the tank rather than into the plant.

Cylindrical steel tanks (riveted-plate construction) are commonly used. They have the advantages of: (1) relative freedom from leaks and easy repair of leaks which have developed, (2) low first cost, and (3) a high salvage value. Concrete storage tanks, which are coming into wider use, have the following advantages: (1) there is no deterioration through corrosion, (2) they are suitable for all locations and can be built in practically any shape, (3) their construction is simple and rapid, (4) no repainting is necessary, as in steel tanks, and (5) the top of an underground concrete tank is useful as a floor for storage space.

Two or more tanks should be installed usually, so that one may be repaired without shutting down the entire plant. The tanks should be located near a spur track and as near the furnaces as possible to simplify unloading from tank cars and to shorten the distance the oil must be pumped to the burners. It is desirable to unload tank cars by gravity, and the fill pipe should be short, of large diameter, and have a steep slope. A flexible hose or hopper fill should be provided for the unloading car.

In addition to the fill pipe, oil-storage tanks should be provided with a suction pipe, a vent pipe for the escape of vapors in case of a fire, a pipe line for the return of oil not used at the burners, a smothering pipe, a heating coil, and a measuring device. In most tanks the fill pipe and the vent pipe may be on the same line. The top of the vent pipe should be goose-necked and should be covered with wire gauze to prevent the flame of accidentally ignited vapors from traveling back into the tank. The smothering pipe (using steam or Foamite for extinguishing purposes) can be used to put out any fires or to flush vapors from the tank prior to inspection or cleaning.

An oil-storage tank of good design is shown in Fig. 180. A hand or intermittent power-driven, gear pump is provided for removing water that collects in the sump due to the slight slope of the tank bottom. The figure indicates connections for a steam coil for heating the oil at the entrance to the suction line. A small hood is used to keep the warm oil near the mouth of the suction line. Excess heated oil from the burners is returned under the hood. This arrangement keeps heated oil under the hood and greatly decreases the amount of steam required for the heating coil. Since hot oil is largely kept at the bottom of the tank, there is only a small heat loss from the top.

Installations should always be so designed that, in case of a break in the pump line or at the burners, oil cannot be siphoned from the storage tank to the furnaces or to the furnace-room floor.*

Strainers.—Strainers must always be provided to remove the dirt and foreign matter which would otherwise be liable to clog or erode the burners. Strainers, which consist of perforated metal baskets, are placed both before and after the oil pump. To eliminate a forced shutdown due to plugging, duplex strainers (two baskets in parallel) are advisable. In strainers of this type the flow of oil through either basket is controlled by a single lever, and the idle strainer can be cleaned while the other is in operation. Sometimes a by-pass (containing a relief valve) around the strainer is provided in order to relieve any pressure that may build up due to clogging.

Large-mesh strainers are used on the suction side and small-mesh strainers on the discharge side. With heavy oil, 16- to 20-mesh screens are used on the suction side of the pump, and 40- to 70-mesh on the pressure side, particularly if the burner has a small orifice. With light oils, 30- to 40-mesh screens are installed on the suction side and 60- to 70-mesh on the pressure side. In addition to the strainers on the main pipe line, it is often advisable to install auxiliary individual strainers at the burners. The frequency of cleaning the strainers depends on the oil used, a heavy oil sometimes requiring inspection and cleaning as often as once a day.

Oil Pumps.—In industrial plants of medium or large size, steam-driven, duplex, reciprocating pumps are most commonly used for delivering the oil from the storage tank to the burners (on account of their low first cost and reliability), though motor-driven triplex pumps or steam-jacketed rotary pumps are also used. For relatively small installations electric-motor-driven rotary and gear pumps are usually used with complete satisfaction. Pumps should be equipped with brass valves, and the packing (usually metallic or braided fiber) should be such that it will not be affected by the oil. A spare pump must always be provided so that the burners will not have to be shut down for pump repairs. A pump governor or regulator is necessary to insure constant pressure on the feed line to the burners, and with reciprocating pumps an air chamber should be provided to

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* A heavy ball cock in the suction line may sometimes suffice. The ball is of sufficient weight so the suction due to any siphoning effect cannot open the cock but the suction of the pump, having a greater pull, lifts the ball.

eliminate pulsations. Pumps must be located near the storage tanks, since the allowable suction head is small (usually not over 16 ft. of oil).

Oil Heaters.*—As pointed out in Chap. VI, fuel oil usually is heated to lower its viscosity and to permit efficient atomization. In general, where there is any appreciable fire risk, the oil is not heated above its flash point. On the other hand, when the fire risk is small, the oil may be heated above the flash point with advantage. It should be noted that the capacity of a mechanical burner will be lowered by heating to too high a temperature, due to the expansion of the oil.

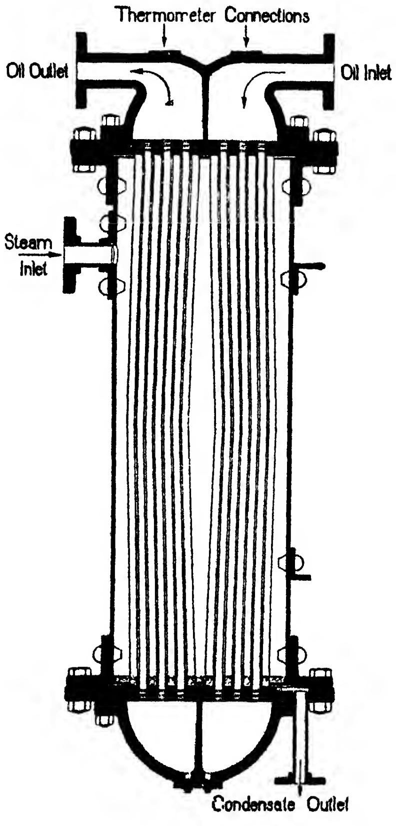

Fig. 181 —Schutte-Koerting oil heater

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* For calculations of heater size required, coefficients of heat transfer, etc., see Appendix II.

Two successful oil heaters are shown in Figs. 181 and 182. In the Schutte-Koerting oil heater, shown in Fig. 181, live steam passes across the tubes and oil passes through the tubes. Baffle plates in the upper and lower headers direct the oil flow and give a long path of oil travel. In the heater[1] shown in Fig. 182 the steam passes through the inner of two concentric pipes, the oil being in the annular space between the two pipes. In these, as well as heaters consisting of a series of tubes or coils, exhaust steam (frequently from the oil pump) is often used. The main resistance to the flow of heat from condensing steam, through metal and into oil, is the resistance between the metal and the oil. Therefore, the oil should be heated in thin films flowing at relatively high velocities in order to get good heat transfer (see Appendix II).

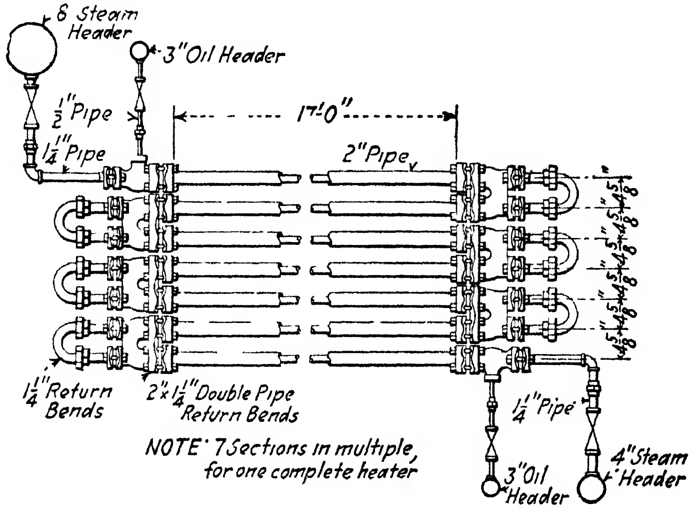

FIG. 182.—Double-pipe oil heater.

Figure 183 shows the variation in the coefficient of heat transfer with changing oil velocity for the heater shown in Fig. 182. The figure also shows how the temperature of the oil in question varies with the output from the heater and it will be noted that a considerable increase in oil through-put causes only a small drop in exit oil temperature, since the increased velocity of the oil increases the heat-transfer coefficient (although not directly proportional to velocity). The same figure shows the coefficient of heat transfer when using exhaust steam as compared with live steam at 100 lb. per sq. in. pressure. The increase in the coefficient when using live steam is due to the oil film next to the metal being at a higher average temperature (and therefore at a lower viscosity) than in the case of exhaust steam. Figure 184 gives the coefficient of heat transfer for the Schutte-Koerting heater shown in Fig. 181.

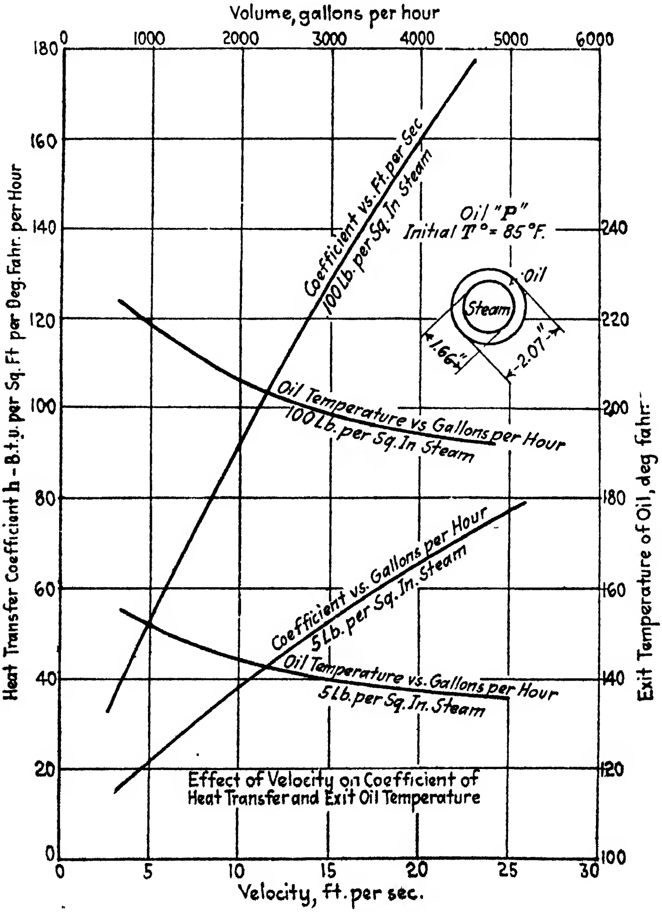

FIG. 183.—Calculated heat-transfer coefficients and exit-oil temperatures for the double-pipe heater shown in Fig. 182.

In order to regulate the temperature of the oil it is well to have a by-pass around the heater so that cool oil may be bled in when the temperature is too high. It is also desirable at times to have both live and exhaust steam connections—live for use at the start, and exhaust when the oil in the tank has been warmed due to the return of hot, excess oil from the burners.

The oil in the tank should not be allowed to get too hot or else the pumps will suck vapor and lose their priming. Most of the heating should be accomplished in the heaters placed between the pumps and the furnace.

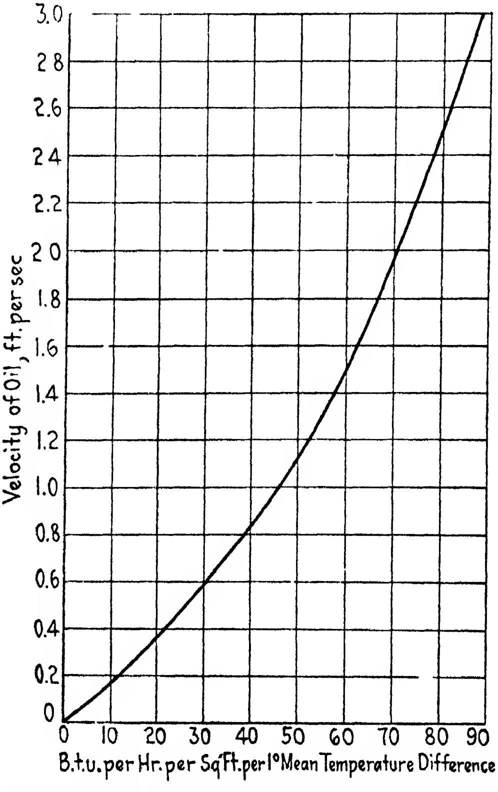

FIG. 184.—Heat transfer rate diagram for Schutte-Koerting oil heater.

In installations that are operated intermittently it is often necessary to put one or two hot-wire electrical-heating elements into the oil lines in order to warm the oil when starting. When these are necessary one such element should be placed as near the oil burner as possible to prevent subsequent cooling of the oil prior to atomization. In place of such heaters a supplementary supply of light, fluid oil may be provided for use until the steam pressure is available for oil heating.

Piping and Valves.—Oil piping must, of course, be absolutely tight. There should be no dead ends in which the oil can congeal and make pumping difficult. As previously mentioned, a return line is usually provided to take excess oil from each burner to the storage tank. Ordinary steel pipe is used in most systems, though extra-heavy piping is specified for the high pressures used with mechanical burners. The small sizes are screwed, litharge and glycerine ordinarily being used as a joint compound. Flanged pipe is used for the large sizes, cardboard or oil paper soaked in shellac serving as the gasket material. In order to reduce the pressure drop through the system, the pipes are usually large enough to keep the oil velocity[2] below 2 ft. per sec.

On the return line to the storage tank a pressure relief valve set for a pressure at least equal to or above the pressure on the burner line must be used to keep up the pressure at the burners and yet permit circulation when desired. Gate valves (on account of reduced friction through them) are well adapted for use on the suction side of the pump. On the pressure side globe valves of the regrinding type are commonly used. In mechanical atomizing systems, or wherever high oil pressures are encountered, it is best to use extra-heavy valves equipped with bonnets which screw over the valve body. A foot valve (ball check) should be placed in the end of the suction line in the oil tank in order to keep the pump primed at all times. A pump governor should be connected so as to regulate oil pressure by the amount of steam sent to the pumps. If electric-driven rotary or gear pumps are used, a by-pass should be installed connecting the pressure side of the piping with the suction side and an adjustable-pressure release valve inserted in the by-pass. This permits the use of a positive displacement pump when the consumption of oil fluctuates.

Meters and Thermometers.—Owing to the ease with which the flow of fuel oil can be determined and the value of data on fuel consumption not only of the whole plant but also of the individual furnaces, meters and thermometers* should be installed at various points.

Oil Burners.—The primary purposes of an oil burner are to atomize the oil thoroughly and to prepare it for intimate mixing

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* Since oil expands with temperature (*i.e.*, the density decreases) thermometers are necessary to determine the true weight of oil from a meter reading.

with air. The condition of the plant, the furnace design, the cost of the installation, and the type of oil used should govern the choice of burner. Burners should be simple in design, difficult to clog, easy to clean, and should require but little expenditure for maintenance. The efficiency of a burner may be considered from either of two viewpoints: first, the amount of steam or power required to atomize a given quantity of oil, and, second, the degree of subdivision produced in the oil. Increased subdivision or atomization usually permits the use of less excess air which, in addition, usually results in lower stack temperatures.

Oil burners may be divided into four main groups: (1) the spray type, in which the oil is divided into a fine spray by the action of an atomizing agent, steam or air; (2) the rotary type, in which the oil drops into the inner section of a revolving hollow cone or plate and is atomized by the action of centrifugal force; (3) the mechanical type, in which the oil under high pressure is atomized by being forced through small slots; and (4) the vapor type, in which the oil is vaporized by heat.

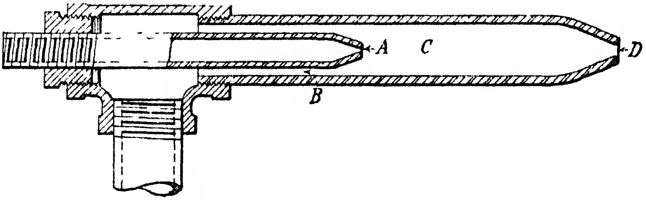

FIG. 185.—Inside-mixing type of oil burner. (*Courtesy of Denver Fire Clay Company.*)

1. *Spray-type Burners.*—The spray-type burners, in which the oil is sheared by the action of an atomizing agent, may be further subdivided into *inside-mixing* and *outside-mixing* types. Either of these types may use high or low-pressure air or dry steam. Also in either case the shape of the spray may be either flat or conical, depending on the design of the burner. Conical sprays are to be preferred, as they permit, to a greater extent, the intermixing of the oil and the air required for combustion. In a given burner, increasing the amount of steam or air used for atomization improves atomization, shortens the flame, and decreases the tendency to smoke.

A burner of the *inside-mixing* type is shown[3] in Fig. 185. The air or steam for atomization enters through the nozzle A and mixes with the oil in the chamber C. The resulting mixture passes out of the orifice *D* in the form of a fine spray, due to the expansion of the steam or air. This type of burner gives best results with the atomization agent at pressures of 40 lb. per sq. in. or above. Although simple to construct and operate, this type is suitable for thin (or high Baumé), clean oils only, due to a tendency for the orifice to clog. The oil supply is regulated by needle-valve adjustment on the oil-feed line. With burners of the steam-atomization, inside-mixing type, the oil does not have to be heated to such an extent as with either the outside-mixing type or the mechanical burner, since a large amount of the heat in the steam for atomization is available for heating the oil

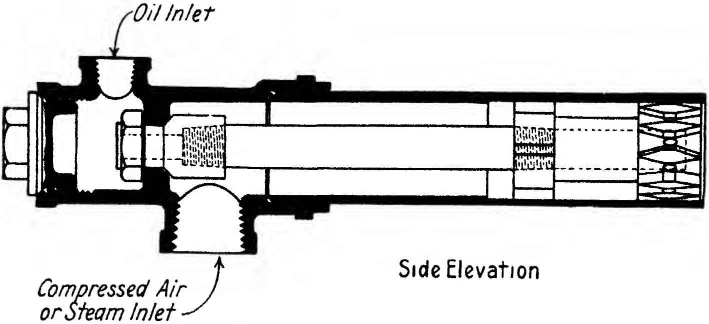

FIG. 186.—Venturi type of oil burner. (*Hauck Manufacturing Company.*)

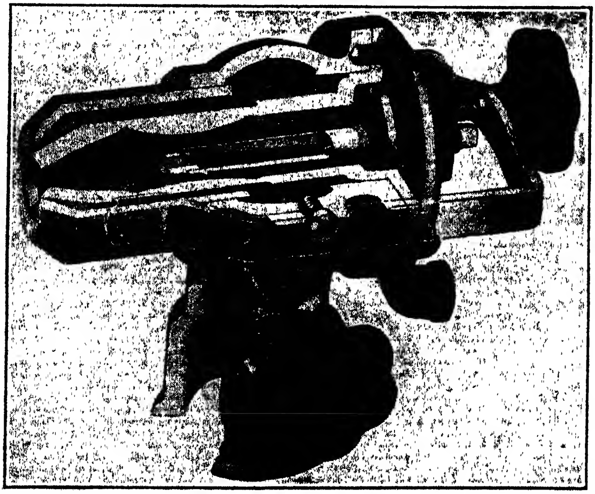

FIG. 187.—Inside-mixing type of oil burner using low-pressure air. (*Hauck Manufacturing Company.*)

before the oil-steam mixture leaves the burner. With this type of burner the ratio of pounds of steam for atomization per lb. of oil is important if good atomization is to result. The temperature of the oil is of lesser importance. Several other designs of inside-mixing burners are shown in Figs. 186, 187, 188, and 189.

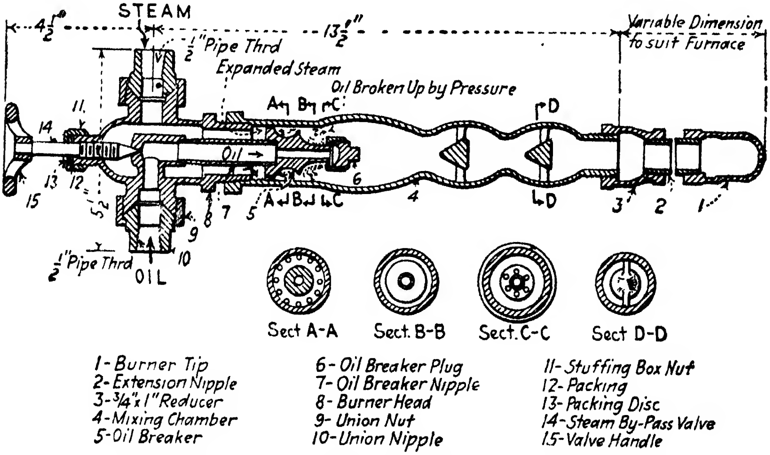

FIG. 188.—Quinn inside-mixing type of oil burner. (*Combustion Engineering Corporation.*)

The Hauck burner, shown in Fig. 186, consists essentially of a casing, within which is a central bore, closed at its outer end and provided with a series of slots which are constricted at the center (like venturi tubes). The constricted parts of these slots are connected with the inner bore of the burner. The oil

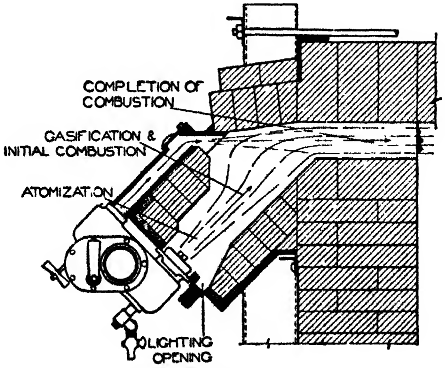

FIG. 189.—Application of an inside-mixing air-atomized burner. (*Mahr Manufacturing Company*).

from the inner bore (under a pressure of 20 to 100 lb. per sq. in.) passes into the slots, meets the atomizing air or steam (under 20 to 100 lb. per sq. in. pressure), and atomization takes place.

The Hauck burner, shown in Fig. 187, uses relatively low-pressure oil and air. Oil and air are partially mixed in an inside chamber, and additional air is added just as the partly atomized mixture passes into the combustion space of the furnace.

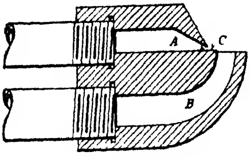

FIG. 190.—Outside-mixing type of oil burner.

In the Quinn burner, shown in Fig. 188, the oil under pressure passes through the passageway 7, reverses its direction in being forced through the small holes shown in section *C-C*. This action, coupled with the effect of expanding steam coming through the openings shown in section *A-A*, atomizes the oil. The atomized fuel then passes to the burner tip (1).

A unique adaptation of an inside-mixing air-atomized burner is shown[4] in Fig. 189. Primary air and oil are mixed and atomized by passing over two knife edges in the path of the air-oil stream. The atomized mixture then passes into a refractory-lined chamber on the furnace front, where initial hydroxylation (see Chap. IX) takes place and all the oil is gasified. Secondary air is preheated by passing around the refractory chamber and meets the hydroxylated and partly burned oil before it passes into the furnace proper. Combustion is practically complete before the gases pass into the main furnace chamber.

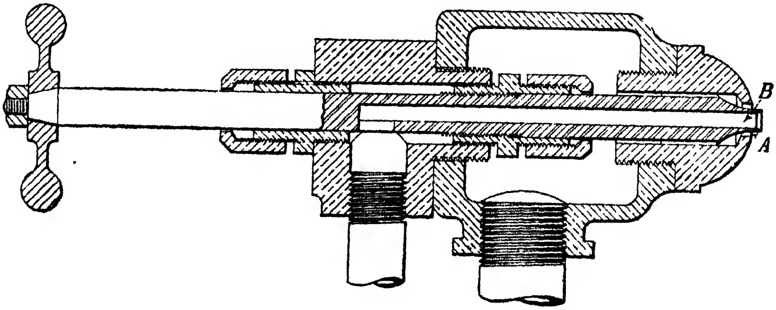

FIG. 191.—Outside-mixing type of oil burner. (*Courtesy of Denver Fire Clay Company.*)

In the *outside-mixing* type of burner, shown[3] in Fig. 190, oil flowing out through the wide slot *B* is sheared by the action of the atomizing agent from the slot *A*. These burners operate best with steam pressures of 25 lb. per sq. in. and up, but with low oil pressures (2 to 5 lb. per sq. in.). This type is well adapted for use with heavy-residue oils as well as with light distillates. Two others types of outside-mixing burners are shown in Figs. 191 and 192.

In the burner shown[3] in Fig. 191 the atomizing agent passing through the circular orifice *A* meets the oil coming through the central oil passage *B* and gives a conical stream of atomized oil. In the Enco burner, shown in Fig. 192, the oil passes through a duct *B* and is atomized by steam coming through the venturi nozzle *C* and is further atomized by jets of steam coming through at *D*. The atomized mixture passes through a mixing chamber *F* and out through the spray nozzles *G*.

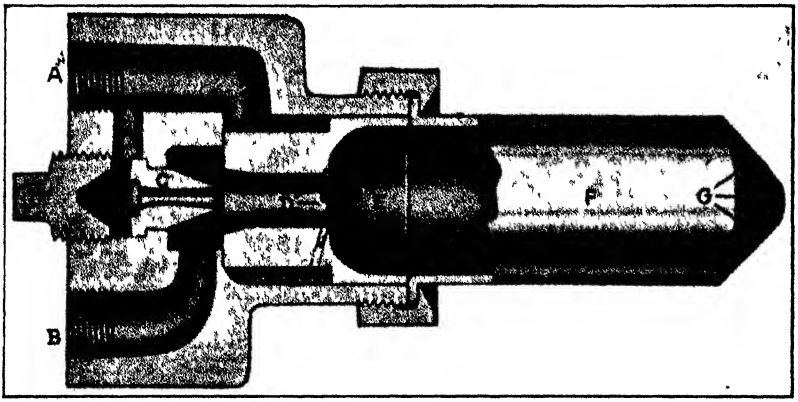

FIG. 192.—Inside-mixing type of oil burner. (*The Engineer Company.*)

The burners shown in Figs. 191 and 192, unlike many steam-atomized burners, give a conical flame similar in shape to the flame from certain mechanical burners. In this way it is possible to control closely and mix thoroughly the air admitted for combustion by means of proper adjustment of a circular air register surrounding the burner.

Except in furnaces of very limited height, the conical spray is better than the flat spray. In the first place, more effective use may be made of the full volume of the furnace. Secondly, the conical spray offers greater possibilities for mixing the atomized oil with the air for combustion coming in through air registers. It is advisable to use air registers, since they not only permit better control of the amount of air by means of vanes, but they also improve mixing, since the air stream may be made to take the form of a hollow cone, which, in turn, intersects the hollow cone of oil in front of the burner. Mechanical burners, to be described later, usually have been designed with conical sprays and many of the advantages of this type of burner thus are due to the better possible mixing of air and oil. Conical steam-atomized burners take one-third to one-tenth the draft of a corresponding mechanical burner.[5]

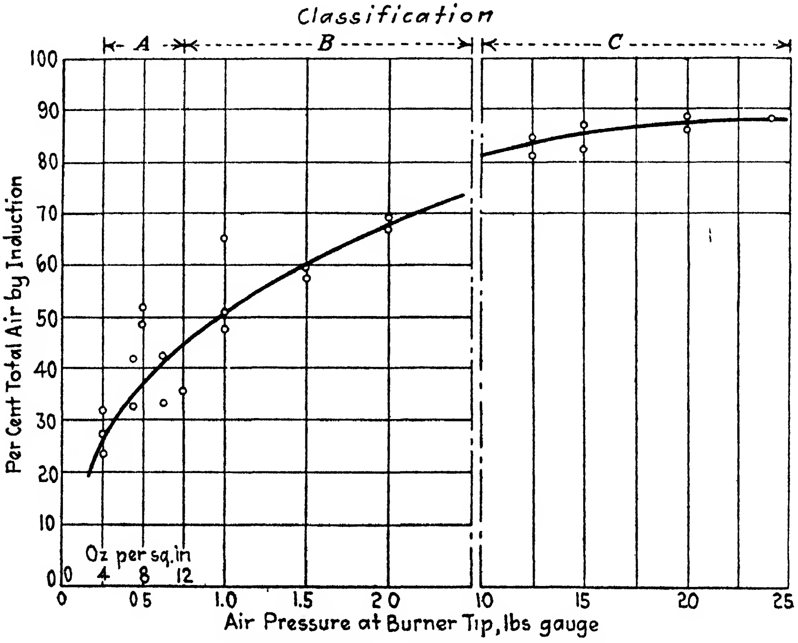

FIG. 193.—Blower performance with different types of air atomization.

(*a*) Air Atomization.—Air-atomizing burners are frequently used where a short, intense flame is required or where ease of regulation is essential. In steam-atomization systems, on account of the high steam pressures required, a small change in the setting of a regulating valve makes a large change in the amount of atomizing agent admitted. With air, on the other hand, particularly with low pressures, a small change in a valve adjustment brings about a relatively small change in the amount of air admitted and close adjustment of the burner is, therefore, possible.

Air-atomizing burners may be divided into low-pressure burners, in which air from a fan or positive blower is supplied at pressures from 8 oz. per sq. in. up to 2 lb. per sq. in., and high-pressure burners using air under pressures from 50 to 80 lb. per sq. in. The low-pressure burners have the advantage that a large proportion of the air for combustion is supplied by the atomization air, as shown[6] by Table XCIV and Fig. 193.

TABLE XCIV

BURNER PERFORMANCE WITH DIFFERENT TYPES OF AIR ATOMIZATION

Type	Velocity of atomizing air, feet per second	Per cent of air for combustion used for atomization
Low-pressure (turboblower) air atomization, 10 oz. per sq. in	300	60
Low-pressure (positive-pressure pump) air atomization, 2 lb. per sq. in	500	50
High-pressure air atomization, 50 lb. per sq. in......................	2500	12

From the third column of the above table may be computed the cubic feet of air per min. required for atomizing a given quantity of oil with the types of burners shown.

The type of blower used for different pressures, and some of the operating characteristics of these machines, are given[6] in Table XCV.

TABLE XCV

OPERATING CHARACTERISTICS OF BLOWERS

Delivery pressure	Type of machine	Operation
4–12 oz..............	Centrifugal fan	Quiet
¾– 1 lb..............	Centrifugal fan	Humming
¾– 2 lb..............	Rotary displacement	Noisy
1– 2 lb..............	Turbine pump	Quiet
15–80 lb..............	Compressor	Quiet for low pressure Noisy for high pressure

The big field of usefulness for burners using low-pressure atomizing air is in small industrial furnaces where the oil consumption is low, where a short flame is required and close control is necessary.

High-pressure air atomizers (40 to 50 lb. per sq. in.) are very efficient, though the compressors are expensive and have high operating costs. High-pressure air is usually used for atomization of oil only when air is already available, due to other requirements of the plant.

In general, air atomization is installed where it is difficult to obtain dry steam, although about the same power is required whether steam is used directly for atomization or for driving compressors or blowers. With air there is no extra loss of sensible heat up the stack, as is the case with steam, but this is balanced by the greater initial expense of the air atomization installations. On the basis of merit it seems as if air atomization with automatic or semi-automatic control should be used more widely than at present.

(*b*) Steam Atomization.—Steam-atomizing burners are generally used in medium-sized installations where steam is available. The steam must be dry, since slugs of condensed water from wet steam interfere with the continuous operation of the burner. The steam consumption varies from 0.8 to 5 per cent of the steam produced (from 0.16 to 1.0 lb. of steam per lb. of oil fired). Steam atomizers have the advantages of simplicity, low first cost, and ease of operation.

The disadvantages of burners using steam atomization are: (1) they cannot be used with small flames (which are liable to be put out by the steam); (2) with wet steam, slugs of water cause a sputtering flame; (3) there is an extra loss of sensible heat up the stack, carried by the uncondensed steam used in atomizing; (4) too much steam is often used, since the extra steam reduces the tendency of the flame to smoke; (5) as previously mentioned under Air Atomization, close regulation with steam-atomizing burners is difficult on account of the high steam pressures involved. It should be borne in mind also that the generation of high-pressure steam requires, under most state laws, the presence, at all times, of a first-class fireman. Therefore, it does not pay to increase the steam pressure of a low-pressure steam installation merely for the sake of using steam atomization on the oil burners.

2. *Rotary Burners.*—In rotary-type burners the oil drops into the inner section of a revolving hollow cone or at the center of a revolving saucer-shaped disc. This cone sometimes forms one end of a hollow shaft on which a fan may be mounted. Oil passes from the pump through the hollow shaft to the revolving saucer or cup. When the shaft rotates rapidly the oil flies out radially in small droplets and is further atomized by a blast of air from the fan. A direct-connected fan is commonly used, and the whole apparatus, including pump, fan, motor and burner, is often mounted as a single unit, as shown in Fig. 194. These burners are adapted for use under low-pressure boilers (where the steam pressure is too low for steam atomization) or where cleanliness and compactness are more important than economy.

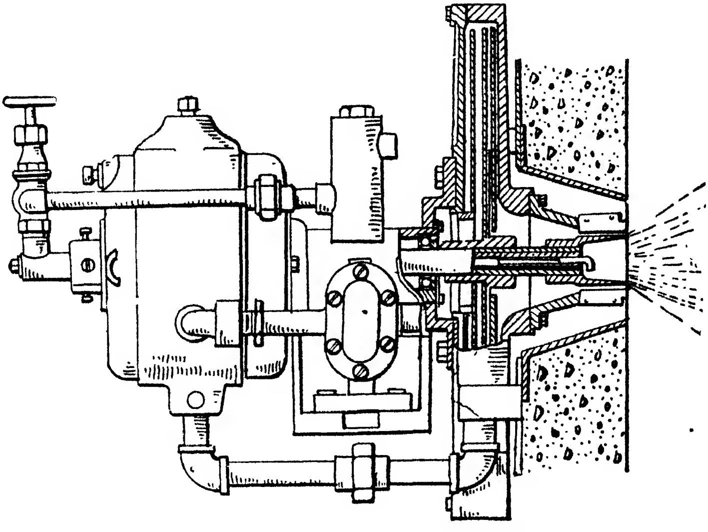

Fig. 194.—Ray rotary-type oil burner.

The efficiency of the rotary-type burner frequently is low, due to the large excess of air required to keep the flame free from smoke (200 to 300 per cent), but the ease of operation and their ability to handle small loads at a low first cost sometimes make their installation advisable. However, some rotary burners are more efficient than just indicated; in fact, their efficiency may be made equal to that of any other type. Their installation often is advisable where the required steam pressure is so low that no trained fireman is needed or where the boiler is shut down nights and Sundays.

3. *Mechanical Spray Burners.*—In the mechanical spray type of burner the atomization is due to the motion imparted to the oil in passing tangentially through small orifices at high pressures and temperatures. The atomization is the result of a series of actions: (*a*) atomization by the friction as the oil comes out the small orifice, (*b*) the "flashing" into vapor of a part of the heated oil as the pressure is suddenly reduced, and (*c*) atomization due to a centrifugal motion imparted to the oil, usually obtained by passing it through tangential slots inside the burner. Because no air is used for atomization and good mixing of oil spray and air is essential, the air necessary for combustion is blown through an air pipe or register surrounding the burner. These registers (see Fig. 195) are used both to control the amount of air admitted and to give the oil a whirling motion. The oil pressures used are from 100 to 300 lb. per sq. in. and the oil is heated to from 140 to 280°F. In the Schutte-Koerting mechanical atomizing burner, shown in Fig. 196, the oil under pressure passes through an oil strainer into a central tube. At the burner tip the actions described in the preceding paragraphs give a hollow cone-shaped spray of atomized oil. In the Peabody-Fisher burner, oil under pressure tangentially enters a central chamber in which a rapid rotary motion is set up. The oil passes into the furnace through the orifice, thereby being atomized, or a portion of the oil may be by-passed back to the oil pump. The object of this

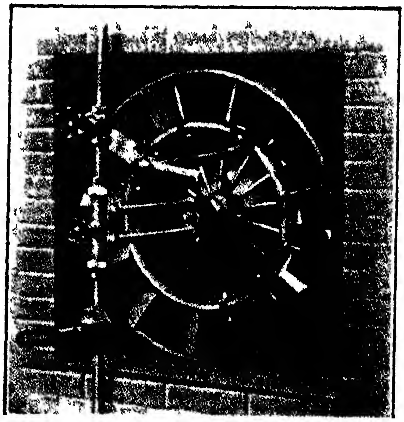

FIG. 195.—Air register for controlling admission of air to an oil-fired furnace. (*The Engineer Company.*)

action is to give satisfactory atomization over a wider range of oil consumption.

A mechanical burner with a given size orifice opening usually has a rather limited capacity range. With a fixed orifice the only method of changing the rate of oil flow is the oil pressure, and Fig. 197 shows how little this affects the output of the type of burner shown in Fig. 196. A material drop in oil pressure decreases the output only to a small extent; for example, with a burner having an orifice 1.75 mm. in diameter, decreasing the oil pressure from 200 to 100 lb. per sq. in. only decreases the burner capacity from 385 to 300 lb. oil per hr. Therefore, the usual method of regulating the flow of oil is either to turn on or to turn off individual burners or to change to an orifice of different diameter.

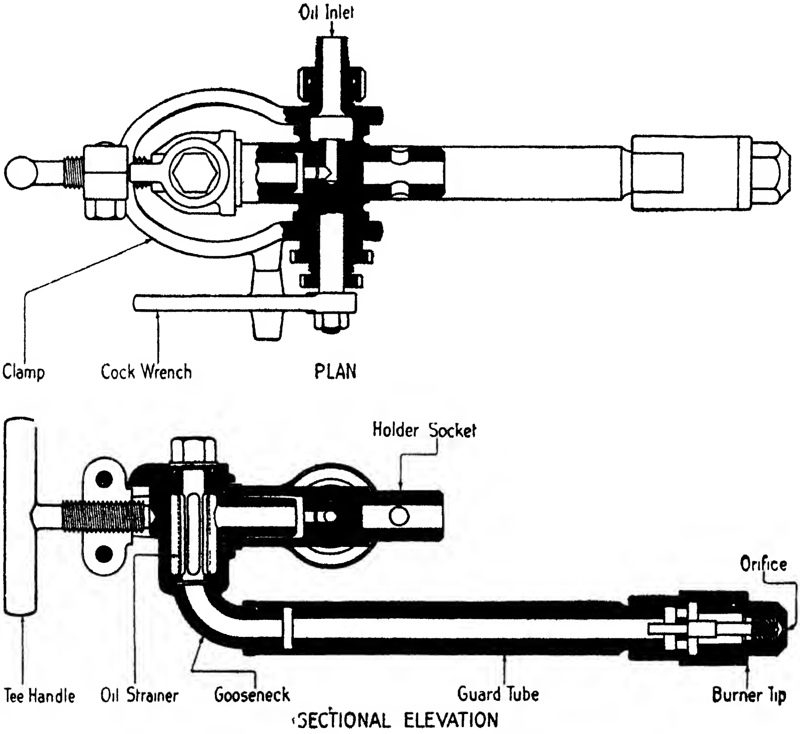

Fig. 196.—Mechanical atomization burner.

Mechanical atomization was formerly used only in marine work where the distilled water used for making steam had to be condensed and recovered. This method is now being used quite widely in land boiler furnaces and high efficiencies are being obtained.

The advantages of the mechanical spray type of burner are: (*a*) No high pressure air or steam lines are required for atomization, (*b*) high efficiencies at high capacities are possible (due to a large extent to proper admission of air through air registers), (*c*) the flame is shorter than with steam-atomized burners and therefore the combustion space may be smaller, (*d*) mechanical atomization requires less work than atomization by either steam or air, (*e*) there is less maintenance cost for the burners. Against these must be balanced the following disadvantages: (*a*) the installation of mechanical atomizing equipment is more expensive than for other types, due to the extra strength of piping, pumps, etc. required to withstand the high pressures, (*b*) they are not as flexible as other types, and (*c*) clogging of the burner tip, due to carbonization of the oil or to the small size of the orifice, is liable to make frequent cleaning necessary. To help overcome this last difficulty extra-heavy strainers with small openings should be installed. The burner tips may be cleaned by blowing steam through them.

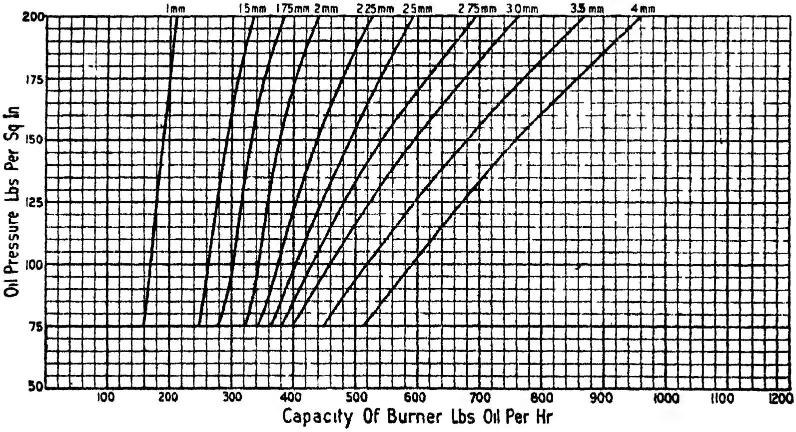

FIG 197.—Pressure-capacity curves of Schutte-Koerting mechanical atomizing burner with different-size orifices.

4. *Vapor-type Burners.*—In the vapor type of oil burners the oil is *vaporized* (*not* atomized) in a stream of preheated air. These burners have been used with forging, annealing, and tempering furnaces, or, in general, where the stack temperatures are high and air recuperation (in order to preheat the air sufficiently for

vaporization of the oil) is possible. This type of burner is not suited for boiler or large furnace work, and, though used somewhat in Europe and in small domestic installations, is relatively unimportant and will not be further discussed here.

GENERAL CONSIDERATIONS IN BURNING OIL

A medium-sized oil-fired furnace should average 5 per cent higher in efficiency than a corresponding coal-fired furnace. Efficiencies of from 75 to 80 per cent are common. These high efficiencies are easy to obtain, but it is also easy to run an oil-fired furnace inefficiently, due to the ease of admitting too much air for combustion and too much steam for atomization in case of steam-atomizing burners. For this reason it is best to install recording instruments, such as oil meters, steam flow-meters, carbon dioxide recorders, and draft gages in order to keep a continuous check on the plant operation. For high efficiencies there should be a slight haze from the stack and the flame should be orange, *not* white. The use of air preheaters where economically possible produces excellent technical results, in that the hot air helps to vaporize the finely atomized oil, and thus produces a shorter flame or permits the use of lower excess air for combustion.

Very heavy oil residues are sometimes difficult to burn on account of their high flash points (300 to 400°F. as compared with 150°F. for the lightest fuel oils). With such fuel, preheated air or a large amount of heat-radiating brickwork is desirable.

Costs of Fuel-oil Equipment and Atomization.—From the results of a questionnaire sent January, 1924, to 24 manufacturers of fuel-oil-burning equipment, the following overall installation costs were calculated:

Horsepower	Steam and air atomization	Mechanical atomization
150	$ 3,000	$ 5,000
300	4,000	7,000
900	9,000	15,000
1,950	14,000	23,000
4,500	30,000	45,000

These figures include cost of burners, strainers, duplex pumps, heaters, underground oil-storage tank, all necessary piping, valves, fittings, and meters, together with the cost of installation under *average* conditions.

In determining the cost of atomization, as given below, a total of 22 per cent per year on the actual initial investment is charged as follows:

	Per Cent
Interest	6
Depreciation, overall	10
Taxes and insurance	3
Maintenance and repairs	3
Total	22

The costs are figured on the basis of 24 hr. per day, 300 days per year. A 150-hp. plant uses 1000 gal. of oil per day when, (1) weight of oil equals 7.85 lb. per gal., (2) oil contains 19,000 B.t.u. per lb., (3) boiler efficiency equals 80 per cent at full load.

TABLE XCVI

Cost of Air for Atomization

	Machine type		
	Fan	Displacement turbine	Piston compressor
Delivery pressure	8 oz.	2 lb.	80 lb.
Anticipated induction of atmospheric air, per cent	36	50	88
Cubic feet of atomization air necessary to burn 1 gal. fuel oil per min.	960†	480	180
Energy necessary to compress above volume of air, kilowatt-hours	3.55	7.6	17.4
Power cost—air for 1000 gal. fuel oil, at 2 cts. per kw.-hr.	$1.16	$2.53	$5.80

† Required 1500 cu. ft. free air per gal. of oil for complete combustion.

For figuring the cost of steam for atomization, 2.5 per cent of the total steam from the boiler was considered as used in atomization,* and is charged at a rate of 80 cts. per 1000 lb. The cost of air for atomization is shown[6] in Table XCVI.

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* While steam-atomized burners often use only 1 to 2 per cent of the total steam, especially in tests, it is felt that 2.5 per cent more nearly represents average plant practice.

No charge for operating labor is included in the cost of atomizing oil by steam, by air, or mechanically.

Table XCVII gives the cost items as calculated on the basis of the foregoing data. From this table, the cost of atomizing 1000 gal. of oil is plotted against equipment capacity in Fig. 198.

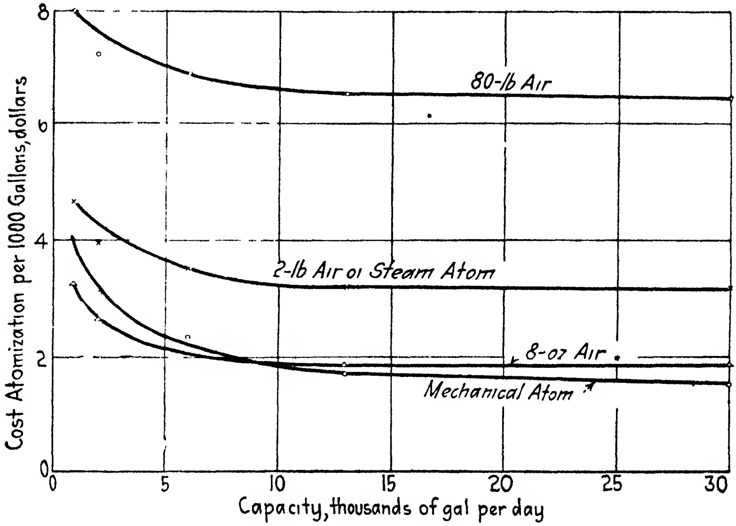

FIG. 198.—Cost of atomizing fuel oil by different methods.

The following conclusions may be drawn from a study of the curves:

1. Capacity has but relatively little effect on the cost of atomizing oil, since large installations are made up of relatively small units.

2. High-pressure air is the most expensive means of atomizing fuel oil.

3. In general, 8-oz. air pressure and mechanical atomization are the cheapest* methods of atomization and the saving by these methods increases as the capacity increases.

4. Air at 2-lb. pressure is cheaper than steam in case steam costs more than $1 per 1000 lb. On the other hand, if steam costs less than this steam atomization is the cheaper.

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* The data in Table XCVII give no consideration to furnace efficiency and deal only with the cost of subdividing or atomizing oil.

TABLE XCVII

COST OF ATOMIZING FUEL OIL

	Mechanical atomization	Steam atomization	Air atomization		
			8-oz. pressure	2-lb. pressure	80-lb. pressure
150 hp. = 1000 gal. per day:					
Air cost	.....	.....	$ 1.16	$ 2.53	$ 5.80
Steam or power cost	$ 0.48	$ 2.48			
Fixed charges	3.67	2.20	2.20	2.20	2.20
Total	$ 4.15	$ 4.68	$ 3.36	$ 4.73	$ 8.00
300 hp. = 2000 gal. per day:					
Air cost	.....	.....	2.32	5.06	11.60
Steam or power cost	0.96	4.96			
Fixed charges	5.13	2.93	2.93	2.93	2.93
Total	6.09	7.89	5.25	7.99	14.53
Cost per 1000 gal. oil	$ 3.05	$ 3.95	$ 2.63	$ 4.00	$ 7.27
900 hp. = 6000 gal. per day:					
Air cost	.....	....	6.96	15.18	34.80
Steam or power cost	2.88	14.88			
Fixed charges	11.10	6.60	6.60	6.60	6.60
Total	13.98	21.48	13.56	21.78	41.40
Cost per 1000 gal. oil	$ 2.33	$ 3.58	$ 2.26	$ 3.63	$ 6.90
1950 hp. = 13,000 gal. per day:					
Air cost	.....	.....	15.08	32.89	75.40
Steam or power cost	6.24	32.24			
Fixed charges	16.87	10.26	10.26	10.26	10.26
Total	23.11	42.50	25.34	43.15	85.66
Cost per 1000 gal. oil	$ 1.78	$ 3.27	$ 1.95	$ 3.32	$ 6.59
4500 hp. = 30,000 gal. per day:					
Air cost	.....	.....	34.80	75.80	174.00
Steam or power cost	14.40	74.40			
Fixed charges	33.00	22.00	22.00	22.00	22.00
Total	47.40	96.40	56.80	97.80	196.00
Cost per 1000 gal. oil	$ 1.58	$ 3.21	$ 1.89	$ 3.26	$ 6.53

Advantages of Oil Fuel.—Oil fuel possesses the following advantages: (1) It weighs 30 per cent less and occupies 50 per cent less space than coal containing the same number of B.t.u.

(2) There is no deterioration during storage. (3) There is no trouble from spontaneous combustion. (4) The oil storage may be located at a distance from the furnaces. (5) The fuel is immediately available and may be stored or removed with practically no labor. (6) High combustion rates per cu. ft. of combustion space are possible. (7) There is great flexibility in oil-fired furnaces, so that peak and valley loads are readily and economically carried. (8) Only a small amount of labor is required for handling the oil at the furnace and for cleaning boiler tubes. (9) No labor is required for cleaning fires or removing ashes. (10) It is possible to run an oil-fired furnace efficiently and with practically no smoke. (11) There is no wear and tear on machinery due to ash and dust. (12) Since there is no fuel bed, the pressure drop through the furnace is small, thus requiring less draft than either a stoker-fired or hand-fired furnace. When the maximum existing draft of a coal-fired furnace is insufficient to maintain the required load, it is sometimes advisable to change to oil fuel, which requires less draft. (13) High efficiencies, due to low excess air required, are possible with oil-fired furnaces, as is shown by Table XCVIII.

TABLE XCVIII

EXCESS AIR IN EFFICIENTLY OPERATED FURNACES OF VARIOUS TYPES

FURNACE	PER CENT EXCESS AIR IN EFFICIENTLY OPERATED FURNACES
Oil-fired	15–25
Powdered-coal-fired	15–30
Chain-grate stoker-fired	35–70
Hand-fired	50–150

Against all these advantages must be balanced the initial cost of handling equipment and the price of fuel oil.

Oil-fired Furnaces.—The *fundamental requirements* for efficient oil-fired furnaces are:

1. *The burner must thoroughly atomize the oil.* This requirement is met by suitable burner design and by heating the oil.

2. *The atomized oil must be thoroughly mixed with the proper amount of air for combustion.* This is accomplished by admitting a controlled amount of air through checkers under the burner, or through air registers and openings around the burner. When checkers are used (mainly with a flat-spray type of burner), the openings for air admission should be distributed approxi-

mately in the shape of the flame as indicated in Fig. 199. There should be no checker openings in between the flames of two oil burners set in the same furnace and the openings should be larger near the burner than at the tip of the flame. Common practice is to allow about 2 sq. in. of air opening per developed boiler hp. (or approximately 1 sq. in. per lb. of oil fired per hr.). The burners should be installed from 8 to 10 in. above the checker work.

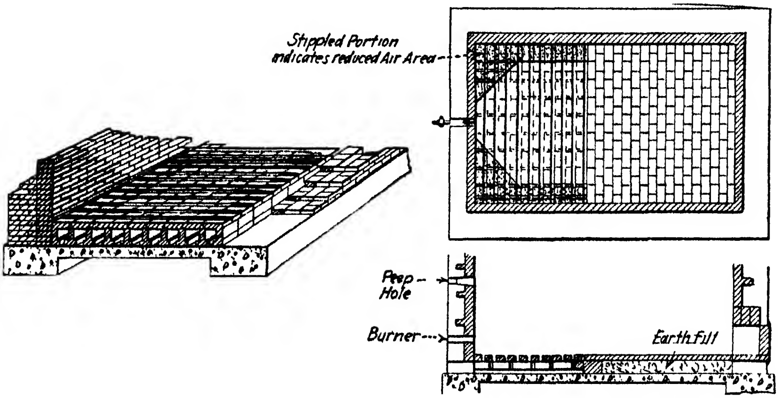

Fig. 199.—Brick checkerwork arrangement in an oil-fired furnace. (*Combustion Engineering Corporation.*)

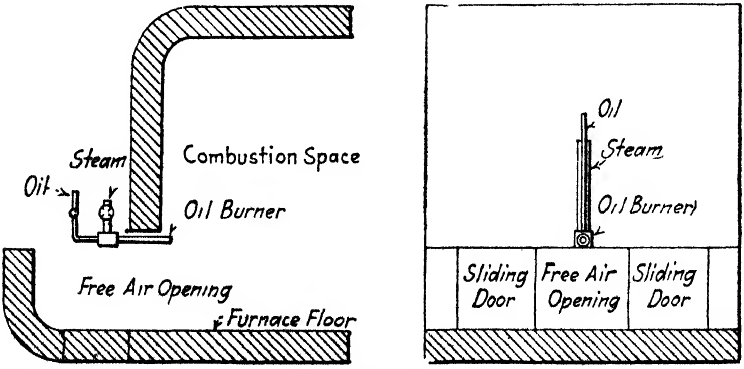

Fig. 200.—Method of air admission to an oil-fired furnace without the use of checkerbrick.

The use of checkers, as outlined above, is representative of general current practice. The best results with mechanical burners are obtained when blower air is supplied through registers which aid mixing by giving a whirling motion to the entering air.

Registers also represent the best means of air admission to furnaces fired with either steam- or air-atomizing burners. In general, with the flat-spray type of burners it is better, however, to discard the use of checkers and admit air through the front wall just under the burner and oil spray, one such method being shown in Fig. 200.

It is usually considered that mechanically atomizing oil burners are better on high overloads than steam atomizing. This belief is largely due to the fact that steam-atomizing burners were almost always of the flat-spray type with the air for combustion admitted underneath through checkerwork openings. Such openings not only give poor mixing of oil and air but the pressure drop is high, particularly at high loads. Consequently, large openings around the burner (for conical sprays) or under the burner (for flat sprays) are to be preferred to checkers. This construction is cheaper than checkerwork both to install and to maintain.

3. *A high furnace temperature is required to sustain combustion.* This requires a sufficient quantity of heat-radiating refractory bricks in the combustion chamber, good practice[7] being to allow from 0.9 to 1.2 sq. ft. of radiating surface per boiler hp. developed. The atomized oil is heated and partially vaporized by this radiation.

4. *Combustion must be complete before the gases are chilled.* This requirement is met by having ample combustion space. Common practice for steam generation[7] calls for about 1 cu. ft. of combustion space per developed boiler hp. From 1.75 to 2 lb. of oil per hr. can be efficiently fired per cu. ft. of combustion space. This corresponds to from 33,000 to 38,000 B.t.u. per cu. ft. per hr. Higher rates of firing (to maintain excessive ratings) with the usual boiler setting are apt to give low efficiencies, due to incomplete combustion, or trouble with brickwork may develop.

5. *Flame action must not be localized on either the heating surface or the furnace brickwork.* If this precaution is not taken, overheating and cutting will result and a deposit of soot will be formed. This blowtorch action is avoided by proper design of furnace and burner, to give a long, unobstructed path of flame travel. One burner[8] should be sufficient for a furnace setting 6 ft. in width; in general, two burners will be required for a setting 10 ft. wide and three burners for a 12-ft. setting. For ease of

ignition and to steady the flame it is well to have the edge of the flame near or very slightly touching a wall or floor.

Under low, short furnaces and particularly under boilers with tubes which incline downwards toward the rear, burners cannot be located at the front of the furnace because the path of flame travel would be too short and the flame would impinge directly on the tubes. Under these circumstances it is customary to use the so-called *"back-shot"* method, in which the burner itself is located at the bridge wall and the flame travel is toward the front of the furnace. The valves controlling the burner are located at the front of the furnace, as usual. This arrangement provides for a better distribution of heat above the boiler tubes and increases the effectiveness of the total combustion space.

The Cannon radiation furnace, shown[9] in Fig. 201, is an interesting and highly efficient type of oil-fired furnace. Combustion takes place in a long two-pass flue chamber constructed entirely of carborundum refractory tiles. The surfaces to be heated may surround the chamber to absorb the heat which is transmitted directly through the incandescent tiles and radiated from the outer surfaces; the furnace may also be arranged without the surrounding heat-absorbing surfaces in order to deliver hot flue gases for heating outside the furnace itself.

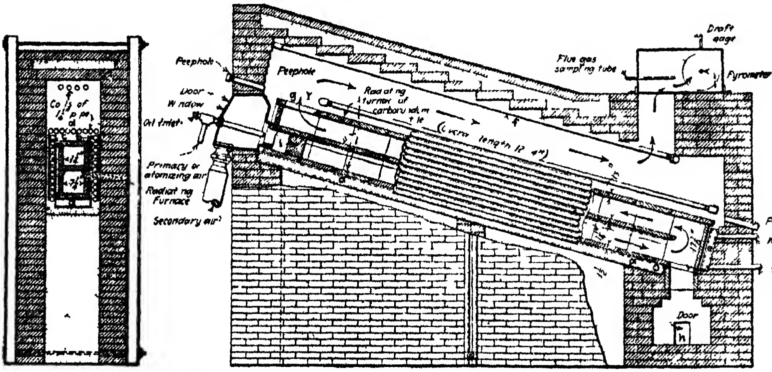

FIG 201 —The Cannon radiation furnace.

The oil burner is located at the left end of the furnace, as shown, secondary air being supplied in the annular space around the burner. The wedge-shaped carborundum block is to facilitate mixing and ignition. The path of flame travel is long and combustion is exceptionally rapid on account of the high temperature level at which combustion takes place. In this furnace it is possible to obtain complete combustion with as little as 3 to 6 per cent excess air when the flame extends nearly to the outlet of the upper part of the combustion chamber.

Size of Combustion Space Required.—The high rate of combustion possible with fuel oil as compared with other fuels is shown in Table XCIX.

TABLE XCIX

AVERAGE RATE OF COMBUSTION IN DIFFERENT TYPES OF FURNACES

TYPE OF FURNACE	B.T.U. DEVELOPED PER CU. FT. COMBUSTION SPACE PER HR.
Hand-fired, coal	10,000
Stoker-fired, coal	30,000
Powdered-coal	13,000–46,000
Fuel-oil	35,000
Fuel-oil (in Cannon furnace)	250,000
Gas (surface-combustion, Bonecourt Boiler)	700,000

To obtain the size of combustion space it is necessary to know the number of B.t.u. that can be developed per hr. per cu. ft. of combustion space and the allowable gas velocities. In general, the velocity of the products of combustion in an oil-fired furnace should not be over 40 ft. per sec. except at overloads. A furnace designed to liberate about 35,000 B.t.u. per cu. ft. per hr. and with velocities below 40 ft. per sec. usually will operate satisfactorily.

The effect of excess air and load on stack losses is shown in Fig. 202. In this figure the stack temperature for two different loads is plotted against per cent CO_2 (or excess air). With large amounts of excess air (low CO_2) the stack temperature (for a given load) is *higher* than with low excess air, as explained on page 363. The increase in stack loss with increasing excess air is, therefore, more than in direct proportion to the increase in the amount of gas going out at the stack. That is (for the case shown in Fig. 202), increasing the excess air from 60 per cent (9.2 per cent CO_2) to 120 per cent (6.7 per cent CO_2) increased the stack temperature from 410 to 440°F. The amount of heat lost out the stack was, therefore, greater than if the losses were only in proportion to the extra quantity of excess air admitted. Moreover, increasing the load also gives an increase in stack temperature. This is due primarily to the greater velocity of the gaseous products of combustion through the boiler. Although the amount of oil burned (and therefore flue gas formed) is practically in direct proportion to the load, the rate of heat absorption by the boiler surfaces only increases with the 0.6 to 0.8 power of the gas velocity and the stack temperature therefore is increased.

Changing From Coal to Oil.—In changing a plant from coal to oil it is necessary to consider: (1) the comparative cost per B.t.u. of coal and oil, (2) the labor cost for operating both methods, (3) the cost of changing from coal to oil, (4) the fixed charges on each kind of equipment, (5) the cost of operating (*i.e.*, the cost of operating coal-firing equipment must be compared with the cost of pumping the oil,) and (6) the dependability of both kinds of fuels.

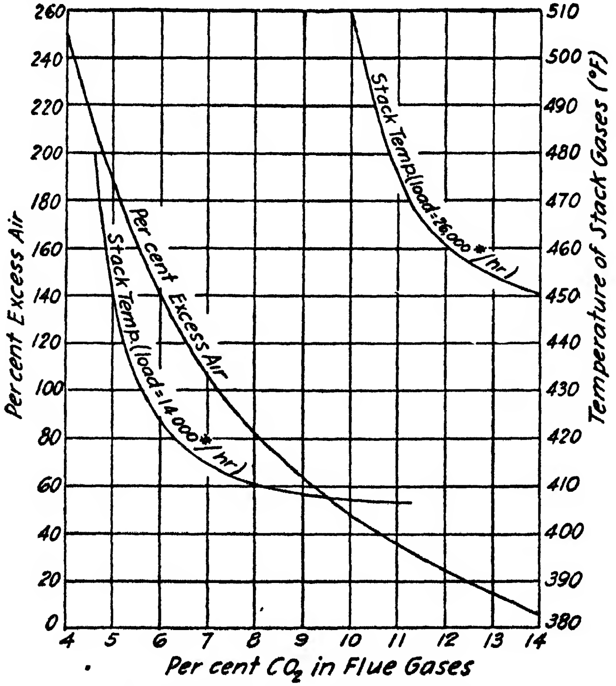

FIG. 202.—Effect of excess air and load on stack temperature at two loads, 14,000 and 26,000 lb. of water evaporated per hr.

It must be remembered that fuel oil is sold by the barrel or gallon, *not* by weight. Therefore more B.t.u. can be purchased per dollar when buying a heavy fuel oil. Furthermore, the cost of a light oil is usually more than a heavy one, so that fuel-oil-burning equipment usually should be capable of burning the heavy residues. In view of the rapid fluctuations in petroleum prices, an oil contract should, if possible, be drawn up to cover a considerable length of time.

Removal of coal-firing equipment is usually necessary, though with high furnace settings it is often possible to leave in stokers and protect them by a sand and brick covering. In hand-fired furnaces it is best to remove the grates, brick up the fire doors, and install the oil burners through the ash-pit doors.

The main cost item in changing from coal to oil is the cost of the oil-storage tanks. In several 50,000-gal. capacity storage systems the costs were 10 cts. per gal. for installation and the modifications in the corresponding boilers cost about $20 per boiler hp. To offset these costs it should be noted that a change from coal to oil firing usually results in an increased efficiency, and that increased capacities can be obtained with oil firing, *without* increase in the size of stack. In a number of new installations provisions are made for both coal and oil firing in order to take advantage of any price differential between the two fuels.

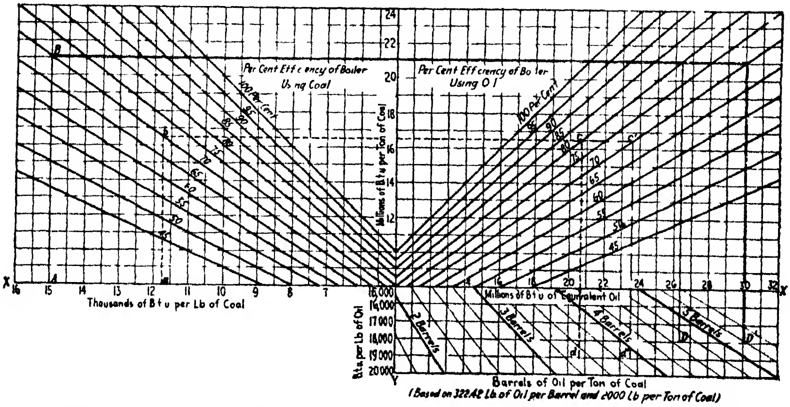

FIG. 203.—Chart for determining the amount of fuel oil required to replace a given amount of coal under different conditions.

Figure 203 (chart and directions for use furnished by A. D. Little, Inc.) may be used to advantage in determining the advisability of changing from coal to oil, or *vice versa*.

For example, in order to find the number of barrels of 18,000 B.t.u. fuel oil equivalent to 1 ton of 15,000 B.t.u. coal with 70 per cent boiler efficiency on coal, and 80 per cent when burning oil, begin at point *A* on the chart, at the lower left-hand corner, follow the heavy line to the 70 per cent efficiency curve, point *B*, to the 80 per cent efficiency line on oil, and then drop down to point *D* on the 18,000 B.t.u. line for oil. Point *D* will be seen to lie between the 4- and 5-bbl. lines, approximately 4.5 bbl.

If conditions are similar to those assumed above, except that the efficiency of the boilers on oil is 70 per cent instead of 80 per cent, follow the heavy line from *A* to *B* and thence to *D′* where the number of barrels of oil equivalent to 1 ton of coal will be found to be 5.22.

With 11,750 B.t.u. coal, 19,000 B.t.u. oil, 70 per cent efficiency of the boilers burning coal, and 80 per cent efficiency burning oil, the dash line can be followed from *a* to *b* to *c* to *d* and the equivalent barrels of oil for 1 ton of coal will be 3.4. In case efficiency of the boilers on oil is 70 per cent and all other conditions are as stated above, follow the dash line from *a* to *b* to *c′* to *d′* and the barrels of oil equivalent to 1 ton of coal will be 3.9. Similar procedure will determine the oil and coal equivalent for any B t.u. coal and oil and any boiler efficiency with either.

Having determined, by use of the coal-versus-oil chart, the equivalent barrels of oil per ton of coal for the desired conditions, figure the dollars advantage for either fuel, using for the cost of coal the cost delivered to the firing floor or coal bunker and in the case of oil the cost of oil delivered to the oil-storage tank.

Credit oil with savings in fire-room and ash-handling labor. Credit coal with repair labor and material, including furnace lining, and the investment charges required to make the oil installation, including oil-storage tanks, piping, pumps, burners, etc. A summary of fuel, labor, repair, and investment charges will determine the advantage of either fuel and the extent of the justification for change. Generally, it is safe to assume a 10 per cent increase, on the average, in boiler efficiency when oil is used as a fuel as against coal. This figure will be decreased in some plants operating stoker-equipped boilers supplied with the necessary instruments and superintendence to control combustion, and will favor oil to a greater extent in a hand-fired boiler plant or in a stoker-equipped plant lacking in instruments and proper superintendence to take advantage continually of the instrument information, or where wide fluctuations in load occur frequently.

Oil and Tar as Open-hearth Fuel.—Within the past few years the use of liquid fuels for open-hearth furnaces has been increasing to a considerable extent, but, except in those localities favorably located with respect to the oil fields, oil has not made so very much progress. Tar, however, used in connection with coke-oven gas, seems to be a favorite combination.

As fuels, oil and tar offer approximately the same advantages and difficulties. One of the difficulties with tar is that effective screening must be provided to remove impurities which may plug the burner. Inasmuch as tar is very difficult to keep in a suitable state, this screening or filtering must be done while hot and *en route* to the burner itself. A good type of oil burner should work almost equally well on a tar which has been properly filtered. With either tar or oil the burner is subject to conditions which differ considerably from those of the ordinary boiler installations. The burner tip must extend into the furnace to a considerable distance in order that the flame may be directed on the bath properly. It is customary to bring the tar or oil in over the gas port in such position that approximately 10 ft. of burner length are exposed to very high temperature due to radiation from the roof and walls. Again, the flame in an open-hearth furnace is reversed about once every 15 min. and the idle burner is exposed to the hot waste gases (2900°F.) and any tar or oil left in the idle burner will be carbonized, especially near the discharge tip. These difficulties are usually overcome by water jacketing the burner as indicated in Fig. 204, by withdrawing the idle burner a short distance into the water jacket at the

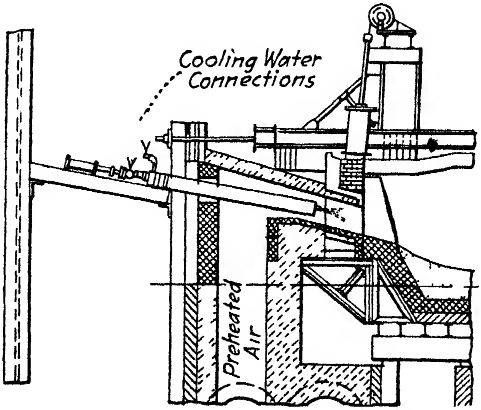

FIG. 204.—Water-cooled oil burner applied to open-hearth furnace. (*Hauck Manufacturing Company*)

time of reversal, and frequently by allowing a small amount of steam or air to bleed through the burner in order to remove any oil or tar in the tip and to help cool the burner. Such arrangements are usually satisfactory for preventing carbonization of the oil in the tip, and for preventing melting or excessive depreciation of the burner parts from the intense heat in the furnace.

The condition of the furnace directly after scrap has been charged is such that, with the burner in its ordinary position above the gas port, the oil or tar impinges upon the cold scrap piled but a few feet in front of it. At such periods it is not unusual to see the entire furnace enveloped in a cloud of dense smoke and actually see tar and oil drops spattering out through the peep holes in the doors. It seems advisable to arrange the burner so that its angle in relation to the bath can be changed during the melting-down period, or be drawn back into the bulkhead far enough to give adequate combustion before the cold scrap is encountered.

So far as is known, the steel is as good with coke-oven gas and tar as it is with producer gas. The presence of a large amount of water vapor, due to the combustion of hydrogen in the gas, and of steam, from atomization from the operation, does not seem to have a serious effect. Only tar or oil with low sulfur content should be used.

Undoubtedly, as each plant more and more approaches independence in coke production (*i.e.*, makes all its by-product coke) coke-oven gas and tar will be more widely used.

The Use of Recirculation of Flue Gas.—Oftentimes it is desirable to heat material to a comparatively low temperature only, a temperature far below the theoretical temperature of an oil or gas flame.* This can only be done with safety by one of two means: first, by using such a large percentage of excess air that the temperature of the resulting flue gas is sufficiently low; and, second, by burning the fuel efficiently with the correct amount of excess air and then diluting the high-temperature gases with recirculated, waste flue gas. The use of a large excess air is wasteful of heat, and, as a rule, it does not result in uniformity in heating. The use of recirculated flue gas is economical, since the excess air used is low, and, by means of a high recirculated gas velocity, it is possible to get good capacity from a given

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* What follows applies to any kind of a flame, whether produced by oil, gas, powdered coal, stoker fire, or other means.

equipment without the danger of overheating. The amount of recirculated flue gas to be used depends on local conditions, particularly the allowable difference in temperature between the flue gas and the material to be heated. The use of this principle is shown in Fig. 205, where starch is to be heated to 500°F. in four rotary cylindrical ovens by the use of an oil fire. The hot gases from the oil-fired furnace are diluted and chilled by the return flue gas before they touch the starch driers themselves. Also, the hot gases are put in at the top and the cooler gases are taken off at the bottom, since the flow of gas in this direction increases the uniformity of temperature within the ovens. By the use of this method high-temperature flames may be used with safety to heat or dry materials that would be damaged if exposed directly to high temperatures.

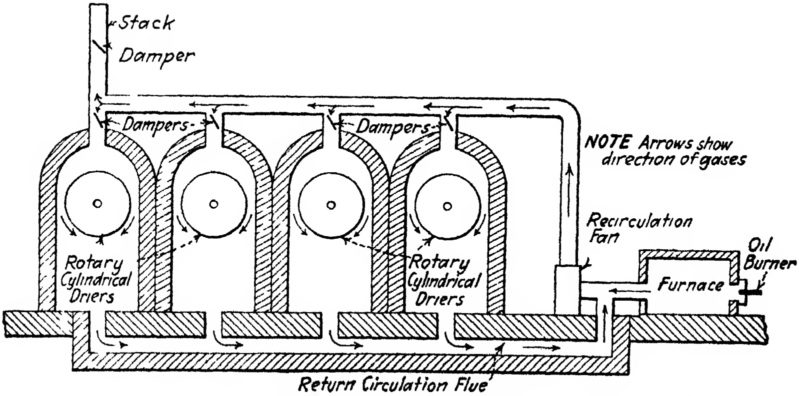

FIG. 205.—Application of flue gas recirculation. Diagrammatic.

DOMESTIC OIL BURNING

During the last four or five years the use of oil for the heating of homes, apartment houses, stores, and office buildings has increased tremendously. At the beginning of 1925 there were in the United States, probably, over 300,000 oil burners in use for this purpose. The chief attraction that has caused this rapid development is that complete oil-burning sets have been developed so that operation is nearly automatic. Thus, the burner maintains the building at a constant temperature regardless of weather, the extent to which the oil burner operates being regulated automatically. In the development of this equipment it was found that this regulation could be secured with greater reliability by the use of an intermittent-type burner rather than by means of throttling or reducing the flow of oil. Consequently, the method of operation, for a steam-heated building with a thermostat (temperature regulator) in the living rooms and a pressurestat (pressure regulator) on the boiler, is as follows: Assuming everything cold, the oil burner starts oil and air into the furnace and this mixture is automatically ignited. Combustion continues until the predetermined pressure on the steam boiler is reached; the burner is then automatically shut down by the pressurestat until the boiler pressure has dropped a bit, when the burner automatically starts up again. This performance is repeated until the living rooms have reached the desired temperature, at which time the thermostat shuts off the burner which then remains off independently of the pressurestat control on the boiler. After the temperature within the living rooms has fallen a definite amount (only 2 or 3° if desired) the boiler pressurestat again takes control until the living rooms have again reached their maximum allowable temperature. In this way no excess of pressure is developed in the boiler and the temperature of the living rooms is kept nearly uniform.

Quality of Oil Used for Domestic Burners.—The type of oil used in domestic heating (apartment houses, hotels, etc) is usually either* 44°Bé. furnace oil (a crude kerosene) or the heavier 28 to 30° Bé. distillate. The latter oil has the advantages of (1) lower price, (2) a wider market, (3) a higher gravity and therefore more B.t.u. per gal., and (4) less fire hazard on account of its higher flash point. No oil with a flash point below 150°F. should be used in this class of service.

Types of Domestic Oil Burners.—Domestic burners may be classified according to the way the air for combustion is taken in either as (1) natural draft, in which the air is drawn in by the natural draft of the chimney, or (2) forced draft, in which the air is blown in by an electrically driven fan. Again, domestic oil burners may be classified in accordance with the manner in which the oil is fed to the burner, either as (1) gravity feed, in which the oil flows into the furnace from a storage tank by gravity, or (2) pressure feed, in which the oil is forced or drawn into the burner by a mechanical action. In pressure-feed burners the oil is either (*a*) *forced* into the burner by a pump (usually a rotary) and atomized mechanically by passing through a fine

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* See Chap. VI.

nozzle, or (*b*) it is drawn into the burner by a *vacuum* induced by the blast of air for combustion, or (*c*) it is pumped to the inside of a *rotary* cup or disc, the oil being atomized and thrown off the rim of the cup by centrifugal force. In general, gravity-feed burners use natural draft only and the oil usually is vaporized, either by heating in a pipe or hollow casting which is in the burner flame, or by dripping on a plate or dome kept hot by direct contact with the flames. Oftentimes combustion is aided by preheating the required air. These burners are subject to trouble from plugging and carbonization, due to the oil "cracking" during the vaporization process. They have the further disadvantage that considerable attention is needed to keep them running properly. These burners usually smoke when starting up cold, and, since the amount of air drawn in is dependent on natural draft, the quality of the flame (freedom from soot), and the efficiency of the burner, are not usually built into the equipment but depend on the draft. These disadvantages are to a certain extent offset by the low cost of the burners and by the fact that low capacities are possible. This is of distinct importance in domestic heating. Furthermore, no electric power is required for operation. The use of burners of this type is usually confined to small homes.

In the vacuum type a blast of air from an electrically driven fan aspirates oil from a carburetor in a manner similar to feeding gasoline in an automobile. This type of burner requires the lighter grades of oil (crude kerosene of about 40 to 44°Bé.).

The forced-feed type of burner depends for its atomization on the garden hose or mechanically atomized oil burner principle, namely, forcing oil out through a small orifice under pressure. Several burners of this type thoroughly premix a small amount of air with the oil and then force the mixture through a fine orifice. This type of burner uses the heavier oils, such as gas oil (35°Bé.) or light-fuel-oil distillate (28 to 30°Bé.).

The rotary type of burner has already been described on p. 485. In general, the lighter furnace oils are most suitable for these burners, although oils as heavy as 12 to 14°Bé. may be used very satisfactorily if the oil is thoroughly preheated.

The Efficiency of Domestic Oil Burners.—Except for the gravity-feed type, practically all domestic burners are intermittent in action, *i.e.*, there is a period during which the fire is on full followed by a period in which the fire is shut off. This introduces into the question of burner efficiency a factor usually not encountered in large-scale industrial oil burning, namely, the heat lost up the stack during the period when the fire is off.

Although little has been published on this phase of burner efficiency, there are available the results of tests[10] on both the pressure type and the rotary type of domestic oil burner. The heat balances on boilers fired by one pressure-feed type of burner (in which the oil is mixed with a small amount of air and then sprayed through an atomizing nozzle into a fire box), and by two different makes of rotary domestic burners are given in Table C.

TABLE C

HEAT BALANCE ON BOILERS FIRED WITH DOMESTIC BURNERS

(Load = 0.0075 lb. of oil per sq. ft. of boiler area per hr.)

	Pressure-feed-type burner, per cent	Rotary-type burner No. 1, per cent	Rotary-type burner No. 2, per cent
Stack losses:			
1. Sensible heat	16.1	33.8	37.4
2. Latent heat of vaporization	6.6	6.5	6.6
3. Potential heat of unburned oil (as CO)	0.0	0.0	0.0
Radiation (assumed)	4.0	4.0	4.0
Unaccounted for (assumed)	4.0	4.0	4.0
Total losses	30.7	48.3	52.0
Boiler efficiency	69.3	51.7	48.0

In these tests the stack temperatures ranged from 515 to 695°F. The excess air was about 90 per cent for the pressure-feed type of burner, while over 300 per cent excess was required with each of the two rotary-type burners to get clean burning without a smoky flame. The high amounts of excess air (inherent in some rotary-type burners) of course result in low boiler efficiencies.

The effect of the length of time the burner is on and off, and of the ratio of $\frac{\text{Time Off}}{\text{Time On}}$, is shown in Fig. 206. The figure shows that the greater the ratio of time off to time on the lower becomes the efficiency, as the boiler is heating the air leaking in through the furnace during the period of idleness. Moreover, the temperature of the stack gas is highest just after the burner is shut down and therefore the efficiency is lower when the burner is on 1 min. and off 1 min. than when it is on 10 min. and off 10 min. It is, therefore, advisable in domestic oil burner installations so to adjust the rate of oil feed as to make burner operation as nearly continuous as possible. Also, it is essential that the furnace be closed tightly, with forced-draft burners, as otherwise the inherent efficiency of this type of burner is lost.

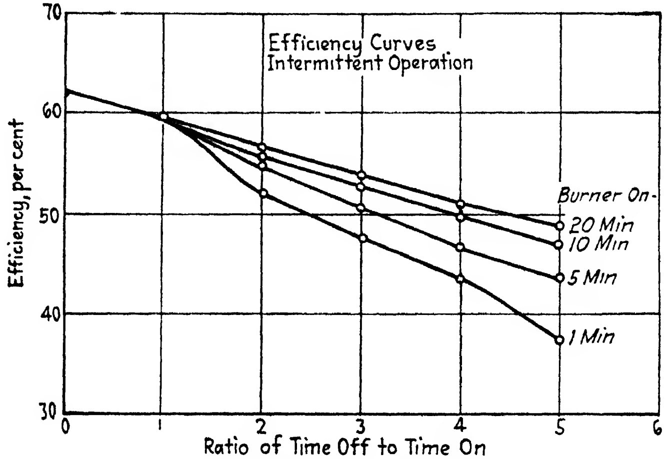

FIG. 206.— The efficiency of a domestic boiler fired with an automatic oil burner —intermittent operation.

Methods of Ignition in Domestic Oil Burners.—The *ignition* of the vaporized or atomized oil and air mixture usually is accomplished by means of a continuously operated gas pilot light. Sometimes the size or length of this pilot is made to increase automatically when the burner starts. Another method, which is suitable only with the higher Baumé, more volatile oils, is to ignite the oil-air mixture with an electric spark. In general, this method is not so reliable as the gas pilot, particularly with the heavier domestic fuels. A third method is automatically to turn on a gas pilot just prior to feeding the oil into the furnace, igniting the gas by means of an electric spark, which, in turn, ignites the oil; after the oil has been ignited, the gas and the spark are automatically shut off. With natural-draft burners the usual procedure is to throttle the oil fire to a small pilot flame of low oil consumption, but not sufficiently low to go out. Natural-draft burners, therefore, in contrast with the on and off pressure burners, operate continuously with a high and low fire

Safety Features of Domestic Oil Burners.—In order to prevent accidents, devices must be installed to shut off the burner if the oil does not ignite or if the oil is fed in too rapidly to be completely burned. The most common method is so to shape the burner or fire pot that any unburned or unignited oil collects and runs through a pipe to a small trip bucket on the outside of the furnace. The weight of the oil actuates a switch, shutting off the motor connected to the air and oil-pumping set. Another method is to place a pyrometer or bimetallic element, usually called a "stack safety," in the stack of the furnace. If the burner, motors, and fans are in operation, the temperature of the stack must be above a predetermined minimum: otherwise the stack safety operates a relay which shuts down the entire apparatus. This same device may, on the other hand, be placed so as to receive radiant heat from the fire box, *i.e.*, if no radiant heat is received by the pyrometer or bimetallic element within a certain time (for example, if the oil failed to ignite) the entire mechanism, fan, pump, etc., shuts down. In another device the safety mechanism is operated by the momentarily increased pressure which results when the atomized oil ignites. If the pressure impulse is not received within a fixed time, the control mechanism shuts down the burner automatically.

BIBLIOGRAPHY

1. Nat. Elec. Light Assoc., "Prime Movers Report," *Combustion*, **9**, 73 (1923).
2. SIBLEY and DELANY: "Fuel Oil and Steam Engineering," McGraw-Hill Book Co., Inc. (1921).
3. Denver Fire Clay Co., *Bull.* 301 (1922).
4. ANDERSON: "The Correct Method of Burning Fuel Oil," Mahr Manufacturing Co.
5. MARTIN: *Mech. Eng.*, **27**, 279 (1925).
6. BUELL: *Proc.* Eng. Soc. Western Pa., **38**, 201 (1922–1923).
7. "Steam Boiler Engineering," Heine Safety Boiler Co. (1920).
8. W. R. QUINN: *Combustion*, **9**, 449 (1923).
9. *Power* (July 17, 1923).
10. HASLAM, MARDEN, and WHITAKER: *Fuel Oil*, **3**, No. 3, 19 (1924).

CHAPTER XVII

FURNACE EFFICIENCY AND DISTRIBUTION OF HEAT LOSSES

No industrial furnace operates at 100 per cent efficiency and in almost all cases the heat losses exceed 20 per cent of the total heat input. In the majoirty of operations, except for boiler furnaces, the efficiency is less than 50 per cent and in extreme cases may be as low as 5 per cent. The amount of heat and money lost through this inefficiency reaches tremendous proportions. Although these losses are large, they are, in some cases, unavoidable. This is true only in furnaces where a serious effort is made to maintain the highest possible efficiency. Unless this effort is made, the efficiency usually will be below the maximum. Fuel saving is more marked in furnaces operating at low efficiencies, for, although an increase in efficiency from 70 to 80 per cent results in a 12.5 per cent decrease in fuel consumption, an increase from 20 to 30 per cent results in a decrease of one-third in the total fuel consumed.

Improvements in the operating efficiency of a furnace can be made best only when the operating efficiency and the magnitude of the various losses are known. In many plants large losses are continually taking place simply because they never have been determined.

Unless heat balances* are made, the only indication of low efficiency is in a high fuel cost. Since the determination of furnace efficiency is usually extremely simple, especially for approximate results, complete records of the characteristics of each furnace in a plant always should be available.

A low efficiency alone merely indicates that improvements in operation are necessary. Distribution of losses shows where, and often how, the improvements should be made. There usually are a number of places where heat losses may occur and the greatest improvement, obviously, can be made where the loss is greatest. All losses should be reduced to the lowest possible

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* See Chap. X on Combustion Calculations.

value, but the greatest savings can be made most easily by decreasing that loss which is furthest above the normal loss for the conditions under which the furnace is operating. For this reason a complete study of the distribution of the heat losses is necessary for a wise revision of furnace design or operation.

Although for purposes of design, maximum furnace efficiency does not guarantee maximum economy of operation, in furnaces already built it generally does. That is, by sacrificing efficiency it is often possible to construct a furnace with such a low investment, and therefore such low overhead charges, that it can be operated at a lower total cost than a larger, more expensive furnace operating at higher efficiency. With a furnace already in operation, however, maximum financial economy is usually at the point of maximum fuel efficiency, since the overhead charges are practically constant.

Furnace Losses.—The major divisions of heat losses in any furnace are shown by the following table:

TABLE CI

FURNACE LOSSES

1. Stack loss:
    (*a*) Sensible heat.
    (*b*) Potential heat.
    (*c*) Latent heat of water vapor.
2. Cinder loss:
    (*a*) Sensible heat.
    (*b*) Undeveloped heat.
3. Radiation.
4. Other special losses.
5. Unaccounted-for losses.

Stack Loss.—When any fuel is burned with air the heat generated is first picked up almost entirely by the flue gases formed on combustion. A large proportion of the heat may then be transferred to the material in the furnace, whether it be steel, molten metal, or the heat-absorbing surface of a boiler. A part of the heat, large or small, may be retained by the gases, pass to the stack and out to the atmosphere. The heat thus lost is considered as stack loss and may vary from 10 to 50 per cent, depending on the temperature and quantity of flue gas.

The heat thus lost may consist of sensible, potential or undeveloped, and latent heat. *Sensible heat* is the heat carried by the gases due to the temperature difference between entrance air and exit stack gas. *Potential* or undeveloped heat is the heat available as unconsumed gas, such as carbon monoxide, hydrogen, methane, or soot. *Latent* heat is the heat of evaporation of any water present in the flue gas.

Sensible Heat.—Sensible heat in the flue gases leaving the furnace is dependent on two factors: (1) the volume of gases, and (2) their temperature, which governs the amount of heat per unit volume. The amount of heat per unit volume depends almost wholly on the exit temperature of the stack gases, while the volume depends on the air used. In order to maintain the minimum sensible-heat loss, it is necessary to cut down the volume of the stack gases by using a minimum excess of air for combustion and to secure a minimum stack temperature by the use of suitable heat interchangers. In general, the proper regulation of air and fuel, with the elimination of leakage into the furnace, and keeping the heat-absorbing surface free from soot, scale, etc., are the most productive methods of securing best results.

Potential or Undeveloped Heat.—Potential heat in the unconsumed gases is determined by the design of the combustion chamber and by the excess air used. With a furnace in operation, the chief means of regulating the loss of undeveloped heat are by increasing the efficiency of mixing and by controlling the air used for combustion.

Air Regulation.—Proper regulation of air does not necessarily mean the use of the quantity of air theoretically necessary for combustion of the fuel. A decrease in air to this amount will ordinarily cause a large loss of undeveloped heat (unburned fuel). The two factors which are affected by a change in air are sensible heat and undeveloped heat. In general, an increase in air used increases the sensible-heat loss and decreases the loss due to unburned fuel. The use of quantities greater than that required to burn all the fuel is obviously wrong. The maximum fuel efficiency is obtained at the point where the sum of sensible and undeveloped heat is a minimum. Since a small amount of carbon monoxide (CO) causes a large undeveloped-heat loss, the point of maximum efficiency is usually found to be where the CO is less than 0.5 per cent and the excess air a minimum for this condition (see p. 364). In any furnace, combustion should be carried on under conditions which result in the minimum total stack loss; that is, at the point where the sum of the sensible-heat loss, the potential-heat loss, and the latent-heat loss is a minimum. Since the latent-heat loss is constant for any given fuel, the air supply should be so regulated that the sum of the sensible-heat loss and the undeveloped-heat loss is a minimum.

Latent Heat.—Latent heat in the stack gases is the heat of condensation of the water in the flue gas, which includes the water formed from the combustion of the hydrogen in the fuel and the water present as moisture in the fuel and air. The largest of these items is often the water from the net hydrogen in the fuel. This loss cannot be controlled except by changing the type of fuel, and it is essentially independent of the amount of air used. If the heat balance were based on the net or low heating value of the fuel, this latent heat would not appear as a loss in the heat balance, but general practice is to use the gross or high heating value, so that the heat of condensation of the water must be considered as a loss in the heat output

Cinder Loss.—The cinder resulting from the combustion of solid fuel may contain a certain proportion of combustible matter caused by improper firing or by the enclosure of carbon within the particles of ash. In hand-fired furnaces operating at relatively low capacity, this loss is usually small. Mechanical stokers, operating at a high overload, unless using preheated air, often carry a large amount of unconsumed coal into the ash pit. Cinders from a mechanically stokered boiler have been found to contain as high as 60 per cent carbon, when the stoker was operating at a very high overload. Although under some conditions the ash loss may be negligible, supervision is necessary to assure minimum loss. In powdered coal furnaces it is necessary to watch the unconsumed carbon in the fine dust particles going out the stack. This is particularly important when low volatile coals are being burned with low excess air.

Sensible heat lost in the cinder is seldom of any importance and is always less than 1 per cent. Usually the heat is returned to the furnace by cold air passing over the hot cinder. For these reasons, sensible-heat loss in the cinders is seldom determined.

Radiation.—The loss in heat from a furnace due to conduction through the walls may be the most important source of loss. In electric furnace work and, in general, in all large furnaces operating at high temperatures, the heat lost by conduction through the walls and subsequent radiation and convection from the surface, may be even greater than the amount of heat actually utilized. Proper construction of the furnace may reduce this loss to a minimum, but to accomplish this, a large initial investment may be necessary. The thickness of furnace walls and the type of insulating materials used are dependent on an economic balance of fuel saved and the increased fixed charges on the furnace. With cheap fuel a simple furnace allowing high radiation losses is more economical than an expensive, well-insulated furnace with low losses. Large furnaces, using large quantities of fuel of high cost, necessitate good insulation for maximum economy. Many times it is necessary to eliminate heat insulation and thus permit a large heat loss by radiation and convection for the purpose of increasing the strength of brick walls and arches.

Other Special Losses.—In furnaces of various types other special losses may be of great importance. In the blast furnace, open-hearth furnace, and gas producer, heat lost in cooling water cannot be neglected In cement or brick kilns heat lost in the product may be relatively large. The loss of heat due to heating the furnace itself may be large in non-continuous furnaces. In determining the distribution of heat losses, all special losses of the above type should be measured if possible

Unaccounted-for Losses.—In most furnaces the total heat input cannot be accounted for in the heat output. That is, the measured output is somewhat lower than the input. This heat is labeled "unaccounted-for loss" and represents a number of minor losses which either cannot be measured or are too small to make their determination advisable There is a tendency, however, to allow too many losses to accumulate in this category. The unaccounted-for loss should never be over 10 per cent, since it should be possible to determine a loss as great as 5 per cent. Usually, on a careful test, it is possible to reduce the unaccounted-for loss down to 3 per cent or under. Often this loss is due to the fact that the temperature of the stack gases was measured under such conditions that the *apparent* temperature of the gas was lower than the true temperature, due to the cooling effect of the walls of the stack on the pyrometer.[1]

Heat Balance.—The heat balance is an expression of the results of a test on a furnace, showing the distribution of heat input and heat output. It is a convenient method of outlining the important factors and summarizing results. Several methods of testing* a furnace are possible, depending on the results desired. Thus either increased capacity or improved fuel efficiency may be the important factor. Again, a material balance or an element balance, such, for example, as a sulfur balance on a blast furnace, may be necessary to determine the distribution of certain materials or elements. For each of these purposes a slightly different method of testing is necessary. The heat balance, such as a balance on an open-hearth furnace, necessitates a material balance and a measurement of capacity and efficiency, and so is a very complete method of testing. An element balance may be made in the same manner as a heat balance, determining amounts of the element involved instead of heat.

Basis of Heat Balance.—In presenting a heat balance, and comparing with others on the same type of apparatus, care must be taken that the *basis* of the balances is the same. Obviously, a balance on a percentage basis is better for comparative purposes than a balance on a B.t.u. per hr. on per ton basis, since the time and quantity factors are eliminated. The basis on which the percentages are reported must be definitely stated. The usual method is to include all the materials in the input and determine the results as a percentage of this total input. On the other hand, the heat in the fuel may be considered as the only input and the results determined as per cent of heat in the fuel. Thus the heat input to a boiler may be considered as the heat in the fuel alone or as the heat in the fuel plus the heat in the feed water. In the first case the useful heat is that used in raising the feed water from its initial temperature and converting it to steam at its final temperature. In the second case the useful heat may be considered as the total heat in the steam. For example, assume that the heat in the feed water is 10 per cent of the heat in the fuel. Heat balances based on the two conditions above are shown in Table CII.

In other words, 60 per cent of the heat in the coal is useful work, as shown by the first balance, while 63.6 per cent of the total heat

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* For accurate tests to determine whether the furnace is operating at guaranteed rating and efficiency, the test codes adopted for the testing of such a furnace must be rigidly followed. Such codes are the Boiler Test Code and Producer Test Code of the American Society of Mechanical Engineers[2] and the Coke-oven Test Code of the American Gas Association.[3] The data obtained by following the outline of these codes are essentially the same as in the heat balance as outlined here, but the arrangement is somewhat different and additional incidental data are required.

TABLE CII

BOILER HEAT BALANCES

	Method 1, per cent	Method 2, per cent
Heat in fuel	100	90.9
Heat in feed water		9.1
Total heat input	100	100.0
Useful heat in steam ..	60	63.6
Losses	40	36.4
Total output .	100	100.0

input is present in the steam, as shown by the second balance. Both bases are used in practice, but the actual basis chosen must be designated to make the results of value.

A difference is often encountered due to the use of high (gross) and low (net) heating values. For example, consider a case where the total heat lost in the stack gases from a boiler fired with coke-oven gas is 25 per cent of the gross heat in the fuel, while the sensible- and potential-heat losses are only 13 per cent of the gross heating value, or 15.3 per cent of the net heating value. Assuming no other heat losses except stack losses, the efficiency of the boiler based on the gross heating value is 75 per cent, while based on the net heating value it is 84.7 per cent.

In specifying the basis used, the *datum line* should also be included. The *datum line* is the starting point for calculations of sensible heat and may be at any normal temperature. The heat input as sensible heat in the air and fuel used will be different if a starting temperature of 32°F. is used than if the temperature is 60°F. On the other hand, if 2 per cent is added to the heat input in this manner it will also be added to the heat output, so the net result will still be a complete balance. This 2 per cent added heat input will appear as in the stack loss as sensible heat in the output, and the efficiency of the boiler in the previous example (gross heating value used) on this basis will then be decreased by 1.5 per cent, or from 75 to 73.5 per cent. The stack loss will be increased from 25 to 26.5 per cent.

From the above values, it is evident that, for comparing the efficiency of equipment, the basis must be the same. Any basis may be used, but it must be clearly shown, and data must be available for calculating the values on any other basis. For convenience in comparisons, it is recommended that the total heat input, high heating values, and a datum line of 60°F. be used in all heat balances. This follows the most widely used practice in the industrial field.

Summary of Heat Balances.—The distribution of heat in furnaces of different types is summarized in Table CIII. No attempt has been made to present best results for each type, but merely to show what results are being obtained in general practice. Considerable variation from these results may be found in many furnaces, but the values of the various heat quantities indicate the usual magnitude.

TABLE

HEAT DISTRIBUTION IN

	Boiler	Gas producer[4]	Blue water gas sets[5]	Carburetted water gas sets[5]	Oil gas sets[6]	Coke oven[4]
1 Heat input (per cent of total heat)						
(*a*) Potential heat in fuel	91 9	96 3	93 3	Coal 57 3 Oil 38 5	96 7	Coal 92 8 Gas 7 2
(*b*) Heat in auxiliaries (air and steam)	0 0	2 7	6 7	4 1	3 0	0 0
(*c*) Heat in raw materials	8 1	0 0	0 0	0 0	0 0	0 0
Total	100 0	100 0	100 0	100 0	100 0	100 0
2 Heat output						
(*a*) Heat in product	71 2	72 7	49 9	60 2	42 8	87 5
(*b*) Heat absorbed in reactions						
(*c*) Furnace efficiency	71 2	Hot 90 2 Cold 72 7	49 9	60 2	42 8	87 5
(*d*) Heat losses, total	28 8	Hot 9 8 Cold 27 3	50 1	39 8	57 2	12 5
(1) Stack loss	21 0	Sensible heat in producer gas 17 5	39 2	19 2	7 2	2 3
(2) Cinder loss	3 9	1 7	7 0	4 0	0 0	
(3) Radiation loss	3 9	8 1	3 9	3 7	1 4	1 2
(4) Unaccounted-for losses					1 3	2 5
(5) Cooling water loss						
(6) Miscellaneous Accounted-for losses					Potential heat in Carbon 9 5 Potential heat in tar 3.1 Potential heat in lamp-black 26 2 Other losses 8.5	Sensible heat in coke 4 1 Sensible heat in gas 2 2 Sensible heat in tar 0.2
Remarks						

In order to correlate and compare the results on all the types of furnaces, the various heat balances were arranged in accordance with the headings of the first vertical column. In some cases it is not possible to separate the losses as given in this column and often several separate losses had to be grouped together. For example, where no unaccounted-for losses occur in the table the radiation loss given includes both radiation and unaccounted-for losses.

Furnace Testing.—The method of testing a furnace depends on the results desired. Since a heat balance includes most of the possible results, and since the same general methods are appli-

CIII

INDUSTRIAL FURNACES

Blast furnace[4]	Open hearth furnace[1]	Billet Furnace		Rotary cement kiln[9]	Rotary lime kiln[4]	Shaft lime kiln, producer gas fired[4]	Down draft brick kiln[10]	Tunnel kiln continuous[11]	Tunnel kiln Dressler[11]	Ring type brick kiln[12]	Blast furnace stove[4]
		non-[7] continuous	continuous[8] with recuperator								
93 1	95 0	100 0	100 0	100 0	100 0	95 9	100 0	100 0	100 0	100 0	100 0
6 8	0 0	0 0	0 0	0 0	0 0	4 1	0 0	0 0	0 0	0 0	0 0
0 0	5 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0	0 0
100 0	100 0	100 0	100 0	100 0	100 0	100 0	100 0	100 0	100 0	100 0	100 0
8 2	Steel 21 6 Slag 3 1	25 0	52 5	14 1	12 3	12 0	23 7	17 2	67 2	51 5	62 0
28 4	−7 9			22 4	28 4	35 0	2 7	7 9	0 0	0 0	0 0
36 6	11 8	25 0	52 5	22 4	28 4	35 0	26 4	25 1	67 2	51 5	62 0
63 4	88 2	75 0	47 5	77 6	71 6	65 0	73 6	74 9	32 8	48 5	38 0
55 0[a] 6 1	62 0	40 0	29 3	48 3	41 2	30 6	33 3	27 9	8 3	38 2	18 6
		2 0				0 7	0 0	3 3	.	.	. .
0 3	21 2	27 0	16 1	15 2	7 9 10 2	12 4	40 3	43 7	5 4	10 3	10 1
2 0		4 0	2 1			4 7					2 6
[a] Potential Heat in. B F Gas is not a real loss because later used		Heat in slag 2 0				Loss in producer gas main 4 6 Overall balance including producers and kiln			Heat in trucks 19.1		6 7

cable to material or element balances, the following discussion will be on the basis of a heat balance.

Before starting a test, the time necessary to prevent error due to accumulation or depletion of heat within the furnace must be decided upon. In a non-continuous process, the test must cover at least one complete cycle and should preferably cover two or more. In a continuous process where there is no possibility of change in the entering raw materials and the finished product, the time factor is not so important. In the majority of commercial applications the time must be long enough so that the raw materials entering will leave as products before the test is completed. Thus a blast-furnace test, to be of value, must be carried over a period of three or more days in order that the entering materials may have time to pass through the furnace and come out as product. When material does not move continuously through the process, and may build up within the apparatus, as, for example, in a boiler, the test must be carried over a sufficient period (usually 24 hr. or more) so that any accumulation, such as a change in the amount of fuel on the grate, will be negligible compared to the total amount consumed.

Quantity Measurements.—For a complete and accurate test, quantity measurements are essential. The amount of solids, liquids, and gases used or formed, and the analyses of each, should be determined as far as possible. Standard methods of measurements and analysis should be used. In many cases it is difficult, or impossible, to weigh or measure quantitatively *all* the streams of material entering or leaving a furnace. By means of balances, using the data from analyses of the various streams, the ratio of one stream to another may be calculated. If one stream is weighed or measured, and analyses of all streams are known, then it is possible to calculate the quantities of each stream. To determine efficiency, analyses without quantity measurements will often be sufficient, but to determine capacity the quantity of at least one stream of material must be weighed or measured. For example, the amount of air entering a boiler furnace is seldom measured directly, but is determined from the weight and analysis of the coal and the analysis of the flue gas and cinder by means of carbon, nitrogen, and ash balances.* Although it is necessary to weigh or measure only one stream of material entering or leaving a system and to *analyze* the others, still, for purposes

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* See Chap. X.

of cross-checking, it is advisable to weigh or measure at least two streams.

Heat Content.—The measurement of the heat content or the heat per unit quantity must also be made. These usually consist of the determination of heat of combustion and temperature. The former may be obtained by means of a calorimeter[13] and the second by means of thermometer, thermocouple, or optical pyrometer [14] Other data on the thermal capacity of materials may be obtained from tables such as those in Chap. X.

Heat Balance.—The heat efficiency of a furnace and the distribution of the heat losses may be best shown by a heat balance. The heat input to and output from the furnace are determined and may be arranged, in general, in the following manner·

TABLE CIV

HEAT BALANCE

Input·
- Heat in fuel (coal, oil, gas, etc.).
- Heat in auxiliaries (steam, air, etc.)
- Heat in raw materials (feed water, molten iron, etc.).

Output:
- Heat in product (steam, steel, etc).
- Heat absorbed in reactions in furnace (for example, $CaCO_3 = CO_2 + CaO - 76{,}500$ B.t u. per lb.-mol).
- Heat losses.
  - Stack loss:
    - Sensible heat.
    - Potential heat (undeveloped heat or unconsumed fuel).
    - Latent heat of water (from the combustion of H_2 in the fuel, etc.).
  - Cinder loss.
  - Radiation loss.
  - Other measurable losses.
  - Unaccounted-for losses.

Heat Input.—The heat input is derived from the heat in the fuel, the heat in the auxiliaries, and the heat in the raw material.

Heat in Fuel.—The amount of heat entering the furnace is the product of the quantity of fuel and the heat per unit quantity, or the pounds multiplied by the B.t.u. per lb. Measurement of quantity is dependent on the type of fuel and special conditions of handling at the furnace. If coal is the fuel it is usually possible to weigh it. Oil is easily metered and gas, natural or manufactured, is easily measured by a displacement meter or by an orifice. In some cases, special methods have to be used, especially if only one of a series of furnaces is being tested. For example, it is usually difficult to measure directly the volume of blast-furnace gas being consumed in a blast-furnace stove. It is possible, however, to blow a known amount of air into the pipe line feeding the gas to the stove, and, by analyzing the blast-furnace gas before and after such dilution, to calculate the volume of blast-furnace gas entering a particular stove.

Determinations of heating value may be made directly by means of a calorimeter or by calculation from analysis. In any case a knowledge of the total carbon content of the fuel is necessary for other engineering calculations. The actual analytical data necessary relative to the fuel are the heating value, moisture, total carbon, and ash. If no other means exist, the carbon in coals may be calculated from the proximate analyses and heating value by formulas given on page 257. Total carbon in gases may be determined by explosion and absorption of CO_2

Heat in Auxiliaries.—In many cases the heat entering the system in auxiliaries (steam and air) is a relatively small fraction of the total heat input and may be neglected. If a large amount of steam, or preheated air is used, however, the amount of heat entering in this manner may be large. The quantity of these materials may be measured directly, although in many cases they are more easily calculated. The methods of calculating the amount of steam used in a producer and the amount of air used in furnaces are discussed in Chap. X. The sensible heat per unit quantity of material is determined from temperature measurements with thermometer or thermocouple and a knowledge of the specific heat.

Heat in Raw Material.—The quantity and the temperature of raw materials must be measured and the heat entering by this path determined from the product of the weight, the temperature difference (above the datum line chosen), and the mean specific heat over the temperature range involved.

Heat Output.—The heat output is divided between heat in the product, heat absorbed in reactions, and various losses. These losses are: (1) the heat lost in the manufactured product, (2) the stack loss, (3), the cinder loss, (4) the heat lost by radiation, (5), other known losses, and (6), unaccounted-for losses.

Heat in Product.—The amount of heat in the product is in many cases a direct measure of the heat utilized efficiently by the furnace. In the boiler furnace or reheating furnace, for example, the object to be obtained is heating the product (steam or steel). The quantity of product may be determined by direct measurement, but is usually calculated by material balances from the amount of raw material. In the boiler or heating furnace, the product is equal in weight to the raw material, and since the raw material may be measured more easily this is usually done. In the lime kiln either the entering raw material or the products may be measured and the other calculated by a calcium balance based on analyses. The gaseous product from a producer is calculated from the weight of coal and the analysis of coal, ash, and gas for carbon. Balances, of materials or elements, may be applied in practically all cases to calculate the ratio of raw material to product, and measurement of the amount of one will permit calculation of the amount of the other.

The heat per unit quantity of product may be sensible, potential (undeveloped), or latent heat, depending on the type of product. In any case it may be determined from analyses, temperature, and heating value of the product. The last is not absolutely necessary, since in many cases it may be calculated from analysis, but when it is a major proportion of the heat output, it should be determined directly to prevent errors in the results. In the boiler a major proportion of the heat is latent heat of evaporation, with some sensible heat of the liquid and of superheat. Producer gas contains most of its heat as potential or undeveloped heat, together with sensible heat due to the gas being at a high temperature. Steel from the open-hearth furnace contains most of its heat as sensible heat, with a small amount of latent heat of fusion and of other chemical changes in the steel. In general, the amount of heat may be determined from the temperature, analysis, and quantity of material.

Heat Absorbed in Reactions.—When any chemical reaction takes place a definite amount of heat is liberated or absorbed. Thus, combustion of fuels with air generates heat, while reduction of CO_2 with C absorbs heat. In the same way the reaction

$$CaCO_3 = CaO + CO_2 - 76{,}500 \text{ B.t.u. per lb. mol,}$$

which takes place in the lime kiln, absorbs heat. Numerous other reactions taking place in furnaces either absorb or develop heat. It will be noted that the actual heat of combustion of the $CaCO_3$ or $CaO + CO_2$, as measured in a calorimeter, would be zero. In other words, the heat of combustion has not changed, but a certain amount of heat energy has been absorbed by the system and converted into chemical energy. The true balance on a furnace should deal with the total energy input and total energy output, but conversion of the various types of energy to a thermal or heat basis permits simple calculations. Thus, instead of adding the total energy in $CaCO_3$ to the input, and the energy in CaO and CO_2 to the output, the difference between output and input alone is used, and is converted to the heat-energy basis. The amount of heat absorbed or developed by any reaction may be found from tables of chemical reactions.[15] [16] The quantity of material taking part in the reaction is determined from the analysis and quantity of raw materials and products.

Heat Loss in Products.—In many cases the heat contained in the products must be considered as heat lost, rather than useful heat. For example, the heat contained in lime from a rotary kiln has no further usefulness as sensible heat in the lime, and should be recovered. It is, therefore, a heat loss. However, the heat contained in iron from the blast furnace may not be lost, since it is necessary to maintain the iron in the molten state for pouring. The heat contained in the steel from the open-hearth furnace is usually not considered as a loss, because one of the major functions is melting of the iron. It is conceivable, however, that the heat in all these products might be recoverable and thus constitute a loss of heat. Common practice often considers the heat in the product as useful work, except where recovery of the heat is perfectly feasible. When comparing the efficiencies of various types of apparatus, it is absolutely necessary to know the basis on which the calculations were made.

Stack Losses.—The amount of heat lost in the stack gases as sensible, latent, and undeveloped heat may be the major source of heat loss. To determine the magnitude of these losses, analysis of fuel and flue gas and temperature of the flue gas are usually necessary. The methods of calculation are outlined in Chap. X, and under some circumstances the analysis of this stack gas and its temperature gives sufficient data to calculate the per cent heat lost up the stack (see Figs. 53 to 57, Chap. X).

Cinder Loss.—For the determination of cinder loss, analyses of fuel and cinder are necessary. In most cases, and especially in boiler testing, the actual weight of carbon lost in the cinder may be more accurately calculated from analysis than from weighing the cinder directly, due to the varying proportion of water present.

The use of the analyses for this calculation is illustrated in Chap. X.

Radiation Loss.—In furnaces having walls which are maintained at relatively low temperatures, the amount of heat lost is small and is usually determined by difference when all other losses have been computed. The loss may be large in furnaces operating at high temperatures and should be calculated directly. If the temperature of the outside walls of the furnace is measured, the coefficients of heat transfer from the walls to the room may be determined, and the heat loss calculated as indicated in Appendix II. When the thickness of the walls and the temperatures on each side are known, the amount of heat lost may be determined from the conduction equation (p. 749). In some cases both methods may be used and the results should agree. The error may, however, be large, due to leakage of cold air or hot gases through the walls, or to errors in the value of the heat-transfer coefficients. A relatively large error will make little difference in the total heat balance, if the percentage of heat lost in this manner is small. In problems of design, radiation loss may be more readily calculated from the conduction loss through the walls, but in furnace testing combined heat flow from the walls by both radiation and convection usually is sufficiently accurate and is much simpler to determine. This is especially true in large and complicated apparatus, such as coke ovens or blast furnaces. For this purpose the furnace may be divided into areas of approximately equal temperature, the average temperature of these areas measured, and the radiation losses from these areas calculated and added together to give the total loss.

Other Known Losses.—The major losses occurring in most furnaces are those outlined above. In some cases other minor, but important, losses may also be determined. For example, in blast-furnace stoves the loss of heat in cooling water may be as high as 5 per cent of the total heat, and may be determined from the quantity and rise in temperature of the water. Leakage of gas out of the furnace, passage of soot from the stack, and loss of product are examples of other sources of heat losses which are usually of minor importance, but in special circumstances may warrant investigation.

Unaccounted-for Losses.—Almost all heat balances on commercial furnaces show a deficiency of heat in the measured heat output. In some cases errors are possible, but even in the most accurate work some heat is not accounted for. Low stack temperature, due to the fact that pyrometers usually read lower than true gas temperature, is one of the most frequent troubles. Loss of heat, due to small amounts of CO, H_2, CH_4, or soot in the flue gas, is usually small but may cause a deficiency of heat in the output. Radiation and conduction loss may be inaccurately measured, or not determined in some instances. For these reasons an item of unaccounted-for losses is usually inserted in the heat balance, determined from the difference between the total heat input and the total known heat output and the losses, to include all losses not measured, and also to include errors. In some cases radiation and unaccounted-for losses are grouped together. This is logical, since the radiation loss may be small compared to unaccounted-for losses, or *vice versa,* so that in such cases the sum may be considered equal to either, depending on which is of major importance. Thus in the boiler furnace, radiation and unaccounted-for losses are usually in the neighborhood of 5 per cent. Calculated radiation loss would not be over 1 per cent, so that the total may be considered as unaccounted-for loss. In the open-hearth furnace, radiation and unaccounted-for losses may be as high as 30 or 40 per cent, of which only 4 to 6 per cent is unaccounted-for loss, while the remainder is due to radiation. The large errors possible in determining radiation loss would not permit an accuracy of better than 20 per cent in this item, so that the assumption that the radiation plus unaccounted-for loss is due to radiation is, in this case, warranted.

Interpretation of Test Results.—When the data from a test have been obtained and the results calculated, the results must be rightly interpreted to be of the greatest value. A heat balance of itself is merely a tool to be used in raising the efficiency of the furnace, and unless the results are used the data are of little value. From a comparison of the various losses with those obtained with other furnaces of the same type, the place where improvements should be made may be found. For example, if a boiler is operating with a stack loss of 30 per cent and a cinder loss of 5 per cent, the greatest possibility for saving is in decreasing the stack loss to 15 per cent. On the other hand, if the boiler were operating with a 20 per cent stack loss and a 20 per cent cinder loss, the greater saving could be made in correcting the cinder loss, although some saving could be made by decreasing the stack loss.

Furthermore, the method of decreasing the stack loss may be seen from the heat balance. For example, a high percentage of sensible-heat loss necessitates the opposite type of regulation from a high percentage of potential- or undeveloped-heat loss. A high stack temperature may mean by-passing in the furnace baffles, a coating of soot or scale on the tubes, or that an economizer or preheater could be used profitably. A thorough study of the data of the heat balance will ordinarily indicate the methods to be employed in increasing furnace efficiency.

For discussion of the interpretation of heat-balance data, combustion devices may be divided into two classes: (1) *final furnaces*, in which the fuel is burned to be used directly for heating purposes; and (2) *preliminary furnaces*, in which the fuel is treated for further use in generating heat. In the first class are boilers, heating and treating furnaces, open hearths, blast furnaces, and all types of kilns. In the second class are gas producers, water-gas and oil-gas sets, coke ovens, and the like. Heat efficiency is of paramount importance in the first class, while quality and quantity of the product, with its final efficiency in the furnace in which it is to be used, are the important factors in the second class. Although heat efficiency should be controlled in the latter type of apparatus, it is often high and little direct saving can be made. The opportunities for improvement may lie chiefly in making a product which will give a higher overall efficiency from raw fuel to final heated product. In other words, stress must be placed on the production of a secondary fuel of highest quality rather than on a high thermal efficiency in preparation of this fuel.

In the following pages the heat distribution in the first type of apparatus is described in considerable detail, while that for the second class is presented in the various chapters on secondary fuels. For complete discussion of the factors involved in the operation of the second class of furnaces, reference should be made to Chaps. XVIII, XIX, XX, in which they are fully discussed.

Hand-fired Boilers.—The hand-fired boiler is the most inefficient type of boiler unless properly controlled. With a fireman well versed in the principles of boiler firing, fairly good results can be obtained, but some control method is necessary to insure their maintenance (for the underlying principles see Chap. XIII).

In testing hand-fired boilers a difficulty is encountered in the varying composition of the gases (see p. 377). In order to avoid this a large number of stack-gas samples must be taken during a firing period. A composite average analysis over the period of several hours does not show the actual conditions, although it is sufficiently accurate for the heat-balance calculations.

The chief sources of loss, with the variations and a normal minimum, are shown in Table CV.

TABLE CV

Losses in Hand-fired Boilers

As per cent of heat in fuel

	Variations, per cent	Normal minimum, per cent
Stack loss (total)	20–40	18
1. Sensible heat	15–35	15
2. Undeveloped heat	0– 5	0.5
3. Latent heat of H_2O	3– 5	3
Cinder loss	3–20	5
Radiation and unaccounted-for loss	3–10	5

From the table it is evident that the greatest variations are in sensible heat in the stack gases and in the cinder. Undeveloped heat in the stack gases may increase to more than 5 per cent in some boilers due to poor firing, but this is not usually the case. Latent heat of water is entirely dependent on the type of fuel used and permits of no variations in the furnace.

If a hand-fired boiler is operating at an efficiency of 50 per cent, with a sensible-heat loss of 30 per cent, undeveloped-heat loss of 1 per cent, latent-heat loss of 5 per cent, cinder loss of 8 per cent, and radiation and other losses of 5 per cent, the greatest difference from the normal minimum is in the sensible-heat loss—30 per cent of the heat in the coal. Although some improvement might be made in the cinder loss, the great opportunity is in decreasing the stack loss. Conversely, if the sensible-heat loss were 20 per cent and the cinder loss 20 per cent, the obvious point of attack for improvement is in the cinder loss, where a saving of 20 minus 5, or 15 per cent, is possible.

Reducing Stack Loss.—Stack loss may be reduced in several ways, depending on the cause of the loss. High excess air, poor mixing, leakage of air into the setting, by-passing of the gases in the boiler, poor firing methods, carrying too thick a fuel bed, and improper damper regulation are some of the causes of high stack loss.* The cause may frequently be determined by means of the data from a heat balance or from the analysis and temperature of the flue gases. When the cause has been determined, regulation is relatively easy in almost all cases.

A stack analysis showing only 50 per cent excess air but an abnormally high temperature (over 500°F.) indicates that the heat is not being efficiently abstracted from the gases, possibly due to by-passing through broken baffles, soot, scale, etc. A high percentage of excess air and a low stack temperature (below 300° F.) indicate leakage of air through the walls of the setting. High excess air with a higher temperature (400 to 500°F.) indicates too much air in the combustion chamber. If under these conditions carbon monoxide is found in the gases, poor mixing is evident. High excess air indicates improper damper regulation. It is impossible to regulate the damper to provide the proper quantity of air at all times, but the amount of air should be such that the total of potential (undeveloped) and sensible heat lost during a firing period is a minimum.

The following table shows a series of gas analyses and temperature measurements with the methods which should be used to decrease the stack loss.

TABLE CVI

Losses in Boilers

	Example No.					
	1	2	3	4	5	6
Per cent excess air	30	150	150	30	150	150
Per cent CO	0 1	0.0	0 5	0.8	0.4	2.5
Temperature of flue gases, degrees Fahrenheit	400	500	300	400	500	400

Probable Method of Improvement.

1. No changes necessary.
2. Inspect and repair baffles. Regulate damper to decrease excess air to 50 per cent.

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* See Chap. XIII.

3. Reduce leakage through setting Increase secondary air over fuel bed. Use thinner fuel bed

4 Increase secondary air immediately over fuel bed

5 Decrease secondary air over fuel bed. Use arch or baffles to promote mixing Operate with thinner fuel bed

6 Change method of firing by firing more frequently Too much distillation of coal at one time Decrease secondary air to combustion space

The results of a series of gas analyses during a firing cycle show more accurately what per cent excess air should be used. If, with a given method of firing, the soot formation and undeveloped heat loss are high, the method or the rate of firing should be changed.

A typical heat balance on a hand-fired boiler is shown in Table CVII.

TABLE CVII

	Per Cent
Input	
Heat in coal	100 0
Output	
Heat in steam (above feed-water temperature)	50 0
Stack loss	
1 Sensible heat	25 0
2. Undeveloped heat	0 0
3. Latent heat	5 0
Cinder loss	15 0
Radiation and unaccounted-for losses	5 0
Total	100 0

The high cinder loss indicates that improvements should be made in the method of firing and cleaning. The high sensible-heat loss indicates too high excess of air or too high stack temperature. The method of improvement depends on which is faulty.

Stoker-fired Boilers.—The efficiency of a stoker-fired boiler almost always is above that of a hand-fired boiler and usually varies from 60 to over 80 per cent. Test runs showing efficiencies slightly higher than this are possible, but hardly represent normal practice except in the large central power houses where efficiencies of as high as 90 per cent have been reached when air preheaters, economizers, water-cooled side walls, etc., were used. The size of the installation, the use of economizers, air preheaters, water-cooled walls, etc., very greatly affect the efficiency. The variations in the chief sources of loss are shown in Table CVIII.

TABLE CVIII

LOSSES IN STOKER-FIRED BOILERS

(As per cent of heat in fuel)

	Variations, per cent	Normal, minimum, per cent
Stack loss (total)	15–35	15
1. Sensible heat	10–25	12
2. Undeveloped heat	0– 5	0.5
3. Latent heat	3– 5	3
Cinder loss	5–30	7
Radiation loss	3–10	5

As in hand-fired boilers, stack and cinder loss are the chief sources of inefficiency and permit the greatest improvement. The major difference is in the cinder loss, which may increase rapidly, due to coke being carried into the ash pit if the boiler is forced. The same principles govern the regulation of stack loss as apply for hand-fired boilers. In addition, high rates of firing cause blowholes through the fuel bed which show up as high excess air in the flue gases and cause excessive clinkering with a consequent high cinder loss. Preheating the air greatly lowers the cinder loss.

Example of Boiler Test.—A heat balance on a stoker-fired boiler furnace is shown in the following table.

TABLE CIX

HEAT BALANCE ON BOILER FURNACE

	PER CENT
Input:	
Heat in coal	91.9
Sensible heat in feed water	8.1
Sensible heat in air and coal	.0
	100.0
Output:	
Heat in steam	71.2
Heat loss in cinder	3.9
Heat loss in stack gas	
1. Sensible heat in dry gas	14.4
2. Potential heat in CO	2.3
3. Latent heat in water vapor	4.3
Radiation and unaccounted-for losses	3.9
	100.0

The above heat balance is the result of a test on a large modern boiler in an industrial plant. No economizers were used, so that the efficiency might be increased easily to 75 per cent with a proper heat-recovery system. With oil firing the efficiency might be maintained as high as 80 per cent, while with hand firing of coal the efficiency might be as low as 55 per cent. The distribution of losses is the important factor to be considered. The most obvious possibility for reducing stack loss is in decreasing the loss of undeveloped heat in CO. Undoubtedly, this could be reduced considerably with no great increase in the loss due to sensible heat. An increase in efficiency could also be obtained by reduction in stack temperature with an economizer. Whether or not this is warranted depends on an economic balance of the factors involved. No appreciable increase in efficiency could be made in this case by decreasing the cinder loss, since it is low at present.

Gas-fired Boilers.—The efficiency of a gas-fired boiler should be greater than that of a coal-fired boiler, since the ash loss is eliminated. On the other hand, with coal all the air used for primary combustion in the fuel bed is rapidly consumed, raising the fuel to a high temperature and thus a large amount of heat is transmitted by radiation. With gas, however, only a relatively small proportion of the total heat is transmitted by radiation. Also the gas-fired boiler requires very complete mixing of gas and air. With low B.t.u. gas, such as blast-furnace and producer gas, a relatively large amount of excess air is required and high stack temperatures are the rule.

The efficiency of a properly regulated boiler fired with a high B.t.u. gas may be as high as 84 per cent and is usually above 65 per cent. When a low B.t.u. gas is used, the efficiency may vary from 50 to 70 per cent, depending on the quality of mixing, the use of checkers in the furnaces, etc.

The latent-heat loss in a gas-fired furnace depends on the hydrogen content of the fuel and varies from 1.0 per cent with blast-furnace gas to 11.0 per cent with coke-oven or coal gas. This loss cannot be diminished for a given gas. The total stack loss may vary from 15 to 30 per cent, of which 10 to 20 per cent is sensible-heat loss, 0 to 1.0 per cent undeveloped-heat loss, and 1 to 11 per cent latent-heat loss. Decrease in excess air to the point which gives maximum efficiency, and promotion of proper mixing, are factors which increase efficiency. With low B.t.u.

gases the introduction of brickwork to promote mixing and aid surface combustion may also be advantageous.

Oil-fired Boilers.—The oil-fired boiler develops the same efficiencies and the same type of losses as indicated for the high B.t.u. gas-fired boiler. In addition, the heat or power necessary for atomization of the oil must be considered as a heat loss so that the normal maximum efficiency is about 80 per cent. The steam for atomization should not be over 2 or 3 per cent, but if regulation is not maintained it may increase up to 5 or 7 per cent.

Heating Furnaces.—In general, the efficiency of furnaces for the heating of billets and for heat-treating operations is quite low, and varies from 15 per cent in coal-fired, non-continuous furnaces to 60 per cent in gas-fired, continuous furnaces with recuperators. In considering the effect of various factors on the efficiency of a furnace of this type, the question of heat-transfer rate must be borne in mind, as outlined in Appendix III. Since radiation losses are high, a decrease in time of heating means a large saving in fuel used. The heat losses from a non-continuous furnace, and from a continuous furnace with a recuperator, are outlined in the following table:

TABLE CX

HEAT LOSSES FROM HEATING FURNACES

	Non-continuous,[7] per cent	Continuous with recuperator,[8] per cent
Input:		
Heat in gas	100.0	100.0
Output:		
Heat in steel	25.0	52.5
Stack losses:		
Sensible heat	40.0	29.3
Potential heat	0.0	0.0
Radiation and other losses	35.0	18.2

The major differences between these types of furnace is in the stack loss and in the radiation losses. In the non-continuous furnace much more heat is lost up the stack than in the continuous furnace. These two tests represent nearly the maximum variation in heating furnaces. A slightly higher efficiency is possible in the continuous furnace with better recuperation.

Certain operations, where the material must be heated slowly, where the supply of material is not continuous, and the furnace is used only occasionally, necessitate the use of a non-continuous furnace. The heat loss due to keeping a continuous furnace at high temperature might, in such cases, be so great as to make the increase in efficiency during operation of no advantage.

Excess air in most heating furnaces is kept at a very low figure. In fact, operators of heating furnaces usually keep a somewhat reducing atmosphere in order to prevent excessive oxidation and scaling of the stack in the furnace. A very luminous flame, produced in this manner, greatly increases the rate of heating in the furnace, but too large a deficiency of air, with lower flame temperature, will cause excessive stack loss as undeveloped heat and as sensible heat. It is essential for maximum economy that the amount of air used be very close to that theoretically required, with a very slight deficiency and not over 1 per cent CO at the point where the gases leave the furnace. When recuperators or regenerators are used, the unburned CO combines with the oxygen from the leakage around the end of the furnaces and in the heat-recovery system, so that determination of the stack loss beyond the end of the furnace will show no undeveloped heat loss. On the other hand, the heat thus generated could be utilized much more efficiently by combustion within the furnace itself. For a study of actual combustion conditions aside from the heat-balance determination, analyses of the gases leaving the furnace are necessary, and the CO or excess air at this point should be kept at a minimum. This is shown by the data given in Table CXI.

TABLE CXI

DATA FROM REHEATING FURNACE

	End of furnace, test No.			After recuperators, test No.		
	1	2	3	1	2	3
Per cent CO_2	7.8	9.2	9.5	4.7	4.8	4.8
Per cent O_2	0.3	0.6	2.5	11.7	11.7	11.7
Per cent CO	4.0	1.6	0.0	0.2	0.1	0.0

Although the CO content of the gases in the furnace varies from 0.0 to 4.0 per cent, the analysis of the gases leaving the recuperators shows practically no CO in each case. Improving conditions on the basis of the stack-gas analysis alone in this case would be impossible. Conditions in the furnace must be determined from the analysis of gases in the furnace. The efficiency in the furnace as operated to give the above results was approximately 20 per cent, while with better control of the air and gas it was later increased to over 30 per cent.

The major reason for the high stack loss encountered in heating furnaces is the high operating temperature employed. Even under the best conditions it is seldom possible to have the gases leave the furnace below 1000°F. For this reason, heating furnaces offer an ideal location for heat-recovery systems. Recuperators or regenerators should be used with all continuous or non-continuous retorts. Calorized iron recuperators give excellent results under these conditions and are frequently used.

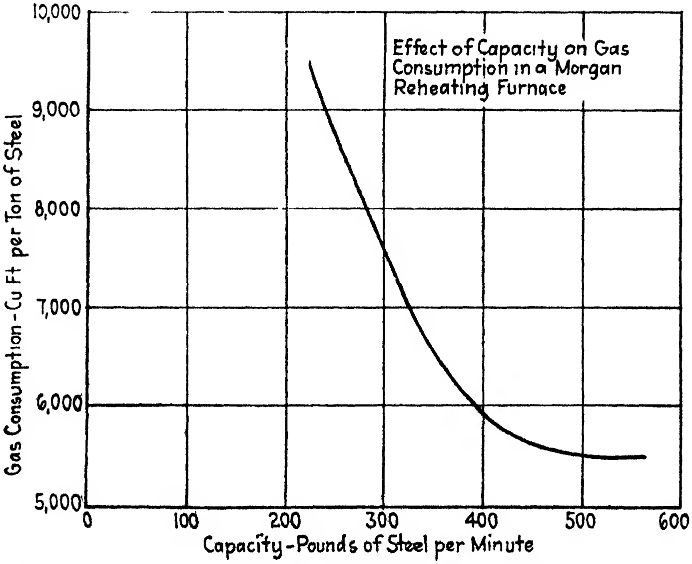

FIG. 207.—Effect of capacity on gas consumption in a Morgan reheating furnace.

Although great effort should be made to maintain the stack and radiation losses at a minimum value, even greater efforts should be made to reduce the time of heating by increasing the rate of heat transfer. Increasing the flame temperature by regulation of air supply, by better distribution of the hot gases, and by preheating the air is the most satisfactory method of accomplishing this. Regulation of air, use of recuperators, and proper insulation have twofold effects in reducing the percentage of losses, and in reducing the time necessary for heating.

Increased capacity in a heating furnace usually causes a reduction in fuel consumption per unit of material heated, primarily because the radiation and cooling-water losses are almost independent of the rate of heating and therefore their percentage decreases with increased capacity. Figure 207 shows the effect of increased capacity on the gas consumption in a Morgan reheating furnace, where the gas* consumption is decreased from 8750 to 5500 cu. ft. per ton as the capacity increases from 200 to 500 lb. of steel per min. Every furnace has a characteristic curve of this general type, and for efficient operation the capacity should be maintained at a point where the fuel consumption is a minimum, *i.e.*, the furnace should operate at high capacity. The economy of the furnace may be increased still further by lowering the characteristic curve, and such improvement often can be made by slight changes in furnace design or methods of control on fuel and air.

Open-hearth Furnaces.—The distribution of heat in a producer-gas-fired open-hearth furnace is given[4] in Table CXII.

TABLE CXII

HEAT BALANCE ON OPEN-HEARTH FURNACE

	PER CENT
Input:	
Heat in metal	5.0
Heat in producer gas	95.0
	100.0
Output:	
Heat in steel	21.6
Heat in slag	3.1
Heat absorbed by reactions	−7.9
Heat in flue gas	62.0
Radiation and unaccounted-for losses	21.2
	100.0

This heat balance may be rearranged, so that the input is the heat in producer gas only, as follows:

* Coke-oven gas, 560 B.t.u. (gross) per cu. ft.

TABLE CXIII

Heat Balance on Open-hearth Furnace

	Per Cent
Input:	
Heat in producer gas	100 0
Output:	
Heat picked up by steel	17.5
Heat in slag	3 2
Heat absorbed by reactions	−8.2
Heat in flue gases	65 1
Radiation and other losses	22.4
	100 0

The heat efficiency in the example cited, which includes heat in steel and slag plus heat necessary for reactions, is thus 21.6 per cent + 3.1 per cent − 7.9 per cent = 16.8 per cent. The results of tests by various investigators indicate heat efficiencies of 17 to 24 per cent on the same basis as above. Heat losses in stack gases vary from 30 to 55 per cent while radiation and unaccounted-for losses vary from 20 to 50 per cent. An average open-hearth heat balance based on the producer gas used may be taken as:

	Per Cent
Useful heat	20
Stack loss	40
Radiation and other losses	40

The major loss which offers possibilities of improvement is the stack loss. All open-hearth furnaces operate with regenerators, but even so the stack-gas temperature is usually in the neighborhood of 1500°F. Most modern plants install waste-heat boilers for the recovery of this heat, cooling the gases to from 400 to 500°F. with considerable saving. The boilers must be properly designed and carefully controlled or a large part of the saving may be lost.

Draft regulation is a very important feature of open-hearth operation. For best results the pressure in the furnace must be as near atmospheric as possible (though maintained slightly positive), and leakage in or out minimized. Regulation of air and gas to give the proper mixture under the extremely varied load conditions is important. This is the source of large stack losses in many plants. Although the gas rate may be changed, the air used remains constant. For best results, both should be regulated together.

Blast Furnaces.—The following table[4] shows the distribution of heat in a typical American blast furnace.

TABLE CXIV

HEAT BALANCE ON BLAST FURNACE

	Per Cent
Input:	
Heat in air	6 9
Heat in coke	93 1
Sensible heat in charge	0 0
	100 0
Output:	
Reactions in furnace	28 4
Heat in metal and slag	8 2
Heat in gas:	
Sensible heat	6 1
Potential heat	55 0
Cooling water	2 0
Radiation (measured)	0 3
	100 0

The blast furnace under present conditions is an extremely efficient unit. In practically all cases the undeveloped heat in the gas is used to good advantage and little thermal improvement may be made in the other minor losses. The important consideration is development of high temperatures at the bosh to increase melting capacity. This is obtained by high temperature and low humidity of the air used for combustion.

Cement and Lime Kilns.—Table CXV illustrates the distribution[9] of heat in a rotary kiln using the dry process for sintering cement.

TABLE CXV

HEAT BALANCE ON KILN FOR SINTERING CEMENT

(Powdered-coal-fired)

	Per Cent
Input:	
Heat in coal	100.0
Output:	
Heat in clinker	14 1
Heat in stack gases	48 3
Heat of decomposition	22 4
Radiation and other losses	15 2
	100.0

The results of tests on rotary lime kilns[4] are given in Table CXVI.

TABLE CXVI

HEAT BALANCE ON ROTARY LIME KILN

(Powdered-coal-fired)

	No. 1, Per Cent	No. 2, Per Cent
Input:		
Heat in coal	100.0	100.0
Output:		
Heat in product	11.2	12.3
Stack loss	43.7	41.2
Decomposition of stone ($CaCO_3$)	25.0	28.4
Radiation and convection (measured)	8.7	7.9
Unaccounted-for losses	11.4	10 2
Total	100.0	100.0

In rotary kilns for both purposes, the major loss is in the waste gases, which carry away from 40 to 50 per cent of the total heat in the fuel leaving the kiln. Radiation and other losses are 15 to 20 per cent and a large proportion of the unaccounted-for loss may be due to inaccurate measurement of the temperatures of the exit gases.

Results of a test on producer-gas-fired shaft lime kilns are given[4] in Table CXVII.

TABLE CXVII

HEAT BALANCE ON SHAFT LIME KILN

(Producer-gas-fired)

	Per Cent
Input:	
Heat in coal	95.9
Heat in steam	4.1
	100.0
Output:	
Heat in product	12.0
Heat of decomposition ($CaCO_3$)	35.0
Stack loss	30.6
Radiation and convection (measured)	12.4
Unaccounted-for losses	10.0
	100.0

The stack loss in the shaft kiln is lower than in the rotary type due to more efficient preheating of the entering stone. The heat used for decomposition of the stone is the useful heat in this process, and varies from 20 to 48 per cent. Heat in the product (10 to 15 per cent of the heat input) may be, and often is, recovered and used.

As in furnaces where a process must be carried out at high temperature, the effect of excess air on the efficiency is very great. The temperature at which the decomposition of stone takes place is above 1400°F. for a high calcium stone. The heat available for the decomposition of stone is, therefore, only that in the gases from combustion above 1400°F. If the temperature of the gases, with theoretical air, is 3000°F., then about one-half the heat is available for decomposition, while if the temperature is lowered to 2100°F. by the admission of 100 per cent excess air the heat available for decomposition is about one-fifth of the heat in the fuel, and the possible efficiency is only 40 per cent of its value when no excess air is used.

In order to operate at high efficiency, the available heat must be a maximum. The temperature of the combustion gases may be increased by preheating the air for combustion, making use of the heat in the product or the heat in the stack gases. The excess air, as stated above, must also be kept at a minimum. The kilns should be thoroughly heat insulated wherever the temperatures are above 1500° F. (for a high calcium stone).

Fuel Consumption of Furnaces.—The following table from the "Calco Handbook of Recuperation"[17] gives the fuel consumption of furnaces of various types, and serves as a basis for comparing the efficiencies of different furnaces making the same product.

TABLE CXVIII

FUEL CONSUMPTION OF TYPICAL FURNACES

Brick kilns:	B T U. PER 1000 BRICK
Hoffman continuous kiln	3,920,000–4,760,000
Ruebon kiln	5,880,000–9,240,000
Grates and trough	4,760,000–6,300,000
Staffordshire kiln	4,760,000–6,300,000
Buhrer continuous tunnel	3,220,000–3,900,000

TABLE CXVIII.—(*Continued*).

FUEL CONSUMPTION OF TYPICAL FURNACES

	B.T.U. PER TON
Metal melting:	
Iron and steel:	
Puddling	14,000,000
Open-hearth—natural gas	5,684,000–6,090,000
Open-hearth—producer gas	7,000,000–9,100,000
Open-hearth—fuel oil	6,500,000
Basic furnaces	10,500,000
Acid furnaces	10,920,000
Air furnace for malleable castings	8,600,000
Copper:	
Reverberatory	4,200,000
Tin:	
Smelting	19,200,000
Rolling-mill Furnaces:	
Busheling	5,600,000–7,000,000
Billet heating	2,240,000–2,520,000
Billet heating (4 by 4)	2,800,000–4,200,000
Continuous furnace (cold billets)	1,500,000–2,250,000
Continuous bloom furnace	1,400,000–2,100,000
Reheating cold blooms	1,960,000–2,800,000
Soaking pits (hot ingots)	1,400,000–1,760,000
Tire furnaces	4,500,000
Wheel furnaces	8,400,000
Sheet annealing	2,800,000
Tin-plate mill	2,500,000
Rivet making:	
Rivet-making furnaces	1,260,000
Galvanizing:	
Galvanizing pots	140,000
Forging:	
Drop forging or bolt heating	6,000,000–7,000,000
Ingot forging	4,000,000–5,000,000
Ingot forging (regeneration furnace)	3,500,000–4,000,000
Brass annealing:	
Brass annealing	1,500,000–2,000,000
Heat treatment:	
Case hardening	4,500,000–5,000,000
Quenching	2,000,000–3,000,000
Tempering	1,500,000–2,000,000

The major source of heat loss in most *final furnaces* lies in the stack gases. Improper regulation of air and fuel is the frequent cause. Where high temperatures prevail heat-recovery systems should be considered, and where economic conditions warrant they should be used. For a complete understanding of the losses involved in operation a heat balance on the furnace must be made, but for rapid improvement determination of stack loss often will suffice. This is the major loss in most cases and the method of reducing it may be determined readily from analysis of the data. Reduction of excess air to give the minimum loss and regulation of the air and fuel so that they give the *optimum ratio* under *all conditions* of operation will usually materially increase the efficiency of the furnace.

BIBLIOGRAPHY

1. R. T. Haslam and E. L. Chappell: *J. Ind. Eng. Chem.*, **17**, 402 (1925).
2. Marks: "Mechanical Engineers' Handbook," McGraw-Hill Book Co., Inc. (1916).
3. D. W. Wilson: "Standard Carbonizing Plant Test Code," *Rep.* of Committee on Carbonization of Coal and Complete Gasification of Coal, 1924 Convention Am. Gas. Assoc., Reprint (p. 13).
4. Mass. Inst. of Tech. School of Chemical Engineering Practice, *Reps*., unpublished.
5. Seventh *Rep.* of Gas Investigation Committee of British Inst. of Gas Engineers. Gas J. **158,** 800–827 (1920).
6. E. L. Hall and S. H. Graff: *Am. Gas. J.*, **115,** 318 (1921).
7. A. Hallback: *Fuels and Furnaces*, **2,** 119 (1924).
8. W. P. Chandler, Jr.: *Iron Trade Rev.*, **71,** 1196 (1922).
9. R. C. Carpenter: *J.* A. S. M. E., **36,** 337 (1914).
10. R. K. Hursh: *J.* Am. Cer. Soc., **1,** 576 (1918).
11. C. B. Harrop: *Trans.* Am. Cer. Soc., **19,** 216 (1917).
12. Z. R. Clauss: *Tonind.-Ztg.*, **33,** 1081 (1909).
13. S. W. Parr: "Fuel, Gas, Water and Lubricants," McGraw-Hill Book Co., Inc. (1922).
14. "Pyrometric Practice," U. S. Bur. Standards, *Bull.* 170.
15. Landolt-Bornstein-Roth Tabellen (1923 ed.).
16. J. W. Richards: "Metallurgical Calculations," McGraw-Hill Book Co., Inc. (1918).
17. "Calco Handbook of Recuperation," Calorizing Co.

CHAPTER XVIII

PRODUCER GAS

Producer gas is generated by blowing a mixture of air and steam up through a thick, hot bed of coal or coke. Essentially, this process results in an oxygen-free gas containing a large proportion of the original heating value of the solid fuel in the form of CO and H_2. The production of producer gas is shown diagrammatically in Fig. 208. The coal or coke is fed into a brick-lined producer chamber, and the producer gas, consisting of CO

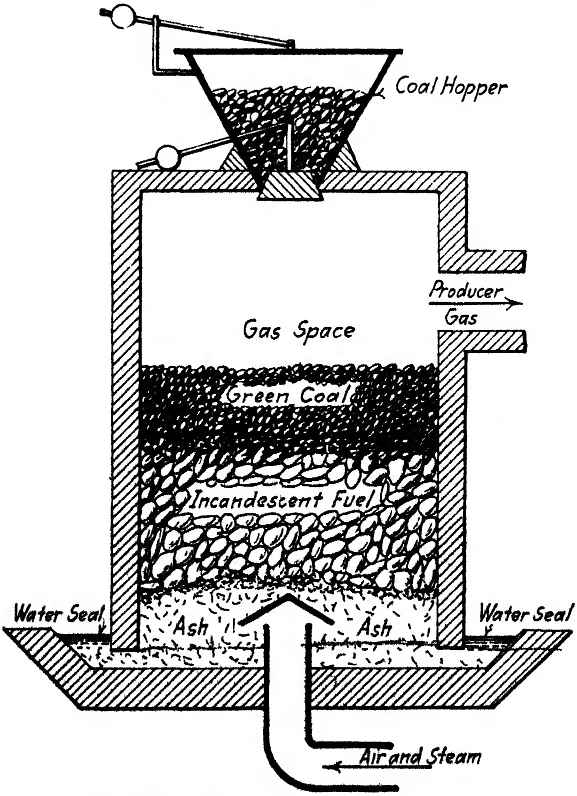

Fig 208.—Gas producer—diagrammatic.

and H_2 together with the volatile matter, tar, soot, (and nitrogen from the entering air), passes out through a discharge main near the top. Producer gas is the cheapest form of industrial gas that can be manufactured. It possesses, in general, all the advantages of gaseous fuels (see p. 260), and, in consequence, it finds wide application in industrial heating.

Historical Outline of Gas-producer Development.—The gases leaving the top of any deep fuel bed resemble producer gas in composition and thus a fuel similar to producer gas has been in use practically ever since coal was first burned on a grate. The first utilization of the true producer principle, however, *i.e.*, having the secondary combustion of the carbon monoxide and hydrogen take place *away* from the top of the fuel bed, must be credited to Faur, who built a gas producer in 1832. Faur's discovery did not bring about any great expansion in gas-producer application, however, until the Siemens brothers further developed producers in connection with their invention of the open-hearth furnace in 1861 The Siemens producer was really a thick fuel-bed, coal-fired furnace with the hearth or crucible separated from the fuel bed That is, the grate was not in immediate contact with the furnace hearth, and the secondary air necessary for the combustion of the gases coming off from the top of the fuel bed was admitted at the hearth rather than immediately over the fuel bed.

Field of Producer-gas Application.—Producer gas is used mainly either where a gaseous fuel is the most economical source of heat or where coal cannot be used and a cheap gas is necessary. In other words, producer gas is used either for economy or in order to improve the quality of product. Indirect or gaseous combustion may increase the economy of production: (1) when fuel is to be used in many scattered furnaces, (2) when power is to be generated in isolated plants, and (3) when high furnace temperatures are required and it is advisable to recover heat by regeneration or recuperation. As previously pointed out in Chap. XI, when utilizing recuperation or regeneration the hot gases going out the stack give up a part of their heat to the incoming air, and in some cases to the producer gas, thereby increasing the furnace efficiency. The higher the temperature of the furnace (and therefore of the hot stack gases) the more desirable it is to use regeneration or recuperation.

The purposes for which gaseous combustion is advantageous as regards quality of product are: (1) when the temperature must be controlled closely, (2) when the atmosphere of the furnace must be controlled, *i.e.*, be kept oxidizing, reducing, or neutral; (3) when long flame combustion is necessary, as, for example, in the open-hearth furnace; and (4) when the ash or sulfur in a solid fuel would spoil the product.

Present Status of Producer Gas.—Although no accurate data on the number of producers in operation are available, it is estimated that there are about 11,000 in the United States The distribution of these producers in the various industries has been estimated[1] to be as shown in Table CXIX.

TABLE CXIX

ESTIMATED DISTRIBUTION OF GAS PRODUCERS IN THE UNITED STATES

INDUSTRY	NUMBER OF PRODUCERS
Steel	6,500
Glass	1,500
Ceramics and lime burning	1,500
Power generation	1,000
Metallurgical and other chemical fields	500
Total	11,000

No figures are available as to the amount of coal gasified in producers, but it is probable that in the United States about 15,000-000 tons of coal are gasified each year by the steel industry alone.

Fundamental Reactions of the Siemens or Air-blown Producer. Although the term "producer gas" generally is used in the United States to define a gas made by blowing a mixture of steam and air up through a thick fuel bed, the term has been used in times past (and sometimes still is so used in other countries) to include the gas produced when *air alone* is blown up through a thick fuel bed. Such gas is also called "Siemens gas" or "air-blown producer gas." Although no producer gas is made in the United States without mixing steam with the air, still, for the sake of clarity, the chemistry of producer-gas formation will be taken up by first discussing the reactions for Siemens or air-blown producer-gas formation from carbon; the reasons for, and the effects of, adding steam to the air blast will be touched upon subsequently.

As brought out previously (p. 336), when a blast of air is forced up through an incandescent fuel bed, the oxygen rapidly burns the hot carbon, forming CO_2, which, in turn, is in part reduced to CO. The reactions are as follows:

$$C + O_2 + 3.76N_2 = CO_2 + 3.76N_2 + 174{,}600 \text{ B.t.u.} \qquad (1)$$
$$C + CO_2 + 3.76N_2 = 2CO + 3.76N_2 - 70{,}200 \text{ B.t.u.} \qquad (2)$$

The thicker the fuel bed the greater is the extent to which reaction (2) proceeds, the resulting gas tending to approach* the composition indicated by the following reaction:

$$\underset{\text{fuel}}{2C} + \underbrace{O_2 + 3.76N_2}_{\text{air}} = \underbrace{2CO + 3.76N_2}_{\text{Siemens gas}} + \underset{\text{sensible heat}}{104{,}400 \text{ B.t.u.}} \qquad (3)$$

which, however, does not represent what *actually* takes place, but merely expresses the overall effect. When this gas is fired in the furnace, additional (secondary) air is added, the combustion taking place and developing additional heat in accordance with the reaction

$$\underbrace{(2CO + 3.76N_2)}_{\text{Siemens gas}} + \underbrace{(O_2 + 3.76N_2)}_{\text{secondary air}} = \underbrace{(2CO_2 + 7.52N_2)}_{\text{stack gas}} + \underset{\text{heat}}{244{,}800 \text{ B.t.u.}} \qquad (4)$$

The nitrogen of the air has been included in these equations to show its diluent effect in producer gas, and to illustrate why the French designate producer gas as "pauvre gas." The composition of the best possible or ideal Siemens gas (using coke as fuel) is shown in Table CXX.

TABLE CXX

IDEAL COMPOSITION OF AIR-BLOWN PRODUCER GAS

Constituent	Per cent by volume	B.t.u. per cu. ft. of producer gas
Carbon monoxide	34.7	110 (317 × 0.347)
Nitrogen	65.3	0
Total	100.0	110 (60°F., sat. with H_2O)

Some of the heat liberated by reaction (1) appears as sensible heat in the Siemens gas ($2CO + 3.76N_2$) and the balance is lost by radiation from the producer, or as heat in the cinders, etc. If it is assumed that the temperature of the Siemens gas leaving the

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* In the discussion immediately following the complete reduction of CO_2 is assumed.

producer is 1600°F. (871°C.), then the total sensible heat in the gases is found by multiplying the number of mols of each gas by the heat content of each mol as read from the total heat chart (Fig. 49, p. 210);

$$\text{Sensible Heat in Gases} = (2 + 3.76)\ 11{,}400 = 65{,}700 \text{ B.t.u.}$$

The potential heat of the Siemens gas (that is, the heat that is available through the further combustion of the CO) is shown by reaction (4) to be 244,800 B.t.u. From these data it is possible to set up the approximate heat balance of an ideal Siemens producer as shown in Table CXXI.

TABLE CXXI

Approximate Heat Balance of an Air-blown Producer

	B.t.u.	Per cent
Heat input:		
From the complete combustion of 2 mols of carbon	349,200	100
Heat output:		
Potential heat in the CO	244,800	70.1
Sensible heat in the Siemens gas	65,700	18.8
Radiation, loss in cinders, unaccounted-for losses, etc. (by difference)	38,700	11.1
Total	349,200	100.0

The efficiency of a gas producer is expressed in either of two ways, depending on whether the sensible heat of the gas can be utilized or not. If the sensible heat can be applied usefully, then the efficiency used is called the *hot efficiency*, which includes as useful heat both the potential and the sensible heat in the gas, and, as illustrated by the above heat balance, is:

$$\text{Hot Efficiency} = \frac{(\text{Potential Heat} + \text{Sensible Heat})}{(\text{Total Heat})} \times 100 = \frac{(244{,}800 + 65{,}700)}{349{,}200} \times 100 = 88.9 \text{ per cent.}$$

In many cases, however, the sensible heat of the gas cannot be utilized completely or efficiently. For example, if the gas is to be cleaned and scrubbed for the removal of tar or soot, the sensible heat is lost in the scrubbing water; again, if gas is to be transmitted through long pipe lines, or if it is to be used in regenerative or recuperative furnaces, much of the sensible heat of the producer gas must be considered as a loss. Under these circumstances the efficiency required is known as the *cold* efficiency, which includes as useful heat only the potential heat of the gas, and, as illustrated from the heat balance of the Siemens producer given in Table CXXI, is as follows:

$$\text{Cold Efficiency} = \left(\frac{\text{Potential Heat}}{\text{Total Heat}}\right) \times 100 = \frac{244{,}800 \times 100}{349{,}200} = 70.1 \text{ per cent.}$$

Troubles with the Siemens or Air-blown Producer.—There are three main troubles with the Siemens or air-blown producer:

1. The gas has a low heating value (110 B.t u. per cu. ft).

2. The cold gas efficiency is low (70.1 per cent is the theoretical maximum) and a large proportion of the heat in the coal is left as sensible heat in the gas

3. Continuous, uniform operation of the producer is impossible due to excessive clinker formation. Clinkering results from the large amount of heat generated in a very small space by the reaction.

$$C + O_2 = CO_2 + 174{,}600 \text{ B.t.u.}$$

The ash in the coal fuses and forms large clinkers which cause blowholes in the fuel bed and stoppage of the producer.

These troubles may be largely overcome, and the cold efficiency of the process increased, by absorbing heat from the fuel and using this absorbed heat for converting an inert gas into a combustible one. Two reactions which absorb heat and which at the same time yield a combustible gas are available for this purpose:

(*a*) The return of flue gases, utilizing the reaction.

$$\underbrace{(CO_2 + 3\ 76N_2)}_{\text{flue gas}} + C = 2CO + 3\ 76N_2 - 70{,}200 \text{ B.t.u} \qquad (1)$$

(*b*) By the use of steam, according to the reaction.

$$C + H_2O = CO + H_2 - 70{,}900 \text{ B.t.u.} \qquad (2)$$

Reaction (1) is not used, because the nitrogen not only dilutes the producer gas, but also carries heat away from the producer and up the furnace stack as sensible heat. Reaction (2), however, introduces no diluent gas and yields combustible CO and H_2. Furthermore, steam is cheap and usually is available in most industrial plants without an increase in investment.

Steam as an Aid to Producer Efficiency.—The low B.t.u. content of Siemens gas is due to the fact that 3.76 nitrogen molecules enter the producer with each oxygen molecule. The decomposition of steam, on the other hand, introduces no diluent gas and is the only commercially possible reaction that yields nothing but combustible gases. The mixture of H_2 and CO formed by this interaction of steam and hot carbon, known as "water gas," has a heating value of about 300 B.t.u. per cu. ft. as compared with 110 B.t.u. per cu. ft. for perfect Siemens gas. The admixture of this 300 B.t.u. gas and the hydrocarbons from the volatile matter of coal, to the 110 B.t.u. Siemens gas raises the heating value so that a gas having 130 to 180 B.t.u. per cu. ft. is obtained from commercial gas producers.

The decomposition of steam by hot carbon is endothermic (absorbs heat), and therefore the reaction converts some of the sensible heat available in the gas producer into chemical or potential heat, which may be utilized in the furnace. In addition to thus raising the cold efficiency of the gas producer, the absorption of heat has an even more important effect on producer operation—namely, in preventing the formation of clinker. The use of steam, as will be shown later, reduces the temperature in both the oxidizing zone and the reducing zones of the producer and thereby allows the ash to pass through the fire without being heated much above its softening point.

Reactions in the Fuel Bed of a Gas Producer.—From the preceding discussion it can be seen that the formation of producer gas, as the term is generally used in the United States, is the result of the simultaneous action of primary air on carbon and the interaction of steam and carbon. The chemistry of the reactions involved (their equilibria and velocities) has been discussed in Chap. VIII and the combustion of carbon by primary air has been described in Chap XII. The reduction of CO_2 and the decomposition of steam by hot carbon have been but briefly treated, however, and a more extended discussion of these reactions as affecting producer operation is necessary.

Reaction $CO_2 + C = 2CO - 70{,}200$ *B.t.u.*—It was shown on page 154 that in equilibrium mixtures only small amounts of CO_2 can exist in contact with carbon at temperatures above 1700°F. (927°C.). The rate at which the reduction of the dioxide progresses has also been shown to increase with temperature and to be affected profoundly by the type and amount of carbon surface as well as the time of contact between gas and hot carbon. The effects of temperature and of time of contact

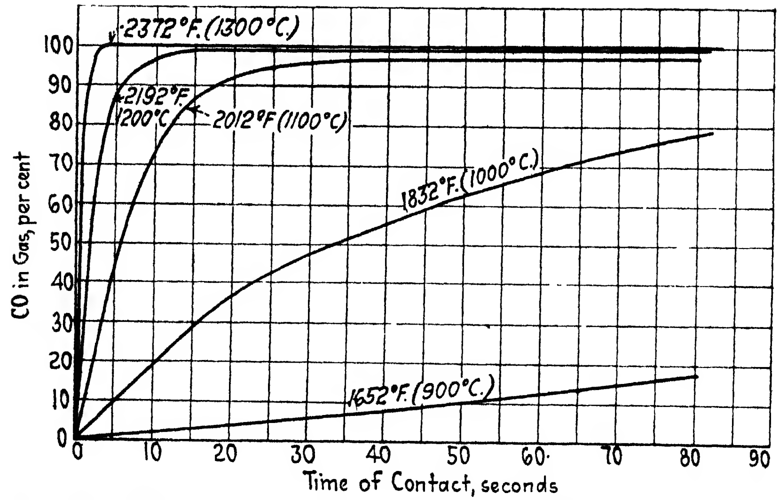

FIG. 209.—Effects of temperature and time of contact on the production of CO from coke and CO_2.

CO FORMED FROM COKE AND 21% CO_2 + 79% N_2 WITH VARIOUS TIMES OF CONTACT

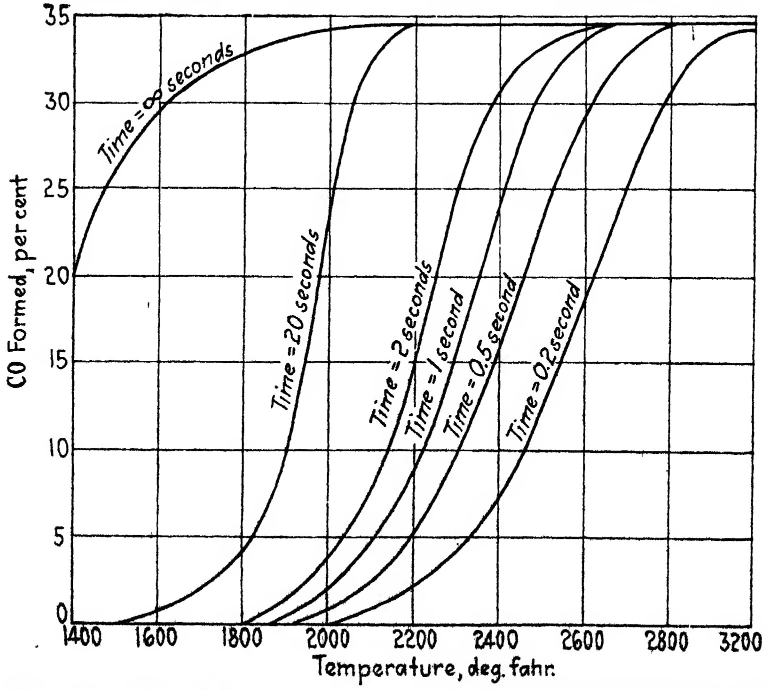

FIG. 210.—Effect of time of contact and temperature on the production of CO from coke and air.

as determined by Clement, Adams, and Haskins[2] are strikingly illustrated* by Figs. 209 and 210. Figure 209 shows the per cent of CO in the resultant gas when CO_2 was kept in contact with coke for varying periods of time at temperatures ranging from 1650 to 2370°F. (900 to 1300°C.). This plot clearly brings out the fact that to obtain a gas of given CO content less time of contact will be required at the higher temperature, or, to obtain the best reduction in a given time, the temperature should be maintained as high as possible. These points are further illustrated by Fig. 210, in which are plotted the results obtained by the Bureau of Mines investigators by passing a gas containing 21 per cent CO_2 and 79 per cent N_2 over hot coke. The ordinates show the per cent of CO formed at temperatures from 1400 to 2700°F. (760 to 1482°C.), each curve representing a definite time of contact between gas and coke. In most gas producers the time of contact is about 1 sec.

The above results show that, in order to obtain a gas high in CO, the time of contact and the area of carbon surface should both be maintained as great as possible and the temperature of the producer run as high as practicable. This latter factor will be discussed later.

The Reduction of Water.—When steam is blown through a bed of incandescent carbon (as in the formation of water gas) the gaseous mixture obtained is considered to be formed by some or all of the following reactions:

$$C + H_2O = CO + H_2 - 70{,}900 \text{ B.t.u.} \tag{1}$$
$$C + 2H_2O = CO_2 + 2H_2 - 71{,}600 \text{ B.t.u.} \tag{2}$$
$$CO + H_2O = CO_2 + H_2 - 700 \text{ B.t.u.} \tag{3}$$
$$C + CO_2 = 2CO - 70{,}200 \text{ B.t.u.} \tag{4}$$

Clement and Adams[2] conclude that, under conditions in the producer, reactions (1) and (2) both take place, although (1) is dominant, accompanied to some extent by reaction (2), and that the effective direction of reactions (3) and (4) is to the right. The results of these investigators concerning the effect of temperature and time of contact on the decomposition of steam by coke are shown in Fig. 211. This figure shows that the extent of the reaction between steam and coke, like the reduction of CO_2, is greatly influenced by time of contact. This same figure

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* Figures plotted from the U. S. Bureau of Mines[2] data.

also shows the very great activity of charcoal as compared with coke.

H. S. Taylor[3] states that the steam is decomposed according to reaction (1), the CO_2 being formed by the establishment of the water-gas equilibrium through reaction (3). He further points out that reaction (3) is catalyzed by ash in the coke.

Gwosdz[4] concluded that even at low temperature (1100°F. or 593°C.) CO_2 is not formed by reaction (2), but that the CO_2 comes from reaction (3) speeded up catalytically by ash in the carbon. He arrived at this conclusion from the fact that with arc-lamp carbon containing 0.1 per cent ash the $CO:CO_2$ ratio is about 11:1, whereas with charcoal containing 1.4 per cent ash the $CO:CO_2$ ratio is about 1:5.

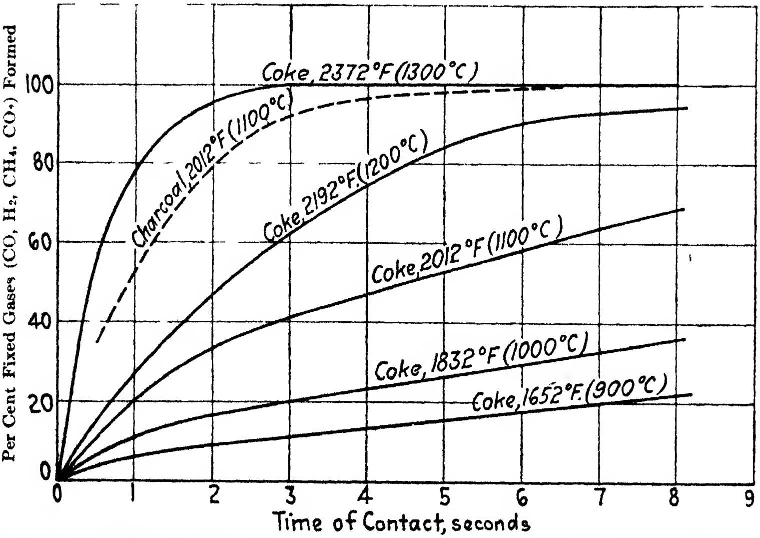

FIG. 211.—Effects of time of contact and temperature on the decomposition of steam by coke.

A more recent investigation,[5] undertaken to determine the mechanism of the reactions and their relative velocities when they are occurring simultaneously, has brought out additional information. It was found under the conditions of the experiments that if the composition of the reaction products be represented by curves such as shown[5] in Fig. 212, the same curves are obtained irrespective of the kind of carbon, the steam pressure, or the time of contact between steam and carbon. All these variables, including temperature, though not affecting the type of curve, have, as previously shown, a decided influence on the *extent* to which the decomposition of steam progresses in a given time. In Fig. 212 the analyses correspond to the volumes of H_2O, CO_2, CO, and H_2 produced from a given unit volume of entering water, plotted against the fraction of undecomposed water, $\frac{H_2O}{H_2O + H_2}$.

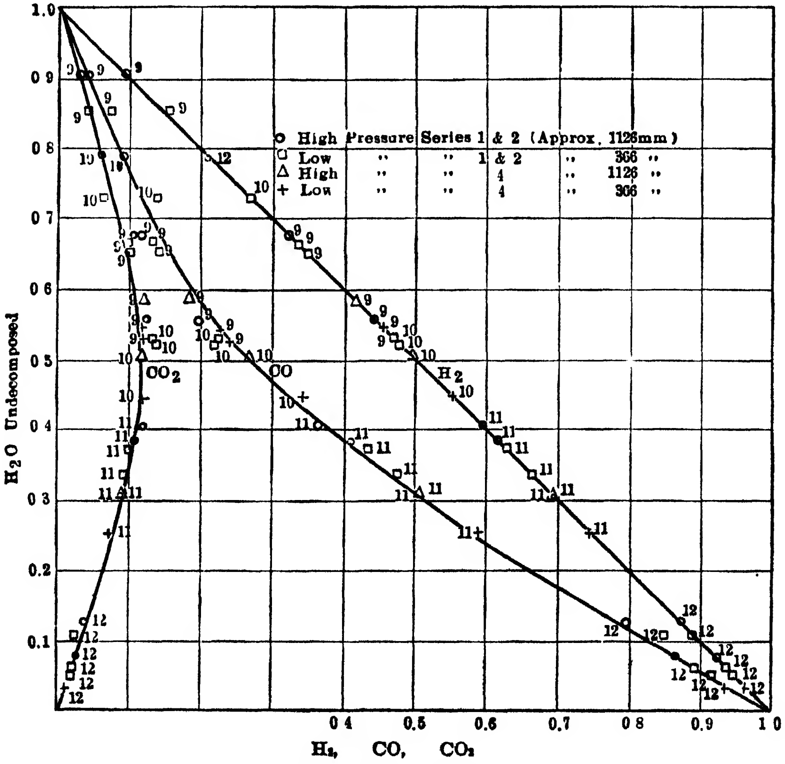

FIG. 212.—Volumes of CO_2, CO and H_2 formed per unit volume of entering steam. Numerals give the temperature in hundreds of °C.

As pointed out in Chap. VIII, this investigation showed that, considering the rate or velocity of reaction (2) as unity, at temperatures much below 1650°F. (900°C.) reaction (1) is practically non-existent and the velocity constant of reaction (4) is 2.0. At temperatures much above 1650°F. (900°C.) the velocity constant of reaction (1) is unity and that of reaction (4) is 2.18.

At temperatures above 1650°F. (900°C.) CO and CO_2 must both be initially formed products of the interaction of steam on carbon, since in Fig. 212 both the CO and the CO_2 curves have finite slopes at 0 per cent water decomposed (or 100 per cent H_2O undecomposed).

Plotting the $CO:CO_2$ ratio for temperatures 1652°F. (900°C.), 1832°F. (1000°C.), 2012°F. (1100°C.), and 2192°F. (1200°C.) against the fraction of water undecomposed (Fig. 213) it is seen[5] that the ratio is substantially constant until the undecomposed

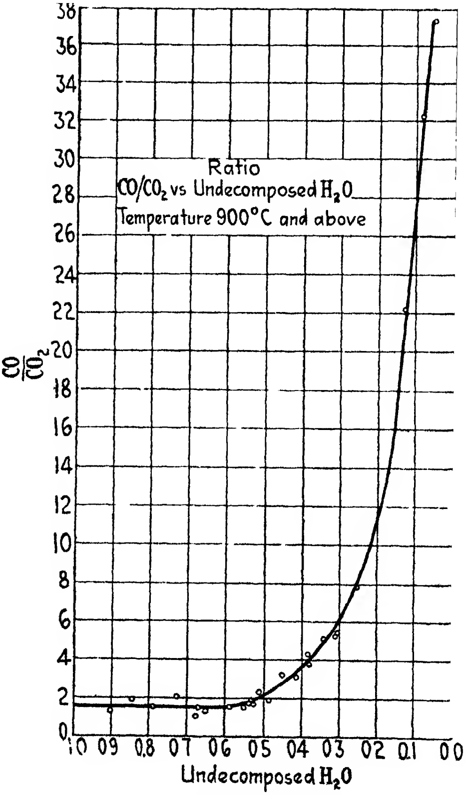

FIG. 213.—Ratio of $CO \cdot CO_2$ vs. undecomposed water in decomposition of steam by carbon.

water is below 60 per cent, at which point the $CO:CO_2$ ratio begins to rise rapidly. In other words, although temperature does exert a small effect, the influence of undecomposed steam is the controlling factor in determining the amount of CO_2 or the $CO:CO_2$ ratio.

The Mechanism of Producer-gas Formation.—Much additional information concerning the reactions which take place in producer-gas formation has been obtained by a study of the composition of the gases as they travel up through the producer bed. If gas samples are withdrawn from closely adjacent sampling points in the fuel bed of a coke-fired producer and the analyses plotted as a function of the distance above the grate, curves are obtained similar to those shown[6] in Fig. 214.

The entering air-steam mixture is unchanged in composition as it passes through the ashes, which merely serve to preheat the mixture. At the end of the ash zone (about 5 in. above the grate in the case shown) the mixture comes in contact with incandescent fuel and combustion begins.

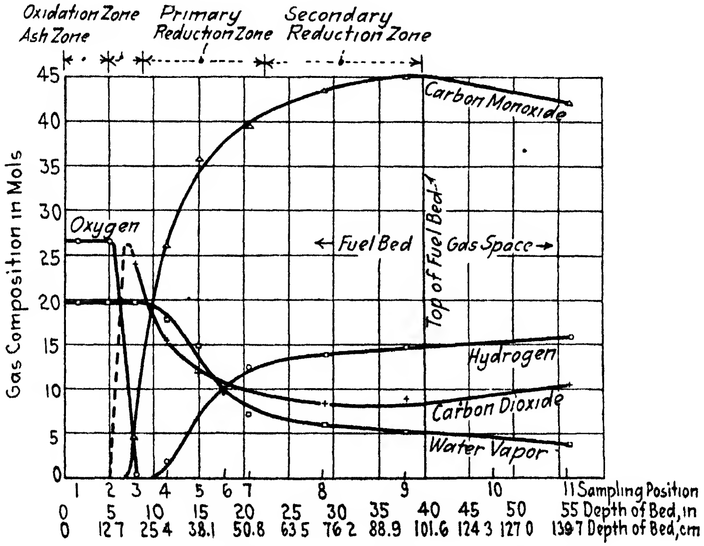

FIG. 214.—Composition of gases in the fuel bed of a gas producer. Basis, 100 mols of N_2. Fuel used, anthracite coal or coke, and therefore no distillation zone.

Just as in the direct combustion of coal on the grate (see p. 236), the oxygen is entirely consumed as the gases traverse the next 3 or 4 in. and the CO_2 content of the gas rises with corresponding rapidity. After this rapid initial increase, the CO_2 is at a maximum about 3 in. above the cinders, and from this point to the top of the fuel bed the CO_2 content gradually decreases.

CO formation does not take place appreciably until all the oxygen has disappeared and the CO_2 content has nearly reached its maximum. After the oxygen has disappeared, the monoxide is rapidly formed in the next 12 in. of fuel bed and from then on shows a gradual increase until the gases pass into the free space over the fuel bed.

As shown by Fig. 214, the water is unchanged until all the oxygen is consumed. Water decomposition starts shortly after the beginning of the CO formation, and the hydrogen content of the gas increases rapidly at first (through a distance of about 1 ft.). Following this rapid initial formation, the hydrogen content of the gas increases relatively slowly.

The actual combustion reactions by which the CO_2, CO, and H_2 are formed may be shown by the changing composition of the gases. On passing through the first 4 or 5 in. above the ashes the oxygen is all consumed and CO_2 only is formed. In this section of the fuel bed, which may be termed the *oxidation zone*, the only reaction is:

$$C + O_2 = CO_2 + 174{,}600 \text{ B.t.u.}$$

On account of the large amount of heat liberated this is the hottest zone in the producer.

As the hot gases pass out of the oxidation zone, CO_2 is reduced to CO, and H_2 is formed by the decomposition of steam. The layer of fuel above the oxidation zone, therefore, is called the *reduction zone,* and in it four reactions take place:

$$C + H_2O = CO + H_2 - 70{,}900 \text{ B.t.u.} \quad (1)$$
$$C + 2H_2O = CO_2 + 2H_2 - 71{,}600 \text{ B.t.u.} \quad (2)$$
$$C + CO_2 = 2CO - 70{,}200 \text{ B.t.u.} \quad (3)$$
$$CO + H_2O = CO_2 + H_2 - 700 \text{ B.t.u.} \quad (4)$$

An inspection of these reactions shows that with reactions (1) and (2) there may be an increase in the per cent of steam decomposed, $\frac{H_2}{(H_2O + H_2)}$, only by a corresponding increase in the amount of carbon gasified, $\frac{(CO + CO_2)}{N_2}$. If steam is being reduced to hydrogen by reaction (4), there will be no increase in $\frac{(CO + CO_2)}{N_2}$. Reaction (3) causes a large increase in the $\frac{CO}{CO_2}$ ratio, although the other reactions also affect this ratio to a slight extent. Figure 215 shows[6] a plot of these ratios versus distance up through the fuel bed.

All three of the ratios $\frac{(CO + CO_2)}{N_2}$, $\frac{CO}{CO_2}$ and $\frac{H_2}{(H_2 + H_2O)}$

increase from the top of the oxidation zone through the next 12 in. of fuel bed. The indications, therefore, are that reactions (1), (2), and (3)

$$C + H_2O = CO + H_2 - 70{,}900 \text{ B.t.u.} \qquad (1)$$

$$C + 2H_2O = CO_2 + 2H_2 - 71{,}600 \text{ B.t.u.} \qquad (2)$$

$$C + CO_2 = 2CO - 70{,}200 \text{ B.t.u.} \qquad (3)$$

are all taking place in the lower part of the reduction zone. This portion of the reduction zone therefore may be termed the *primary reduction zone,** *since all the reduction of steam by carbon* and most of the reduction of CO_2 by carbon takes place in this zone, which is fairly thin. In twelve experimental runs, in which the rate of gasification varied from 10 to 75 lb. per sq. ft. per. hr. and the total depth of fuel bed from 1½ to 4½ ft., the primary reduction zone was never over 14 in. in depth.

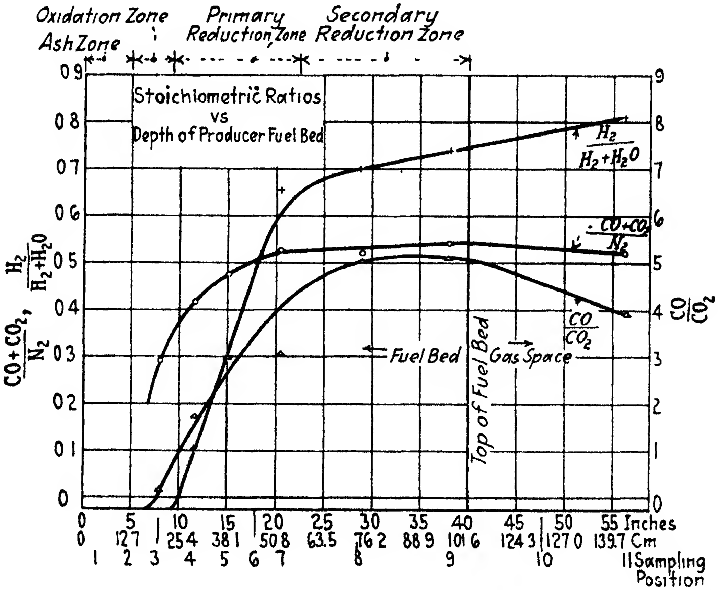

FIG. 215 —Gas ratios, useful in showing the course of reactions within a producer fuel bed.

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* The fact that this *primary reduction zone* is the only part of the producer fuel bed in which appreciable amounts of steam are decomposed by carbon must be borne in mind in applying the Bureau of Mines data (p. 549) to actual producer problems. Thus, for the data here quoted the time of contact between steam and carbon must be figured for 1 ft. of fuel, and the total depth of bed cannot be used, since the reactions take place *only* in the primary reduction zone.

Above the *primary reduction zone*, Fig. 215 shows that, while the ratio $\frac{CO + CO_2}{N_2}$ is substantially constant, the ratio $\frac{H_2}{H_2 + H_2O}$ continues to increase, showing that steam is still being decomposed. This indicates that any steam that is decomposed in this portion of the fuel bed is being decomposed by reaction (4). The increase in the ratio $\frac{CO}{CO_2}$ without a substantial increase in the ratio $\frac{CO + CO_2}{N_2}$ shows that reaction (3) also is taking place in this zone. These facts indicate that in this portion of the reduction zone only the two following reactions are taking place, and these to a relatively slight extent only:

$$C + CO_2 = 2CO - 70{,}200 \text{ B.t.u., and} \qquad (3)$$
$$CO + H_2O = CO_2 + H_2 - 700 \text{ B.t.u.} \qquad (4)$$

Since the reduction of H_2O by carbon does not proceed in this zone and since the total reduction is small, this portion may be designated as the *secondary reduction zone*.

The main function of the *secondary reduction zone* is to act as a heat interchanger, in which the hot gases give up some of their sensible heat to the incoming fuel This heat exchange raises the temperature of the primary reduction zone, thus increasing the amount of steam decomposed and aiding producer operation. Increasing the depth of the secondary reduction zone may increase the temperature within the fuel bed[6] by as much as 600°F.

Above the *secondary reduction zone* the coal is distilled and its volatile matter (hydrocarbons, tar, etc.) is added to the gas This addition is extremely important in producer operation, since the volatile matter from the best producer coals increases the heating value of the gas as much as 40 per cent. In American practice this zone of distillation is thin (usually less than 1 ft.).

Above the top of the fuel bed, as shown in Fig. 214, the gases change in composition, and their heating value decreases. In the free gas space above the top of the fuel bed, the gas is no longer in contact with hot carbon and it is possible for CO to revert to CO_2, according to the reaction:

$$2CO \rightarrow CO_2 + C. \qquad (3)$$

The extent to which this decomposition progresses depends on the speed of reaction (affected greatly by the temperature of the gases as they leave the top of the producer) and on the time which the gases are in the gas space over the fuel bed (determined by the gas velocity). Furthermore, a certain amount of CO_2 and undecomposed steam always are added to the gas at the top of the fuel bed, due to leakage around the edge of the fuel bed, and through blowholes and channels. These additions alter the gas composition considerably. This reversion was noticed first by Neuman,[7] who from actual gasification data found that the loss occasioned in this way may amount to between 10 and 27 per cent. Since this loss is increased by high gas temperatures (due to high velocity of the reaction $2CO = CO_2 + C$ at high temperatures) it may be reduced by using a deep fuel bed, therefore lowering the gas temperature.

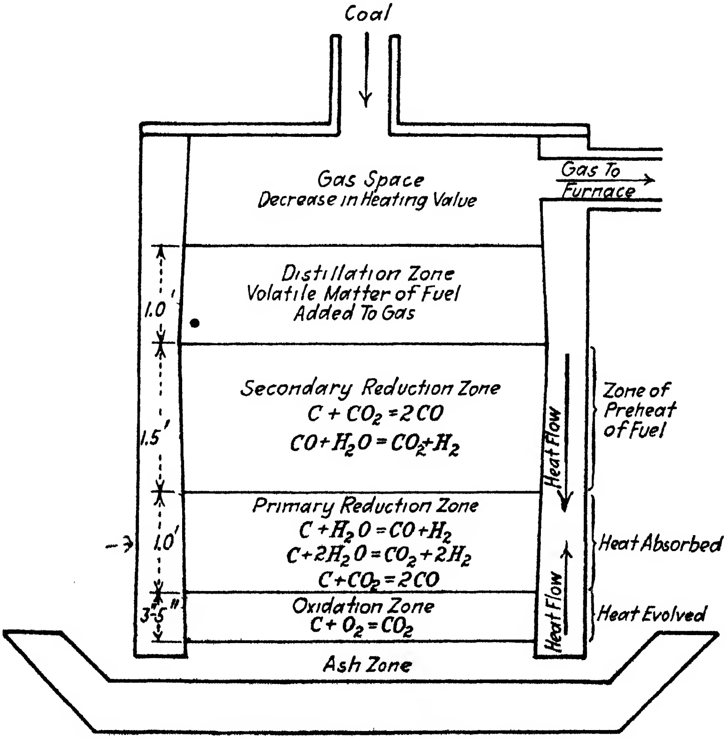

FIG. 216.—The reaction zones in the fuel bed of a gas producer.

The various zones in the fuel bed of a gas producer and their functions in producer operation are shown diagrammatically in Fig. 216 and Table CXXII.

TABLE CXXII

REACTIONS WITHIN A GAS PRODUCER FUEL BED

Zone	Function	Reactions
Ash	Cooling of ash and preheating of air-steam blast	
Oxidation	1 Combustion of fuel 2 Production of heat 3 Formation of CO_2	$C + O_2 = CO_2$
Reduction Primary	1 Reduction of CO_2 and decomposition of steam by carbon	$C + H_2O = CO + H_2$ $C + 2H_2O = CO_2 + 2H_2$ $C + CO_2 = 2CO$
Secondary	1 Some reduction of CO_2 by carbon 2. Slight reduction of H_2O by CO 3 Interchange of heat from gas to fuel	$C + CO_2 = 2CO$ $CO + H_2O = CO_2 + H_2$
Distillation	1 Separation of the fuel into coke and gaseous hydrocarbons	

Actual Gasification Results.—The application of the preceding considerations to producer practice can be illustrated by giving the results of a number of gasification tests made on producers in actual operation. Bone and Wheeler[8] carried out a comprehensive series of tests on a Mond producer with a 3.5 and a 7-ft. fuel bed, using varying proportions of steam in the air-steam blast. Their results with the 7-ft. fuel bed are shown graphically in Fig. 217, in which the gas composition, heating value of gas, gas production, steam decomposition, and efficiency are plotted against the pounds of steam admitted per lb. of coal fired. Clement[9] has also carried out a series of tests with a 5- and a 3.5-ft. fuel bed. Clement's results, plotted on the same basis as those of Bone and Wheeler, are shown graphically in Fig. 218 and 219. It will be observed in all three figures that the CO content of the gases decreased and the CO_2 content and H_2 content increased with increasing proportions of steam. The heating value of the gas and the gas temperature (see Clement's tests) decrease with increasing amounts of steam. Although more pounds of steam are decomposed as the steam content of the blast is increased, the per cent steam decomposition (100 minus the per cent steam undecomposed) decreases, thus giving a producer gas high in moisture. The efficiency of the producer is shown to be at a maximum at from 0.3 to 0.4 lb. of steam

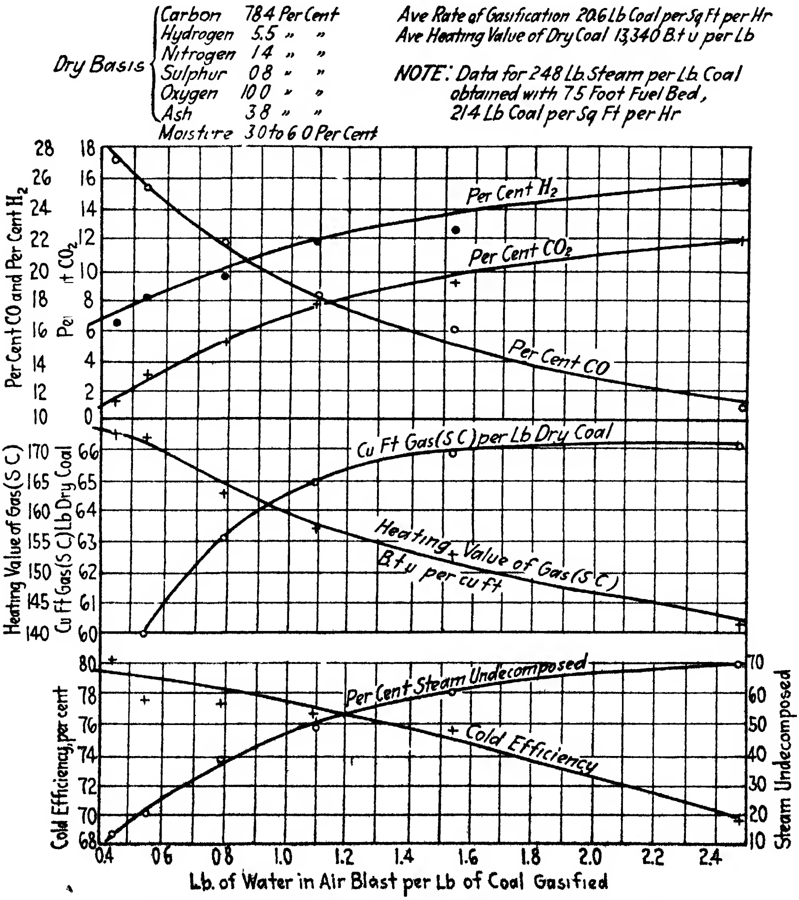

FIG. 217.—Results of gas-producer tests using a 7-ft. fuel.

per lb. of coal fired, but, as will be shown later, this figure may have to be exceeded in order to prevent trouble from clinker formation.

The cold gas efficiency in the generation of producer gas usually ranges between 65 to 80 per cent, (see Figs. 217, 218, and 219).

Radiation and cinder loss is usually not over 9 per cent. Consequently, the sensible heat in the producer gas usually is from 11 to 26 per cent of the total heat input. About 97 to 98 per cent of the total heat input usually comes from the fuel itself, the balance being the heat in the steam used in the air blast.

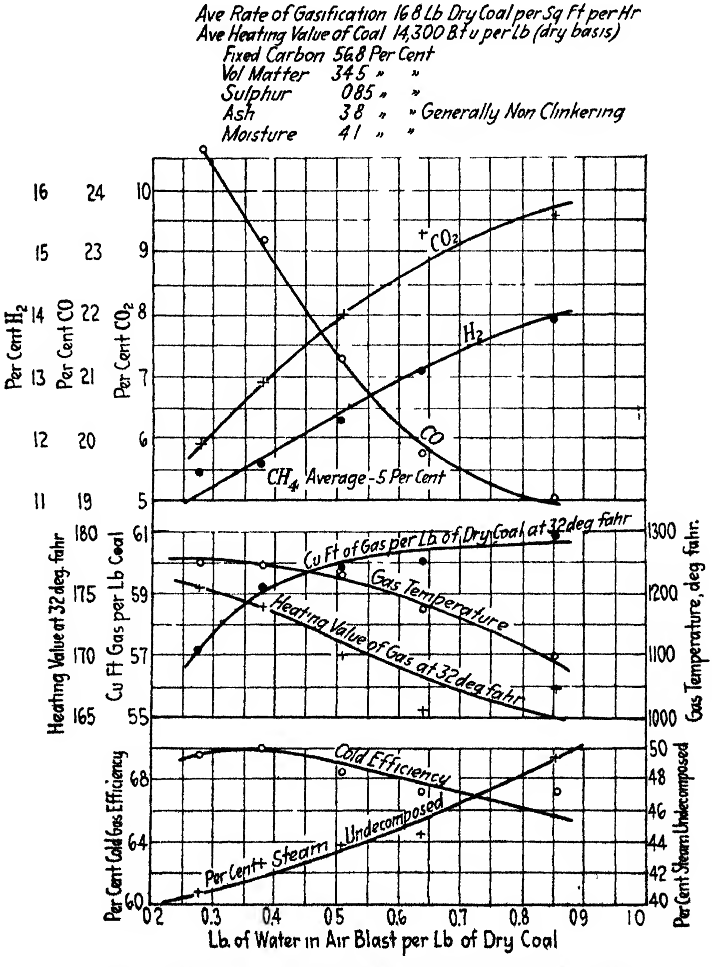

FIG. 218.—Results of gas-producer tests with 5-ft. fuel bed.

A further discussion and comparison of these results is given below.

The Effect of Depth of Fuel Bed on Gas Composition and Amount of Water in Gas.—If the pounds of steam decomposed per lb. of coal fired are plotted against the pounds of steam in the air blast per lb. of coal for the three series of tests outlined above, the influence of depth of fuel bed on gas composition is clearly illustrated. Thus in Fig. 220 it will be seen that for a given

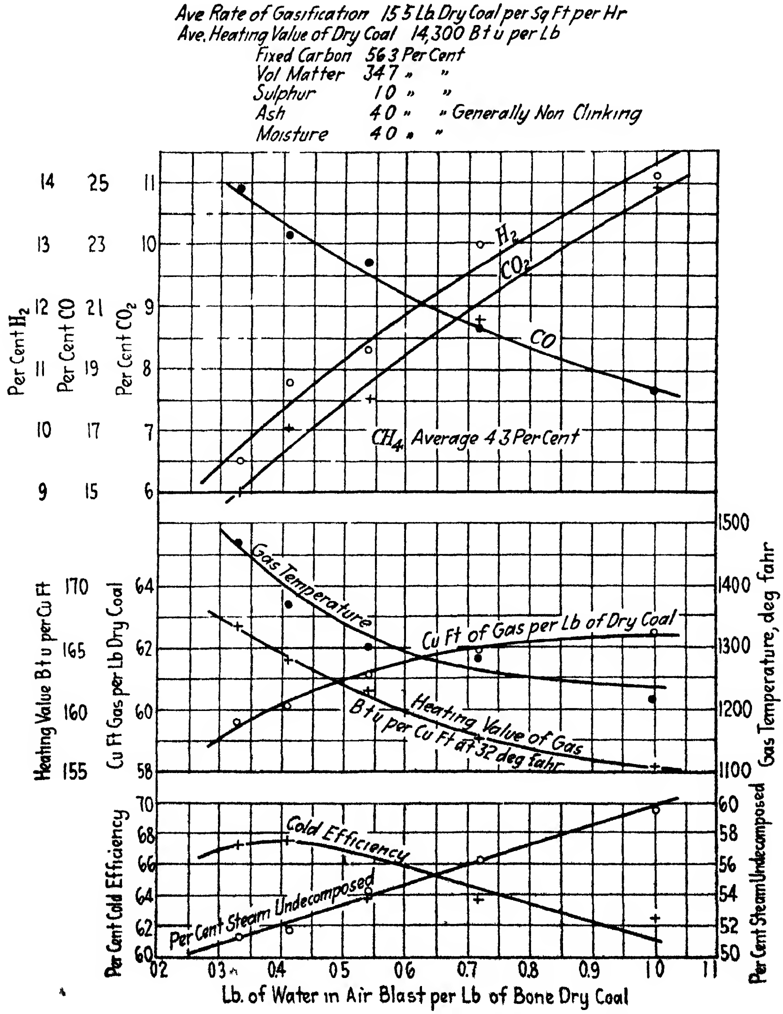

FIG. 219.—Results of gas-producer tests with 3.5-ft. fuel bed.

ratio of steam to coal, increasing the depth of fuel bed results in markedly increasing the amount of steam decomposed. For example, at 0.4 lb. of H_2O in the air blast per lb. of coal (*i.e.*, in the range of optimum per cent steam) the decomposition* is as shown in Table CXXIII.

TABLE CXXIII

EFFECT OF DEPTH OF FUEL BED ON STEAM DECOMPOSITION

Depth of Fuel Bed, Feet	Pounds H_2O Decomposed per Lb. Coal Fired
7	0.36
5	0.23
3.5	0.19

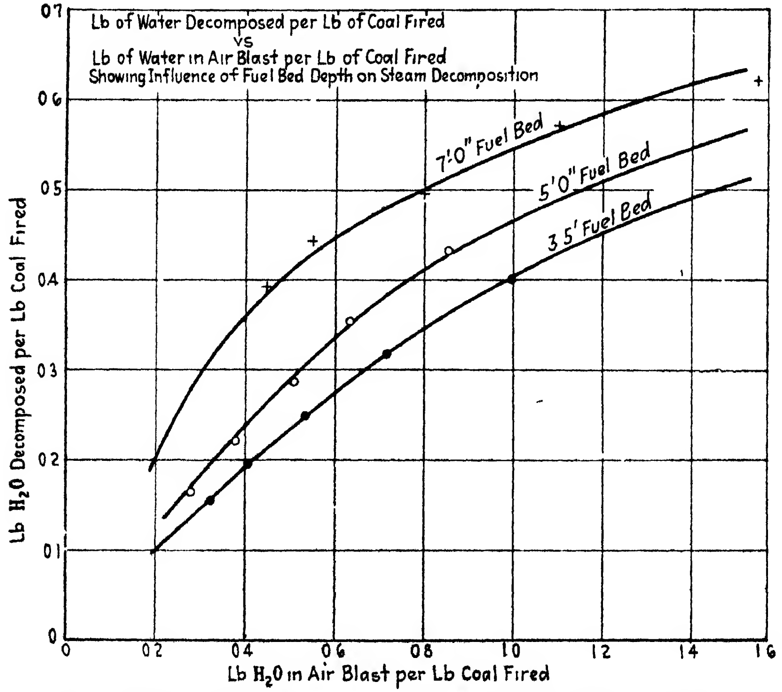

FIG. 220. Effect of depth of fuel bed and moisture in air blast on the amount of steam decomposed.

As previously pointed out, practically all the decomposition of water takes place in the *primary* reduction zone, which is about 1 ft. thick. Now for any given ratio of steam to air the only means by which the amount of steam decomposed can be

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* It should be noted that BONE and WHEELER preheated the primary air blast, and consequently some of the big increase in water decomposed as the depth is increased from 5 to 7 ft. is due to this factor. The extent of this effect is unknown.

increased is by raising the temperature, or by increasing the depth of this zone. The temperature (and therefore the steam decomposition) is increased with thicker fuel beds, since the secondary reduction zone becomes thicker and the fuel coming into the primary reduction zone is, therefore, preheated to a higher temperature.

In addition to influencing the amount of steam decomposed, the depth of fuel bed further affects the composition of the resultant producer gas. It has been found that the constituents of producer gas come to an apparent equilibrium value $K' = \frac{(CO_2)(H_2)}{(H_2O)(CO)}$ dependent[10] on the thickness of fuel bed alone, and independent of gas velocity (rate of firing), ratio of pounds of steam to pounds of coal, or temperature of the exit gas.

Table CXXIV gives[10] the apparent equilibrium constant for each depth of fuel bed, the various values having an average deviation of between 5 and 10 per cent.

TABLE CXXIV

Apparent Equilibrium Constant vs. Depth of Bed

Investigators	Overall depth of fuel bed	Apparent equilibrium constant: $K' = \frac{(CO_2)(H_2)}{(H_2O)(CO)}$	Average deviation of five to seven experiments, per cent
Bone and Wheeler[8].	7	0.66	10
Clements[9]........	5	0 48	6
Hunt, Johnson, and Willis[11].........	4.25	0.405	9
Clements[9]...	3.5	0.34	5

In Fig. 221 the average value of K' is plotted[10] against the depth of fuel bed in feet and it is to be noted that the value

$$K' = \frac{(CO_2)(H_2)}{(CO)(H_2O)}$$

bears a linear relationship to the depth of fuel bed which may be expressed by the equation:

$$K' = 0.096L = \frac{(CO_2)(H_2)}{(CO)(H_2O)}$$

where L = depth of fuel bed in feet.

(H_2O) = volume of total moisture (undecomposed steam plus the free and combined moisture in the coal) per 100 volumes of dry producer gas.

(CO_2), (H_2), and (CO) = the per cents of CO_2, H_2, and CO in the dry gas as determined by analysis of the gas in an apparatus of the Orsat type.

The range covered by the above investigators was as shown in Table CXXV.

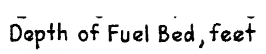

FIG. 221.—Effect of fuel bed depth on apparent equilibrium constant of producer gas.

TABLE CXXV

VARIATION IN CONDITIONS COVERED IN PRODUCER INVESTIGATIONS

Variables	Minimum	Maximum
Depth of fuel bed, in feet	3.5	7
Pounds coal per sq. ft. per hour	8.9	29.2
Pounds steam per lb. coal	0.28	1.57
Per cent CO_2	5.25	13.25
Per cent CO	12.9	27.3
Per cent H_2	8.6	22.65
Per cent H_2O decomposed	40.4	87
Volume total water per 100 of dry gas	5.6	28.1
Exit-gas temperature	1095°F. (591°C.)	1481°F. (805°C.)

In general, the equation from the above data appears to be accurate to within 10 per cent and affords a means of determining the amount of water in producer gas from the ordinary gas analysis to within 10 per cent by the following equation:

$$H_2O = \frac{(CO_2)(H_2)}{(0.096\,L)(CO)}$$

The use of this equation sometimes avoids the necessity for the direct determination of water in producer gas, a relatively difficult analysis to carry out in the plant. An example is shown in the producer calculations on page 230. The equation, together with Fig. 220, is also valuable in predicting the result of changes in producer operation. For example, suppose it is planned to gasify the coal used in the producer-gas example on page 227 with air containing 0.5 lb. of steam per lb. of coal fired and with a 4-ft. fuel bed. From Fig. 220 it would be estimated that about 0.265 lb. of steam would be decomposed per lb. of coal fired.

The coal analysis, on the dry basis, is as follows: C = 76.0, O_2 = 5.8, H_2 = 5.4, N_2 = 1.5, S = 1.5, ash = 9.8, and moisture = 4.0 per cent. For the purpose of these calculations, the ratio of the number of mols of each constituent is more useful. The composition of the coal, based on one mol (12 lb.) of carbon, may be determined as follows:

	Pounds per 100 lb. coal	Mols per 100 lb. coal	Mols per mol carbon
C	73.0	6.08	1.00
O_2	5.57	0.174	0.0286
H_2	5.18	2.56	0.421
Moisture	4.0	0.222	0.0365

The amount of steam used, 0.5 lbs. per lb. of coal, is $\frac{0.5}{6.08} \times 100$ = 8.23 lb. per mol carbon, or 0.457 mol per mol carbon. The steam decomposed is $\frac{0.265}{0.50} \times 0.457$ = 0.242 mol per mol carbon and the undecomposed steam is 0.457 − 0.242 = 0.215 mol per mol carbon in the coal.

Although not absolutely accurate, the O_2 may be assumed to be combined with the H_2 as water. The water formed in this manner is 0.0572 mol, requiring 0.0572 mol H_2 and leaving 0.421 − 0.0572 = 0.364 mol H_2 in the gas. In addition, the assumption may be made that 10 per cent of the carbon combines with H_2 to form CH_4. This is approximately true for most gas-producer coals.

The calculations are to be made on a basis of 1 mol (12 lb.) of carbon. The moisture in the gas on this basis is

$$\overset{\text{Moisture in coal}}{0.0365} + \overset{\text{Water from } O_2 \text{ in coal}}{0.0572} + \overset{\text{Undecomposed steam}}{0.215} = 0.308 \text{ mol } H_2O$$

Assuming 0.10 mol CH_4, equivalent to 0.20 mol H_2, then the H_2 in the gas is

$$\overset{H_2 \text{ from coal}}{0.421} + \overset{H_2 \text{ from steam}}{0.242} - \overset{H_2 \text{ in } CH_4}{0.200} = 0.463 \text{ mol } H_2$$

If we let x = mols CO_2 formed per mol carbon, then $1 - 0.1 - x$ = mol CO formed per mol carbon (0.1 = mol carbon as CH_4). Substituting these values in the equation,

$$0.096L = \frac{(CO_2)(H_2)}{(CO)(H_2O)}, \text{ we have } 0.096 \times 4$$

$$= \frac{(x)(0.463)}{(0.9 - x)(0.308)} = 0.384 = \frac{0.463x}{0.277 - 0.308x}.$$

Solving, $x = 0.1832$ mol CO_2, $0.9 - x = 0.717$ mol CO.

The O_2 in the CO_2 and CO is 0.1832 + 0.358 = 0.542 mol O_2. The O_2 from decomposed steam is 0.121 mol, so that 0.542 − 0.121 = 0.421 mol O_2 must have come from air, bringing in $0.421 \times \frac{79}{21} = 1.583$ mols N_2. The composition of the gas as calculated is shown in the following table.

TABLE CXXVI

Calculated Composition of Producer Gas

	Mols per mol carbon	Per cent of dry gas
CO_2	0.1832	6.0
CO	0.717	23.5
H_2	0.463	15.2
CH_4	0.100	3.3
N_2	1.583	52.0
Total	3.0462	100.0
B.t.u. per cu. ft. (60°F.)	.....	155.9
Moisture	.	10.1

The Most Desirable Ratio of Steam in Air Blast to Coal as Fired.—Inspection of Figs. 217, 218, and 219 shows that the heating value of the producer gas reaches a maximum at about 0.4 lb. H_2O in the air blast per lb. of coal and then decreases. Furthermore, with larger amounts of steam, the proportion of undecomposed steam increases rapidly and acts as an inert diluent material in the gas. This is shown by the following table of results calculated from Clement's[9] tests with a 3.5 fuel bed.

TABLE CXXVII

The Diluent Effect of Undecomposed Steam

Pounds steam per lb. coal	Mols N_2 per 100 mols combustible gases	Mols CO_2 per 100 mols combustible gases	Mols H_2O per 100 mols combustible gases	Total mols inerts per 100 mols combustible gases
0.31	141.8	15.4	19.0	176.2
0.41	141.0	18.3	24.1	183.4
0.54	143.3	19.7	31.6	194.6
0.72	143.6	23.6	43.8	211.0
1.00	143.6	29.8	65.5	238.0

The table emphasizes the increasing amounts of inert water and CO_2 in the gas as the amount of steam is increased. When the producer gas is burned in a furnace these inert gases lower the flame temperature and furnace efficiency by carrying heat out the stack.

The tests made by all the investigators here reported (Figs. 217, 218, and 219) show that the efficiency of the producer is seriously lowered with increasing ratios of steam to coal. The use of too large an amount of steam in the air blast has the following disadvantages: (1) the calorific value of the producer gas is lowered and the amount of inerts is increased, (2) the efficiency with which the gas can be used in the furnace is decreased, (3) the thermal efficiency of the producer is lowered. Consequently, in order to minimize losses from these effects, not more than 0.35 to 0.40 lb. of steam should be used for each pound of coal gasified. Continuity of producer operation outweighs any question of efficiency and, consequently, enough steam must be used to keep the producers reasonably free from clinker trouble. Therefore, although from the standpoint of efficiency only about 0.35 lb. of steam should be put into the air blast per lb. of coal, it is often necessary to increase the steam up to about 0.5 or even 0.6 lb. per lb. of coal in some cases in order to reduce clinker formation to such an extent that continuous operation is possible. Most plants, however, have a tendency to use too much steam because it is easier for the producer operator.

With coals of high-fusion-point ash, less steam is required than with coals having ash of low fusion point. The fusion point and the amount of ash, therefore, greatly affect the efficiency of a gas producer and it is necessary to watch this point in purchasing a coal for a gas producer.

The Effect of Fuel-bed Condition.—The discussion on page 553 outlined conditions in the fuel bed of a well-operated producer. In poorly poked producers, however, there is a tendency for the formation of channels and for the leakage of air around the edges of the producer where resistance to the flow of the gases is low. This is strikingly illustrated by the results of Terres and Schierenbeck[12] who took four samples across the fuel bed of a 5-ft. producer from which the cinder was removed once in 24 hr. Their results are shown in Table CXXVIII.

The table shows clearly the wide variation in gas composition due to channeling and to edge leakage that may occur even on the same level in the producer. The admixture of "edge gas" lowers the heating value of the resultant producer gas, and may well be as harmful in this respect as the Neuman reversion

TABLE CXXVIII

VARIATION IN PRODUCER GAS DUE TO CHANNELING

Time after ash removal	Gas composition		
	Constituent	Per cent at wall	Per cent 20 in from wall
7 hr	CO_2*	9.5	3.6
	CO	19.4	29.8
	H_2	11.3	6.0
8 hr	CO_2	13.8	4.3
	CO	12.4	27.3
	H_2	10.7	8.3
10 hr	CO_2	14.4	3.9
	CO	10.0	29.1
	H_2	4.8	11.0

referred to on page 557. Moreover, where a channel forms, "hot spots" are also liable to result and clinker formation usually follows. Troubles due to clinker and to the leakage of edge gas may be greatly reduced (1) by uniformly covering the bed with green coal, (2) by proper poking, either by hand or mechanically and (3) by continuous removal of the cinder.

In poking, the bar should be shoved down along the side wall of the producer, and then pulled toward the center with a slight upward motion. Care must be taken not to insert the bar too deep and lift the incandescent fuel of the oxidation zone into the reduction zone. This causes clinkering, due to the introduction of the hot ash into a reducing atmosphere. It is best to poke reguarly at short intervals, rather than to poke vigorously at long intervals.

Rate of Gasification.—The rate of gasification or the rate at which coal can be gasified is dependent on the size and type of fuel, its moisture, and the amount and fusibility of the ash. The amount and nature of the ash are most important, since they limit the temperature of the reduction zone. The usual rates of gasification are 8 to 30 lb. per sq. ft. per hr., the low rate of 8 lb. sometimes being the maximum for a hand-operated producer, using poor coal (*i.e.*, a caking coal containing easily fusible ash). The average rate of gasification in American producers is about 15 lb. per sq. ft. per hr., 25 lb., however, being possible with good coals. In some of the newer producers, which have continuous ash removal and very efficient mechanical poking, the combustion rate is about doubled and gasification rates of as high as 50 lb. per sq. ft. per hr. have been obtained.

Results[6] on an experimental, anthracite coal gas producer, with fuel beds of three different depths when the steam in the air blast was around 0.4 lb. per lb. of coal, show the importance of a high gasification rate when the fusion point of the ash is high enough not to cause trouble. The following table shows the marked increase in fuel-bed temperatures which brings about a high percentage of decomposed steam and a high cold gas efficiency as the rate of gasification (and also as the thickness of the fuel bed) increases. These figures also show the importance of using fuels with ash having as high a fusion point as possible, since this determines producer capacity.

TABLE CXXIX

Effect of Rate of Gasification and Depth of Fuel Bed on Fuel-bed Temperature, Per Cent Steam Decomposed, and Cold Gas Efficiency

	Rate of gasification, pounds coal per sq. ft. per hr.	Depth of fuel bed		
		1.5 ft	3.0 ft	4.5 ft.
Maximum temperature in primary reduction zone, in degrees Fahrenheit	10 lb.	1600	1600	1800
	40 lb	1700	2200	2300
	70 lb.	2100	2400	2400
Per cent steam decomposed	10 lb	55	67	77
	40 lb.	68	73	99
	70 lb.	84	97	90(?)
Cold gas efficiency, per cent	10 lb.	54	63	67
	40 lb.	64	66	74
	70 lb.	72	74	72(?)

Judging the Quality of Producer Gas.—The quality of producer gas may be judged in two general ways: first, by the appearance of the gas and of the producer bed, and, second, by analysis of the gas and calculations based on the analysis. Judging *by appearance* is quick, and, to the trained eye, is often very satisfactory. The best gas is a slightly smoky (sooty) yellow and the top of the producer fuel bed should be maintained a slight dull red. A dark sooty producer gas with no yellow gas from the distillation of the volatile products of the coal indicates that the top temperature of the producer is too high due to (1) blowholes in the producer bed, (2) too thin fuel bed, and (3) forcing the producer too hard for a given depth of bed. On the other hand, if the gas is free from all soot, the producer is usually being operated below capacity or with a very high steam-coal ratio.

Good indications as to the condition of the fuel bed may be obtained by observing not only the general color of a bed but also the uniformity. If blowholes exist they usually manifest themselves by the appearance of bright spots due to localized high temperature. The thickness of the bed may be judged by pushing a thick poker bar straight down through the fuel bed into the ash pit. After ½ to 1 min. the poker is withdrawn. The length of the red-hot portion of the poker corresponds to the hot zone of the fuel bed and the thickness of the distillation zone is usually evident by the tar and soot stuck to the poker. Knowing the distance from the top of the fuel bed to the top of the producer shell, the poker test also gives the depth of the ashes. Whether the producer is being pushed too hard usually can be determined by the condition of the clinkers and by a comparison of the rate of gasification with the normal rates.

In judging the quality of producer gas *by analysis*, either the per cent CO_2 only may be considered, or else the complete analysis may be made. The best practice in producer operation yields a gas containing 2 to 5 per cent of CO_2. If the CO_2 content of the gas is over 5 per cent, the operation usually is at fault in one or more of the following ways:

1. Too much steam is being used. Therefore, the temperature of the reduction zone is too low and a large amount of unreduced CO_2 and undecomposed water are passing up through the fuel bed.
2. The fuel bed is too thin. This does not allow sufficient time of contact for reduction of CO_2 and also does not permit sufficient preheating of the fuel in the upper zones of the fuel bed.
3. The rate of gasification (pounds of coal gasified per sq. ft. of producer-grate area per hr.) is too high, and as, a result, clinkers

formed in the oxidation zone cause a large proportion of the gas to flow up through the fuel bed in channels, thus decreasing the contact between the gas and the fuel.

4. Air may be leaking in at the top of the producer (if it is of the suction type) and a portion of the CO thus burned to CO_2.

5. The bed may contain channels through which the CO_2 can travel without coming into contact with the coal. These channels may result from poor blast distribution, from too little poking, from too deep poking (which brings the red-hot ash up into the reduction zone, where it fuses, due to the reducing atmosphere) or from the clinkers formed, through high temperatures in the fuel bed, due to too high a rate of gasification.

The remedy for each trouble is obvious.

The use of the complete gas analysis as a method of judging the quality of the gas has been ably presented by Husson.[13] He pointed out that the heating value of the gas (B.t.u. per cu. ft. of dry gas) is not a sufficient criterion of gas quality, since this value does not take into account either the diluent effect of water vapor, or the enrichment due to tar and soot. Furthermore, the usual method of determining the heating value at 32°F. (0°C.) with the products cooled to 32°F. (0°C) is not in accordance with the way the gas is usually utilized, as in the open-hearth furnace, for example, where gas and air enter the furnace at about 1800°F. (1000°C.) and the products of combustion leave at about 2900°F. (1600°C.).

Although heat transfer is a function of the velocity of the gas, there is a limiting velocity which cannot be exceeded. This limiting value of gas velocity is not a question of the rate of heat transfer, but is determined by stack draft and by the rate of combustion of the gas. Due to this latter consideration, the time that the gas must be in the combustion space or hearth is at least 1 sec. Therefore, the important criteria of a producer gas to give good performance (*i.e.*, high rate of heat transfer) are: (1) the flame temperature T, and (2) the heat available in 1 cu. ft. of the products of combustion in cooling from the flame temperature to the minimum temperature at which the heat can be utilized (2900°F., or 1600°C. for the open-hearth furnace).

Figure 222 illustrates the result of Husson's* calculations. The curve marked L represents B.t.u. per cu. ft. of gas (including

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* HUSSON assumes equilibrium between gas and coal, and although this is probably not true in most cases of producer operation, the equilibrium concentrations represent conditions beyond which it is not possible to go at any given temperature.

soot and tar), P is the producer efficiency in per cent, and V is the cu. ft. of wet gas produced per lb. of coal fired. The maximum producer efficiency occurs when the temperature of the gas leaving the producer is about 1300°F.

In Fig. 223 the B.t.u. of heat available above 2900°F. (1600°C.) per cu. ft. of flue gas at the flame temperature is designated by N; T is the flame temperature in degree Fahrenheit, and P is the furnace efficiency with air and gas entering at 1800°F., and leaving at 2900°F. (1600°C.). The values of these factors are plotted against the temperature of the gas leaving the top of the producer. The furnace efficiency is at a maximum when the producer-gas temperature is about 1300°F. (700°C.), and the amount of heat available above 2900°F. (1600°C.), and the flame temperature of the gas is greatest when the gas temperature is around 1400°F. (760°C.). From this figure the criteria

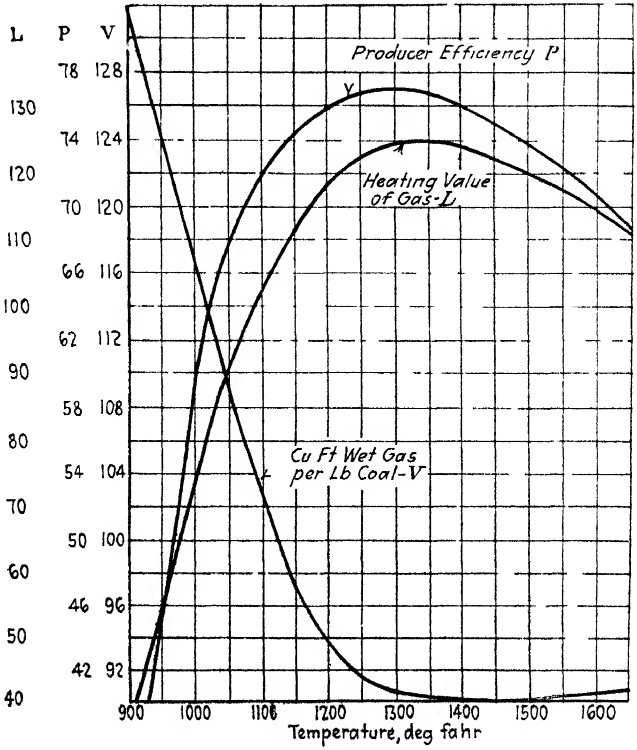

FIG. 222 —Effect of producer-gas temperature on producer efficiency heating value of the gas, and volume of gas produced

for the best producer gas for furnace work are that N equals 4.1 B t.u per cu. ft. of gaseous products at the flame temperature, and that T (the flame temperature) equals about 4100°F. (2250°C.). If heat is being absorbed at high temperatures, such as through the melting of metals or through the endothermic decomposition of a product, such as the formation of lime from limestone, the heat available above this temperature in each cubic foot of the products of combustion is of the greatest importance.

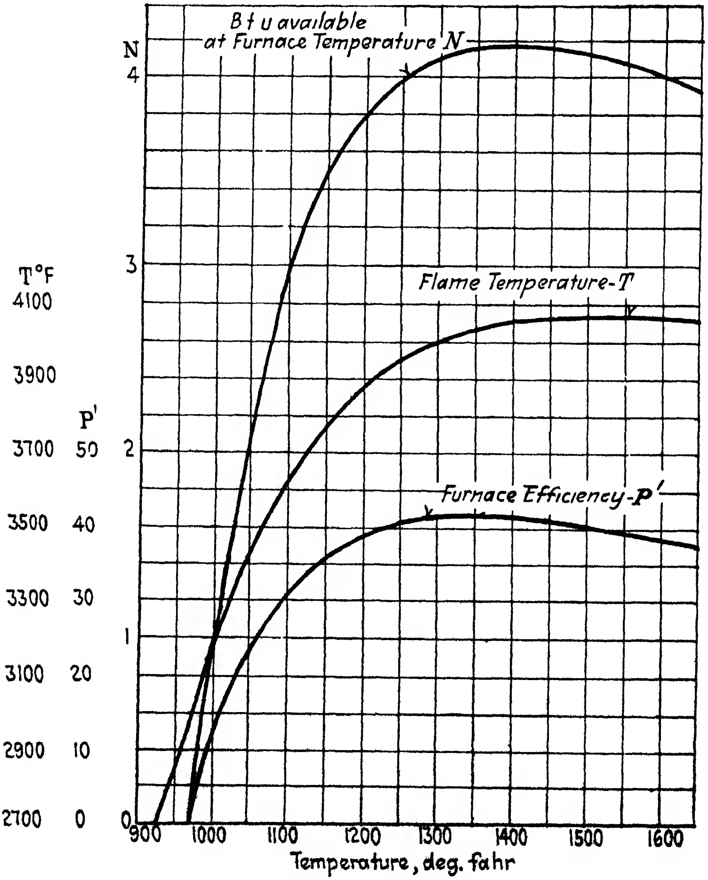

Fig 223 —Effect of producer-gas temperature on the amount of utilizable heat in producer gas above 2912°F. (1600°C)

These plots show why gas producers used in connection with open-hearth furnaces usually are operated at rather high gas

TABLE CXXX

INFLUENCE OF METHOD OF OPERATION ON CHARACTERISTICS OF PRODUCER GAS

Kind of operation	Gas analysis			Characteristics of gas		Appearance of fuel surface	Value of gas obtained and causes of poor operation
	Per cent CO_2	Per cent CO	Per cent H_2	Temperature, degrees Fahrenheit	Appearance		
Cold fire.	Over 7	Below 20	13–17	1200–1300	Golden yellow much tar	Dull red or pale cherry	Very bad, combustion too slow; upper part much hotter than lower, too much steam or grate choked
Good fire	3–5	About 24	About 15	1300–1500	Yellow, still much tar, often a little smoky from decomposed tar	Cherry to bright red	Excellent operation
Good fire with channels in bed	Over 5	Less than 24	15 or less	1300–1500	Smoky; tar decomposed	Cherry to bright red	Upper part much hotter than lower, poor operation, not enough poking, coke too sticky, or too much fusible ash
Hot fire. . . .	3–4	Over 25	5–10	Well above 1500	Smoky; tar entirely decomposed	Bright red to pale yellow	Not enough steam; operation passable, and sometimes to be recommended with rich- or low-ash coals, channeling changes this to next type
Hot fire with channels in bed.	Over 3 or 4	Less than 25	5–10	Well above 1500	Smoky; tar entirely decomposed	Bright red to pale yellow	Not enough steam; too much ash, ash too fusible, not enough poking; very poor gas, distinguished by low H_2

temperatures. At high gas temperatures the change in efficiency (of the furnace) is small, with a small change in gas temperature, while at low gas temperatures (below 1250°F., or 680°C) a small change in gas temperature causes a large decrease in efficiency.

The influence of method of operation on the characteristics of producer gas is shown[13] in Table CXXX.

Fuels Suitable for Gas Producers.—Almost any fuel from anthracite coal to wood waste can be and has been successfully gasified in gas producers By far the greater proportion of producer gas, however, is manufactured from bituminous coals, although coke is widely used in the gas industry. Pea-size anthracite coal finds application in small producers (equivalent to about 10,000 cu ft. of city gas per hr).

For power purposes a clean gas is necessary and very low-volatile, low-sulfur fuels, such as coke, anthracite, or charcoal, are used. This explains the small number of producer-gas power plants, since the above fuels are expensive, and for economic reasons cheap bituminous coal must be used as the real source of power

For the production of a gas for the heating of furnaces it is necessary to obtain the greatest amount of heat in the form of gas per dollar. Therefore it is advisable to strike an economic balance, which includes the cost and heating value of the available coals, together with the cost of gasification. This latter factor includes depreciation, labor cost, and cost of cleaning with the different coals.

As regards size, run-of-mine coal, all of which will pass through a 4-in ring, usually is cheapest. Dust-free coal which will all pass through a 2-in. ring, though not the cheapest, usually is best on account of the increased ease and economy of operation. Any run-of-mine coal can be used, but a closely sized fuel makes operation far easier, since the finer the coal the higher the air-steam pressure must be (with steam-jet blowers) to drive the producer at a given rate.

The coal should be non-caking (non-coking). The caking coals are almost impossible to handle, because cracks in the coked fuels are liable to result in blowholes forming in the fuel bed. For this reason the ratio of hydrogen to oxygen (see p. 44) in the moisture- and ash-free coal usually should be less than 0.55 and preferably around 0.50. High-volatile coals give a rich gas, practically a coke producer gas enriched with coal gas. This fact is of distinct importance, since, in addition to giving a rich gas, coals of high-volatile matter (36 to 38 per cent) usually are non-caking and, therefore, well adapted for use in gas producers. High-oxygen coals, being non-caking, make satisfactory producer-gas coals.

Ash in the best producer coals should be non-clinkering, *i.e.*, of high fusibility. Coals with ash having a fusing point* above 2600°F, or 1427°C. (*i.e.*, class A), give a higher B.t u. gas, since the producer can be run hotter. Although coals with ash of lower melting point can be used, the coals with class A ash allow greater capacity with less operating trouble. The sulfur content of the coal may seriously affect the ash-fusion point, and for best results in producers the coal should not contain much over 1 per cent of sulfur, though 1.5 per cent is possible and even as high as 4 per cent sulfur coal *may* be gasified. The actual per cent of ash is relatively unimportant if it is non-clinkering and low in sulfur; fuel containing as high as 50 per cent ash has been gasified successfully. If the producer is to be forced, however, low-ash fuel is safest and does not require as much cinder removal. Furthermore, in poorly designed or inefficiently operated gas producers, a high percentage of ash may mean a considerable loss of unburned fuel in the cinders. The iron content is also important as affecting the fusing point of the ash.

If the fuel contains a high percentage of moisture, a considerable heat loss results. This loss is due to three effects: (1) heat from the producer gas is utilized in evaporating the moisture in the fuel at the top of the producer, (2) the sensible heat in this water vapor is lost out the stack, and (3) water vapor, whether from undecomposed steam or from moisture in the fuel, brings about an oxidation of CO to CO_2 in regenerators according to the reaction

$$CO + H_2O \rightleftarrows CO_2 + H_2.$$

Thus moisture in the fuel not only increases the sensible-heat loss out the stack, but also increases the per cent CO_2 in the producer gas.

Classification of Producers.—Gas producers may be classified in a number of different ways, such as (1) direction of draft, (2)

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* The fusion point of the ash from the large number of coals from different parts of the United States is given in the tables on pp. 27–29.

pressure, (3) method of feeding and poking, (4) method of ash removal, and (5) cleanliness of gas.

1. *Classification According to Draft.*—In the *up-draft* type of gas producer the air-steam mixture passes up through the bed of descending fuel and the gas is taken off at the top of the producer. This type is simple and is the one in most common use. It has, however, the disadvantage that gas made from bituminous coal contains tar and soot. In *down-draft* producers, the green coal and the air-steam mixture are admitted at the producer top, the gases pass down through the fuel bed, and are taken off at the producer bottom. Producers of this type give the cleanest gas, since the tar and volatile matter given off from the green coal are cracked or decomposed to fixed gases as they pass through the incandescent fuel bed. Thus the gas from down-draft producers does not contain admixed coal gas. Further, moisture from the fuel may react in the fuel bed, which is not the case in producers of the up-draft type, in which the moisture is driven off from the fuel and into the gas. In spite of their advantages, down-draft producers have not been adopted widely on account of the following mechanical difficulties: (1) igniting the fuel is difficult, (2) the location of the zones of oxidation and reduction is liable to vary considerably, (3) grates burn out rather frequently, and (4) it is difficult to maintain uniformity of operation and gas quality.

2. *Classification According to Pressure.*—Gas producers may be run either under minus (suction) or plus (positive) pressure. Suction gas producers usually are provided with a grate and the air-steam mixture is drawn up through the fuel bed of the producer by means of an exhauster or a stack. In some plants, in which the gas is used in internal-combustion engines, the producers are direct connected through coolers to the suction side of the cylinder. In producers of the suction type, the gas flow is proportional to the consumption, and they are used mainly in power work, for heating gas retorts, or when the gas for long-distant transmission is to be compressed and cleaned. The moisture content of the entering air is satisfactorily controlled by drawing it through a humidifying tower previous to entering the gas producers. The saturating water in the tower is heated by steam, and the air passing to the producer is saturated at 110 to 130°F. (43 to 54°C.), depending on the ratio of water to fuel desired.

In pressure producers the pressure is usually produced by steam injectors or by a steam-driven turbo-blower. The steam injector uses the steam which must be fed to the producer for injecting the necessary air. The injector has the lowest first cost and lowest upkeep charges. It is, however, impossible (by changing the steam pressure) to change the amount of air injected without changing the ratio of air to steam in the producer. This is, of course, undesirable, since this ratio should be kept substantially constant for a given fuel. With an injector the only method by which the steam-coal ratio can be kept approximately constant is to change the size of steam nozzle. The Koerting steam injector, which is typical of modern design, is shown in Fig. 224. Thorough mixing of steam and air is secured by the admission of air at each of the series of nozzles shown in the diagram. Each design of injector has different pressure and volume characteristics, and the amount of air delivered with a given air-shutter setting and steam pressure will depend on the type of injector employed.

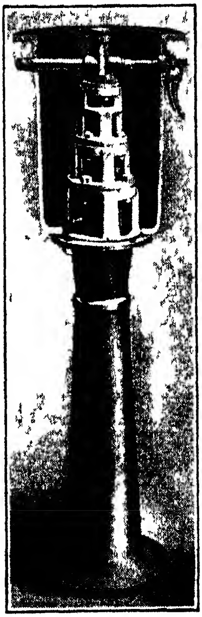

FIG. 224.—Schutte-Koerting steam injector

A positive blast from a fan or turbo-blower may be used to overcome the disadvantages inherent in steam-injector operation. A single- or multi-stage fan may be used when the air pressure is under 2 to 3 lb. per sq. in. For higher pressures, in conjunction with gas-producer plants, a turbo-blower is commonly used. The speed of the blower can be adjusted to meet the pressure or load demands on the producer and the proper amount of steam admitted (1) by bleeding in a thermostatically controlled amount of exhaust steam from the blower itself and (2) by humidification at the proper temperature of all the entering air. Thus the air-steam ratio may be readily kept constant and in most cases, in spite of the greater first cost, a steam-driven blower will prove to be more economical than a steam injector, particularly if the load on the producer is subject to wide variations. This method of steam admission is coming more and more into use.

3. *Classification According to Method of Feeding and Poking.*—Hand-fired and hand-poked producers have the lowest first cost, but they result in a non-uniform gas, the operators must have skill and experience, and hand operation is hard work. The choice between hand and mechanical producers depends mainly on the size of the producer, hand-operated equipment being suitable in small plants only.

Mechanical producers give a uniform gas, but are considerably more expensive than the hand-operated type. To offset this high first cost, however, they frequently bring about a saving of 20 to 25 per cent of the fuel when a poor coal is being used. Mechanical poking relieves the operator of considerable hard work and aids in the elimination of clinkers and blowholes in the fuel bed. On account of the installation cost, producers with mechanical feeding and poking are used mainly in large-capacity equipment where the labor and fuel saved are considerable items.

4. *Classification According to Method of Ash Removal.*—Suction gas producers and certain types of pressure producers are equipped with grates, the ash being removed by movement of the grate bars. It is essential not to remove more ash from the edges than from the center of the bed. Otherwise blowholes and edge leakage will increase. In most pressure producers the ash stands in a water-sealed pan and no grates are employed. The ash is accessible from all points on the producer circumference and may be drawn out either continuously or intermittently.

The removal of ashes by hand is cheap, but often is unsatisfactory, since the depth and the position of the fuel bed are not kept uniform and the resistance to the flow of the gas-air mixture is distributed unequally, thus giving a non-uniform gas.

Automatic ash removal is of two types: (1) intermittent, and (2) continuous. In the intermittent type, the ash is removed once or twice in 24 hr. During the ash-removal period the fire zone is disturbed and, for half an hour or more, regular gas production is interrupted. This method is somewhat similar to ash removal in hand-operated producers and labor only is saved. Since ash is removed infrequently, large storage space for the ash must be provided at the bottom of the producer. This type is most common in the United States.

Continuous ash removal permits steady production of a uniform gas, since the fire zone remains in the same position; there is no interruption during cleaning, and there is less tendency for the formation of blowholes and clinkers. Continuous ash removal is used more in Europe than in the United States.

5. *Classification According to Cleanliness of Gas.*—Gas from an up-draft producer is hot (about 1300°F., or 700°C.), and if the coal is bituminous the gas contains considerable quantities of soot, dust, and tar. The carbonaceous and tarry matter is not detrimental (and may even be advantageous) if the gas is to be used in a recuperative or regenerative furnace near at hand, but when the gas is to be used at a distance from the gas producer it cools and the tar and the soot settle out of the gas, clogging valves and obstructing piping.

Clean gas is necessary for the operation of internal-combustion engines or furnaces scattered at a distance from the gas producer, and the tar, soot, etc. are removed from the gas by scrubbing. The usual scrubbers, such as sawdust, glass wool, coke-filled spray towers and the like, are used for this purpose. With bituminous coals this cleaning process is difficult.

Prevalent Types of Producers.—The choice between hand and mechanical feeding, poking, and ash removal depends on the size of the installation and the difficulties in handling the fuel. With fuels containing a clinkering ash and having a tendency to cake, efficiency is almost impossible without using mechanical operation. Most small plants, especially if using coke or anthracite coal, resort to hand operation on account of the low first cost.

For power, as previously stated, the primary requisite is for a clean gas, so up-draft producers of the suction type are used with coke or anthracite, which, on account of their low content of volatile matter, give a gas relatively free from tar or soot. For furnace heating, on the other hand, it is necessary to produce large quantities of cheap gas, and cleanliness is not so important. Therefore, pressure, up-draft producers using bituminous coal are generally used and no attempt is made to scrub the gas, since raw gas proves satisfactory and is much better in both cost and thermal efficiency.

Description of Modern Producer Types.*—Modern producer design and the various devices used for fuel feeding, poking, and ash removal are illustrated by the following descriptions of a number of producers of both American and European

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* For more complete descriptions of the Wellman, Wood-Morgan, and Chapman producers the reader is referred to CHAPMAN.[1] RAMBUSH[14] gives detailed descriptions of these and numerous other producers not described in this chapter.

manufacture. Six American producers are described: (1) the Wellman, (2) the R. D. Wood, (3) the Morgan, (4) the Chapman, (5) the Smith, and (6) the U. G. I. pressure producer. Three producers of European design are: (1) the Mond, (2) the slagging-ash type, and (3) the Wollaston.

Producer Types Used in the United States.—The Wellman Producer.—The Wellman (earlier models known as the Hughes) was the first successful mechanical producer installed in America. The fuel bed is mechanically poked by a vertical water-cooled finger attached to the stationary top of the producer (see Fig. 225). The finger is made to oscillate between the center and the side of the fuel bed as the producer body and its contents revolve. The producer body and the fuel bed make one revolution in 8 min. and thus the entire top of the bed is stirred frequently. Automatic ash removal is accomplished by a stationary bar which sweeps the ash from the revolving grate. This bar is inserted into the ash pan about once in 24 hr., though in

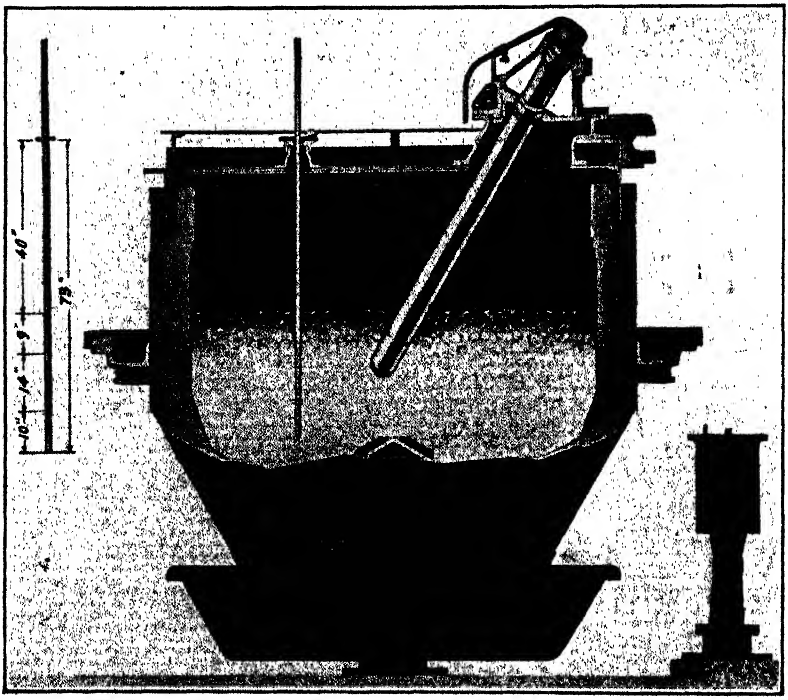

FIG. 225.—Wellman mechanical gas producer.

the newer types of Wellman producers ash removal is continuous, and in consequence the combustion rate is about doubled (30 lb. per sq. ft. per hr.). An automatic double-quadrant feed device gives exceptionally good distribution of coal on the fire bed.

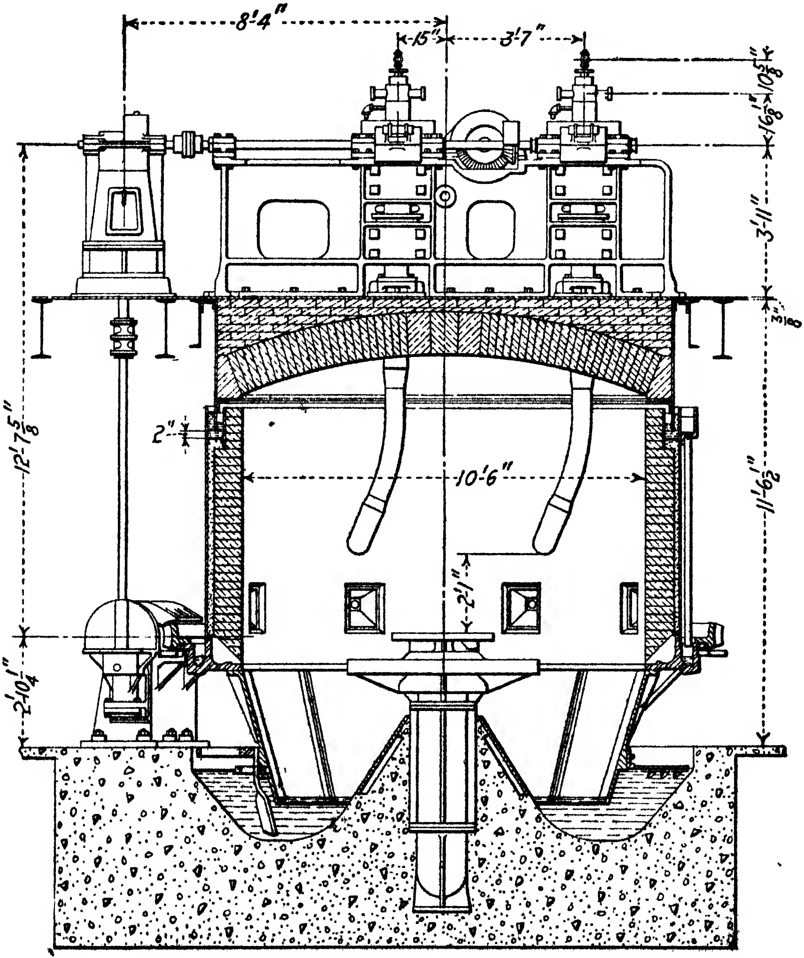

Fig. 226.—Wood gas producer.

The R. D. Wood Producer.—In the R. D. Wood producer the bed is poked by the two water-cooled, J-shaped vertical stirring arms shown in Fig. 226. Unlike the Hughes poking device, these arms do not oscillate, but revolve about their vertical axes and stir the bed as the producer revolves. The producer makes one revolution in 30 min. A double, automatic feed located eccentrically distributes coal over the revolving fuel bed. The ash is removed by a plow attached to the lower edge of the producer shell. The plow will remove ash slightly more rapidly than it is formed, and therefore it cannot be used quite continuously. On account of the practically continuous ash removal and the two poking arms, this producer has a large capacity (about 35 lb. per sq. ft. per hr.).

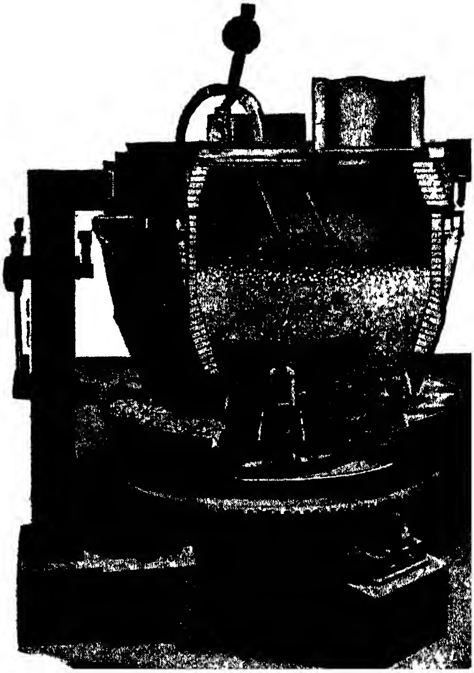

FIG 227.—Morgan gas producer.

The Morgan Gas Producer.—In the Morgan producer shown in Fig. 227 the fuel bed is not disturbed by poking, but is *leveled* by a horizontal U-shaped arm which rides on the top of the bed. The arm is stationary and the leveling process is brought about by the rotation of the producer body once in 12 min. With caking coals this method does not give as good results as actual poking.

The ash-removing device shown in the figure consists of a horizontal spiral arm which normally revolves with the producer body. From one to three times each 24 hr. the outer end of this arm is held stationary and the relative motion of the arm and pan causes the ash to be carried to the outer edge, where a plow removes it. After ash removal the fuel bed is "broken down" by hand and drops to replace the ash which has been removed. These producers can gasify up to 35 lb. of coal per sq. ft. per hr.

The Chapman Producer.—The body of the Chapman producer (Fig 228) is stationary and the bed is poked by a revolving floating agitator. This agitator is in the form of a rake with water-cooled teeth. This rake makes one revolution in 7 min., thus stirring the fuel bed every 3½ min. The teeth extend from 8 to 11 in. into the bed, or about half way between the surface agitation of the Morgan and the deep agitation of the Wood or Hughes producer. The agitator is so arranged that if the resistance to its passage becomes too great (as when it is buried under a heavy charge of fresh coal or comes in contact with a clinker

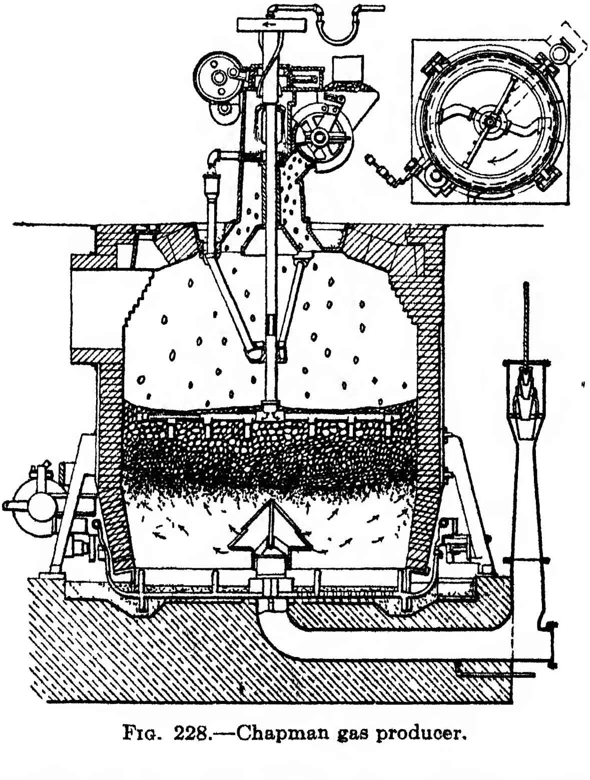

Fig. 228.—Chapman gas producer.

fused to the wall), the driving head screws up and the agitator is drawn up until the resistance again becomes normal. Thus, the agitator continuously floats near the top of the bed.

This producer is equipped with an automatic coal-distributing device and ash removal is continuous. The ash-removal device consists of a revolving beam with fingers which agitate the ash bed from beneath and thus lessen the tendency for blowholes. The beam speed is adjusted until the ash is forced to the outside of the pan as fast as it is formed

The Smith Producer —The Smith gas producers are either poked by hand or use a mechanical poking device which gives a punching action similar to hand poking.

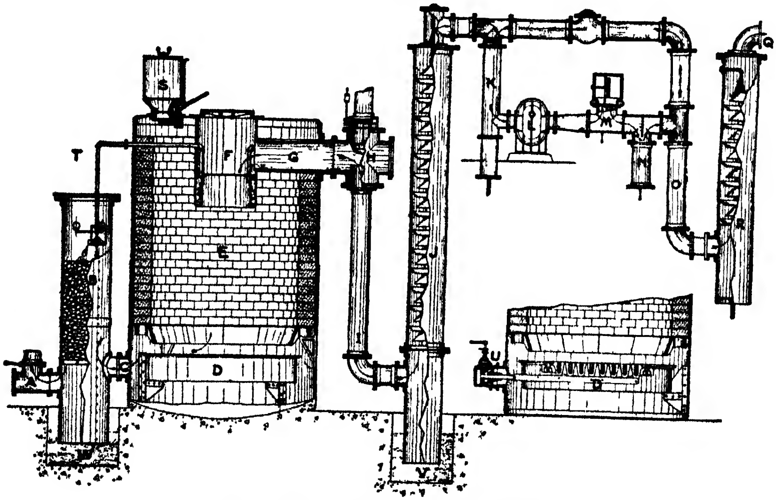

Fig 229 —Smith suction producer

In the Smith suction producer (Fig. 229) the steam is supplied by saturating the entering air with water at a thermostatically controlled temperature. These producers are designed to deliver clean gas, the tar removal being accomplished primarily by glass-wool extractors across which there is a 1 to 3 lb. per sq. in. pressure drop. Positive displacement blowers draw humidified air through the producer, cooler, and primary scrubber, compress the producer gas, and force it through the glass-wool tar extractor and through the secondary scrubber. A pressure relief valve connected in a by-pass around the blower and glass wool extractor maintains a constant pressure (2 lb. per sq. in.) on the distributing lines.

The U. G. I. Pressure Producer.—The U. G. I. rotary grate pressure gas producers (Fig. 230) are designed to operate with fuel (often coke) having a large percentage of fines without a material reduction in capacity or efficiency. The fuel is charged into the top of the producer from a hopper on a revolving water-sealed table. The space around the upper gas outlet cell is always kept full of fuel. This space contains about a 2-hr. supply. In charging by this method the large lumps of fuel tend to segregate and flow toward the middle of the producer and the fines

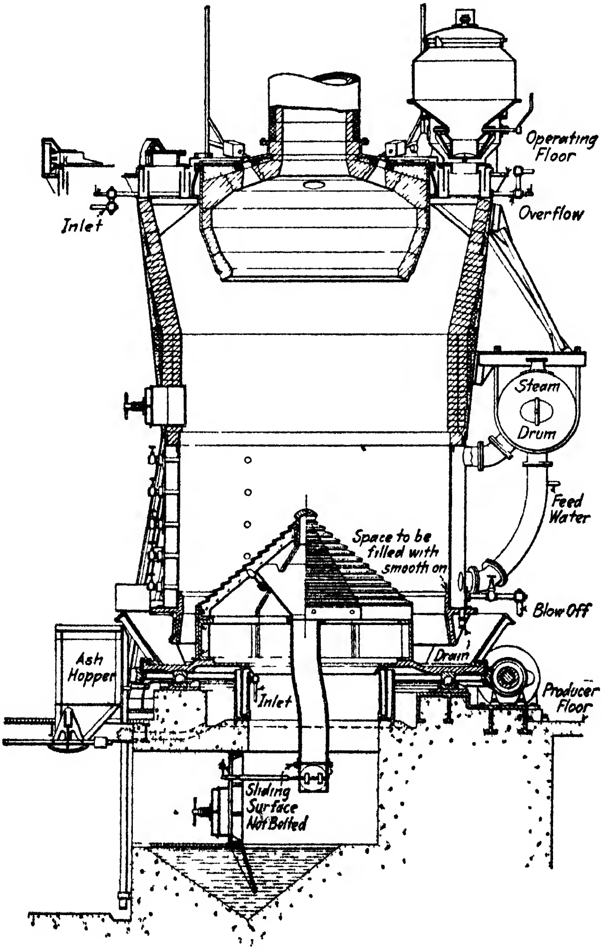

Fig. 230.—U. G. I. pressure producer.

sift down the sides. This distribution of fuel gives good operating conditions, since channels along the walls are reduced to a minimum.

A water jacket and a steam drum are provided on the lower part of the producer, which serves to generate sufficient steam for the producer blast. Primary air is blown under pressure (10 to 12 in. H_2O) into the producer. Where a cold, clean producer gas is desired, the gases passing out of the producer are led through a waste-heat boiler and a further supply of steam is obtained in this manner.

Fuels containing up to 30 per cent ash are easily handled with this producer. With a fuel of 20 per cent ash containing a mixture of half and half breeze and small coke, the producer is claimed to deliver 88 per cent of the heat content of the fuel in the form of hot gas. The capacity of this producer is from 25 to 40 tons of fuel per 24 hr , depending on the kind of fuel and the size producer. It will be noticed that the heavy steam distributor is set eccentric to the shell and thus in rotating crushes any large lumps of clinker. This producer is somewhat similar in design to the Koppers Kerpeley.

European Gas-producer Types.—Three gas producers of European design—(1) by-product recovery producers, (2) slagging-ash producers, and (3) the Wollaston producer—are described to show some of the differences between European and American design although many of the European producers are similar in construction to those discussed above.

By-product Recovery Producers.—In addition to supplying producer gas it is possible so to operate a producer that nearly two-thirds of the nitrogen content of the original fuel is recoverable as ammonium sulfate. It is also possible to obtain considerable quantities of tar and tar products. The pioneer successful producer designed for ammonia recovery was built by Mond in 1883–1885. Since most by-product recovery producers are similar to the Mond design, the discussion of the type will be limited to a description of the Mond system.

In order to obtain large yields, the ammonia must be kept at relatively low temperatures, as otherwise the NH_3 is broken down into N_2 and H_2. In addition, a high H_2 content of the gas also tends to force back the dissociation of NH_3. Therefore, when operating to obtain the maximum ammonia recovery, a very large amount of steam is used with the air. This results in a large

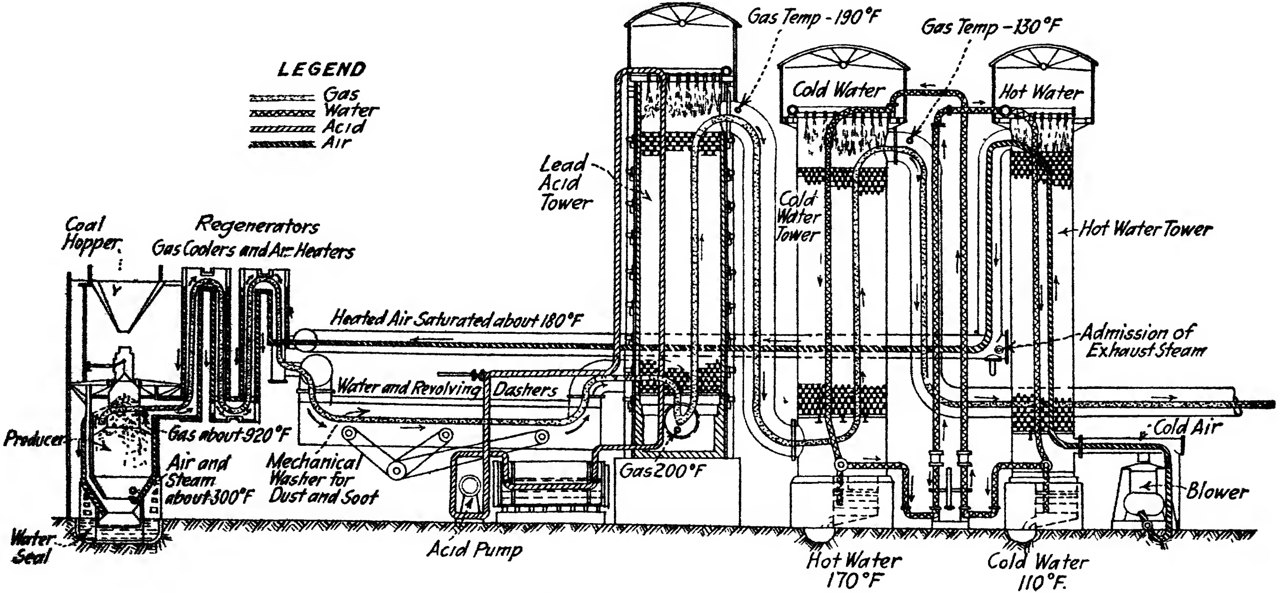

FIG. 231.—Mond by-product recovery gas producer. (*Courtesy of the American Gas Association*)

amount of undecomposed steam (see p. 567) in the producer gas, so that it is advisable to cool the gas and condense the water before using the gas in furnaces. The hot gas on leaving the producer therefore passes through a heat interchanger which preheats the entering air to about 480°F. (250°C.).

The preheated air-steam blast passes through an annular space around the producer (see Fig 231) and is further preheated before reaching the grate. The grate itself is in the form of an inverted truncated cone, with spaces for the admission of the air. These producers are run with very deep fuel beds and are provided with a large bell in the upper part of the producer to aid in the distillation of volatile matter from the coal.

As mentioned above, it is best to use a large amount of steam, the usual amount being about 2 lb. of steam to 1 lb. of coal gasified. According to Bone,[15] with coal containing 1.2 to 1 6 per cent nitrogen it is possible to recover from 1 ton of coal about 80 to 90 lb. of ammonium sulfate and to generate up to 150,000 cu. ft. of gas of 150 B t u. (high) per cu. ft.

In view of the large amount of steam used and the resultant low temperature of the fuel bed, the CO_2 content of the producer gas is very high, the representative analysis of Mond gas given by Bone being as shown in Table CXXXI.

TABLE CXXXI

ANALYSIS OF MOND PRODUCER GAS

	PER CENT
CO_2	17 0
CO	11 0
H_2	24 0
CH_4	3 0
N_2	45 0
	100 0
B t u per cu ft	144

As shown in Fig. 231 the hot gas leaving the producer (at a temperature of about 1000°F., or 540°C.) passes through heat interchangers or superheaters, where it gives up a portion of its sensible heat to the incoming air. The dust and a portion of the tar are removed by mechanical washers. The gas, which has thus been cooled to about 212°F. (100°C.), is then passed up a lead-lined tower in which the ammonia is absorbed by a mixture of sulfuric acid and ammonium sulfate. In passing through the cooling tower shown in the diagram the gas temperature is lowered and the hot water obtained is used in humidifying the entering air.

Producer plants of this type give a relatively low B.t.u. gas, occupy a large ground space, require considerable capital investment and are subject to high labor charges. Although successfully used in Europe before and during the war, the above disadvantages, coupled with the low market price of ammonium sulfate (resulting from the development of the synthetic ammonia processes), have brought about a gradual decrease in their use.

Slagging-ash Producers.—Since the ash from many coals fuses and clinkers at relatively low temperatures, and because the large quantities of steam required to prevent this clinkering result in a gas of low calorific value, high CO_2, and high moisture content, a number of producers have been developed to fuse the ash, withdrawing it in molten form. In these producers, which are similar in construction to a blast furnace, no steam is used and the gas produced is air-blown producer gas (see p. 544), consisting mainly of CO and N_2. Since the ash is withdrawn in the molten state, no trouble is caused from clinkering and very high rates of gasification are possible.

The S. F. H. producer, Fichet, Heurty and Sepulchre, and the Wurth producer have held first place among the grateless producers in which the ash is fluxed. The S. F. H. producer is similar to a small charcoal blast furnace with three to six water-cooled tuyères and two tapping holes for removing the fluid slag. The producer has a double charging bell for the fuel and the air is introduced at high pressure without steam, The fuel is fed in with limestone, sand, or blast-furnace slag to flux the ash. A thick fuel bed is maintained in order to preheat the fuel, thereby increasing the temperature in the combustion zone and making ash fusion easier. The producer gives a gas containing essentially CO and N_2 and works as well with lignite or brown coal as with hard coal and coke briquettes. Very high combustion rates (up to 200 lb. per sq. ft. per hr.) have been maintained.

The Würth and Georgs-Marienhuette producers are similar in construction and use the sensible heat in the gases to preheat the entering air, thereby raising the temperature of the combustion zone and increasing the thermal efficiency. The Georgs-Marienhuette producer (Fig. 232) also resembles a small blast furnace with water-cooled tuyères arranged axially. Blast-furnace slag is employed as the flux and since no steam is used the gas is essentially dry and consists chiefly of CO and N_2 with very small contents of H_2, CH_4, and CO_2. The blast-furnace slag melts quickly and dissolves the coal ash before it becomes pasty. It has been found that about 18 per cent of the coke is a suitable weight of slag to use for certain coals and the quantity of iron produced as a by-product varies from 0 5 to 1.25 tons per day. The gas temperature leaving the producer is about 1500°F.

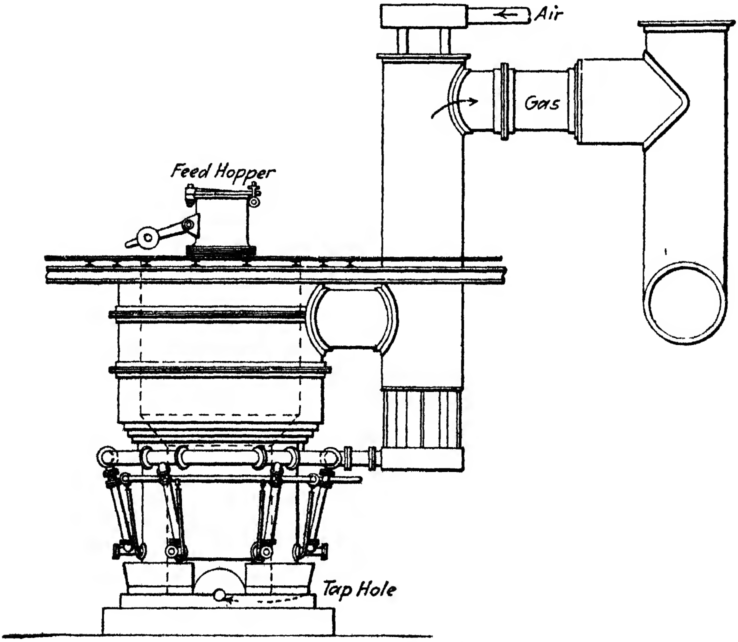

FIG. 232.—Georgs-Marienhuette slagging-ash type gas producer.

(815°C.)—200°F., or 110°C., higher than the Würth producer. Both the Wurth and the Georgs-Marienhuette producers work at high rates of gasification (with consequent high air and gas velocity) and considerable fuel (as high as 7 to 8 per cent at the higher rates) may be carried out mechanically with the escaping gas. Producers of this type take up but little room, the greatest diameter of a 100 ton per day producer being but 12.5 ft. The lining, if well constructed, will last about one year and after this time the supports and the bosch must be renewed.

The Wollaston Producer —A producer of recent English design embodying several rather novel features is shown in Fig 233. The coal is automatically fed into the retort *B* on the top of the producer. As the fuel falls on the top table *m* a curved plow gradually forces it to the edge, from which it is discharged to a cone-shaped plate *n* with a central discharge hole *e* below Th

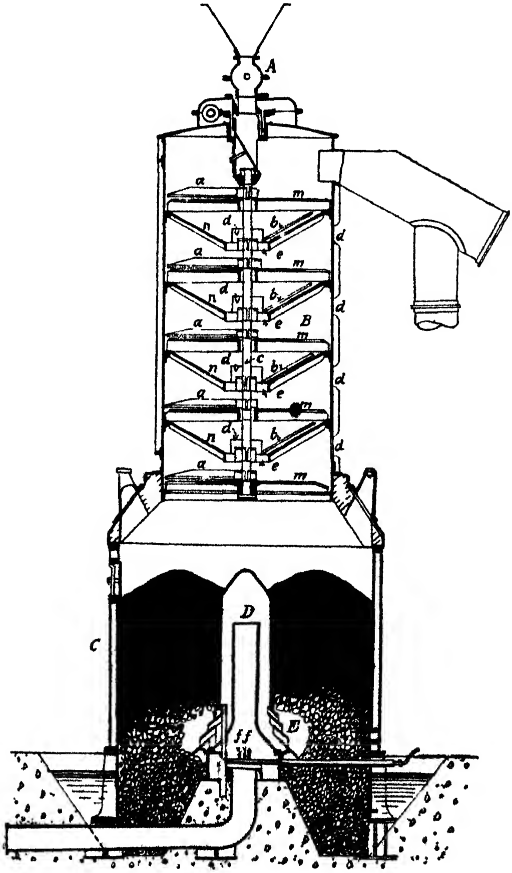

Fig. 233 —Wollaston gas producer.

incoming fuel is thus coked at low temperature by hot gas rising from the producer proper. The gas is enriched with tar and ammonia from the fuel. The central shaft carrying the pushers makes about six revolutions per hr. The average time (coking period) in the upper retort is 1¼ hr. The heating of the coal in the upper retort renders it non-caking and thus aids in the maintenance of a uniform producer bed with this type of fuel.

The producer shell consists of an annular steam boiler, which is claimed to give an evaporation of 1.5 lb. of water per lb. of coal gasified. The cast steel core *D* serves to preheat the entering air, evaporate the water admitted through the vaporizers *f*, and preheat the steam formed, thus aiding the shell boiler in maintaining low fuel-bed temperatures without the excessive use of steam. The ash is removed by rotating crusher rolls (not shown in Fig. 233) and falls into an ash pit, from which it may be removed by water circulation. This producer is claimed to be essentially free from clinker troubles (due to the cooling of the producer bed) and to be especially adapted for the recovery of by-products without giving a high CO_2 gas The gases are said to leave the producer at about 500°F. (260°C.), the cold efficiency being about 80 per cent.

Summary of Precautions Necessary for Efficient Producer Operation.—To maintain the steady production of a uniform, good producer gas the following precautions should be observed:

1. Coal should be fed continuously and uniformly over the fire bed.
2. Any channels that may form should be kept closed mechanically.
3. When poking to close blowholes or channels care must be taken to prevent mixing of ash with hot coal.
4. Ash should be removed uniformly and mechanically.
5. The air-steam mixture should be distributed evenly over the bottom of the fuel bed.
6. The amount of steam used should be under independent control in order to permit the maintenance of a constant Air: Steam ratio with varying load.

The Use of Oxygen-enriched Air in Producer-gas Manufacture.[16,17]—Producer gas as at present manufactured contains slightly over 50 per cent of nitrogen, which acts as a diluent, causing a low heating value and decreasing the possible flame temperature. The quality of the gas can, of course, be greatly improved by enriching the air blast with oxygen, and with the cheap separation of large quantities of oxygen from the atmosphere this process will undoubtedly come into use, although at present the question is mainly of theoretical interest. (At the present time development along this line is going on in at least one plant in the United States.)

A number of good gas coals have the following average proximate analysis (dry basis):

	Per Cent
Fixed carbon	60
Volatile matter	34
Ash	6
Total carbon	80
Heating value	14,240 B.t.u. per lb.

With these values as a basis and using the results of current gas-producer operation, Jeffries[17] has calculated that it will be possible to manufacture a gas of the following analysis by using oxygen in the producer instead of air:

TABLE CXXXII

Comparison of Producer Gas Made by Using Air and Oxygen

Constituent	Using air, per cent	Using oxygen. per cent
CO_2	3.5	3.0
C_2H_4	0.8	1.4
CH_4	2.7	7.6
H_2	12.0	28.0
CO	29.1	58:0
N_2	51.9	2.0
	100.0	100.0
High heating value (B.t.u. per cu. ft. at 60°F., 30 in. Hg, and saturated with water)	176 B.t.u.	383 B.t.u.

The table shows the great increase in heating value of the gas manufactured with oxygen instead of air. With oxygen it becomes possible to produce a gas of high heating value and high CO. The producer may be operated as a slagging-ash producer, or, sufficient steam may be used to keep the ash from clinkering, thus utilizing what practically amounts to a continuous water-gas process (see Chap. XIX). The calculated analyses of the gas produced by this latter method are shown in Fig. 234 (for the conditions outlined on the figure), it being assumed in each case that, as the per cent of oxygen in the atmosphere is increased, enough additional steam is admitted with the air blast to keep the temperature of the fuel bed constant. It will be seen from the figure that it is possible to produce a gas of exceptionally high heating value in this way. The extra heat generated by using oxygen can be converted into potential heat as H_2 and CO through the reduction of additional steam, thus making the apparatus, in effect, a continuous water gas generator. It may be necessary to operate such a producer as a slagging-ash type since all the oxygen will be consumed in the oxidation zone before any of the steam will be decomposed.

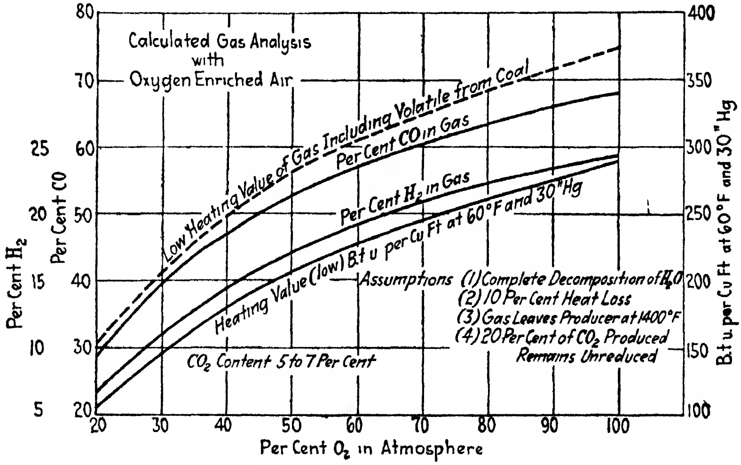

Fig. 234.—Effect of increasing oxygen in air on calculated composition and heating value of producer gas.

BIBLIOGRAPHY

1. Chapman: *Mech. Eng.*, **43,** 717 (1921).
2. Clement, Adams, and Haskins: U. S. Bur. Mines, *Bull.* 7 (1911).
3. H. S. Taylor: "Fuel Production and Utilization," D. Van Nostrand Co. (1920).
4. Gwosdz.: *Z. angew. Chem.*, **31,** 137 (1918).
5. Haslam, Hitchcock, and Rudow: *Ind. Eng. Chem.*, **15,** 115 (1923).
6. Haslam, Entwistle, and Gladding: *Ibid.*, **17** (1925). Also, Mackie and Reed: Mass. Inst. Tech., *M. S. Thesis* (1925).
7. Neuman: *Stahl u. Eisen*, **33,** 394 (1913); and *Z. Ver. deut. Ing.*, **58,** 1481, 1501 (1914).
8. Bone and Wheeler: *J. Iron Steel Inst.* (London), **76,** 126 (1907).
9. Clements: *Ibid.*, **107,** 97 (1923).
10. Haslam: *Ind. Eng. Chem.*, **16,** 782 (1924).

11. Hunt, Johnson, and Willis: Mass. Inst. Tech., *Thesis* (1923).
12. Terres and Schierenbeck: *Gas u. Wasserfach,* **67** (1924).
13. Husson: *Rev. ind. minèrale,* **2,** 373, 406 (1922). Translated and condensed by Haslam and Forrest: in *Blast Furnace Steel Plant* **11,** 264, 328 (1923).
14. Rambush: "Modern Gas Producers," Benn Brothers (1923).
15. Bone: "Coal and Its Scientific Uses," Longmans, Green & Co. (1922).
16. Cobb: *Gas J.* (June 8, 1920), p. 640.
17. Jeffries: "Gas Making with Cheap Oxygen," New Engl. Gas Assoc (1921).

CHAPTER XIX

WATER GAS AND OIL GAS

Water gas is the mixture of permanent gases obtained by the action of steam on incandescent carbon. It is, therefore, a constituent of all ordinary producer gas, of the gas from steamed coal gas retorts, and of the gases obtained from several processes of complete gasification and low-temperature carbonization. But aside from these, water gas is in itself one of the most important industrial gases, being widely used as such and after being enriched by high B.t.u. oil gas produced by the cracking of petroleum oil. The unenriched gas is called *blue water gas*, from the blue color of its flame. When it is enriched, it is called *carburetted water gas*. This chapter describes the properties of these gases and the methods employed in their manufacture. A discussion of *oil gas* (which is made by the cracking of petroleum oils and is similar to coal gas in composition) also has been included since considerable quantities of this gas are manufactured in localities where oil is cheap.

BLUE WATER GAS

Applications and Uses of Blue Water Gas.—Blue water gas is not a gas of low calorific intensity in spite of its low heating value (about 310 B.t.u. per cu. ft. gross). It is slightly superior to coal gas and natural gas in flame temperature, and the fraction of its heat available at high temperatures is at least the same. It shares with these gases in many cases the advantage of not requiring regeneration or recuperation in order to attain a high temperature.

The chief use of blue water gas is in the manufacture of carburetted water gas (see p. 618). It has, however, two* further important fields of use: (1) as an auxiliary to city gas plants making coal gas, and (2), as an industrial gas where high calorific intensity is essential.

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* The use of water gas as a raw material for the manufacture of wood alcohol, motor fuel and other commodities is just being developed and the future possibilities along this line are great (see p. 635).

1. *The Use of Blue Water Gas as an Auxiliary to City Gas Plants.* Blue water gas has three major disadvantages for use as city gas: (1) the distribution cost is greater than for the higher B.t.u. gases (though not in proportion to the reduction in B.t.u.); (2) the high rate of flame propagation would permit only a narrow range of adjustment of air shutters on domestic appliances without causing back-firing; furthermore, air shutters designed for present city gas would have to be readjusted and this readjustment would be troublesome; and (3) its high CO content and lack of odor carry an element of danger due to poisoning.

In spite of the disadvantages, blue water gas is widely used in England as an auxiliary to coal gas. English plants of this type use coke from the coal retorts, thus tending to make the profits independent of the state of the coke market. At the same time the capital expenditure is low per million feet of gas made per year, and the ground area required is moderate, often an important consideration in making extensions. However, the addition of water gas results in a heavier load on the mains, as already remarked, so that water gas used in this way is often lightly carburetted.

In this country blue water gas is being used to compensate for a diminishing natural-gas supply at Buffalo, N. Y., Elmira, N. Y., and in other cities. Just how much blue water gas is used as an addition to city gas throughout the world would be difficult to say. In England in 1920, for example, 44,000-000,000 cu. ft. of water gas were produced by public gas companies. See p. 648 for a comparison with coal gas.

2. *The Use of Blue Water Gas as an Industrial Fuel.*—Blue water gas is also used quite widely in cases where a manufacturing plant finds it advisable to produce for itself a gas of high calorific intensity (for welding, etc.). The plant involved is simple, is available in small units, and can be started and stopped easily. The purification offers no serious difficulty (since there is no tar), and the gas can be transported easily over plant distances. It requires, to be sure, the constant attendance of a gas maker. If purified from H_2S it is quite odorless. Like all gases containing large amounts of CO, water gas is poisonous. This last fact is one of the reasons against distributing blue gas as city gas, but this difficulty has been overcome by "perfuming" the gas with some distinctive odor—usually the liquids condensed from Pintsch or oil gas.

Though no accurate statistics are available, industrial water gas is certainly a fuel of secondary importance, the special conditions calling for its use being comparatively rare.

Formation of Blue Water Gas.—The primary reactions by which water gas is formed are reactions (6) and (7) of Chap. VIII (See pp. 162, 164 and 549):

$$C + H_2O = CO + H_2 - 70{,}900 \text{ B.t.u} \qquad (6)$$
$$C + 2H_2O = CO_2 + 2H_2 - 71{,}600 \text{ B t.u.} \qquad (7)$$

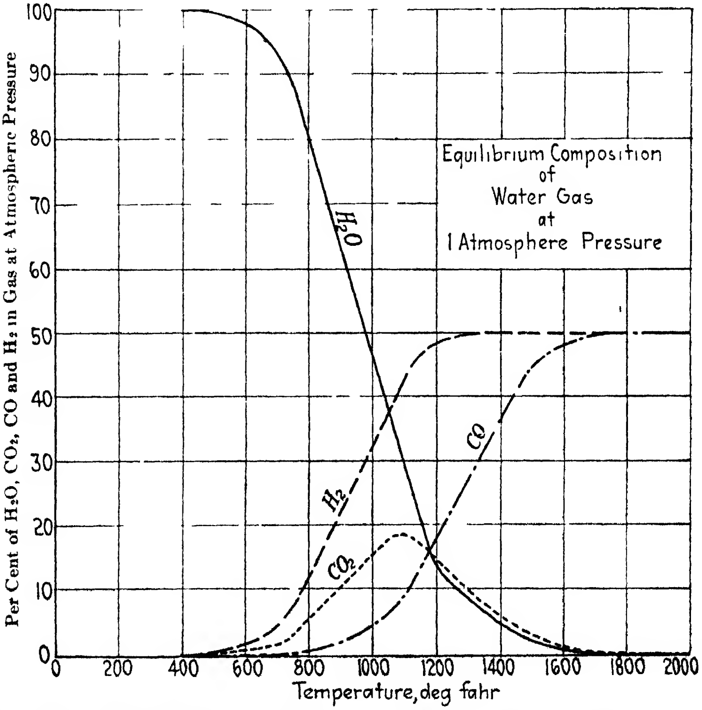

FIG 235 — The equilibrium composition of water gas at 1 atmosphere pressure and at various temperatures

For every temperature and pressure there exists one definite mixture of water gas and undecomposed steam for which the components are in equilibrium with respect to these two reactions. The equilibrium composition at one atmosphere and various temperatures as given by Juptner[1] is shown in Fig. 235.

At all temperatures at which coke will react with steam to any considerable extent (say above 1475°F. or 800°C.) the greater part of the steam will be decomposed if the steam, carbon, and gases come to equilibrium. In commercial practice, however, equilibrium is rarely approached.

Both of the steam reactions involve the absorption of a large amount of heat. Unless heat is provided from some source, an incandescent fuel bed through which steam is being blown will fall in temperature until finally no reaction whatever occurs. Early experimenters tried to add this heat to the fuel bed from external sources through the surrounding walls. It is quite impractical, however, with the present stage of development, to get enough heat into any large commercial apparatus in this manner, the heat transfer being too poor unless the temperature outside the wall is higher than is commercially possible with present materials of construction. The actual method adopted in practice is to stop the production of gas when the fuel-bed temperature has fallen to a fixed minimum, and then blow air through the bed, thus highly heating the bed and storing in the fuel and in the walls of the generator the heat necessary for the subsequent decomposition of additional steam. The operation of a water-gas generator is, therefore, an intermittent process; during one part of the cycle the generator functions like the fuel bed of a furnace or an air-blown producer; during the other part steam is decomposed to form water gas, which is collected separately.

The heat absorption of the steam-carbon reactions does not in itself represent a loss, since all this heat is recovered when the water gas is burned. In fact, it will be found that the heat of combustion of 1 mol of CO plus 1 mol of H_2 is greater by 70,900 cal. than the heat of combustion of one atom of carbon. This is just equal to the heat absorbed by the reaction

$$C + H_2O = CO + H_2.$$

A similar remark applies to reaction (7). This does not mean that the water-gas process is exceptionally efficient, but rather that the losses are due to other causes. The losses in the process are occasioned, for the most part, by the fact that the steam decomposition must take place at a highly elevated temperature.

The Composition of Blue Water Gas.—For reasons to be discussed later, almost all water gas, blue or carburetted, has in the past been made from fuels containing only a little volatile matter, that is, from coke or anthracite coal. When made for use as an industrial fuel gas, the following[2] may be regarded as a typical analysis, though the relative proportions of CO and CO_2 may vary considerably.

TABLE CXXXIII

Composition of Blue Water Gas

	Per Cent
CO_2	4.2
O_2	0.1
CO	41.4
H_2	49 2
CH_4	0 8
N_2	4 3
	100 0

Specific gravity (air = 1.0) = 0.55
Gross heating value = 307 B.t.u. per cu. ft.

The nitrogen comes from the incidental admixture of the products of combustion from the air blow, or, possibly, to some extent, from the nitrogen in the coke. The CH_4 is always present. It may arise from residual volatile matter in the coke, but it is more likely due to the interaction of CO and H_2 catalyzed by the coke or the ash in the fuel.

In addition to the constituents given, water gas always contains H_2S, the amount present depending largely on the sulfur in the coke or coal used. It is always present in smaller amounts than in coal gas, 1 gr. per cu. ft., or 0.16 per cent by volume being a typical figure.

Blue water gas always contains minute amounts of iron carbonyl, which is objectionable when used in connection with incandescent mantles, as it gradually deposits iron on the mantle. A more serious impurity is an exceedingly fine dust composed largely of silica and sulfur. The sulfur is supposed to be due to the interaction of H_2S and SO_2 and the silica to the decomposition of gaseous organic silicon compounds formed in small amounts from the ash. It deposits in the pipes and meters, and is very difficult to eliminate. Although sometimes troublesome, these impurities are only present to a slight extent.

Description of a Blue-water-gas Plant.—The elevation of a typical blue-water-gas set are shown in Fig. 236. It should be noted here that practically the entire discussion of blue-water-gas equipment (with the exception of changes referred to later) applies equally well to carburetted-water-gas sets. The generator is almost always a steel shell, lined with fire brick, which may vary from 3 to 10 ft. in internal diameter, depending on the capacity required. The sides are usually made plain and vertical, An iron charging door in the center of the top slides on a gas-tight ground seat, and is closed with a screw clamp. A double-door charging bell may be used, so that charging can be done without interrupting operations, but this does not give the same opportunity for the observation of the fuel bed, and in large machines it is difficult to secure proper distribution of the fuel. Its use is not common.

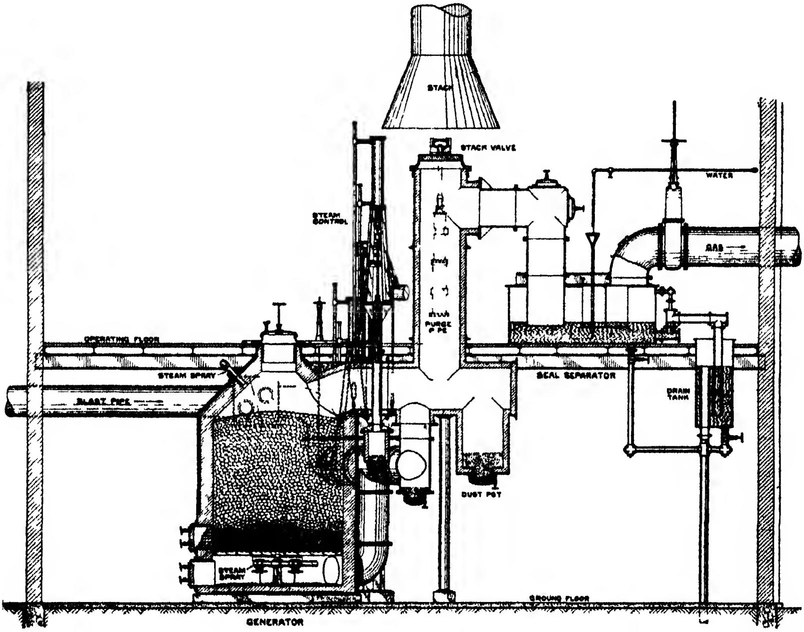

FIG. 236.—Blue-water-gas generator. (*The Gas Machinery Company*)

The fuel bed is usually supported on grates, which may be arranged to dump. In some small generators grates are omitted and the fuel lies on the floor, the bottom of the generator being wider to admit the air and steam around the lower part of the fuel bed. On generators of any size it is impossible to secure adequate distribution of steam and air in this way.

The cleaning doors, which usually swing on vertical axes, are on a level with the grate. They are made tight by asbestos packing and are closed with screw clamps. For small sets there may be only one door; on large sets there are usually four.

The generator has, of course, an opening below the grates for the admission of the air blast, and also a steam nozzle in the same space for the admission of steam. But practically all sets are arranged so that steam may be put through the fuel bed from top to bottom also. Hence there will be an additional steam nozzle in the top of the generator. Both nozzles must be provided with suitable quick operating valves. Two exits are provided for the products, one at the side of the top and one below the grate; these lead into a common pipe. One of these exits is always open, and only one, so that there is needed at the junction point the equivalent of a three-way valve. A great deal of ingenuity has been expended on the design of this valve, which is called the hot valve. Sometimes it is water sealed; again the one valve may be replaced by two valves; sometimes a single gate cuts off one or the other of the openings alternately Beyond the hot valve there is ordinarily a dust trap for catching the breeze and cinders blown along with the gases.

Beyond the dust trap the gas passage again divides. An exit is provided for the blast gases to escape in the chimney, this exit being closed by the stack valve. The other branch leads to the scrubber, where the end of the pipe dips an inch or two under the water in the bottom of the scrubber. Hence, when the stack valve is open, water rises in the dip pipe, and the gases escape to the air; when the stack valve is closed, the water gas escapes under the edge of the dip pipe and enters the scrubber.

In addition to acting as a check valve, the water in the seal serves to lower the temperature of the gases leaving the set. This lowering is caused by the evaporation of the water in the seal into the gas. The action is remarkably effective; with a dip pipe 30 in. in diameter, and dipping into the water only 2 in., the gases leaving a (carburetted) water-gas set at 1400°F. (760°C.) may be cooled to 170°F. (77°C.), though more than 300 lb. of mixed steam and water gas may be passing per min. Comparatively little heat is removed from the gas in this way, since the heat is merely transformed into latent heat of water. This heat is later removed in the various scrubbers and condensers in the purifying system.

The large increase in gas volume produced by the water evaporated from the seal has to be considered in designing the piping for this part of the plant, especially in the case of carburetted gas.

The air blast may be provided by a considerable variety of apparatus. Perhaps the favorite arrangement in this country is a centrifugal blower driven by a direct-connected steam turbine. Reciprocating engines also are used, as well as positive pressure blowers. Turbines have the great advantage that they can be partly shut off and started again simply by opening the steam valve.

In very small generators the valves may be operated by simple levers, but in sets of any size it is necessary to use a hand wheel and rachet arrangement. Most modern plants are equipped with hydraulic apparatus for opening and closing the valve. This produces quicker and more economical operation and greatly lightens the work of the gas maker. In the case of hydraulic apparatus the valves may be mechanically linked, so that the gas maker cannot operate the valves in a wrong order.

Operation of a Blue-water-gas Generator.—Consider the generator after the fire has been cleaned. It is first charged up to the proper level, the usual depth of fuel bed being 7 to 9 ft., though there are advocates of thinner beds. The stack valve is open. The gas maker opens the air valve, and air is blown from below through the bed. What happens has already been discussed in the chapter on the gas producer (Chap. XVIII). The air burns the coke in a narrow zone near the grates; this zone becomes intensely hot. Immediately above the zone of combustion the CO_2 is reduced partly to CO; above this point chemical action nearly ceases, and we have a zone where the coke is being preheated by the hot gases. At first, when the bed is still cool, the reduction is small; as the blow proceeds, the proportion of CO rises and if there is a pilot flame at the stack the gases will be observed to burn. The temperature of the exit gases must also rise. Hence after a certain length of time the heat stored in the fuel bed per lb. of coke burned falls off, and it becomes uneconomical to proceed with the air blast. The air valve is now shut, the stack closed, and steam is blown in below the grate.

As the steam passes up through the grates, it encounters first the intensely hot coke of the oxidation zone. Here it is highly heated and almost instantly and completely reduced to a good water gas, thereby absorbing heat very rapidly from the bed. Above this point it is cooled, thus serving further to warm the coke in the upper zones of the fuel bed.

This rapid extraction of heat from the lower layers of the bed causes a rapid fall of temperature at that point. The coke quickly becomes incapable of completely decomposing the steam. As has been pointed out on page 600, this inability is not because the equilibrium composition requires any considerable residual steam, but because there is not sufficient time of contact between coke and steam. The lower temperature and lower steam decomposition mean that less heat is being removed from the active zone; hence, as the run progresses, the decomposition falls off more slowly. The type of curve observed is illustrated[6] in Fig. 237

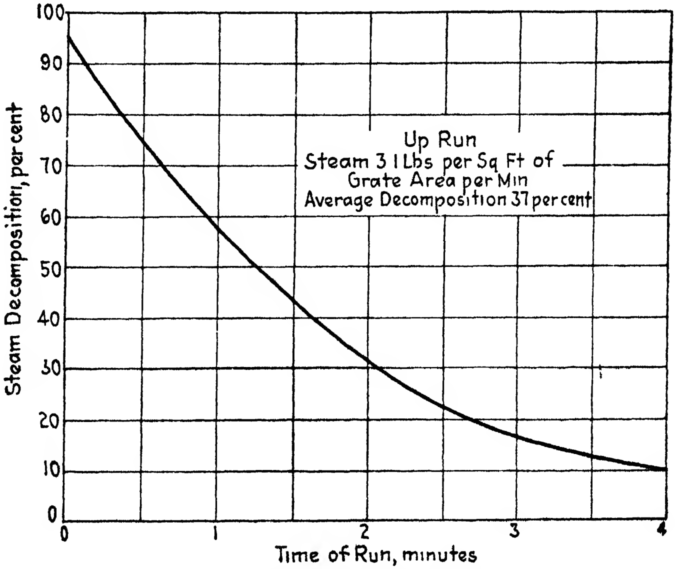

FIG 237 —The variation in steam decomposition during the up run on a blue water-gas set.

The CO_2 in the water gas made increases as the steam decomposition decreases. Thus the following was the composition of the gas made during successive thirds of a 6-min. run.

TABLE CXXXIV

COMPOSITION OF GASES DURING WATER-GAS RUNS

	First 2 min. per cent	Next 2 min. per cent	Last 2 min. per cent
CO_2	2.95	5.30	8.55
O_2	0.45	0.30	0.10
CO	45.60	39.50	34.20
H_2	44.95	51.20	53.00
CH_4	0.10	0.05	0.10
N_2	5.90	3.60	4.05
Total	99.95	99.95	100.00
Heating value, B.t.u. per cu. ft.	289.7	289.7	279.1

After the decomposition has declined to a certain extent, the losses of undecomposed steam, and the heat it carries off, become so large that it is uneconomical to proceed further. The steam is shut off, the stack valve opened, and the air valve opened. The position of the valves and the path of the gases are shown diagrammatically in Figs. 238 and 239 for the "air blow" and for the conditions just described, called the "up run."

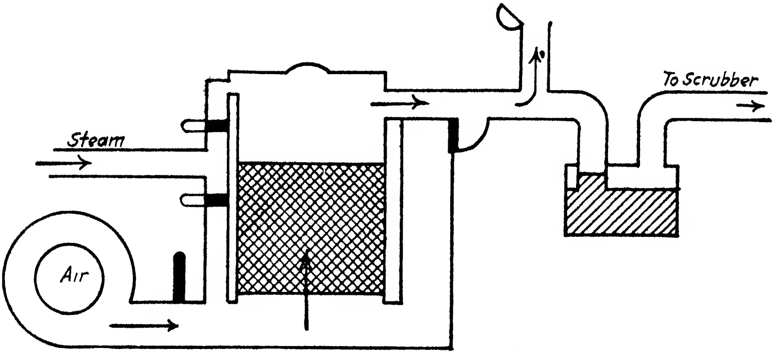

FIG. 238.—Path of gases and position of valves during the air flow of a water-gas set. (Diagrammatic.)

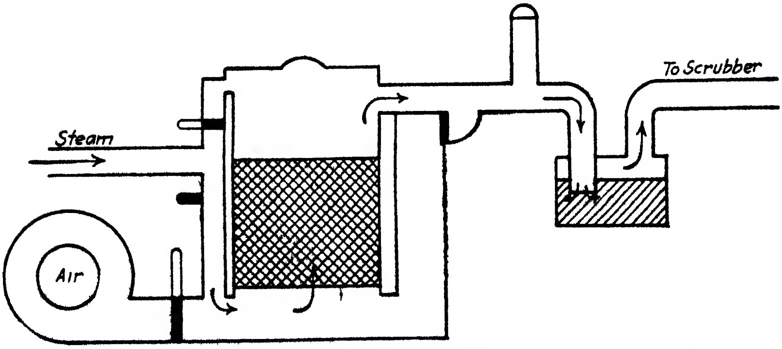

FIG. 239.—Path of gases and position of valves during the up run of a water-gas set. (Diagrammatic.)

The valves are changed on a time schedule, which is set as a result of experience, the main criteria of successful operation being good capacity and low CO_2 in the gas made, on the one hand, and low CO in the blast gases and an easily cleaned fire on the other. The former imply a high fuel-bed temperature, the latter a low temperature, so there is room for considerable divergence of opinion as to the best cycle. Of course, the size of the turbine available and the rate of steam supply will affect the relative duration of air-blow and steam-decomposition run. The whole cycle of blow and run will, in general, be not longer than 15 and not shorter than 4 min.

Automatic devices are now on the market for use with hydraulically operated generators. These will open and close the valves according to any predetermined schedule. They reduce the attention required of the gas maker and enable him to take care of more than one machine. Being automatic, they eliminate the variations due to the human element.

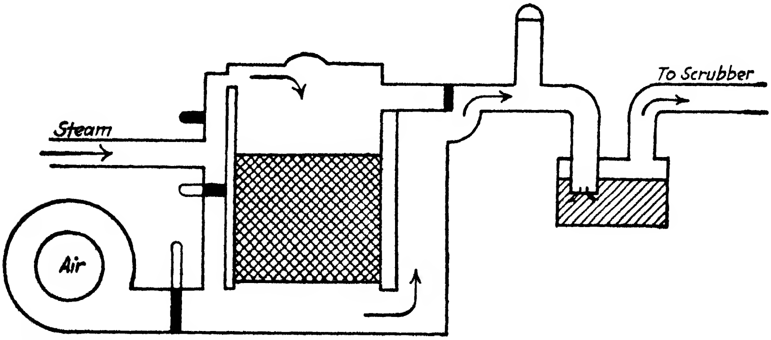

Fig. 240 —Path of gases and position of valves during the down run of a water-gas set (Diagrammatic.)

If the cycle of blow and run always proceeded in the way described, both steam and air being admitted below the grates, it is evident that the lower part of the active zone, where the ash is the highest, will be much chilled by the relatively cold steam, so that, unless the fuel is of exceptionally high grade, it will not be kindled again when the blast is readmitted. In this way the oxidation zone would tend to creep upward in the fuel bed, leaving behind an excessive amount of combustible that would be lost with the cinder. Besides, as one cycle followed another, the upper part of the fuel bed, which was cold when charged, would become quite hot, thereby causing a greater loss of energy as sensible heat and as CO in the gases leaving the generator.

Consequently, during part of the run, the steam is admitted at the top of the fuel bed, and passes down through it, the hot valve being changed accordingly. This is called the "down run." The path of the gases and the position of the valves during the down run are shown in Fig. 240. The down run is usually not made immediately after charging, since it is desirable to warm up the fresh coke somewhat. It is also common to omit it for a few runs before cleaning, in order that the bottom of the fuel bed may be cooled.

The down portion of the run cannot begin immediately after the blow. The water gas formed would mix with the air below the grate, and a dangerously explosive mixture would be formed so that some seconds of up run must precede the introduction of the down steam. Similarly, every run must end as an up run.

Another danger to be guarded against is the chance of getting any of the down-run water gas past the valve into the air line. Any failure of the air valve to seat, as by a bit of coke getting in it, may have disastrous consequences, and explosions in the air line are probably responsible for a good percentage of all gas-plant accidents. As a rule, the air line is provided with tees blanked off with heavy pasteboard. These serve as safety valves. In some cases a vent passes from behind the air valve to the outer air; the blower, while idling, blows sufficient air through this to carry off any gas that may get into the line. On the blow this vent is closed by a valve linked to the air valve (see Fig. 241).

The relative proportions of up and down steam are determined largely by the amount and the nature of the ash. The object is to keep the oxidation zone as low down as possible without damaging the grate, as in this way the clinker is kept in an accessible position. In the case of some cokes having a very large amount of fusible ash, too much down run produces a solid paste on the grates that chokes off the blast almost completely.

Water-gas generators almost always are charged discontinuously. The period between charges varies with the rate of running; every 45 to 75 min. would be usual intervals.

The cleaning of water-gas machines is one of the most important items in their operation, this being due to several factors. In the first place, the amount of cinder made is large. A generator with a powerful blower and good steam supply may consume a column of coke of its own diameter and more than 35 ft. high every 24 hr. In the second place, during the air blow the temperatures attained in the oxidation zone are very high, since there is no steam to cool the fire at this time. In the third place, during the run the fire is very rapidly cooled, so that there is no possibility of making the action slagging (*i.e.*, tapping out liquid slag). The result is the formation of a solid mass of clinker, enclosing pieces of partly burned coke, If the oxidation zone is not kept well down by sufficient down-run steam, the clinker tends to build out from the walls at a point which it is difficult to reach during cleaning.

Mechanical ash removers have not been very successful in the water-gas industry, though several types have been tried. The operation of the generator exposes any such apparatus to rapid alternations of temperature, and to steam, air, and water gas successively. To this must be added the condition of the ash. With cokes made from western coals, the ash is pasty when hot and hard when cold. Hence the work of cleaning is done mostly by hand. With typical cokes from midwestern coal, cleaning will be done by four to eight men. The procedure is to open the doors, drop the grates, and bar out the large pieces of clinker. The grates are then raised and the small material raked out from beneath them. Naturally, in this process some of the fuel above the clinker comes out with the clinker. This is raked up and charged back into the generator. With bad cokes this returned material may easily amount to 15 per cent of the new coke charged. As it is partly burned, and therefore of high ash content, it tends to aggravate the clinker difficulty. With such cokes the cleaning time may last an hour and a half, three times in 24 hr. (Both the length of cleaning time and the interval between cleanings vary widely under other circumstances.)

Of course, with cokes of better ash quality, conditions are much better (some anthracites do not form clinker at all), but it emphasizes the desirability of securing coke with low ash of high melting point, other things being equal. However, other things rarely are equal, and on a cost basis a poor coke may turn out to be the most economical. It would seem, however, that where a carbonization plant is run in connection with a water-gas plant this point is worth close consideration in buying coal. It should be remembered that blast-furnace and foundry men prefer a low melting ash, and that cokes for them are in this respect not well suited for water-gas plants.

Energy Distribution in the Blue-water-gas Process.—*Heat Balance of a Water-gas Set.*—In only a few industrial processes is the attainment of a heat balance so difficult as in the water-gas generator, primarily because the process is cyclical, the length of the cycle being a matter of minutes. During the 8 or 10 min. of one cycle, two different products are made—blue gas and blast gas—and the composition of these products, as well as their temperature, passes through a wide range of values. The large quantity of fuel always in the bed adds to the uncertainty of the tests. Accordingly, while the preparation of a heat balance is a commonplace part of an ordinary boiler test, a similar test on a water-gas generator is an elaborate affair, and there are very few reliable tests reported in the literature. The nature of the energy distribution may be seen from the following figures (Table CXXXV) taken from the Fuel Research Board[7] and from the Gas Investigation Committee of the (English) Institution of Gas Engineers.[2] To the balance are added some figures on temperatures which will be useful in comparing them with similar balances for carburetted water gas.

TABLE CXXXV

HEAT BALANCE OF WATER-GAS SET

	Fuel Research Board, per cent	Institution of Gas Engineers, per cent
Heat input:		
Coke	92.52	93.41
Steam	7.35	6.50
Air	0.13	0.09
	100.00	100.00
Heat output:		
Water-gas, heat of combustion (potential heat)	57.81	52.19
Water gas, sensible heat	3.23	2.04
Blow gas, heat of combustion (potential heat)	18.95	22.59
Blow gas, sensible heat	9.33	5.30
Clinker and unburned coke, heat of combustion (potential heat)	4.54	6.94
Dust, heat of combustion (potential heat)	0.33	
Sensible heat of clinker, unburned coke and dust	0.33	0.42
Uncondensed steam, sensible and latent heat	2.20	2.32
Radiation and unaccounted-for losses	3.28	8.20
Total	100.00	100.00
Approximate average temperature of blow gas	1280°F. (700°C.)	625°F. (330°C.)
Approximate average temperature of water gas	935°F. (503°C.)	690°F. (366°C.)

These tests are in some respects curiously different, but they will serve to bring out a few well-established facts.

1. A blue-water-gas plant will deliver, as water gas, from 50 to 60 per cent of the heat input, reckoning the heat both in coke and steam.

2. The percentage of heat input as steam is considerable, when the thermal losses in generating this steam are considered

3 Compared to many combustion processes, the loss in unburned fuel is large.

4. By far the most important loss of heat is in the potential and sensible heat in the blow gases, more than one-fourth of all the heat input being lost in this way

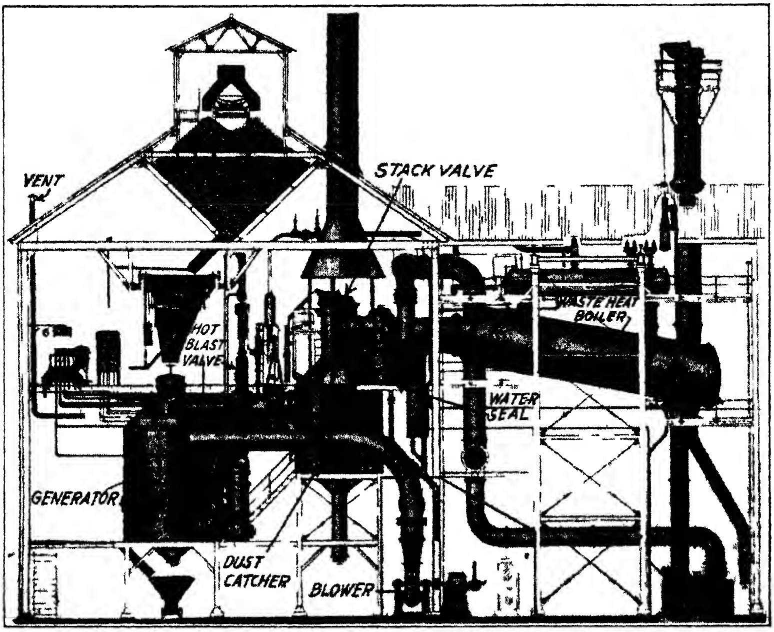

Fig 241 Blue-water-gas apparatus with waste-heat boiler (*U G I Contracting Company*)

Waste-heat Boilers.—The serious loss of heat in the blow gases (shown in Table CXXXV) suggests the desirability of installing waste-heat boilers, and boilers suitable for this purpose are available. One successful type is illustrated in Fig. 241 In order that the CO in the blast gases may be completely burned before they encounter the cold tubes, it is necessary to have a combustion chamber ahead of the boiler. This chamber will have a stack valve and gas off-take for use when the boiler is not to be employed. In carburetted-water-gas practice the carburetted gas is not passed through the boiler, as it would deposit a carbon layer that would be baked on the tubes during the blow and seriously diminish the capacity of the boiler. The type illustrated is a horizontal boiler, but vertical boilers are used also.

According to Morris,[8] a waste-heat boiler on a blue-water-gas set used for heating coke ovens generated 55 lb. of steam at 150 lb. pressure per 1000 cu. ft. of blue water gas made. This is equivalent to 12.5 per cent of the heat input to the machine. According to the *Seventh Report*, Institution of Gas Engineers,[5] a waste-heat boiler attached to a carburetted-water-gas plant evaporated an average of 73 lb. of water* from 70°F. (20°C.) to 95 lb. pressure per 1000 cu. ft. of carburetted water gas made. This latter figure corresponds to an efficiency in the boiler of 45 per cent; but the relatively low temperature of the entering gases would make an efficient boiler uneconomically large. Waste-heat boilers, in spite of the heat economy they make possible, are not widely employed in the United States, though they are gradually coming into more general use. Certain features of waste-heat boilers, which should be given consideration, are: (1) They make steam for part of the cycle only, so that they furnish a fluctuating supply. (2) They are larger and costlier than other boilers evaporating the same amount of water. (3) They complicate the operation of the generator, since the secondary air, the feed-water supply, etc. require attention. (4) The steam they raise usually supplants steam made by coal or coke breeze and not by the more costly lump coke; hence, a 20 per cent heat saving by a waste-heat boiler is *not* the cost equivalent of a 20 per cent coke saving, but something a good deal less.

Recuperators and Regenerators.—In principle, it would be better to use the heat in the blow gases to preheat the blowing air and superheat the steam, thereby saving coke instead of coal. Regenerators to accomplish this result require such bulky checkerwork, so many large valves subject to bad heat conditions, and such complicated operation that no such arrangement has been successful. The major problems in recuperator design for this purpose lie in selecting the proper materials of construc-

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* This figure for the amount of water evaporated in the waste-heat boiler is considerably higher than would be obtained in American practice. This is due to the fact that much more CO was allowed to escape during the Institution tests than is allowed in the United States.

tion to withstand the temperature fluctuations. It is possible that some of the more modern recuperators or regenerators, such as the rotary (see p. 421), will prove of value in water-gas work. A certain amount of the heat in the blow gases is recovered in the "back-run" system, to be described under carburetted water gas (p. 634).

Superheated Steam.—It is common practice to superheat the steam going to the generator in a separate furnace, thereby replacing a certain amount of coke by coal. An early attempt to do this failed because the varying steam demand caused too rapid changes of temperature in the tubes of the superheater, but this difficulty has been overcome and, although the fluctuating load makes attention necessary, superheated steam is now widely employed.

Accumulators.—So far as heat is concerned, the exhaust from a non-condensing engine (containing per lb. only 10 or 15 per cent less heat) is nearly as good as 150-lb. steam. For the generation of power in a condensing engine, however, such steam is only half as valuable as 150-lb. steam. It is obviously advantageous, therefore, to use exhaust steam, which has already been used for power, in water-gas generators. In many gas plants, moreover, power units are not run condensing. In such cases, to use the exhaust steam in the generators is nearly clear gain, subject only to the necessity of burning a little more coke in the set to make up for the reduction in total heat of the steam.

Unfortunately, the greatest single power consumer in the set is the blower engine, and this is producing exhaust steam just when it is not needed, whereas during the steam-decomposing run the steam supply is much reduced. Hence, exhaust steam is very little used, except in the larger plants where there are a number of machines—one being supplied with steam while another is blowing. Even here much live steam must be added at times, while at other times large amounts of exhaust steam are blown into the air.

The use of accumulators to remedy this condition is now exciting much interest. Essentially, an accumulator is nothing but a large tank of hot water, attached to the exhaust steam line. When there is an excess of exhaust steam, the pressure rises in the line and the exhaust steam condenses in the water, heating it. When the pressure falls, the water boils, supplying the deficiency of steam. Accumulators have been widely adopted in a number of industries and a half dozen or more are already in use in gas plants; more are being considered.

Principles of Blue-water-gas Operation.—*Proper Depth of Fuel Bed.*—The Dellwik-Fleischer system, devised in Germany, consists in using a generator of large grate area, a thin fuel bed (3 to 4 ft.), a high blast velocity, and a low steam velocity. The effect of varying the depth of fuel bed has been studied both by the Fuel Research Board[7] and by the Institution of Gas Engineers.[9] The main results of the Institution work are given in Table CXXXVI.

TABLE CXXXVI

EFFECT OF DEPTH OF FUEL BED ON WATER-GAS PRODUCTION

(*Institution of Gas Engineers*)

	Fuel-bed depth				
	4 ft.	4.25 ft.	5 ft.	6 ft.	7 ft.
Thermal efficiency, per cent.. ..	50.7	47.0	53.3	56.1	52.2
CO_2 in blow gas, per cent... ..	13.6	16.2	11.8	12.1	12.2
CO_2 in water gas, per cent... ..	4.7	8.2	5.5	4.5	4,2
Gas made per day, cubic feet..	364,400	307,700	564,900	614,400	564,200

The experiments of the Fuel Research Board resulted as shown in Table CXXXVII.

TABLE CXXXVII

EFFECT OF DEPTH OF FUEL BED ON WATER-GAS PRODUCTION

(*Fuel Research Board*)

	Fuel-bed depth			
	3 ft.	3.5 ft	4.0 ft.	7.0 ft.
Thermal efficiency, per cent............	54.7	57.8	59.9	57.8
CO_2 in blow gas, per cent..............	19.5	19.03	19.06	14.55
CO_2 in water gas, per cent............	8.30	5.30	6.00	4.50
Gas made per day, cubic feet..........	237,000	230,000	219,000	378,000

A study of these tables shows that very different results may be obtained with the same depth of fuel bed. The cause of these differences is undoubtedly a variation in the temperature of the fire, which, in turn, is affected mainly by the cycle of operations employed. The two tests at 4 and 4¼ ft. by the Institution show how difficult it is to draw safe deductions as to the effect of a single factor, even from the very best available data. It seems, however, from these results, that just as good gas can be made with a thin fuel bed (as in the Dellwik-Fleischer system), but that the tendency is toward more CO_2 both in the blow gas and in the water gas. The capacity is certainly reduced very materially with a thin fuel bed, but the efficiency falls off only slightly, if at all, the differences being within the range of variation in results.

Velocity of Blow Gas.—High efficiency in a water-gas generator requires as high CO_2 as possible in the blow gas. An increase in air velocity, by allowing a shorter time of contact in the reducing zone, has this effect. At the same time the heat transfer between the hot ascending gas and the cooler fuel in the upper layers of coke is nearly as good, since the increased turbulence almost makes up for the decreased time of contact. Hence efficiency is decidedly increased by a high blast velocity. At the same time, of course, capacity goes up. From both points of view, therefore, as high a rate of blowing as possible is desirable. On the other hand, the power required for blowing goes up approximately as the square of the blast rate, so that eventually a blast rate will be reached at which the saving in coke and increased capacity will be overbalanced by the loss in blower steam. Modern American practice, however, is toward the limit imposed by the tendency of the coke to be blown out of the generator, blast pressures of 30 to 40 in. of water column being in use. No actual test data on the effect of varying blast rates appear to be available.

Rate of Steam Supply.—Figure 242 shows the effect of halving the rate of steam supply during the course of an ordinary up run.[6] The steam rate was reduced 50 per cent after 1 min., restored to its original value after 2 min., and again reduced 50 per cent after 3 min. It is evident that a large reduction in steam rate produces only a small increase in decomposition, so that the net result is a large decrease in the rate of gas making. Strache[10] obtained a similar result. In practice, a balance must be struck. If the steam rate is very high, production will be great, but the losses due to undecomposed steam will be high. If the steam rate is very low, the production will be small and the radiation losses will be high. There is a great variation in practice in the amount of steam actually used. British practice seems to lead to the decomposition of about 70 per cent of the steam. In most American water-gas plants, the average decomposition probably is not far from 50 per cent.

Temperature of Fire.— The temperature of the fuel bed is controlled by the relative amounts of steam and air put in per cycle. If, other things remaining the same, the length of the blow is slightly increased, and the length of the run decreased, the blow will be putting more heat into the fire than before and the run will be taking less heat out. The temperatures of the fire, therefore, will increase. This will, as already explained, reduce the heat kept in the fire by the air, since the CO in the blast gases will rise. The heat taken out of the fuel bed by the steam will be increased, owing to the higher steam decomposition. Hence a new balance will be reached at a higher temperature. Under the new conditions CO in the blast gases (and the blast-gas temperature, to a small extent) will be higher, so that heat loss from this source increases; the loss from undecomposed steam will be

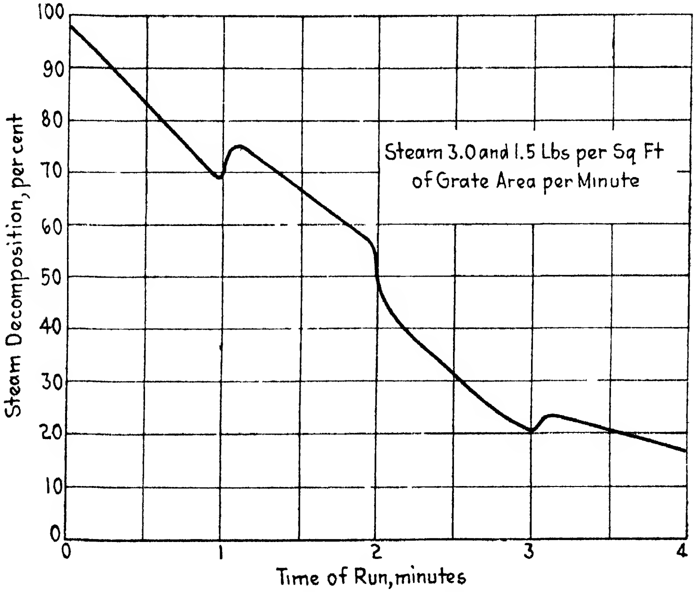

FIG. 242.— Effect of rate of steam supply on the formation of blue-water gas during the up run.

less; the production will be greater; the water gas made will be better, containing less CO_2; the radiation losses and clinker trouble will be probably a very little worse. If the relative amount of air is decreased, the reverse changes will occur. To determine the proper conditions, a balance should be struck. As a matter of fact, no rational tests on this point are ever made. The cycle is adjusted by trial and error to secure a good capacity, low CO_2 in the gas, and reasonable coke economy.

Due to fluctuation throughout the fuel bed, the temperature of the fire, as defined here, is a somewhat vague term, but a rise in temperature probably means that both the oxidizing and the primary (active) reducing zone rise in temperature throughout. There is no such thing as assigning one definite temperature as the fuel-bed temperature.

Length of Cycle.—The shorter the cycle of combined blow and run the smaller the variations in conditions, and the more nearly the optimum conditions may be maintained On the other hand, short cycles impose more labor on the gas maker, and if the cycle is very short there will be an appreciable loss in capacity, due to the time occupied in changing the valves; at the same time there will be an increased loss occasioned at the end of every run by the purging into the air of the blue water gas in the machine at the end of the steam-decomposing run. Similarly, the blast gas introduced into the water gas at the beginning of every run will increase (in percentage). Throughout the evolution of the industry the tendency has been toward shorter cycles. The hydraulically operated valves have been of assistance in this respect. With automatic operation the whole cycle may occupy only 4 min., or even less.

CARBURETTED WATER GAS

As has been said in Chap. XII, it is practically impossible to blow air through a fuel bed of commercial thickness without the production of a considerable proportion of CO. The heat content of this gas is utilized in the *carburetted*-water-gas generator. The water-gas generator proper is connected by a pipe to a large chamber full (usually) of checkerwork. In this the CO made during the air blast is burned with additional secondary air, the temperature of the chambers thus being maintained at about 1350°F. (732°C.). During the steam-decomposition run petroleum oil is sprayed into the blue water gas as it goes into these chambers and is cracked by the heat into lighter hydrocarbon compounds, which at ordinary temperatures are fixed gases or quite volatile liquids, which have a large number of heat units per cu. ft. These compounds are largely ethylene, propylene, benzene, and other unsaturated compounds which go under the name of "illuminants," since, when burned in a fishtail burner, they impart to the blue, non-luminous water-gas flame a high illuminating value. Blue water gas which has been enriched in this manner through the decomposition or cracking of oil is called *carburetted water gas.*

Applications of Carburetted Water Gas.—Carburetted water gas is used almost exclusively as a city gas, in the same way as coal gas, very many towns distributing mixtures of the two. The abundance of oil has made the United States the great field for this gas. In 1923 the production of carburetted water gas was 230,206 million cu. ft. as against a total of 379,590 million of all kinds of gas manufactured for public distribution. As compared with coal gas, carburetted water gas requires less ground area and less capital expenditure. A plant may be started up and shut down within a few hours at most, and may be kept under fire without very great expenditure of fuel. Hence, it requires less gas storage capacity, and carburetted-water-gas equipment gives a gas plant great flexibility both as to output and heating value.

Composition of Carburetted Water Gas.—With certain reservations, to be discussed later, carburetted water gas may be looked upon as a mixture of a rich oil gas and blue water gas. Its composition, therefore, will vary with the proportions of these two gases, that is, with the amount of enriching oil used. Indeed, one of the peculiar advantages of this gas to a city gas plant is the possibility of producing with the same equipment gas varying in heating value all the way from blue water gas at 300 B.t.u. up to a gas of 800 B.t.u. per cu. ft. or even higher, and making the change almost instantly. The heating-value standard most widely used in the United States is in the vicinity of 530 B.t.u. per cu. ft. (gross). The present tendency, however, is toward lower standards.

The "oil gas" is composed of hydrocarbons and it appears in the usual technical analysis as "methane" and "illuminants." Burrell, Seibert, and Robertson[4] give the values shown in Table CXXXVIII as a complete analysis of a carburetted water gas.

TABLE CXXXVIII

ANALYSIS OF CARBURETTED WATER GAS

(After Seibert and Robertson)

	PER CENT
CO_2	4.8
CO	29.8
H_2	32.0
CH_4	13.1
C_2H_6	2.9
C_3H_8	0.3
C_2H_4	9.8
C_3H_6	2.8
C_4H_8	1.7
Higher hydrocarbons (mostly benzene)	1.5
O_2	0.0
N_2	1.3
	100.0

This gas would have a gross heating value not far from 710 B.t.u. per cu. ft.—much higher than the general run of gas today. The *amount* of the hydrocarbons present will vary with the amount of oil used, but the relative *proportions* of the various compounds probably will not vary a great deal. Of course, the analysis just given must be looked upon as merely representative.

Some English tests[5] give the values shown in Table CXXXIX as the analysis of a gas of 483 B.t.u. per cu. ft. (gross).

TABLE CXXXIX

ANALYSIS OF CARBURETTED WATER GAS

(After Institution of Gas Engineers (English))

	PER CENT
CO_2	5.3
Illuminants	5.9
CO	34.7
H_2	38.3
CH_4	10.0
O_2	0.3
N_2	5.5
	100.0

These compositions are for purified gas, as distributed. Many plants use up to 4 or 5 per cent of air in the gas to aid in the removal of the H_2S by the iron oxide purifiers; in such cases the nitrogen content of the gas may be increased by 4 per cent, the oxygen also being higher.

Due to the sulfur in ordinary gas oils, the carburetted water gas will have a higher H_2S content than the blue gas from which it is made. It also will contain "organic sulfur" compounds (CS_2, COS, mercaptans) which are not present in uncarburetted gas, and which are not removed in the purifiers. With the advent of high-sulfur gas oils these compounds have become a serious problem.

Among the higher hydrocarbons whose vapors are present are the unsaturated compounds indene and styrene. These possess the property of polymerizing, especially under the influence of free oxygen and of metallic compounds, to hard resins. They are most abundant in the gas when the oil used has not been completely cracked. In such cases they tend to deposit in consumers' meters during cold weather and to harden seriously, thus interfering with the operation of the meter. Under normal conditions they do not cause trouble.[17]

Description of Carburetted-water-gas Plant.—A carburetted-water-gas apparatus is essentially a blue-water-gas apparatus in which one or more large chambers (usually filled with checkerbrick) have been inserted in the path of the gases between the dust catcher and the stack. Figure 243 shows a section through a typical plant. It consists of three cylindrical steel shells lined with fire brick, all of about the same diameter. The first, called the generator, contains the fire, and has exactly the same connections as the generator in a blue-water-gas set. From the three-way valve the gas passes to the top of the carburetor. This is filled nearly full of fire brick, called checkerbrick, laid in courses in the form of a sort of grid, no binder of any kind being used. In the top of the carburetor there is an oil spray through which the enriching oil is spread on the top of the checkerbrick during the run. Steam is sometimes used to atomize the oil, although in most plants the oil is simply admitted under high pressure. The oil atomizer often is constructed so that during the blow it is drawn up into a recess in the top of the carburetor, where it will be more or less protected from the hot gases, a small amount of steam running through it continuously to prevent clogging. The oil admitted is usually measured by a meter placed conveniently to the gas maker's board.

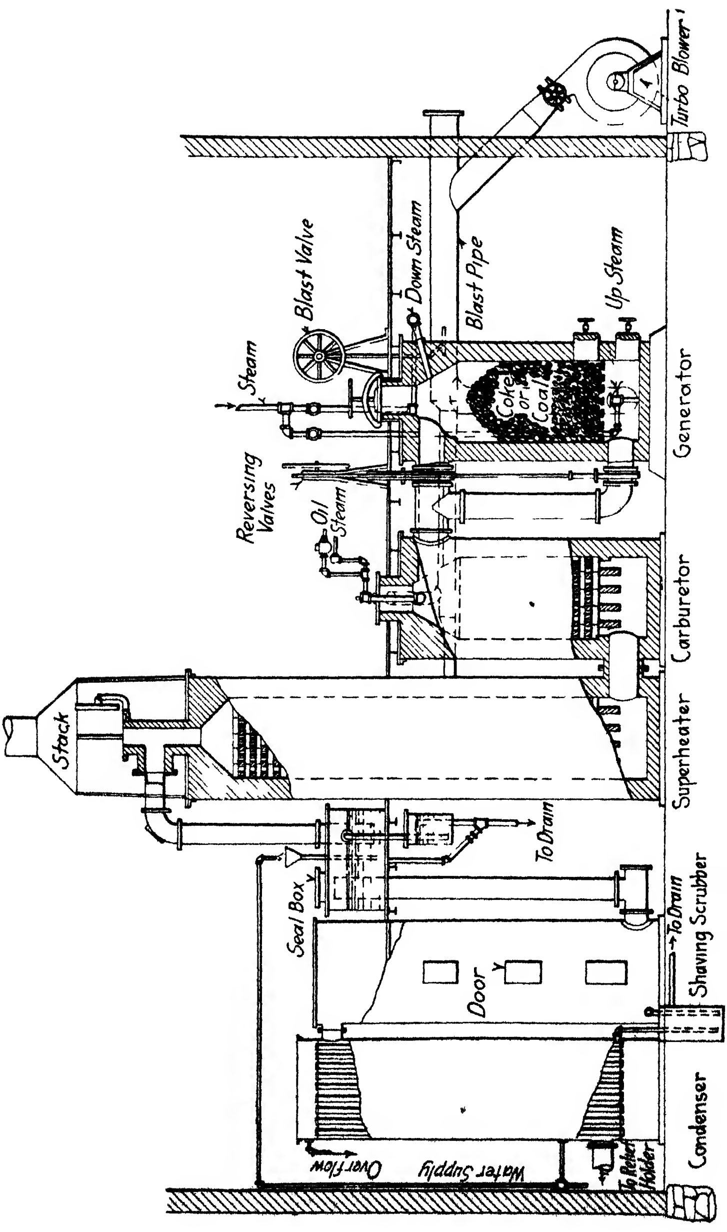

FIG 243 —Carbureted water-gas apparatus—diagrammatic layout

A branch line from the main air line leads either to the pipe leading to the carburetor, or else to the top of the carburetor itself. This admits the secondary air for burning the CO in the blast gases and is closed by a separate valve.

The superheater is similar to the carburetor, except that it is taller. This makes a positive draft which prevents gases from coming out the charging door of the water-gas generator. The stack is a short section set on the superheater; the gas off-take goes not to the scrubber, but to a separate seal pot, having a dip pipe as described for blue water gas. The tendency of the hot, tarry gas in the off-take to deposit carbon makes it necessary to have only a short, easily cleaned pipe at this point. Separate seal pots are also sometimes used on blue-water-gas sets.

Every water-gas set will contain one or two pyrometers, for controlling the temperature of the checkerbrick. These are base-metal thermocouples protected by a metal casing, and projecting 2 or 3 ft. into the shell. If there is one it will be at the base of the carburetor or the superheater. If there are two, the second will be near the top of the superheater. The indicating or recording device will be placed conveniently for the gas maker; if recording, the pyrometers furnish a valuable check on his efficiency.

The arrangement of the carburetted-water-gas apparatus in three shells (generator, carburetor, and superheater) is not necessary. In the Williamson apparatus the carburetor and the superheater are set on top of the generator, and are contained in a single large shell divided vertically into two hemicylinders by a brick wall. The generator is charged from side doors and the blue water gas enters the carburetor through a long, vertical pipe external to the shells. This arrangement saves floor space.

In the "twin machine" two generators are used, each generator being equipped with a single carburetor. During the blow the two generators are blown separately as in any other machine. The up runs are also made separately. In the down run, however, steam is admitted at the top of one carburetor, passes down through it and down through the fuel bed of the first generator. From the first generator the mixture of steam and partially formed water gas passes *up* through the bed of the second generator and up through the second carburetor. The enriching oil is added at the base of this second carburetor. The next time the steam is admitted at the top of the second carburetor and the action is reversed. The object of this arrangement is to provide a double depth of fuel bed and thus increase steam decomposition during the run. The *Kramer and Aarts process* is somewhat similar and is used to a certain extent for the production of blue water gas and carburetted water gas in Europe. In this process the fuel beds (5 to 6 ft. deep) of twin generators are blown in parallel and the gases burned in a regenerator consisting of two vertical shafts connected at the upper end and then into the atmosphere. Steam passes up through one generator, is superheated in passing up one side and down the other side of the regenerator, and then travels down through the fuel bed of the second generator. In making carburetted water gas by this process oil is admitted into the second generator. After the run, the blow is repeated; in the next run the steam is admitted to the *second* generator and the previous process is reversed.

Both of these variations are now mostly survivals and practically all new construction (in the United States, at least) is of the three-shell type.

Oil Cracking for Enrichment.—*Kinds of Oil Used.*—The oil used for enrichment is a petroleum oil. Originally, "gas oil" was a high-boiling petroleum distillate of gravity 35 to 40°Bé., distilling mostly between 200 and 700°F., with usually less than 2 per cent heavy oil residue. The development of oil-cracking processes for the production of gasoline made this oil more valuable as a source of motor fuel, and the products of corresponding boiling range that came on the market for gas making were highly unsaturated and much inferior as gas oil. The industry then turned to "topped crudes," that is, crude oils from which the gasoline and kerosenes have been removed by a simple distillation. These oils are heavy and viscous, boil from 300 to 750°F., and leave 5 to 10 per cent of coke in the still. Most of the gas oil now in use consists of such "topped crudes." The best gas oils belong to the paraffin series (see Chap. VI). The more unsaturated oils, in general, are not so good for gas-making purposes as the oils which are more nearly saturated. Thus the aromatics are inferior to the olefines. The use of either gas oil or topped crudes by the gas industry depends largely on the price of gasoline. When petroleum production is high and gasoline cheap, gas oil comes on the market. At times of high gasoline prices, gas oil is "cracked" to increase the supply of gasoline (see Chap. VI) and topped crudes are more generally used in gas making.

The Enriching Process.—Since the advent of heating-value standards for city gas (in place of candlepower standards) the primary aim in the enriching operation is to produce the greatest possible number of gaseous heat units from each gallon of oil, and, in addition, to produce the smallest amount possible of naphthalene, carbon, and gum-forming constituents.

The carburetting of water gas is a vapor-phase, thermal cracking process, in which heavy hydrocarbons are broken down to obtain lighter products which are fixed gases at ordinary temperatures. In its general character it is no doubt the same as the processes used in the oil industry, but it differs in several respects. In gas making, the process is carried further, as gases and not volatile liquids are wanted. The temperature is higher, because the process must take place in a shorter time, and to a greater extent. Also the cracking takes place in an atmosphere of a reactive gas. All the constituents of the water gas affect the cracking, but the hydrogen in particular combines with the oil vapors to the extent of several per cent and the steam reacts with carbon and tar to form blue gas.

The products formed by the carburetting process are threefold: gaseous, liquid, and solid. The gaseous products have already been described. The solid consists of a coke or carbon. The amount of this material formed varies greatly with the kind of oil and the conditions of operation, but it is usually between 5 and 15 per cent. The liquid is called water-gas tar. It is lighter than coal tar (specific gravity about 1.05 to 1.10), black, with less pitch when distilled than coal tar, and contains less free carbon. The constituents are less valuable than those of coal tar. Ordinarily, the tar formed will range from 20 to 30 per cent by weight of the oil used. The gas formed from the oil will represent, under modern conditions, a contribution to the total gas make of about 100,000 B.t.u. per gal. of oil—rather less than more.

If the temperature of the checker chambers is too high, objectionable amounts of naphthalene ($C_{10}H_8$) will be formed. This substance, though a solid, has an appreciable vapor pressure, and is capable of causing great trouble by depositing in the distribution system. If not too abundant it remains largely in the tar.

The tar has a good effect in that it removes from the gas the dust and the iron compounds which cause trouble with uncarburetted gas.

Factors Affecting Cracking.—*Temperature.*—The temperature fluctuations in the top of the carburetor are violent (from 1700 to 1100°F. (927 to 600°C.) or even lower), but this variation does not matter, because in this part of the brickwork the oil is merely vaporized. These fluctuations can only be allowed in the top of the carburetor, however, and for the successful operation of a carburetor and superheater very close regulation of the temperature at the latter end of the checker chambers is required. The temperature varies somewhat with conditions, but it is usually between 1350 and 1400°F. (732° and 760°C.). If the temperature is too high, more naphthalene and carbon are formed. If the temperature is too low, the heat contributed to the gas by the oil falls off, and at the same time the tar made contains uncracked oil, and forms an emulsion with the condensed water that is very difficult to separate; at low temperatures the gum-forming constituents, styrene and indene, also increase.

If the cost per B.t.u. contributed to the gas by the coke were much lower than for the oil, there would be an advantage in favor of the oil which gave an oil gas of the highest heating value per cu. ft., since less oil would be required to produce a gas of a given heating value. At present prices, however, this question is not important. The volume of gas produced by 1 gal. of oil is of the order of 50 to 70 cu. ft.

Effect of Time of Contact.—According to the experiments of Downing and Pohlmann,[12] times of contact ranging from 1¼ to 10 sec. varied the heat contributed by the oil to the gas only about 3 per cent.

Effect of the Brick Surfaces.—It is often assumed that the brick surface is the active agent (catalyst) in bringing about cracking. There is no direct proof of this, however. In the carburetor the brick is needed to store the heat for vaporizing the oil and raising the gas to the cracking temperature; but the cracking process proper proceeds almost without absorbing or giving off heat, and it is doubtful whether the superheater brick has any other function than to keep the gases at the proper temperature. The presence of the brick, of course, reduces the time of reaction.

A number of variations of checkerbrick arrangement have been tried—increasing or decreasing the ratio of brick to free volume, or changing from staggered to flue arrangement. These changes, however, have apparently not been made on any discoverable set of established principles, but merely according to individual experience, judgment, or guess.

Effect of Atmosphere.—The atmosphere in which the cracking is carried out has, as Downing and Pohlmann showed, a great effect on the products of cracking. Figure 244 shows the heat contributed per gal. of oil as determined in the laboratory.[12] As it happens, blue gas containing steam is as favorable as any of the atmospheres investigated.

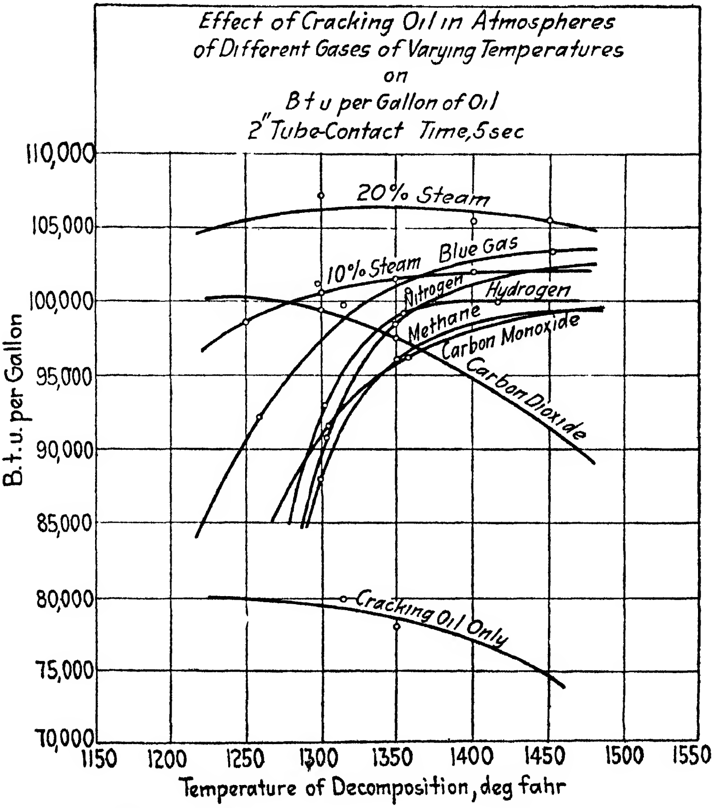

FIG. 244.—Effect of different gas atmospheres on cracking oil for water-gas enrichment. Improvement due to (1) flushing action of gases, (2) hydrogenation, and (3) formation of water gas from carbon and tar.

Operation of a Carburetted Water-gas Set.—At the beginning of the blow the gases contain practically no CO, so the secondary air valve is left closed. As the blow progresses, a slight blue flame of CO appears at the stack, and the gas maker gradually opens his secondary air valve so as just to prevent this flame from appearing. He is then operating at maximum efficiency, there being no CO and no excess air. The very intimate mixture of producer gas and air in the checker chambers makes it possible to attain almost theoretical combustion. Due to the high temperature in the brick, even a gas containing very little CO will ignite. If for any reason the checkerwork is exceptionally cold, as when first starting the machine, it may be necessary to wait until the percentage of CO is fairly high before opening the secondary air valve. Otherwise an explosive mixture may be formed which, when it does ignite, will cause considerable damage.

When the burning of the CO has raised the checkerbrick to the proper temperature, as determined by the pyrometer, the blow is ended, and the gas maker begins the run. As the protected pyrometer has a lag of the order of a minute, some experience is required to determine the proper time for stopping the blow. As soon as the steam is turned on, the gas maker will begin to admit the oil. As the oil strikes the hot brick at the top of the carburetor,* it flashes into vapor, which cracks in its passage through the checkerwork. In vaporizing, it leaves a layer of coke on the top of the carburetor. At the same time it violently cools the top checkerbrick. Hence, the distillation of the oil goes more and more slowly, and the oil drips further down into the carburetor. The gas maker, therefore, arranges to put in the predetermined amount of oil in about a minute less than the full time of the run, so that all the oil may vaporize and the oil gas be purged from the machine and not lost when the stack is opened.

On the subsequent blow, part (but only part) of the coke deposited in the carburetor is burned off. During cleaning and charging also the draft through the machine helps to remove this deposit. Much coke remains, however, and is mixed with the fragments of the top carburetor brick, which disintegrate under the severe temperature fluctuations they undergo. Thus the top of the carburetor gradually chokes up, and after a certain time this seriously reduces the capacity of the apparatus, and interferes with the proper cracking of the oil. It is necessary,

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* In addition to the oil vaporized at the top of the carburetor, all sprays make some mist and the streams of oil forcibly striking the brick make more vapor. This is carried down to the lower courses of brick.

therefore, to shut down the set, cool it, and rechecker the carburetor, mostly with new brick. The frequenty of recheckering varies—every 1000 hr. of running is a common practice. The superheater brick lasts much longer.

The poor quality of oil now used has made the problem of coke deposits a serious one. A recent development is the use of a carburetor without any checkerbrick. The oil is very finely atomized by an efficient spray, and the radiant heat of the walls is relied on to vaporize it. The superheater is checkered as usual. The coke formed collects in the bottom of the carburetor, where it may be hoed out easily. This process makes it possible to use cheaper oils. It is stated[13] that a number of plants are now operating in this way in the United States.

The Chrisman down-run process also has been developed to increase the capacity and efficiency of carburetted water gas sets (see Report of Water Gas Committee, Am. Gas Assoc., 1925). In this process the gas made on the down-run is taken from the bottom of the generator direct to the wash box (as when the reversed air blast system is employed). The gas made on the down-run leaves the generator at a lower temperature than either the up-run or the blow gases, and this temperature is below that necessary for efficient oil results in the carburetor. Since these relatively cool gases are bypassed to the wash box, they do not cool the checkerbrick in the carburetor and superheater, and less generator fuel need be burned to maintain the brickwork at the proper temperature for oil decomposition. Enriching oil is, of course, admitted only on the up-run. Since less heat must be absorbed by the carburetor and superheater during the blow, this part of the cycle may be shortened, thus increasing the capacity of the set by virtue of the greater number of runs possible per hour. The actual fuel savings observed with this system have been 2 to 3 lb. of generator fuel per 1000 cu. ft. of gas made.

Energy Distribution in Carburetted-water-gas Manufacture.—The production of carburetted water gas is really a double process—the production of blue water gas and the cracking of oil. However, since the blue-water-gas process must supply the heat for the checkerbrick, and since the oil reacts with the water gas, it is not practicable to separate these processes and present heat balances for each. However, it may be said, in general, that the cracking of gas oil is comparatively an efficient process, so that the average efficiency for carburetted water gas is higher than for blue water gas. Two sets of results are presented in Table CXL. One series of values was obtained by Evans[11] for a rich gas made with anthracite on a 6-ft external diameter set. The other figures are from tests by the Institution of Gas Engineers[5] made on a 7- or 8-ft. internal diameter set, in which a relatively lean gas was produced from coke.

TABLE CXL

HEAT BALANCE OF CARBURETTED-WATER-GAS SET

	Seventh Report, per cent	Evans, per cent
Input		
Heat of combustion of coke or anthracite	57.3	40.1
Heat of combustion of oil	38.5	55.8
Heat in air	0.1	0.2
Total heat of steam	4.1	3.9
	100.0	100.0
Output		
Heat of combustion (potential) of gas made	60.2	68.1
Sensible heat in gas made	4.0	3.3
Total heat of undecomposed steam	1.9	4.0
Heat of combustion of tar	6.5	8.8
Sensible and latent heat of tar	0.3	0.5
Sensible heat, blow gases	12.9	6.6
Heat of combustion, blow gases	6.3	0.1
Heat of combustion of clinkers, dust	4.0	1.5
Sensible heat in same	0.2	0.1
Undetermined (radiation, carbon deposits, etc.)	3.7	7.0
	100.0	100.0
B.t.u. of gas made per cu. ft.	485 B.t.u.	656 B.t.u.
Temperature of blast gases leaving set	1340°F (727°C.)	1280°F (693°C.)
Temperature of gas leaving set	1330°F (721°C.)	1280°F (693°C.)

The greater thermal efficiency of the oil-cracking process is obvious. It may be remarked that the heat of combustion in the blow gases, as observed in the Institution tests, is more than American practice would allow, the per cent of CO in the blow gases being 3.3.

Principles of Carburetted-water-gas Operation.—It is a cardinal principle, in American plants at least, that no CO flame shall be visible at the stack of a well-operated machine, except momentarily at the beginning and end of the blow. The reasoning on which this is based is obvious. If the blow is continued after the checkerbrick is at proper temperature, a large portion of the heat in the coke burned will escape at the stack, since the CO in the gases is high at the end of the blow. Also, the gases escape at checkerbrick temperature, so that a further large loss as sensible heat results.

Admitting the principle, several important consequences follow. In the first place, it is not possible to operate a carburetted-water-gas apparatus on an arbitrary time schedule, with fixed steam and air rates, for if all these things are fixed, the temperature of the fire is fixed, and, therefore, also the amount of CO obtained on the blast and the amount of water gas made on the run. Since the heating value of the gas is fixed, the oil to be admitted is fixed, and hence the heat to be abstracted from the checkerwork on each run is fixed. If, therefore, the fixed schedule does not happen to provide exactly the right fire temperature, there will be either too much or too little CO to provide the heat needed in the brick. Too little is not allowable, for the checker temperatures must be maintained. There must not be too much, by the principle given. Hence either the time schedule, or the steam or air rates, must be adjusted until there is a balance between the CO made and the oil consumed.

It follows from the above (with fixed steam and air rates) that for every heating value of gas made there corresponds a certain fuel-bed (fire) temperature, and that the higher the heating value the higher the fuel-bed temperature. The more oil is used per cu. ft. of gas made the more CO will be required in the blast gases, and this means a higher fire temperature.

The higher the heating value of the gas made the more oil will be used in the set; the more oil is used the more CO must be made on the blow to reheat the brick; to make more CO requires a higher fire temperature; a higher fire temperature means a better steam decomposition and a higher capacity for making water gas. If the heating-value standard of a carburetted water gas is raised, the plant making the gas will be capable of making a larger volume of gas, for two reasons: First, the blue-water-gas generating capacity goes up for the reasons stated above. Second, the increased oil input also increases the oil gas made. The production in heat units will, of course, be increased even more than the production as measured by volume.[6]

Another peculiarity of the carburetted-water-gas apparatus is the constancy of the sensible-heat losses. Both blow gas and water gas are discharged substantially at checkerbrick temperature. It is, therefore, immaterial, from this point of view, at what temperature these gases leave the generator.

Altogether, the variations which it is possible to produce in the operation of a carburetted-water-gas apparatus are far more restricted than in the case of the uncarburetted gas.

Coal as a Generator Fuel.[15] Since bituminous coal is decidedly cheaper than coke, especially in the Middle West, it would be advantageous as a substitute for coke, even if the efficiency were somewhat lower In recent years a great deal of work has been done along this line, and there are now quite a number of plants operating solely on coal, and others using mixtures of coal and coke.

The difficulties in the use of coal arise from two causes—the coking of the coal and the volatile matter, with the dust from the coal as a further source of trouble The tendency of the coal to cake, in the deep fuel beds used, makes it difficult to put through the necessary air and steam, and increases the tendency to form holes in the fire Accordingly, it is always found that the use of coal alone decreases the capacity of the generator to make water gas. Smaller amounts of coal may be mixed with coke without much trouble in this respect.

In so far as the volatile matter in the coal distills off during the run, it is advantageous, reducing the amount of oil required. But the distillation goes on during the blow also. It is quite impossible to utilize this coal gas by burning it in the checkerbrick. Every per cent of coal gas in the blast gases would mean that some 2 per cent CO may be dispensed with. This means that, from a thermal-balance standpoint, the temperature of the fire can be reduced; and experience shows that the reduction is so great that the fuel becomes incapable of decomposing steam. If burned in the air, the blast gases produce more or less smoke, which is objectionable in cities.

To overcome these difficulties, the "blow-run" method of operation has been devised. This consists in closing the stack valve toward the end of the blow, and discharging the products (producer gas, since the *secondary* air will be turned off) into the water-gas system for a fraction of a minute. This procedure makes the efficiency of utilization of the coal* very good indeed. At the same time, since the producer gas has a low heating value, more oil will be required. At the present prices for oil this is not a great disadvantage. The capacity of the apparatus is very much increased, and the smoke nuisance reduced. This procedure is only applicable, however, where the pressure of the blow gases is greater than the pressure in the gas line.[11]

The "reverse air blast process" (see Howard, Am. Gas Assoc. Monthly, October 1925) was also devised especially for improving the operation of water-gas sets using bituminous coal as fuel. As the name of the process implies, during part of the blow the air blast passes *down* through the fire. The cycle is as follows: (1) up blast, (2) reversed blast, (3) up-run (all the oil is admitted during this period), (4) down-run, (5) up-run, and (6) air purge. The reversed blast replaces the blow run in the ordinary cycle, air being admitted into the top of the superheater, passing up through the carburetor and then down through the fuel bed. The reversed blast gases pass directly from the bottom of the generator to the wash box or scrubber. During the down-run the blue gas also passes directly to the wash box through the same pipe (called the "bypass pipe"). In this process the hot valve is eliminated, two reversing valves being located in the wash box.

When the reversed blast is used the up blow is so short that the lower part of the fuel bed is not heated to the very high temperatures attained in ordinary practice. During the reversed blast, combustion takes place in the upper part of the fuel bed, and the reduction of the resulting CO_2 by the fuel lower down in the generator absorbs heat, the net effect being a lower maximum temperature, a higher average temperature, and a more uniform distribution of heat. Results from the initial installations seem to indicate the following advantages: (1) the capacity is increased, (2) the generator fuel is reduced, (3) there is a slight reduction in gas oil, (4) the life of the checkerbrick is prolonged, and (5) on account of the lower temperatures at the base of the fire, the clinker is soft, brittle and easily removed. The set has

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* This system is also quite generally used with coke as well as coal. Furthermore, the "blow purge" is also being used. That is, the stack valve is not raised until some time after the air valve is opened, and air has therefore started to go through the set at the beginning of the blow. In this way the water gas which is left in the set at the end of the run is blown into the holder instead of being wasted into the air.

more valves, however, and there is a slow corrosion of the pipes and valves through which the reverse blast gases are passed.

Another new departure of considerable importance in sets using coal as fuel is called the *Pier* process (see Report of Water Gas Committee, Am. Gas Assoc., October 1925). It has been found that in the fuel bed of a bituminous coal fired water-gas set, activity is confined to a ring extending from 1½ ft. to 2 ft. from the generator walls toward the center, and that the fuel in the center is only partially consumed. This condition is caused by the resistance to air flow offered by coked coal in the center of the charge, and by the relatively free passage of air up along the generator walls, where the larger pieces of coal tend to collect. This not only results in a loss of unburned fuel at the center, but also seriously lowers the capacity and causes excessive clinker formation in the hot active ring near the generator walls. Moreover, in attempting to force the generator, the blast velocity along the walls is unduly high, and considerable quantities of fuel are blown over into the carburetor. The Pier process (a refractory cylindrical pier built in the center of the generator) has been adopted to remedy this situation. The insertion of a pier cores out the inactive section of the fire, gives additional radiating wall surface and results in a fuel bed of practically uniform temperature across horizontal sections. The use of the process with Illinois and other bituminous coals has indicated the following advantages: (1) the gas making capacity is materially increased, (2) the fuel consumption is reduced, (3) the clinkering time is reduced by 30 to 40 per cent, (4) oil consumption is decreased because of the better quality of blue gas, and (5) less fuel is blown over.

Back-run Process.—A recent development of much interest, now operating in a good many plants with success, is the back-run process. In this process the down-run steam is put in, not at the top of the generators, but at the top of the superheater. It passes down through the superheater, up through the carburetor, into the top of the generator, and out of the bottom of the generator through an additional back-run line to the seal pot. There is no three-way valve, but valves are placed in the back-run line and in the regular gas off-take. In passing through the superheater the down-run steam cools the upper courses of brick. Hence on the subsequent blow the gases leave the machine at a lower temperature, effecting a saving in generator fuel. Also, since the down-run steam does not pass through the checkerwork, it also leaves the apparatus at a lower temperature, effecting a further saving. The expensive and troublesome three-way valve is also done away with.

It is necessary, in this process, to put in the carburetting oil during the up-run periods. This makes for a slightly lower oil efficiency. Actual tests, however, seem to indicate that the loss is not serious, and that a saving in generator fuel is obtained, together with an increase in capacity. Actual tests show a reduction of from 1½ to 3 lb. (out of perhaps 30 lb.) of generator fuel per 1000 cu. ft.

Enrichment by the Catalytic Production of Methane from Blue Water-gas.—The prospect that the price of enriching oil would become very much higher in the future has led to a great deal of work on the production of methane from blue water gas. With the aid of catalysts (usually metallic nickel on a porous support) various reactions producing methane occur. The principal reactions are:

$$CO + 3H_2 = CH_4 + H_2O + 107{,}700 \text{ B.t.u.} \qquad (1)$$
$$2CO + 2H_2 = CH_4 + CO_2 + 107{,}000 \text{ B.t.u.} \qquad (2)$$
$$CO_2 + 4H_2 = CH_4 + 2H_2O + 108{,}400 \text{ B.t.u.} \qquad (3)$$

Of these reactions (1) and (3) would be the most desirable, since the water produced will condense from the mixture, thus improving the enrichment. In practice, however, reaction (2) predominates, when blue gas of normal composition is used.

Though the equilibrium for these reactions becomes better at low temperatures, they do not proceed at measurable rates, even with the catalyst, below 392°F. (200°C.). As the temperature is raised, the rate becomes sufficient for commercial operation; at the same time, however, the reaction

$$2CO = CO_2 + C \qquad (4)$$

is also catalyzed by the nickel, and at too high a temperature this reaction begins to deposit carbon on the catalyst, injuring its activity. As a result, the reaction must take place in a very narrow range of temperature in the neighborhood of 620°F. (325°C.).[18]

It will be noted that all the reactions involved are exothermic. Since the heat is evolved at a low temperature, it would not be practical to recover it, and the enrichment would, therefore, involve the loss of some 20 per cent of the heating value of the gas converted to methane.

It has been stated that the nickel catalyst is very sensitive to poisoning, especially with organic sulphur compounds. How serious this would be in commercial operations is uncertain.

The large amount of CO_2 introduced into the gas is an objection; but this could be removed, and the enriching effect much increased, by scrubbing with water under pressure.

At the present prices for gas oil, this process does not seem to have any chance for commercial success. It has been estimated[18] that for material costs alone the catalytic process (with scrubbing) with coke at $9 a ton corresponds to an oil price of 6.9 cts. per gal. If the cost of catalyst, of scrubbing, and of heating to reaction temperature be considered, it will be seen that the method can have commercial significance only if the price of oil rises above present levels.

Blue Water Gas as a Raw Material for Liquid Fuels and Chemicals.—During recent months articles[19] [20] have appeared in the literature and patents have been taken out tending to show that in the future blue water gas may be the raw material for chemicals such as wood alcohol (methanol), as well as for liquid fuels suitable for the automobile. For example, water gas can be used to manufacture wood alcohol (methanol) in accordance with the following reaction by compressing the gas to 2000 to 3000 lb. per sq. inch and passing the mixture over zinc oxide as a catalyst at a temperature of 750°F.

$$CO + 2H_2 = CH_3OH.$$

If the same process is carried out with iron filings impregnated with a strong base, such as potassium carbonate, as a catalyst instead of zinc oxide, then the product is a mixture of higher alcohols called Synthol by Franz Fischer. Approximately 97 per cent of the mixture of these higher alcohols distills off at temperatures below 392°F. (200°C.), thus making this material of interest as a motor fuel, especially since it should have a higher thermal efficiency than gasoline, due to the anti-detonating effect of alcohols. The properties of that fraction of Synthol that distills below 392°F. (200°C.) are as follows:[19]

Color: Light yellow, not darkening on standing.

Specific gravity 20°/4°C. = 0.8289.

Carbon, 69.3 per cent; hydrogen, 12.3 per cent.

Higher heating value: 14,750 B.t.u. per lb.
Lower heating value: 13,570 B.t.u. per lb.
Thickening point: Non-viscous at −22°F.
Boiling range:

Temperature, degrees Fahrenheit	144 176	212	248	284	320	356	392
Per cent	11	37	64	76	84	91	97

It is interesting to note that if hydrocarbons are desired from this type of a process equal proportions of H_2 and CO are used, thus:

$$2H_2 + 2CO = CH_4 + CO_2.$$

An excess of hydrogen over CO causes the alcohols to be formed. A variety of compounds, such as methane and higher hydrocarbons, ethylene, methyl alcohol, and the higher alcohols, have been found by these processes, all of which show the possibilities of this type of synthesis. It would seem as if the gas works, with water-gas equipment already installed, could make use of some of these processes to even out their fluctuations in load.

OIL GAS

Oil gas is made by the thermal decomposition (cracking) of petroleum oils and is very much like coal gas in composition. Considerable quantities of this fuel, which is essentially the gas used for the enrichment of carburetted water gas, are produced in localities where oil is cheap and coal or coke expensive, or in localities where it is advisable to safeguard against uncertainties in solid-fuel delivery. Oil gas is made either by the checkerwork process (such as the Jones) or, in very small plants, by the Dayton process. In the Jones process a mixture of oil and steam is passed through generators filled with checkerbrick, the brick having previously been heated by the combustion of an air-oil spray. In the Dayton process the heat necessary for the formation of the oil gas is supplied by the combustion of part of the oil inside a gas-making retort.

Applications of Oil Gas.—Although employed to a limited extent for industrial heating, the main outlet for oil gas is in place of coal gas for the distribution through city mains. On the

Pacific Coast, for example, where oil is the only cheap fuel, large quantities of oil gas are used as city gas. In the eastern United States, however, oil prices are higher and oil gas cannot compete with coal gas or carburetted water gas except in localities near tidewater, (where relatively cheap oil is available). Oil gas manufacture has many advantages as a stand-by process to meet peak loads.

The advantages of oil gas are: (1) the plant can be built in such large units that labor charges can be greatly reduced, (2) there is no trouble from clinkering, (3) the heating value of the gas produced can be varied at will between quite wide limits, (4) on account of the lower gravity of oil gas from the Jones process (as compared with water gas), the capacity of the distribution system may be somewhat increased, and (5) both the Jones process and the Dayton process can be relied upon to produce gas practically continuously and therefore smaller holders are required.

There are certain disadvantages, however, in using oil gas to replace a city gas. In the first place, as pointed out for blue water gas on page 599, the air shutter adjustment for present city gas in most localities is different from the adjustment required with oil gas. The necessary changes in shutter adjustment would be expensive. Furthermore, the rate of flame propagation with Dayton gas is very low on account of the high percentage of hydrocarbons (see Chap. XI). There would be a danger of the flames from existing burners blowing out, since the gas velocity through the ports might be greater than the speed of flame propagation.

Composition of Oil Gas.—The composition of oil gas varies greatly according to the process of production. Jones process gas contains but little nitrogen, since no air is admitted to the machine during the gas make, and is relatively rich in hydrogen, which results partly from the action of steam on the carbon deposited during the blow. Dayton process gas, on the other hand, is relatively low in hydrogen, and contains all the nitrogen from the air used to partially burn the oil. The composition of the gas also varies with the type of oil used. These points are brought out by the analyses of typical gases made by the Jones[21,22] process and by the Dayton[23] process given in Tables CXLI and CXLII.

TABLE CXLI

ANALYSIS OF JONES PROCESS OIL GAS

Constituent	San Jose, California, per cent	Gas from heavy California crude, per cent
CO_2	2.8	2.6
C_nH_{2n}	2.7	7.0
O_2	0.1	
CO	10.6	9.2
CH_4	27.0	34.6
H_2	53.5	39.8
N_2	3.3	6.8
	100.0	100.0
B.t.u. per cu. ft.	535 B.t.u.	680 B.t.u.
Specific gravity	0.391	0.482

TABLE CXLII

ANALYSIS OF DAYTON PROCESS OIL GAS

CONSTITUENT	ANALYSIS OF 450 B.T.U. GAS, PER CENT
CO_2	6.1
Unsaturated hydrocarbons	14.7
CO	5.6
Saturated hydrocarbons	7.8
H_2	1.7
O_2	0.9
N_2	63.2
Total	100.0
B.t.u. per cu. ft.	450 B.t.u.
Specific gravity	1.02

Apparatus and Operation of an Oil-gas Set.—Since the Jones process and the Dayton process for the production of oil gas differ greatly, the two methods are discussed separately.

*The Jones Process.**—The Jones improved oil-gas set, as shown in Fig. 245, consists of two generator shells in the form of a U, with a wash box and scrubbing equipment. The wash box, though different in shape, functions similarly to the wash box in a water-gas set. Both generators are lined with fire brick,

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* For a more complete discussion of oil-gas production by the Jones process see LINDSAY and PEASE,[24] and ROGERS.[21]

and are divided into various chambers partly filled with fire brick supported by arches. The heating oil is burned in the first chamber at the top of the primary (shorter) generator. Below the arches supporting the checkerbrick in the first chamber is the primary make chamber into which the primary make oil is sprayed. The primary make checkerbrick extends into the bottom of secondary (taller) generator. Secondary make oil is sprayed into a secondary make chamber above the

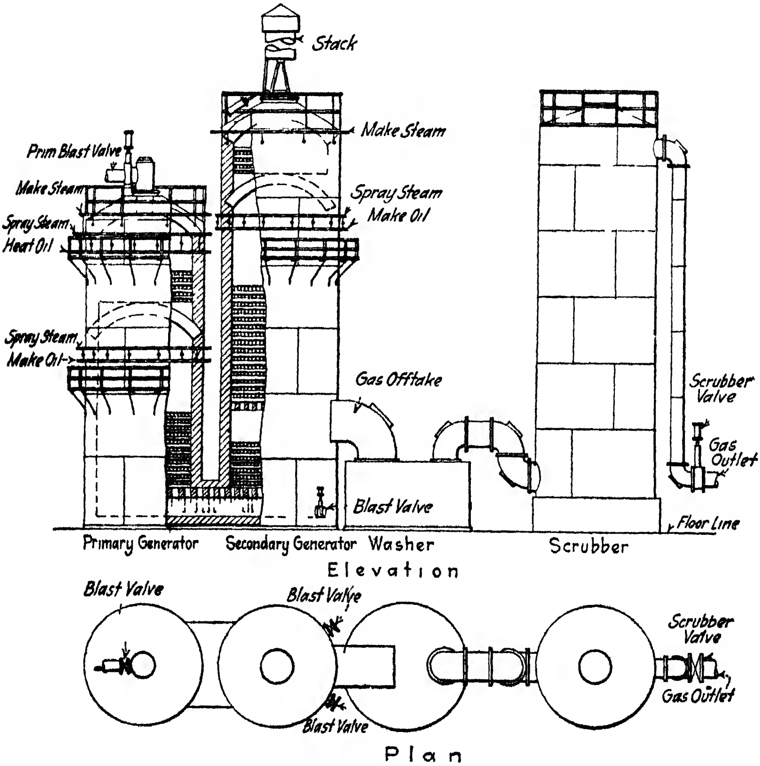

FIG 245.—Jones process oil-gas set.

top of the primary make checkers and during the gas make steam is admitted through secondary superheat steam checkers which are above the secondary make chamber. The stack valve is located in the top of the secondary generator.

The gas off-take is part way up the secondary generator (on the level of the top brick of the primary make checkers) and is equipped with large doors for cleaning out the considerable quantities of lampblack formed in the process. The gas offtake leads into the wash box, to which rapidly circulating water is constantly added. This water circulates through a continuous filter which removes the lampblack formed in the generators.

The usual cycle of the Jones process is about 20 min., divided as follows: (1) one-minute primary air blast with stack valve open. This clears out the machine, burns out carbon deposits in the primary generator, and thus aids in raising the heat for the next run. (2) One-minute secondary air blast with reduced primary blast (stack valve open). This burns out carbon deposits in the secondary generator. (3) Eight-minute heating period with primary blast and from 4 to 6 gal. of heating oil per min. (stack valve open). (4) Ten-minute gas make (stack valve closed). At the end of the heating period the blast and the heating oil are shut off, the stack valve closed, and steam admitted at the top of the primary generator at a rate of about 2200 lb. per hr. Steam is also admitted at the top of the secondary shell at about half this rate. The purpose of the steam is to react with carbon deposited during the heating period. About 1½ min. after the stack is closed, primary make oil (25 to 30 gal. per min.) is sprayed in and secondary oil admitted (15 to 20 gal. per min.). The addition of oil is continued for from 6½ to 7½ min. After the make oil has all been admitted the oil valves are shut and steam left on for 2 or 3 min. to purge gas out of the generators and to reduce carbon deposits. The stack valve is then opened and another heat started.

It is stated[21] that in large Jones process sets the heating oil amounts to about 16 per cent of the total oil used, which averages about 8.5 gal. per 1000 cu. ft. of gas made. In older oil-gas sets very large quantities of lampblack were produced (over 30 lb. per 1000 cu. ft. of gas made), but under modern conditions this figure can be kept down to about 20 lb. of lampblack per 1000 cu. ft. The lampblack does not represent a direct loss, however, since after reasonable draining and air drying (from 60 to about 25 per cent water) it can be used as a boiler fuel. One plant[21] is obtaining satisfactory results with a mixture of 75 per cent lampblack and 25 per cent anthracite screenings on a Coxe stoker. In certain other plants the lampblack is briquetted.

The Dayton Process.—In the Dayton process the combustion of a part of the oil furnishes the heat necessary to "crack" the remainder of the oil in the generator unit. A section through a

Dayton process unit is shown[23] in Fig. 246. In starting up the generator enough oil is burned in the combustion chamber surrounding the retort to bring it to the proper temperature. When this temperature has been reached, the amount of heating oil is greatly decreased, and a mixture of air and gas make oil admitted through the atomizer. The air (which is preheated by hot gases leaving the retort) and the oil pass into the retort, in which partial combustion and thermal decomposition take place. In this process it is possible to produce a gas continuously with any desired heating value between 300 and 560 B t.u. per cu. ft with the use of but little heating oil in the combustion chamber surrounding the retort. For the production of gases having values above 560 B.t.u. per cu ft. it is necessary to use considerable heating oil, because the air supplied for this heat content does not give enough partial combustion to maintain the retort at sufficiently high temperatures.

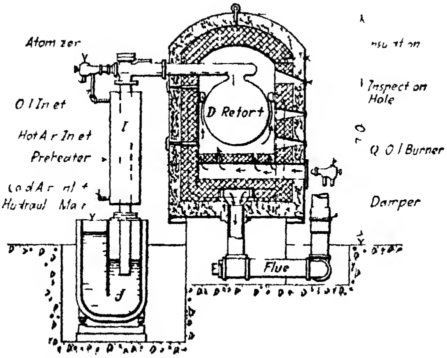

Fig 246 —Dayton process oil-gas set.

It is stated that about 4 gal. of oil are used in the production of 1000 cu. ft. of 450 B.t.u. gas. The net cost (not including interest on the investment, depreciation, and overhead) is said to be about 35 cts. per 1000 cu. ft. of gas. A small amount of tar (about 0.3 gal. per 1000 cu. ft.) is recovered in the process. The gas made is relatively clean and free from sulfur and the equipment required occupies but little floor space. It is possible to bring the equipment to capacity from complete shutdown in less than an hour. Unless properly regulated, carbon is apt to form in the retort.

Energy Distribution in the Formation of Oil Gas.—Heat balances on Jones gas-oil sets and on Dayton process sets of course vary greatly, considerable quantities of lampblack being made in the Jones process. Heat balances of the "single-shot" process[25] (somewhat similar to the Jones process), in which the oil gas is made in a single shell, and of the Dayton process are given in Tables CXLIII and CXLIV.

TABLE CXLIII

HEAT BALANCE OF A "SINGLE-SHOT" OIL-GAS SET

(*Portland Gas & Coke Co.*, July, 1920, to June, 1921)

	PER CENT OF TOTAL HEAT INPUT
Heat input:	
Heat equivalent of make oil	89.9
Heat equivalent of heating oil	6.8
Sensible heat in oil (above 60°F.)	0.3
Heat in steam used	2.9
Sensible heat of blast air (above 60°F.)	0.1
Total	100.0
Heat output:	
Potential heat in purified gas (heat of combustion)	42.8
Potential heat in lampblack	26.2
Potential heat in tar	3.1
Total heat lost	27.9
	100.0

Table CXLIII shows that the heating oil supplies about 13 per cent as much heat as the make oil. The gasification efficiency is not high, only 72.1 per cent of the heat being recovered in usable form (26.2 per cent being in the form of lampblack). The major part of the item, "Total heat lost," is accounted for by heat lost in the stack gases (7.2 per cent of the heat input) and by heat in carbon deposited but not burned (9.5 per cent of the total heat input).

The heat balance of a Dayton process oil-gas set[26] is shown in Table CXLIV.

Table CXLIV shows the high efficiencies attained in the Dayton process, over 90 per cent of the total heat input being recovered in usable form as potential heat in the gas made and as potential heat in the tar.

TABLE CXLIV

Heat Balance of a Dayton Process Oil-gas Set

(Data obtained at Plymouth (Mass.) Gas Light Co.)

	Per Cent of Total Heat Input
Input:	
Potential heat in oil	93.5
Sensible heat in oil	0 0 (nearly)
Potential heat in pilot gas	6.5
Sensible heat in pilot gas	0.0 (nearly)
Sensible heat in pilot air	0 0 (nearly)
Sensible heat in retort air	0.0 (nearly)
	100.0
Output:	
Potential heat in gas made	74.1
Sensible heat in gas made	2 6
Radiation and convection	0 6
Potential heat in tar	20 0
Sensible heat in tar	0.6
Sensible heat in stack gas	0.4
Latent heat in stack gas	0.1
Water from reaction	2.4
Total	100.8

BIBLIOGRAPHY

1. Juptner: "Lehrbuch der Chem. Tech. der Energien," Fig. 87.
2. Sixth Report of the Research Sub-Committee of the Gas Investigation Committee of the Institution of Gas Engineers, *Gas J.*, **154**, 619–626, 672–676 (1921).
3. *Trans.* Inst. Gas Eng., p. 366, (1922–1923).
4. Burrell, Seibert, and Robinson: U. S. Bur. Mines, *Tech. Paper* 104 (1915).
5. Seventh Report of the Research Sub-Committee of the Gas Investigation Committee of the Institution of Gas Engineers, *Gas J.*, **158**, 800–827 (1922).
6. Benesh: *Proc.* Am. Gas Assoc., Tech. Sec., pp. 9–38 (1921).
7. Fuel Research Board, *Tech. Paper* 6 (1923).
8. Morris: *Proc.* Am. Gas Assoc., Tech. Sec., pp. 28–34 (1922).
9. Tenth Report of the Research Sub-Committee of the Gas Investigation Committee of the Institution of Gas Engineers, *Gas J.*, **163** (1923).
10. Strache: "Gasbeleuchtung und Gasindustrie," p. 800 (Braunschweig, 1913).
11. "American Gas Works Practice."
12. Downing and Pohlmann: *Proc.* Am. Gas Inst., **11**, 588–636 (1916).
13. Wolf: *Proc.* Am. Gas Assoc. (1924).
14. Evans: *Proc.* Am. Gas Inst., **7**, 730 (1912).

15. For a discussion with bibliography of work on bituminous coal, see *Proc.* Am. Gas Assoc., Tech. Sec. (1924).
16. SMITH: *Proc.* Am. Gas Assoc., Tech. Sec. (1924).
17. BROWN: *Proc.* Am. Gas Assoc., Tech. Sec. (1924). (This has a good list of references.)
18. HASLAM and FORREST: *Gas Age-Record* (Nov. 17, 1923). (This has a good list of references.)
19. FRANZ FISHER: "Liquid Fuels from Water Gas," *J. Ind. Eng. Chem.*, **17–6,** 574 (1925).
20. CHARLES LORMAND: *J. Ind. Eng. Chem.*, **17–4,** 430 (1925).
21. ROGERS: "Manual of Industrial Chemistry," D. Van Nostrand Co. (1921).
22. Second Progress Report on Efficiency and Economy of Gas, R. R. Commission of California (1922).
23. BINNALL: *Ind. Eng. Chem.*, **13,** 242 (1921).
24. LINDSAY and PEASE: "Oil Gas Manufacture at New Bedford," Paper read before New Engl. Assoc. of Gas Engineers, (Feb. 18 and 19, 1925).
25. HALL and GRAF: *Am. Gas J.*, **115,** 318 (1921).
26. WELLING and HAMILTON: Mass. Inst. Tech. *Thesis* (1922).
27. Other references on oil gas are: W. S. YARD, *Proc.* Pacific Coast Gas Assoc., **12** (1919–1920); *Gas Age* (Nov. 25, 1920); "Oil Gas," issued by Pacific Gas & Electric Co.; DEMOREST: *Chem. Met. Eng.*, **32,** 329 (1925).

CHAPTER XX

THE CARBONIZATION OF COAL

The carbonization or destructive distillation of coal in the absence of air has for its primary objective the conversion of the initial fuel into the superior forms of coke, coal gas, tar, and the benzols. Coke, the hard cellular residue left on carbonization, is essential to the metallurgical industries and is being adopted to an increasing extent where a smokeless fuel is advantageous. Coal gas, the gaseous product of carbonization, forms the backbone of the manufactured-gas industry and its application in domestic and industrial heating is expanding rapidly. The by-products of carbonization, such as tar, ammonia, and the benzols, have become exceedingly important in chemical industry.

At the present time, practically all carbonization is carried out at relatively high temperatures. Recently, however, particularly in countries where a shortage of smokeless and liquid fuels exists, a great deal of attention has been paid to low-temperature carbonization. This process has for its object the production of a smokeless solid fuel from high-volatile coals, obtaining simultaneously the maximum yield of liquid fuel. Since most low temperature processes yield a high B.t.u. gas, low temperature carbonization may offer an attractive method of gas enrichment. Complete gasification of coal, in which coal is coked by the sensible heat of producer gas in the top of a special type of gas producer, and the resultant hot coke converted into producer or water gas in the lower part of the same apparatus, also has been receiving considerable attention. This process allows the recovery of a portion of the liquid by-products and ammonia.

This chapter deals with the mechanism of carbonization as applying to coke ovens and gas retorts, takes up fundamental considerations in carbonization for coke production and for gas manufacture, and includes descriptions of various types of carbonization equipment. Our knowledge of the composition of coals, already discussed in Chap. II, is reviewed briefly in the light of the application of this knowledge to the carbonization process. Low-temperature carbonization and complete gasification of coal, although still more or less in the experimental stage, are discussed on account of their probable future industrial importance.

Importance of Carbonization.—When it is realized that nearly one-seventh of all the coal mined in the United States goes to form metallurgical coke, that the metallurgical industries of the United States consume annually in the neighborhood of 50,000,000 tons of this fuel, and that about one-third of the gas distributed through city mains is coal gas, it becomes evident that coal carbonization is one of the most important industries.

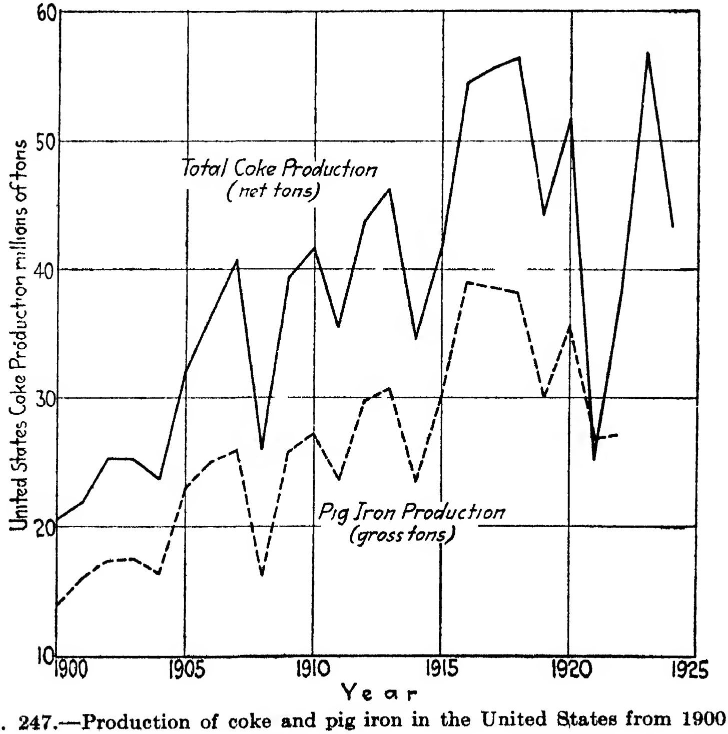

FIG. 247.—Production of coke and pig iron in the United States from 1900 to 1924

It is well known that the activity of the iron and steel industry is an excellent barometer of business conditions in general. The close parallelism between the production of coke and the production of iron and steel is shown[1] by Fig. 247, the interrelation between pig iron and coke production being clearly marked. In addition to bringing out the close relation between coke production and industrial activity, Fig. 247 also emphasizes the growth in coke production from 1900 to 1923—an increase from slightly over 20,000,000 tons in 1900 to over 50,000,000 tons per annum during the period from 1916 to 1918, and in 1923 an even greater coke output than at the war-time peak.

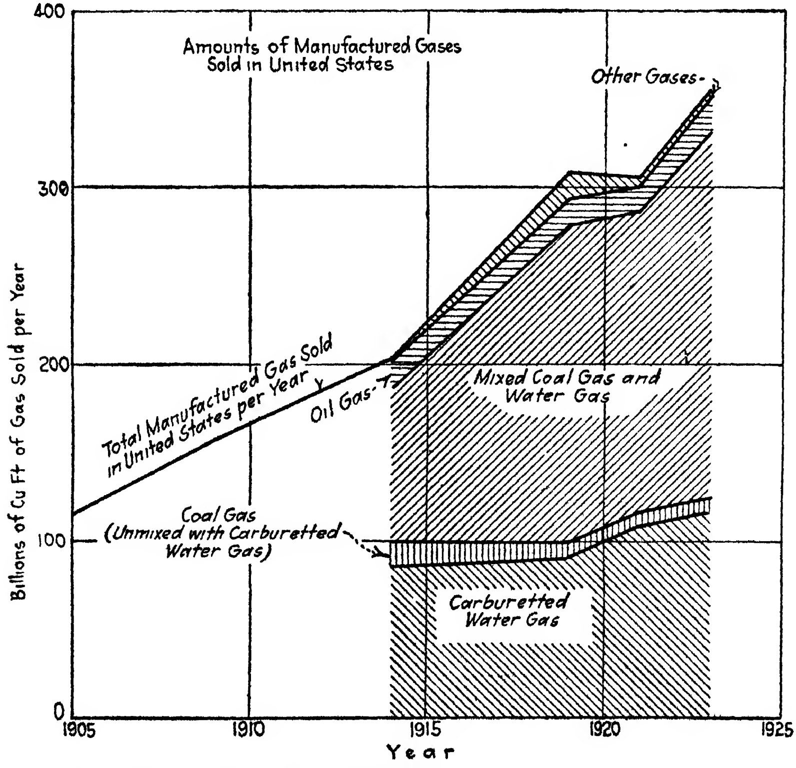

FIG. 248.—Manufactured gases distributed by public utilities in the United States.

The gas industry, although carbonizing less coal than the metallurgical coke industry, directly affects a large proportion of the community, furnishing as it does the gas used for cooking, heating, and lighting. The amounts of manufactured gases distributed by public utilities in the United States, over the period from 1905 to 1923, and the proportions of the different gases going to make up the total are shown[2] in Fig. 248. The tremendous growth of the industry, from slightly over 100,000,-

000,000 cu. ft. in 1905 to over 400,000,000,000 cu. ft. in 1924, and the large amount of coal gas and mixed coal gas and water gas sold during this period of expansion, indicate the important part played by carbonization in the gas industry. In the future, as the many advantages of gaseous fuels cause their adoption in domestic and industrial heating to an increasing extent, it is probable that as much (if not more) coal will be carbonized for gas-making and domestic coke purposes as for the production of metallurgical coke.

In addition to the gas made and the coke produced for metallurgical purposes, both the gas and the coke industries are the source of highly important by-products. Most of our supplies of ammonia come from the carbonization of coal, although in the future, atmospheric nitrogen fixation processes will be direct competitors. Practically all the benzols (benzol, solvent naphtha, etc.), naphthalene, and the coal-tar products like anthracene, etc., from which are manufactured high explosives and dyes, are derived from carbonizing plants, and these by-products are a significant source of revenue to the industry. Tar, already important as a source of by-products, road materials, and roofing materials, also is being widely adopted as a liquid fuel in steel plants and in other industries which produce their own coke.

Scientific coal carbonization, since it conserves the tar, ammonia and other valuable by-products, and produces higher-grade fuels, must be regarded as a factor in fuel conservation. It has been said, for example, that the use of raw coal as a fuel ought to be, and some day probably will be, prohibited by law. This is doubtful, however, since there are many purposes where carbonization can show no substantial advantage. For some purposes, such as domestic heating, considerable fuel economy can result from the substitution of gas for coal. It is in fields of this nature, as well as in industrial furnaces where gas gives improved quality of finished product, that the adoption of gas (and therefore carbonization) may become an important fuel-conservation measure.

Development of Carbonization.—The practice of carbonizing coal in order to produce metallurgical coke, although known in the latter part of the sixteenth century, was not utilized on any large scale until about 1730. The second type of carbonization, with coal gas as the desired product, was instituted in 1792 by

Murdoch in England. In the early stages of the development of these two processes, they were considered fundamentally different, and this idea persisted until relatively recent years.

In the first attempts to produce metallurgical coke, the methods employed were those which had previously been used in the manufacture of charcoal. The coal was coked in earth-covered heaps, the combustion of part of the coal furnishing the necessary heat. This practice gradually gave way to beehive coking, in which a charge of coal is carbonized to a strong coke in a dome-shaped brick oven. In this process, as in the earlier earth-covered heaps, or meilers, the combustion of most of the volatile matter and part of the coal itself furnishes the heat necessary for carbonization. The first by-product coke ovens were constructed by Knab in France about 1856, and since that time, first in Europe and later in the United States, the trend has been away from the wasteful beehive ovens and toward the type of equipment in which the by-products are saved. The earlier by-product ovens, although recovering the by-products from the gas, required practically the entire output of gas for heating the carbonization ovens themselves. With the adoption of regeneration and recuperation, and with modern ovens, only from 35 to 40 per cent of the coal gas must be burned to supply the heat needed for carbonization, and the remainder is available for distribution through city mains or to meet other fuel requirements around the plant. Thus, although the developments in the metallurgical coke industry have been directed primarily toward the production of better coke, progress also has been made along the lines of more complete recovery of by-products and conservation of the coal gas given off during carbonization.

In the early stages, the manufactured-gas industry desired as high a yield of rich gas as possible and the quality of coke produced was considered to be of little importance. The first use of coal gas was for illuminating purposes, and gas standards were drawn up on the basis of candlepower requirements. To produce a gas of the required standards, gas plants were forced to use high-volatile coals which would give the maximum yields of rich gas. These coals, moreover, in order to give the requisite yields quickly, were heated rapidly and this dictated the use of shallow, low-capacity retorts. Murdoch, in his first plant, built in the early part of the nineteenth century, used cast-iron pots, but these were soon superseded by horizontal iron retorts. These, in turn, were displaced by the ⌓ shaped fireclay retorts still so widely in use. With the development of the Welsbach mantle, however, and with the consequent reduction in illuminating power and heating-value standards, the trend was toward the larger inclined retort and finally to the present-day vertical retorts together with by-product ovens similar to those used in the metallurgical coke industry. Practically all the new large construction going on at the present time is of one of the latter two types. Thus, the gas industry, although believing in the early days of the art that satisfactory gas making and coke manufacture could not be carried out in the same plant, is gradually adopting the carbonization equipment in which both processes are possible. The coke oven, producing as good quality of gas as is made in retorts, operates with less labor cost, and produces coke of the highest quality. The type of coke is highly important, even in a plant primarily engaged in gas manufacture, because coke sales may be, and oftentimes are, as great a source of revenue as the gas distributed through the city mains.

It is evident from this discussion that the gas and the coke industry gradually are merging as far as equipment requirements are concerned, and that the tendency is toward the same type of plant for both branches.

THE MECHANISM OF CARBONIZATION

Carbonization is essentially a distillation of the volatile matter of coal and, consequently, an understanding of the steps in the process and of the effect of various factors, on the resultant coke and coal gas, is of great importance. The subsequent discussion of the mechanism of carbonization outlines the essential characteristics of coals for carbonization, describes the progress of carbonization as it takes place in the coke oven or retort, and points out the effect of numerous variables.

Essential Characteristics of Coals for Carbonization.—As already pointed out in Chaps. II and III, certain coals soften or melt on heating, give off volatile matter, and finally solidify into a solid cake, leaving a hard, cellular, coherent mass on further heating. The kind of coke left, the degree of fusion, the amount of gas evolved, the porosity and strength of the resultant coke—all depend to a large extent on the coal used. The purpose for which carbonization is carried out, and economic factors, such as the price and availability, determine the coal selected. Coals giving the best metallurgical coke in by-product ovens are not necessarily the coals best suited for the ~~production~~ of manufactured gas, though the simultaneous production of city gas and metallurgical coke is being carried out in a number of instances.

The chemical analysis of a coal (see Chap. II) does not give a definite indication as to its coking quality. Certain ratios, however, such as the ratio of hydrogen to oxygen proposed by White (this is discussed in Chap. II, p. 44), may be applied to obtain a preliminary idea of the value of a coal for carbonization purposes. According to the "Coal Catalog,"[3] coals for by-product coking and for gas manufacturing purposes should be able to meet the following conditions:

Constituent	Coals for by-product coking, per cent	Coals for gas making, per cent
Volatile matter	15–38*	32–37
Ash.	4–8	6–8
Sulfur	Not more than 1.5	Not more than 1.25
Phosphorus	Not more than 0.02	
Gas yield		10,000–12,000 cu. ft. per ton
Coke yield		65–70

* This range provides for mixing.

No analytical method derived thus far, however, can be relied upon, and in spite of the valuable indications that can be obtained in the laboratory from small-scale coking tests, the only satisfactory method of determination is to coke a considerable quantity of the coal in the actual oven or retort and carefully interpret the results obtained.

Several laboratory methods[4] have been devised for determining readily the quality of coke which may be produced from a given coal. These methods usually consist of coking the coal when mixed with varying proportions of sand or anthracite, and observing the properties of the resultant coke. The greater the fraction of sand or anthracite that can be mixed with the coal, and still produce a coherent coke, the better the coking quality is supposed to be. Conclusions from such tests, like deductions based solely on analyses, should be viewed with caution, since the conditions prevailing during the small-scale test (such as rate of coking, temperature, etc.) vary so widely from actual operating conditions that misleading results are quite possible.

As pointed out in Chap. II, considerable information relative to the coking characteristics of coals has been obtained in the various researches on coal constitution. Thus, Wheeler, Illingworth, and other English workers in the field of solvent extraction have found that the gamma (γ) compounds, portions of the coal soluble in both pyridine and chloroform, are well fused and coked on heating, and that coals deficient in these compounds do not yield a satisfactory coke. Fischer and his co-workers in Germany have had pronounced success with the extraction of coal with benzol at high pressures and relatively high temperatures. The benzol-soluble extract of coal contains the compound responsible for the coking characteristics (*i.e.*, similar to the γ compounds). The compounds soluble in benzol (called bitumens), when further treated with petroleum ether, are separated into a soluble oily substance (supposed to impart the cementing properties to the coke), and a hard waxy residue (to which is attributed the swelling of the coal during carbonization). It is interesting to note that, even though a coal loses its coking properties on extraction, if the extracted material is again added to the coal its original behavior on coking is restored.

These compounds, on which the coking property depends, in addition to being present in the proper amounts, must also be of such constitution that they melt below the temperature at which their decomposition takes place.[5] If this condition is not met, the bonding constituents will distill off, leaving a granular, non-coherent, unsatisfactory coke. Furthermore, if the cementing substances are present in large amount, or if large quantities of gases are evolved from them while they are in a plastic state, the resultant coke is liable to be light and porous, and the chances for a high degree of swelling are greatly increased.[6] It should also be emphasized that the compounds responsible for the coking properties of coal are profoundly altered by oxidation or weathering, and even a relatively short period of exposure to air may be sufficient to impair seriously the quality of the coal for coking purposes.

To summarize present knowledge of the essential characteristics of coals for carbonization,* it may be said that a great deal of fundamental information has been gained in recent years and that a number of general considerations have been quite well established. Much work remains to be done, however, and although valuable preliminary information as to the behavior of a coal on carbonization may be obtained in a number of ways, the only entirely satisfactory criterion is the behavior of the coal in the full-scale oven or retort in which carbonization is to be carried out.

Effect of Heat on Gas and Coking Coals.—The major researches undertaken to determine the effects of heat on coals have been discussed in detail in Chap. II. The conclusions from them are summarized on page 39. These conclusions outline the order of evolution of the volatile matter in coal, and have been augmented by the following valuable deductions made by Warner[7]

1. If the removal of volatile matter from coking coal were the only consideration, the high heats commonly used in carbonization are unnecessary. That is, practically all the volatile matter in coal could be driven off below 1050°F. (566°C.), but the necessary conditions for effecting this (prolonged, slow heating, for example) destroy the coking properties. The reasons for high temperatures lie in the necessity for producing a special type of solid residue and for completing carbonization within a reasonable time.

2. Although the thermal decomposition of tar and other materials in the oven or retort increases the gas yield, the increase in gas is far from being proportional to the degree of decomposition, since carbon is one of the main products of such a decomposition.

3. The less the cracking and the more the distillation of condensible hydrocarbons occurs the less will be the intensity of heat required to drive off the volatile matter of coal. In other words, considerable heat is absorbed by cracking reactions, and if this heat could be conserved it would be possible to decrease the amount of heat required for carbonization.

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* In a recent paper, LESSING [*J. Soc. Chem., Ind.* **44**, 345T (1925)] has pointed out the possible importance of the inorganic ash constituents of coal. For example, small percentages of aluminum sulfate or iron chloride increase the amount of coke residue and the percentage of fixed carbon.

4. The less the heat through which the non-condensible hydrocarbons must pass in leaving the oven or retort the richer will be the gases. In other words, the reactions of thermal decomposition or cracking speed up markedly at high temperatures, and therefore the lower the temperature or the shorter the time of contact between these hydrocarbons and hot coke or the heated oven walls the less will be the cracking. High temperatures and long time of contact in practice are also responsible for troublesome naphthalene deposits. However, a very small amount of carbon deposited from the thermal decomposition of the gases sometimes will increase greatly the strength and density of the resultant coke residue.

The application of these simple, fundamental considerations may be seen from the subsequent discussion of carbonization in the oven or retort, and of the effect of various factors on the products of coking.

Progress of Carbonization in the Oven or Retort.—The accepted explanation of what takes place when coal is charged into a coke oven or retort, and of the progress of carbonization in the coking chamber, is substantially as was first postulated* by von Hilgenstock.[8] When coal is placed in the hot oven or retort, that portion of the charge in immediate contact with the hot walls is heated very rapidly to a high temperature, a thin layer softens, becomes plastic, and melts. This layer of plastic material (usually called the fusion zone) travels toward the center of the oven† as the temperature of the charge is raised. In the fusion zone itself (as indicated in Fig. 249) and on either side of it destructive distillation of the coal takes place, some of the gaseous products forcing their way out of the plastic material, thus imparting the cellular structure which appears in the final coke. On the wall side of the fusion zone, the plastic layer hardens into a cellular residue, and the volatile matter left in the coke is driven off gradually (yielding a gas rich in hydrogen) as the temperature rises during the remainder of the coking period. At the inner face of the fusion zone, as it travels away from the hot wall, the relatively cool, raw coal from the

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* For more complete description of the mechanism of carbonization than is given here, the reader is referred to von Schiewind,[9] Simmersbach,[10] Cooper,[11] Ruebel,[12] Ramsburg and Sperr,[13] Sperr,[14] O. B. Evans,[15] Porter,[16] E. V. Evans,[17] Wilson,[18] and Ryan.[19]

† For purposes of explanation the following discussion pictures the coal being coked in a coke oven.

center of the charge is melted and coked. Finally, if the heating from both sides of the oven or retort has been uniform, the fusion zones initially formed at the two walls will meet at the center —the line of meeting appearing as a center crack in the coked charge.

The gases evolved from the fusion zone (and the volatile constituents driven off at each side of it) force their way either to the walls of the coking chamber or through the cooler raw coal. Thus, in the initial stages of carbonization, the volatile matter driven into the center of the charge will have the tar, oils, and water condensed out, to be distilled again, as the heat penetrates across the oven. The gases forcing their way through the hotter wall face of the fusion zone, however, have to pass through cracks in the coked coal up along the incandescent oven or retort walls, and finally through the hot free space above the top of the charge. During this passage they are subjected to heat and are thermally decomposed. During the latter part of the coking period this process is intensified, for the path of the gases becomes longer and the coke hotter as the fusion zone travels further from the wall. Thus, thermal decomposition becomes more severe as coking proceeds, the hydrocarbons of the gas coming from the retort become less complex in structure, the gases become richer in hydrogen, and the coke becomes silvery-gray in color, due to the deposited particles of retort carbon formed by the breaking down of the hydrocarbons and tars. As pointed out by Evans,[17] a knowledge of chemistry does not help very much in the interpre-

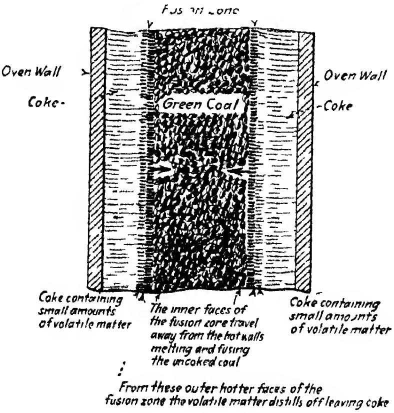

FIG. 249.—Process of carbonization in a coke oven—diagrammatic.

tation in this picture—in fact, it rather tends to complicate matters because the better appreciation of the complexity of the reactions is offset by the confusion which results from a consideration of the almost limitless range of reactions that may take place.

With this brief explanation of the coking process as background, it becomes possible to examine more in detail the rate of travel of the fusion zone, the rate of heat transfer through the charge, and to point out the influence of these factors on the design of carbonizing equipment.

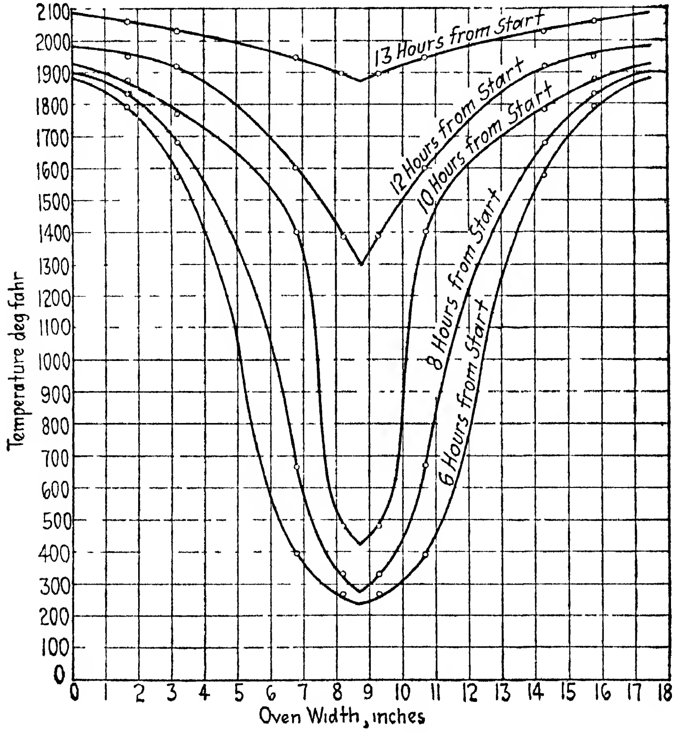

Fig. 250.—Temperature gradients across a coke oven at different intervals during the coking period.

Rate of Travel of the Fusion Zone.—From the difficulty encountered in getting heat to flow into coal rapidly and thus change the coal to coke there has arisen a belief that the fusion zone is a very poor conductor of heat and that across this zone a large drop in temperature occurs during coking. The well known facts apparently correspond to this theory. For example Fig. 250 shows the temperatures throughout a coke oven at 6, 8, 10, 12 and 13 hrs. after charging and even at 10 hrs. after starting, or considerably more than half way through the coking period, the center of the oven is only slightly over 400°F. (204°C.). At the tenth hour there is a difference of about 900°F. (500°C.) between points of 6.8 and 8.2 inches in from the wall. The fusion zone at this time (in the oven studied and with the flue temperatures used) lies between these two points and as a result, it would naturally be supposed that the drop in temperature was caused by a highly resistant fusion zone. Ryan[19] has shown that the above facts are exactly in accord with what is to be expected when heat flows through any material of relatively high thermal conductivity (such as coke) into any material of low thermal conductivity (coal) and that the fusion zone has essentially the same thermal conductivity as coal.

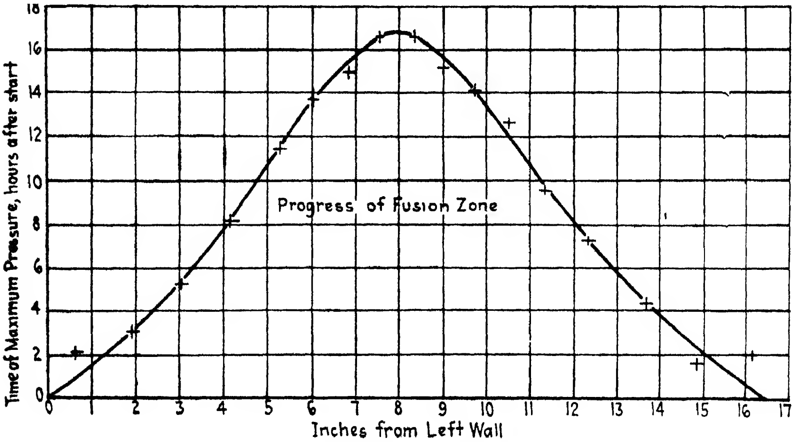

FIG. 251.—Time of passage of fusion zone across a by-product coke oven as shown by the location of the point of maximum pressure.

The rate of travel of the fusion zone through an oven or retort may be determined conveniently by measurements of the pressures developed within the coal at various points as carbonization progresses. The rapid evolution of the volatile constituents of the coal during fusion causes localized high pressures as the gases escape from the zone and make their way through the coal or coke to the collecting main. This method, substantiated by the valuable researches of Foxwell,[20] has been used by Ryan[19] to obtain the type of curves shown in Fig. 251. This figure shows the time required for the fusion zones to pass across a 16.5-in. oven. The time of passage of the fusion zone is indicated by time required for maximum pressure to be developed. The data given in Fig. 251 were recalculated to show the average rates of travel of the fusion zone for each ½ in. across the oven, and the results[19] plotted in Fig. 252. This figure clearly brings out the way in which the fusion zone crosses the oven from the two hot walls.

When a fresh charge of coal is dropped into a hot oven, the rate of heating at the wall is extremely high, and as long as the fusion zone is near the wall, its rate of travel will be fast because the heat has such a small resistance through which to flow in order to reach it. As the fusion zone recedes from the hot wall, heat flows into it more slowly because heat is absorbed by the coal already coked and by the gases escaping to the foul gas main, consequently decreasing the rate of travel. Moreover, during this period while the zone is receiving less heat from the wall, it is also being cooled by the absorption of heat by the cold charge ahead of it. When the heat wave ahead of the fusion zone reaches the center of the oven, it encounters a plane across which no heat is transmitted, for, if carbonization has been uniform on both sides, just as much heat is reaching this plane from one side as from the other. The chilling action of the cold coal on the fusion zone, therefore, begins to be checked and

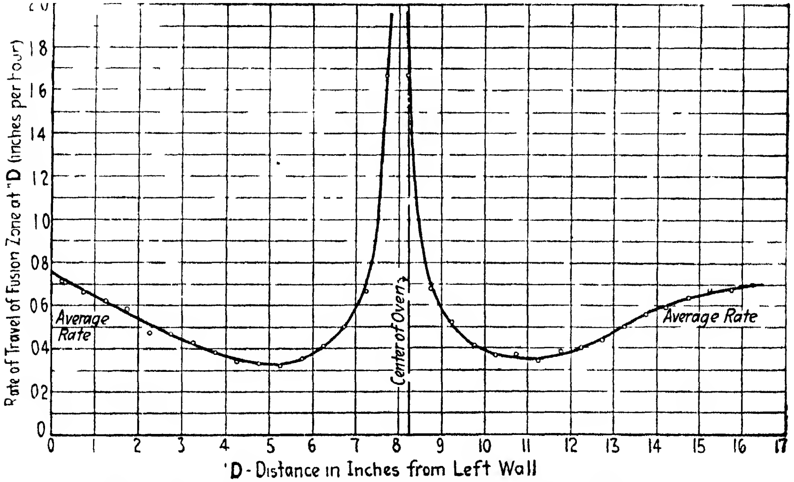

FIG. 252.—Rate of travel of fusion zone across a by-product coke oven

as the temperature at the center line rises, this chilling action is more and more diminished. Although less heat from the wall is flowing into the fusion zone, this is more than compensated for by the decrease in chilling action, and the rate of travel actually increases again.

It is interesting to note the effect of increase in oven width on the rate of travel of the fusion zone and on the time required for coking. Figure 253 illustrates the differences in the characteristic rate curves for a 16-in. oven (shown with full lines) and for a 20-in. oven (shown with dotted lines). The figure is drawn as if the 16-in oven had been split in two and 4 in. additional

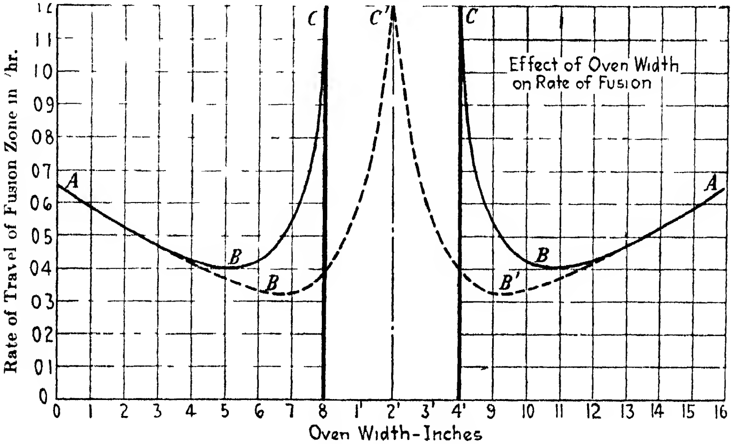

FIG 253 —Diagram illustrating the effect of increasing coke-oven width on the rate of travel of the fusion zone

inserted at the oven center in order to make the oven 20 in. wide. In the 20-in. oven the fusion zones would move inward from the wall at a steadily decreasing rate, past the previous minimum points in the 16-in. oven, and continue on for a distance of 2 in., as shown by the curves $AB'C'$. This means that the additional 4 in. of oven width are traveled at a rate *lower than the minimum reached in the* 16-*in. oven.* In other words, the result is to increase the time required for carbonization much more than in direct proportion to the increase in oven width.

After the fusion zone has reached the center of the oven the bulk of the volatile constituents have been driven off, and heat is being transferred through material of higher conductivity than before. As a result of these factors the temperature across the oven or retort rapidly becomes uniform, as shown by Fig.

250. For example, 13 hr. after charging the temperature at the center of the oven (in the case shown) was only about 200°F. (111°C.) lower than the wall temperature, although at the end of 12 hr. the center was about 800°F. (444°C.) below the wall temperature. It may be concluded, therefore, that the rate of travel of the fusion zones is the controlling factor in the rate of coking of coal and that the rate of travel of the fusion zone is dependent on the rate at which the heat flows through the coke into the coal. No large amount of heat is absorbed when coal fuses, nor is the fusion zone a large thermal resistance in the path of heat flow. The effect of increasing oven width is to increase the time required for carbonization out of proportion to the increase in capacity. There are other economic factors (such as cost of construction, the properties of coke produced, etc.) which fix the minimum width of oven, but the maximum width is determined by the rate of travel of the fusion zones. The trend of American practice is now toward narrower ovens, which, by virtue of the increased speed of coking possible in them, allow greater outputs and make available a wider range of coals for coke manufacture.

Path of Volatile Products.—The distillation which takes place in the fusion zone determines the characteristics of the coke formed, while secondary reactions occurring elsewhere to a large extent control the nature of the liquid and gaseous products. Only slight changes occur in the coke after the passage of the fusion zone, while the final gaseous and liquid products differ widely from those leaving the distillation zone. Although distillation of the gases may take place at only 700 or 800°F. (371 or 427°C.), the amount of thermal decomposition of the gases may be so great that the products reaching the gas off-take more nearly correspond to those which would be distilled off at much higher temperatures. Such secondary reactions take place during the passage of the volatile products from the fusion zone to the outlet of the carbonizing chamber, and it can be seen that the path taken (*i.e.*, through hot coke or through raw coal) has a decided practical significance.

Most investigators believe that the major part of the gases pass through the hot coke. A thorough consideration of the action of the fusion zone as a screen and the large pressure drop across it indicates that little or no gas passes through it. Those products distilling at low temperatures would be expected to pass through the coal side, while the products evolved at the hot-wall side of the zone (*i.e.*, at high temperatures) would be expected to pass through cracks in the hot coke, and up along the small free space formed at the oven walls by the shrinkage of the coking charge. Furthermore, all analyses of liquid and gaseous products from high-temperature carbonization (and the presence of retort carbon in the final coke) show excessive decomposition of the primary products of distillation. That the gases pass through the hot coke, therefore, seems most reasonable, although undoubtedly some of the gases (mainly water, oxides of carbon, and tar) pass up through the coal.

Foxwell,[20] for example, from measurements of the resistance of coal in the plastic state, has calculated the fraction of volatile constituents which pass through the coal and coke. His results on a Durham coal are shown in Table CXLV.

TABLE CXLV

PERCENTAGE OF TOTAL GASES PASSING THROUGH UNCOKED PORTION OF CHARGE

(*After Foxwell*)

Time after charging, hours	Height above oven floor, inches	Percentage of total gases passing through coal zone
9	39.4	Nearly all gases pass into coal
9	100	
14	39.4	69.8
14	100	54.0
18	39.4	41.8
18	100	31.7
20	39.4	21.6
20	100	17.5

The table shows the large fraction of the total gases that pass through the coal during the early part of the coking period, and also brings out the fact that, as coking proceeds, more and more gas passes through and is partially decomposed by the hot coke. Gases given off near the bottom of the oven or retort have a longer path of travel (and therefore are cracked more) than gases given off near the top of the charge. This effect is compensated for to a certain extent, however, by the fact that a greater percentage of the gases given off near the bottom of the oven tends to pass through the cooler coal in the center of the charge (see Table CXLV.) The influence of the temperatures to which the gases are subjected is discussed more in detail in a later section.

Changes in Composition during Carbonization.—Both the amount and the type of gas obtained vary greatly as carbonization proceeds. This is well illustrated by means of thermal models, such as have been used by Evans[17] (see Fig. 254). These figures are constructed by combining the hourly volume and the calorific-value curves, so that the length *A* in Fig. 254 represents the coking time the height *B* shows the rate at which the gas is being evolved, and the width *C* gives the heating value. Just after coal is charged into an oven or retort, large quantities of gas rich in methane and other hydrocarbons are evolved, and in the case shown in Fig. 254 (typical of gas-retort practice) 5 per cent of the total gaseous thermal yield was produced in the first 10 min. Gradually, the gases are forced to travel over a longer and hotter path as more of the coal is coked and the temperature of the coked portion is raised. This results in increased thermal decomposition of the volatile products; methane and other rich hydrocarbons therefore form but a small fraction of the gases finally evolved, hydrogen being the chief constituent of the lean gases given off at the end of the coking period.

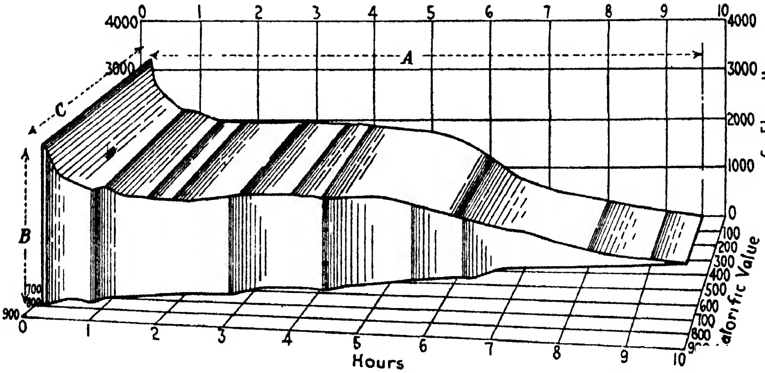

FIG. 254.—Quantity and heating value of gases evolved during the coking of coal.

In view of the importance of the reactions of thermal decomposition in carbonization, a number of investigations have been carried out in this field. The results and conclusions from these investigations may be summarized* as follows:

1. The volatile hydrocarbons distilled from the coal break down into unsaturated groupings, such as $CH_2 =$ or $CH \equiv$, which, depending on the temperature and other conditions, may combine with each other (to form ring compounds, for example), may pick up hydrogen to form saturated hydrocarbons, or may further decompose into their elements, H_2 and C.

2. Methane, which is the chief hydrocarbon in coal gas, is the most stable of the volatile organic constituents and is not decomposed appreciably below 1472°F. (800°C.). Methane, as such, may be formed in the early stages of carbonization, but is mostly produced by decomposition of higher hydrocarbons. Above 1472°F. (800°C.) the decomposition of methane into carbon and hydrogen becomes marked.

3. Unsaturated hydrocarbons, such as ethylene, are formed by the decomposition of heavier hydrocarbons, the decomposition reactions being particularly affected by hot surfaces. Between 932 and 1472°F. (500 and 800°C.) ethylene tends to produce aromatic hydrocarbons, such as benzol.

4. Hydrogen, although perhaps formed to a limited extent from the initial decomposition of the coal itself, mainly results from cracking of the hydrocarbons.

5. Nitrogenous compounds in the coal break down to ammonia, HCN, and free nitrogen, although about half the nitrogen in the original coal still remains in the final coke. Little ammonia is formed below 1000°F. (538°C.) and the most favorable temperature range for its formation is probably between 1350 and 1550°F. (732 and 843°C.), since above the latter temperature decomposition into nitrogen and hydrogen becomes rapid. Nitrogen appearing in the final coal gas comes from this decomposition and from air leakage into the retort.

6. Sulfur compounds in coal gas result from the partial decomposition of organic and pyritic sulfur in the original coal (see Chap. II) and all the sulfate sulfur is retained by the coke. From 20 to 50 per cent of the total sulfur of the coal is driven off by carbonization. In the gas the sulfur is found mainly as H_2S and to a slight extent as CS_2, the former existing mostly in the early part of carbonization and the latter (formed by the

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* For complete details of the thermal-decomposition reactions the reader is referred to BONE,[22] PORTER,[28] SIMMERSBACH,[10] and EVANS.[17]

interaction of H_2S with hot coke) during the second half of the coking period. Small traces of such sulfur compounds as thiophene are also formed.

Effect of Miscellaneous Factors.—In addition to the quality of the coal itself and the type of oven or retort used, a number of other factors exert a considerable influence on the coking process. Among these are the temperature and the pressure in the carbonizing chamber, the rate of coking, the size of the coal particles, and the compression of the charge.

As just shown, the varying temperature during a coking period profoundly affects the nature of the volatile constituents given off on carbonization. A number of tests[23 24 17] have been carried out to determine the effect of temperature variation over the ranges (1600 to 2000°F., or 871 to 1093°C.) commonly employed in high-temperature carbonization. These investigations showed the following results with increasing temperatures:

1. There is an increase in gas volume per ton of coal carbonized.
2. There is a decrease in the heating value of the gas.
3. The thermal yield of gas (*i.e.*, B.t.u. per cu. ft. times the cubic feet evolved) is increased.
4. Tar production is decreased. In other words, the gas yield is increased at the expense of the tar yield.
5. The ammonia yield is lowered.

In addition to these influences, higher temperatures have the decided advantage of a more rapid rate of heat transfer and therefore a better coking rate with consequent greater capacity per retort. There is another advantage of rapid coking which is extremely important in low-grade coals. As shown by Warner (see p. 654), slow heating destroys the coherency of the resultant coke, and, conversely, with higher heats and more rapid coking, the bonding material is more efficiently used. With a slow rate of heating distillation of the hydrocarbons before thermal decomposition is greater than with rapid heating. Consequently, with rapid heating the amount of volatile carbon is less, with the result that more fixed or decomposed carbon is available for bonding or cementing purposes. Since the better coking coals usually contain excess cementing material, high temperatures do not offer any decided advantage in their carbonization (except as regards increased retort capacity), but with high temperatures and a rapid rate of coking a strong coke may be made from a relatively poor coking coal. The coals of Illinois, for example (see Table XXIX, p. 68), contain such a high percentage of oxygen that they would not be expected to form a good coke under ordinary normal conditions. With the rapid coking rates possible in the narrow Becker oven (see p. 681), ~~good blast-fur~~nace coke has been made[25] on an 11-hr. coking time.

The *pressure* under which coal is carbonized has considerable effect on the quantity and the quality of the products formed. It is impossible in practice, however, to maintain any large pressure or vacuum in the retorts, due to the difficulty in keeping settings tight at high temperatures. In almost all cases the retorts are maintained under a slight vacuum of from 0 to 0 2 in. of water in order to prevent gas being excessively decomposed in the carbonizing chamber. Even with the slight vacuum commonly employed, any leaks in the setting will result in a lowering of the heating value of the coal gas produced. Leaks from the flues will introduce diluent stack gases, and leaks in or around the door will admit air, which, if it passes through the charge, will give a low B.t.u. producer gas, or, if it goes directly into the gas space, will burn the richest constituents of the coal gas.

Influence of Coking Properties of Specific Coals.—The appearance and other properties of coke vary greatly with coals from different localities. Coking coals from the Connellsville district in Pennsylvania, for example, yield a dense, hard, lustrous coke, while the Westmoreland and Kanawha gas coals yield a light, porous, friable coke. These facts form the basis for the widespread practice of mixing coals prior to carbonization. It is well known that a coal which gives a light porous coke on account of a high excess of cementing or bonding constituents, if mixed with a low-volatile, poor or non-coking material, such as even anthracite or coke itself, will yield a denser, stronger coke. This method seems to leave numerous small channels through which evolved gas can escape, thus, by preventing excessive disruption of the plastic layer by gas bubbles entrapped during the plastic stage, reducing the swelling of the charge. The practice of mixing coke or other non-coking material has not yet been employed to any extent on a commercial scale.

American coke-oven operators for many years have mixed high- and low-volatile coking coals in order to improve the peratures may be necessary to produce a satisfactory grade of coke for a blast furnace, and the higher temperatures may result in an appreciably leaner gas. The sacrifice in gas quality is justified if the overall cost of manufacture is decreased (in this case perhaps by making a cheaper local coal available for coking purposes).

In any event, in the metallurgical coke plant the factors influencing the type of coke are judged most important, and close supervision is kept over the coking time, the structure, and the properties, physical and chemical, of the resultant coke.

General Considerations in Carbonization for Gas Manufacture. Although the by-product oven and the beehive oven are the only forms of carbonization equipment which will produce satisfactory metallurgical coke, several other types of apparatus, such as the vertical and horizontal retort, become available when the production of gas is the prime consideration. In the manufactured-gas industry, a gas meeting strict legal standards must be made at all times, and the quality of coke and by-products may fluctuate to a certain extent This does not mean that coke quality is unimportant, nor that its characteristics must not be controlled carefully, but rather that a different type of coke from that required in metallurgical work can be used for the purposes for which gas-works coke is used (for domestic heating and for water-gas manufacture).

The cost of gas in the holder and flexibility in operation are the factors of prime importance in the gas industry. To keep the cost low, gas must be conserved around the plant, and there must be a ready market for coke. Thus, although part of the gas from metallurgical coke ovens is used to heat the carbonizing chambers, in the gas industry coal gas for this purpose is largely replaced by producer gas. With regard to the coke market, it has already been pointed out that coke sales often represent as important a source of revenue as the gas actually sent through city mains. Until recently, however, many plants were to a certain extent independent of the coke market, since practically the whole coke output could be used in the plant as a water-gas fuel. The great flexibility of the water-gas process has a decided advantage for meeting peak-load conditions, but, due to its lower thermal efficiency (as compared with bulk carbonization) and to the certainty of increased prices of the oil used for enrichment, the trend is more and more toward carbonization. Water-gas equipment is low in first cost, however, and, on account of the flexibility, blue water gas and carburetted water gas will probably continue to be used to carry peak loads, even though the gas itself may be somewhat more expensive. The expansion of carbonization in city gas manufacture depends on the development of the domestic market for coke or on some process for the complete gasification of coal. That the use of coke for domestic heating has progressed materially in certain localities is evidenced by the fact that some gas plants are finding it profitable to sell their own coke for domestic consumption and to purchase the necessary fuel for water-gas generators from other sources.

In deciding on the type of equipment, whether horizontal or vertical retorts, or by-product ovens, it is necessary to investigate carefully the markets and the market prices of all the products—gas, coke, ammonia, and tar—and to balance the revenue obtainable from the yields of these products against the first cost, fixed charges, and operating costs on each type of equipment.

Considerations in Low-temperature Carbonization.—Low-temperature carbonization undoubtedly will become of importance in the future, in spite of the relatively small degree of commercial success attained thus far by this method, both in Europe and in the United States, and it must be realized that low-temperature carbonization has not yet grown entirely out of the experimental stages.

The widespread adoption of this type of carbonization depends on the reduction of equipment and operating costs to a point which is at least on a par with those of the present high-temperature processes, and on the opening up of new markets for the special form of by-products obtained. A comparison of costs and returns in coal carbonization at both low and high temperatures, as drawn up by Porter,[28] is shown in Table CXLVII.

Thus, with the present prices obtainable for by-products, low-temperature processes have to operate with a smaller margin of profit than the modern high-temperature methods. With the depletion of our best seams of anthracite, and the consequent increased demand for a free-burning, smokeless domestic fuel, with the discovery of new fields for the application of low-temperature tars and oils, with higher yields of gas, and by utilizing the lower grades of coal, it seems probable that low-temperature carbonization may attain a sound economic footing. Low-temperature carbonization offers a possible method of pro-

TABLE CXLVII

Costs and Returns in Coal Carbonization

(Based on 1 lb. coal)

	High-temperature process, cents	Low-temperature process, cents
Costs:		
Coal at $5 per ton	0.250	0.250
Conversion expense, fixed charges, etc.	See below	See below
Returns:		
Coke at $6 per ton	0.174	0.204
Gas at 5 cts. per 100,000 B.t.u.	0.165	0.090
Tar and oils at 5 and 6 cts. per gallon	0.030	0.072
Motor fuel at 20 cts. per gal.	0.026	0.026
Ammonia at 3 cts. per lb. sulfate	0.038	0.012
	0.433	0.404
Balance, available for conversion costs, fixed charges, and profit	0.183	0.154

ducing a high B.t.u. gas, of high density, which will be of increasing value to gas companies as the advancing price of oil makes the enrichment of water gas uneconomical.

CARBONIZATION EQUIPMENT

The types of equipment used for coal carbonization consist of: (1) beehive coke ovens, applicable only for the manufacture of metallurgical coke; (2) by-product coke ovens, which may be used for the simultaneous production of metallurgical coke and city gas; and (3) the retorts used primarily in the manufactured-gas industry, prominent among which are continuous vertical retorts, intermittent vertical retorts, and the familiar D-type retorts.

The relative sizes and shapes of these ovens and retorts for carbonizing coal according to Porter are shown[30] in Fig. 255, and general data concerning them are collected[30] in Table CXLVIII in order that the relationships between the various types may be kept in mind during the subsequent descriptions of individual retorts.

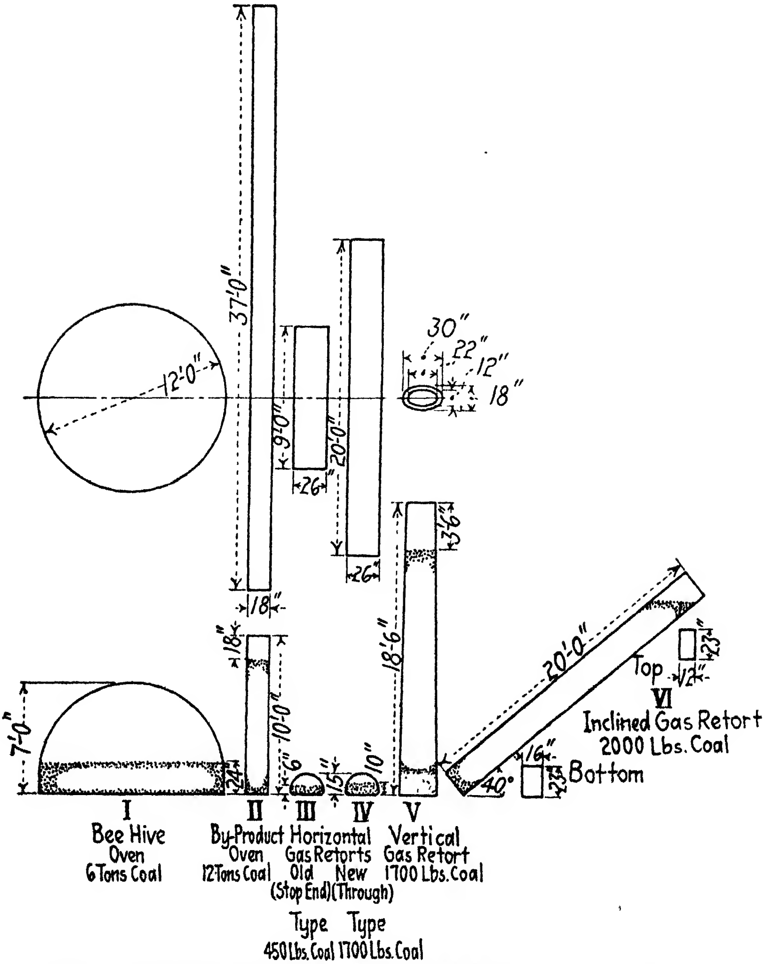

FIG. 255.—Comparison of dimensions and charges of different ovens and retorts for carbonizing coal plans and elevations.

TABLE CXLVIII

Comparative Data on Various Carbonizing Processes

	Beehive coke oven	By-product coke oven	Vertical gas retort*	Horizontal gas retort	
				Old†	New‡
Coal charge, tons	6.5	12.0	0.85	0.23	0.85
Coking period, hours	48–72	11–18	11	4	6
Gas made per hr, 1000 cu ft	.	7.3–12	0.87	0.65	1.63
Coking velocity, inches per hr	0.5	0.6	0.7	1.1	1.0
Heated wall area, square feet	..	795	97	53	130
Wall area per 1000 cu ft gas per hr		105	112	82	80
Volume of charge, cubic feet	290	472	35.5	9.4	34.5
Depth of charge, feet	2.2	8.5	15.5	0.5	0.82
Open space above charge, cubic feet		83.3	5.2	10.3	10.4
Open space per 1000 cu ft gas per hr		10.0	6.0	15.8	6.4
Approximate temperature in free space, degrees Fahrenheit		1750	1450	2100	1900
Approximate time of gas contact with free space, seconds		39	21.5	58	23

* U G I intermittent type, 18½ ft long, 12- by 22-in top, 18- by 30-in bottom, filled within 3 ft of top

† Old "stop-end" type, 15 by 26 in by 9 ft, charged to 6-in depth

‡ Modern "through" type, 15 by 26 in by 20 ft charged to 10-in depth

The characteristics of these types of equipment,§ as well as descriptions of some of the processes for low-temperature carbonization and complete gasification, are given below.

Beehive Coke Ovens.—In the ordinary type of beehive coke ovens (shown in Fig. 256) coal is charged through a door in the top and leveled to a height of approximately 2 ft. Air is admitted through the door at the side for the combustion of the volatile products in the dome-shaped space above the coal bed. The heat thus generated serves to keep the brickwork hot and radiation therefrom assists in coking the coal. Additional heat is available for this purpose from the actual combustion of a portion of the fixed carbon in the coal.

After the oven has been in continuous operation for some time, the heat retained by the walls and surrounding materials is

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§ It is beyond the scope of a work of this kind to attempt the description of all the types of ovens and retorts used. This chapter describes only those ovens and retorts which are of major importance in current American practice. For descriptions of equipment not discussed here, see references 14, 16, and 31 to 37 in the Bibliography.

sufficient to heat the coke to between 575 and 750°F. (302 to 400°C.), when the actual evolution of combustible gas takes place. The gas comes out slowly at first and then more rapidly as the temperature rises. After a few hours the gas is of sufficient quantity and quality to maintain a continuous fire with the air admitted above the charge. The whole retort gradually becomes hotter until after a period of two or three days the entire content of volatile matter is expelled. The coke is then quenched inside the oven and raked out through the door at the side This process is evidently wasteful, the only recoverable product being coke, and from 10 to 15 per cent of this is usually consumed by the air admitted to the coking chamber.

The more modern types of beehive oven are arranged in double rows with doors on opposite sides for unloading. The loading tracks usually are constructed on the earth mound between the two banks of ovens, so that the retort does not have to withstand the strain. The oven itself is usually from 12 to 13 ft. in diameter and 6 to 8 ft. high. The walls and the bottom of the oven are constructed of fire brick, while silica brick is generally used for the dome.

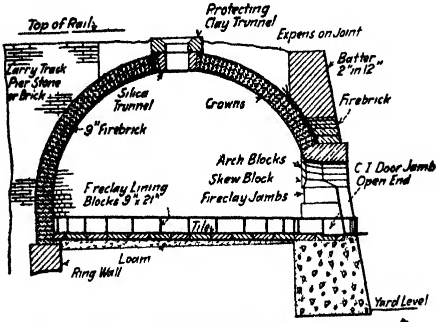

Fig 256 —Typical beehive coke oven

Among the most important improvements in the beehive oven is the use of the hot gases for heating the bottom of the oven and for the generation of steam in boilers. The temperature of the gases leaving the ovens in most cases is in the neighborhood of 2000°F. (1093°C.), and the steam generation by the use of waste-heat boilers[29] may be as high as 1 to 2 lb. of water per lb. of coal carbonized. Even with the recovery of this heat, however, the beehive oven still loses a very valuable portion of its by-products and is extremely inefficient. Except for the low first cost and the fact that blast-furnace men considered beehive coke tremendously superior to by-product coke, the beehive oven would long since have disappeared.

As may be seen from[1] Fig. 257, the wasteful beehive oven is rapidly being replaced with the more efficient by-product type. From only about 5 per cent of the total United States production in 1900, the percentage of by-product coke has increased until in 1924 it represented about 78 per cent of the total. The time is not far distant when the beehive ovens will be practically eliminated from the carbonization field, and they are rapidly becoming more and more an auxiliary source of supply, to be called on only in boom times when by-product oven capacity is insufficient to meet the coke demand.

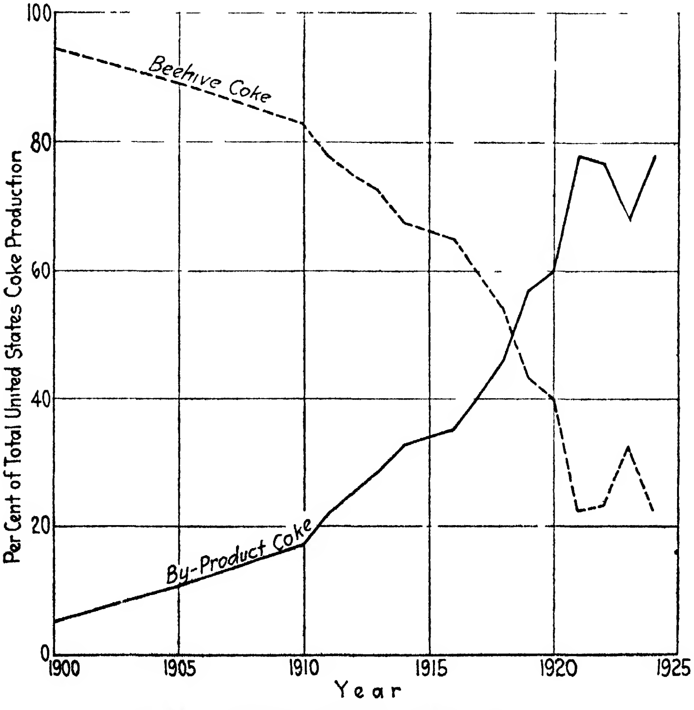

FIG. 257.—Percentages of coke produced in beehive and in by-product coke ovens in the United States.

By-product Ovens.—The modern by-product oven is essentially a long, narrow, rectangular chamber, filled with coal and heated from without by gas passing through flues in the sides of the oven. For metallurgical coke this chamber may be 30 to 40 ft. long, 11 to 22 in. wide, and 6 to 13½ ft. high. Coal is charged through a series of holes in the top of the oven and the coke discharged by means of a ram or pusher, ejecting the coke from the end of the oven. The oven is tapered from the pusher to the coke (discharge) side in order to increase the ease of pushing.

In general, the materials of construction are the same in almost all ovens. A highly siliceous brick is used for the oven linings and the flues. British manufacturers prefer a brick containing 80 per cent SiO_2 and 14 per cent Al_2O_3, while American practice is to use a brick containing 92 to 96 per cent SiO_2 because of its higher heat conductivity and greater durability.

The fundamental difference between the various types of by-product ovens lies in the method of heating and heat recovery. A division can first be made between ovens with vertical and horizontal heating flues, and each of these may be further subdivided into waste-heat, recuperative, and regenerative types. The waste-heat type is not efficient and has not found favor with American operators. The most important types of by-product ovens in American and British practice are listed below in accordance with this method of classification.

HORIZONTAL FLUE
Recuperative:
  Semet-Solvay
  Rothberg

Regenerative:
  Semet-Solvay

Waste-heat:
Huessner
Simon Carvés

VERTICAL FLUE
Recuperative:
  Otto-Hilgenstock

Regenerative:
  Otto Hoffman
  Koppers
  Simon Carvés
  Coppee
  Collin
  Bauer
  Piette
  Wilputte

In order to produce a uniform coke, the rate of heating throughout the charge must be the same. Uniformity of heating is the aim of all oven designers and superiority in this direction is claimed for both vertical- and horizontal-flue types. But the temperature must not only be uniform—it must also be high. The heating flues are separated from the coking chamber by a wall of brick of relatively low conductivity, so that a high flue temperature (usually in excess of 2000°F., or 1094°C.) must be maintained in order to insure rapid heat transfer through this wall. With these high temperatures, a large fraction of the heat in the gas is not utilizable unless recuperative or regenerative methods are employed (see Chap. XI). With the use of these heat-recovery methods, the amount of heat available from the combustion of the gases is greatly increased and the heat loss up the stack materially lessened. When the ovens or retorts are run primarily to produce gas, it is advisable to substitute producer gas or water gas in place of the coke-oven gas used in most metallurgical coke ovens. With the lean producer gas, preheating of both gas and air is necessary in order to maintain high efficiency.

As will be seen from the following description of the important ovens, practically all employ the regenerative system of heat recovery.

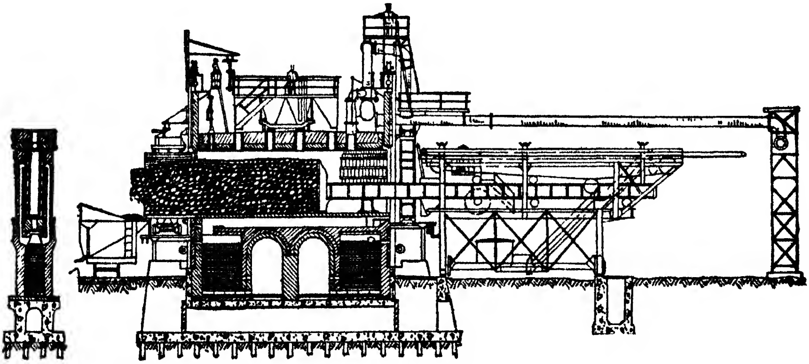

FIG. 258.—Semet-Solvay standard regenerator coke-oven

In all the varied types of retorts the basic design of the oven chamber itself is the same. A rectangular chamber, varying slightly in length, height, and width, but with the same general characteristics, serves as the coal container. The factors which affect the actual length and height of the oven are the method of construction and the desired capacity of the individual retort. The width of the oven is dependent on the type of coal to be carbonized and the size of coke required. For poor coal a narrow oven is essential, so that the rate of heating may be rapid. With a better coal, the rate of heating may be slower and a wider oven may be used, producing a larger size coke.

The Semet-Solvay By-product Coke Oven.—The Semet-Solvay oven is the chief exponent of the use of horizontal flues, and was the first type of by-product oven built in the United States. A longitudinal section through the regenerative oven, showing the pusher machine, coal larry, coke car, and the regenerators beneath the oven, is shown in Fig. 258. A diagram of the heating system of these ovens is shown in Fig. 259. The figure shows the path of the gases on "down draft," or when flowing from the top to the bottom of the oven. Air enters from a blower into the waste-heat flue, passes through the regenerator, where it is heated to 1300 to 1500°F. (704 to 816°C.), then up the "riser" into the top flue (No 1). It then passes down through the six side flues into the "sole flue" under the oven floor, meeting on the way and burning the gas which enters at the five points indicated All the air for combustion thus travels through the entire flue system and there is considerable excess air at each of the first four burners. Complete and rapid combustion, with a high rate of heat transfer due to the high gas velocity, is thus assured After a period of 15 to 20 min., the air is reversed by a simple manipulation of the stack and air-inlet dampers so that "up-draft" combustion takes place. The gas need not be reversed, since the direction of air flow regulates the direction of combustion and on up draft the gases are burnt in the flue above the burner while on down draft combustion takes place in the flue below.

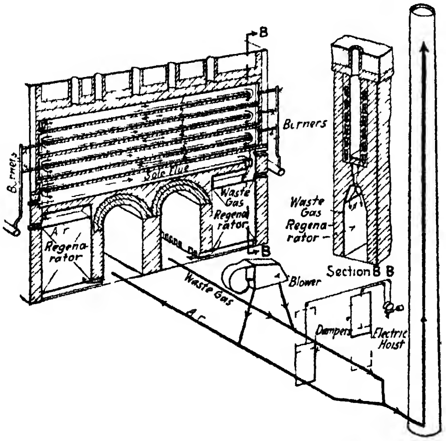

FIG 259 —Gas flow in a Semet-Solvay coke oven on the down draft — diagrammatic

Koppers By-product Coke Oven.—The Koppers by-product ovens are of the vertical-flue type and have been adopted in the United States more widely than any other make. Koppers ovens are of two general types: (1) the older so-called cross-regenerator oven, and (2) the new Becker-type oven.

A cross-section through the ovens, flues, and regenerators of the cross-regenerative type is shown in Fig. 260. The ovens are separated from each other by a series of vertical flues and have individual regenerators running beneath the ovens parallel to the oven axis. There are 30 vertical flues on each side of the oven. The fuel gas enters through a special nozzle A, burns with preheated air from the regenerators B, passes up through the flue C of half the oven, the products of combustion meet in a common horizontal flue D above, and then pass down through the flues in the other half of the wall, giving up part of their heat as they leave through the regenerator. Reversals, of both gas and air (compare with Semet-Solvay ovens, which only require reversal of air) take place each half hour. The design of the Koppers oven permits the regulation of the fuel-gas and air supply to each flue. The fuel-gas supply is regulated by the interchangeable nozzle A and the draft or air supply is regulated by the individual damper blocks or "sliding bricks" E.

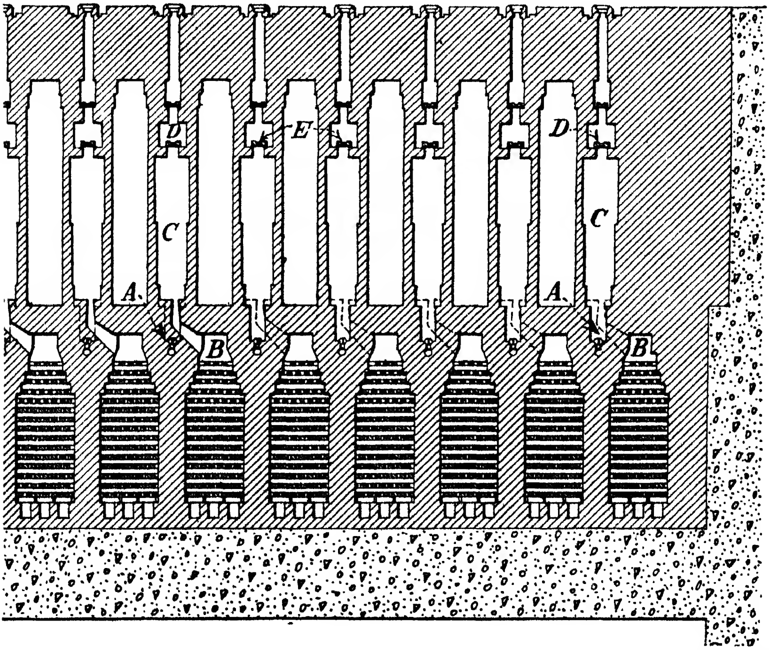

Fig. 260.—Cross-section through Koppers cross-regenerator coke oven. (*The Koppers Company.*)

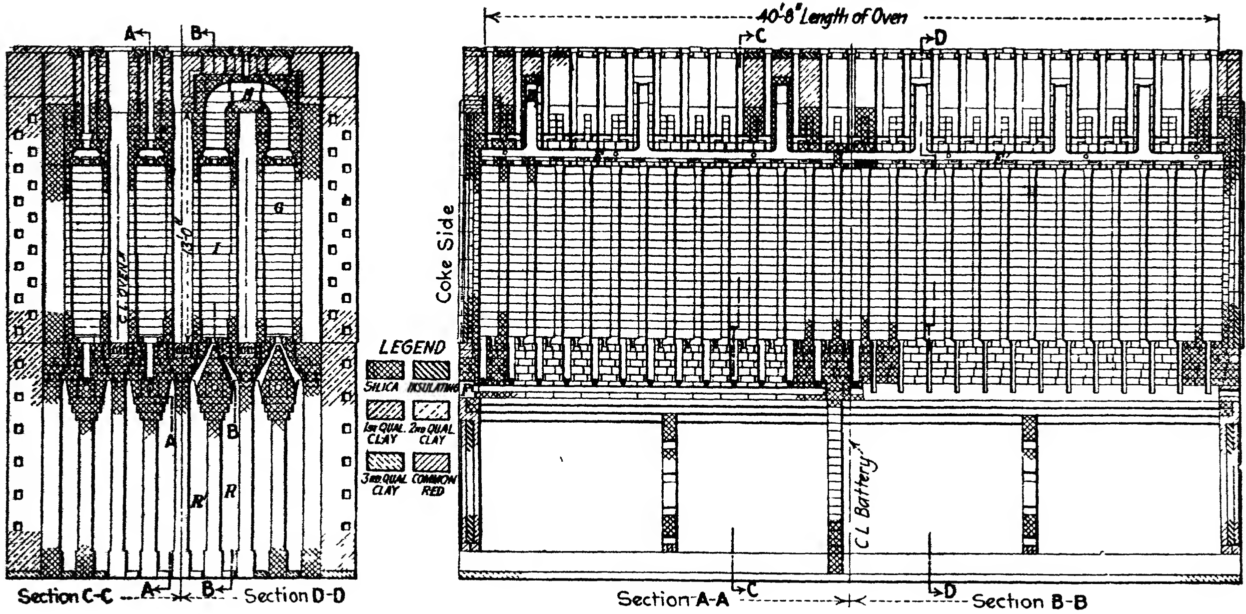

FIG. 261.—Cross-section through a Becker-type by-product coke oven. (*The Koppers Company.*)

A cross-section through the Becker-type oven built by the Koppers Company is shown in Fig. 261. This oven is usually, but not necessarily, narrower than the older type (an average of about 14 to 16 in. as compared with 18 to 20 in.) and, instead of using a common horizontal top flue extending the length of the oven, cross-over flues (connecting the smaller common flues *E* and *E'* used in the new type) are provided. The gas enters at *F*, burns in the flues *G* on one side of the oven, passes through the cross-over flues *H*, and goes down the flues *I* on the other side. Divided regenerators, *R* and *R'*, are provided, so that when it is advisable to conserve coal gas in the plant preheated producer gas may be used for oven heating. A diagrammatic repre-

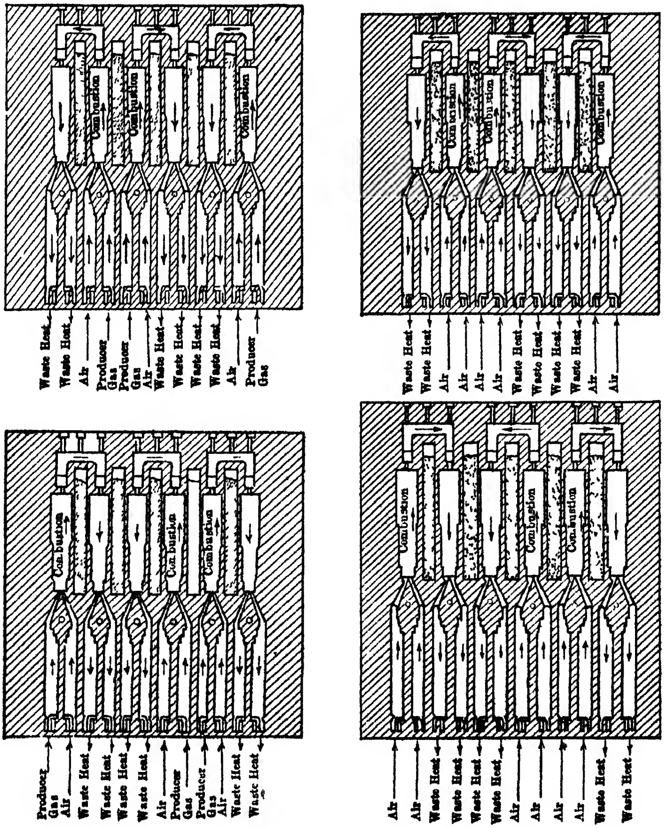

FIG. 262.—Gas flow in the Becker-type Koppers coke oven. The figures on the left indicate the flow in each reversal when producer gas is used. The figures on the right indicate the flow when a rich gas is used for oven heating and the air alone is preheated.

sentation of the flow of fuel gas and air in the Becker oven is shown in Fig. 262. It will be noted that there is another important difference between the older Koppers and the Becker ovens. In the Becker type *all* the flues on one side of the oven are operating either with the gases burning as they pass up through the vertical flues, or, when the flow is reversed, they serve to take the downward-flowing products of combustion. This means that there is very little pressure difference between adjoining flues and the tendency for leakage is, therefore, at a minimum. When producer-gas heating is used (see Fig. 262) one of the regenerators serves to preheat air, and one is for producer gas; when the richer coke-oven gas is used, the air alone is preheated. This arrangement allows great flexibility in the coke-oven gas-send-out. When coal gas is in great demand, the ovens may be heated entirely with producer gas; when the coal gas demand is light, the supply of producer gas may be shut off and the heating requirements met with coke-oven gas alone. It is evident that intermediate adjustments are possible between these two extremes, one part of a battery of ovens being heated with coke-oven gas, and the remainder using producer gas, blast-furnace gas, or blue-water gas.

The Koppers Company also build a smaller oven for use in city gas plants. This type of oven is from 10 to 13 ft. high, 19 to 24 ft. long, and has an average oven width of 13½ in. These ovens are of the Becker type and were designed primarily for the smaller gas plants where the gas send-out is relatively low.

The Wilputte cross-regenerative oven is similar to the Koppers type above the level of the oven floor, but differs from the Koppers in the construction of regenerators and in the method of supplying preheated air. Individual regenerators are supplied for each flue, and air under pressure from a blower (as contrasted with the natural draft in the Koppers oven) is admitted through adjustable openings at the bottom of each regenerator.

The Roberts by-product ovens, of which a few batteries have been built in the United States, differ radically from the horizontal- and vertical-flue types just described. In the Roberts type, each oven is provided with independent heating walls (*i.e.*, two ovens are separated by two heating walls, as in the Semet-Solvay ovens). These walls are "flueless," being made up of a special type of staggered checkerbrick. Half of the gas and all of the preheated air enter at the top of the heating wall, and burn as the mixture passes downward through the checkerbrick. The second half of the fuel gas is admitted about 6 ft. below the oven top. The waste gases pass into a sole flue at the base of the oven and then are drawn through a bus flue into recuperators. There is no reversal in this type of oven, since recuperators are employed instead of regenerators.

Energy Relations in Coke Ovens.—Heat balances on two important types of by-product coke ovens are shown[38,39] in Table CXLIX. Both sets were operated with coke-oven gas as the heating medium. Obviously, the major source of heat input is the potential heat in coal, while the major sources of output

TABLE CXLIX

COKE OVEN—TOTAL HEAT BALANCE

	Type of oven	
	Semet-Solvay[38]	Koppers[39]
Number of ovens	60	106
Duration of test	54 hr.	
B.t.u. of coal	14,470	
Per cent volatile in coal	26.0	25.77
Input:		
Potential heat in coal	92.82*	91.0*
Sensible heat in coal	0.013	0.01
Potential heat in gas	7.15	9.1
Sensible heat in gas	0.013	0.01
Sensible heat in air	0.000	0.00
Output:		
Potential heat in coke	61.60	64.50
Sensible heat in coke	4.14	3.72
Potential heat in gas	21.04	20.55
Sensible heat in gas	1.24	1.30
Water in gas	0.97	0.87
Potential heat in tar	4.84	4.02
Sensible heat in tar	0.22	0.06
Sensible heat in stack gases	1.29	3.30
Latent heat in stack gases	1.03	
Radiation and convection	1.19	1.20
Unaccounted-for losses	2.54	0.48
Per cent of gas used in ovens	36.5	44.2
Efficiency	87.4	89.0

* Figures expressed as per cent of total.

are potential heat in coke and gas. These values are so large as completely to overshadow the losses, and the efficiency of the process is very high, 87 to 89 per cent. By omitting the latent heat in the raw material and products a *sensible-heat* balance may be set up as in Table CL.

TABLE CL

COKE OVENS—SENSIBLE-HEAT BALANCE

	Type of ovens	
	Semet-Solvay, per cent	Koppers, per cent
Input:		
Sensible heat in coal	0.18	0.11
Sensible heat in gas	0.18	0.10
Sensible heat in air	0.03	0.00
Potential heat in gas	99.60	99.80
Output:		
Sensible heat in coke	57.70	41.2
Sensible heat in gas	17.24	14.5
Water in gas	13.50	9.7
Sensible heat in tar	3.08	0.6
Sensible heat in stack gases	17.97	36.6
Latent heat in stack gases	14.30	
Radiation and convection	16.66	13.3
	140.45	115.9
Heat of reaction (B.t.u. per lb.)	452	210

A diagrammatic presentation of the sensible-heat distribution in the Koppers oven is shown[39] in Fig. 263. The same data are presented as in Table CL, but the diagram serves to make this type of sensible-heat balance more clear.

In the sensible-heat balance the input is essentially the potential heat of the gas used for oven heating. The distribution of this heat is of more significance than that of the total potential heat of the gas and coal, since little change is made in the latter. The losses due to sensible heat in the coke and as sensible heat in the stack gases are seen to account for 80 to 90 per cent of the heat in the fuel gas. Moreover, in both tests there was a surplus of sensible heat in the output, indicating that part of the potential heat of the coal has been converted to sensible heat. The difference in the excess observed in the two tests may be due to the use of different specific-heat constants, or to the cumulative error in all the other measurements, but in all probability it is due either to the coal burning by air leakage or else to the oxidation of the coal by absorbed or entrapped air. This extra heat appearing in the sensible-heat balance is called the exothermicity of the carbonization process, and has been shown by Davis and Place[40] to be dependent largely on the amount of oxygen retained in the coal. Under certain conditions (heating in nitrogen) the coking reactions themselves were found to be only very slightly exothermic.

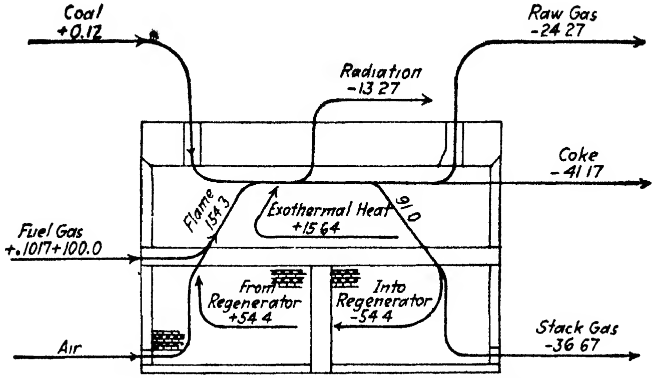

FIG. 263.—Sensible heat balance on a Koppers coke oven.

D-type Gas Retorts.—In the early days of the manufactured-gas industry, small D-shaped horizontal retorts (capable of taking a charge of from 400 to 500 lb.) stopped at one end were widely used. These retorts were inefficient, had high operating charges, and were subject to excessive naphthalene deposits; on account of these disadvantages they have become practically obsolete. Newer, larger types of through D-shaped retorts, both horizontal and inclined, are subject to the same disadvantages, though to a lesser degree. They are retained mainly in plants which have a relatively small gas send-out and where their low first cost and flexibility make it inadvisable to replace them with more efficient but costlier equipment.

A section through a relatively new (1922) installation of horizontal through D-type retorts at the Lowell (Mass.) Gas Light Company is shown in Fig. 264 and is typical of this kind of retort setting. The D-shaped silica retorts extend through the entire setting, and have a capacity of 1500 to 1800 lb. of coal

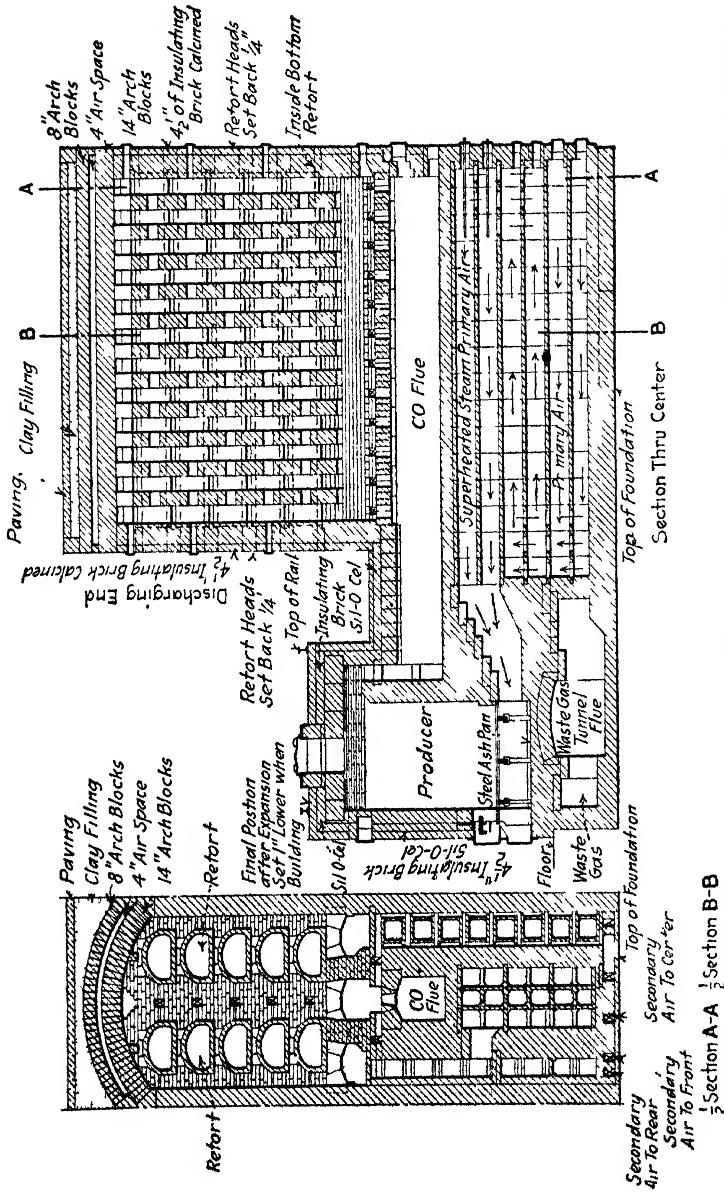

FIG. 264.—Horizontal D-type retorts at the Lowell (Mass.) Gas Light Company (The Russell Engineering Company.)

per charge. The carbonizing period varies from 8 to 12 hr. In this, as in other modern installations, the retorts are mechanically charged and pushed and therefore labor charges are lower than with the old stop-end retorts.

Primary air is drawn through and preheated by a recuperator in the base of the setting. Steam is added in the recuperator, is superheated and mixed with the primary air, and the mixture passes into the base of a producer, which is fed with hot coke from the retorts. The hot producer gas enters a flue beneath the bench (over the recuperators), burns with secondary air (preheated by passing through the setting, as indicated) up around the retorts, and the waste gases are drawn through the regenerator. In this particular installation, there are ten retorts per bench.

Inclined retorts, which have the obvious advantages of ease in charging and gravity discharge, have been quite widely used. Like the horizontal D-type, however, their use is restricted at the present time to relatively small or old plants, and most new big construction is of the vertical-retort or coke-oven type.

Vertical Retorts.—Following the adoption of inclined retorts, with the attendant ease of charging and pushing, the next step in the evolution of carbonizing chambers for gas manufacture was the application of vertical retorts. These retorts, which may be either intermittent or continuous in operation, permit automatic feed and gravity discharge of the finished coke. Both of these types effect a considerable saving in labor over the older horizontal retorts, and can compete on even terms with the by-product coke oven as far as thermal efficiency and recovery of by-products are concerned. Although the coke produced cannot be used for metallurgical purposes, for a gas company this disadvantage is offset by the great flexibility of the vertical retort. As will be shown later, the thermal yield of gas is greatly increased by passing steam through the incandescent coke, and vertical retorts are particularly adapted for "steaming." This type of retort is small, and therefore possesses the advantages of rapid rates of carbonization. Due to the constant movement of coke in continuous vertical retorts, and the consequent breaking up of the coke formation, coke from these retorts is apt to be of relatively small size.

If strongly swelling coals are used in vertical retorts (either continuous or intermittent) trouble is experienced due to the charge "hanging up" or sticking in the retorts. Hand poking then has to be resorted to in order to remove the charge. The exact cause of swelling is not certain, but the best work seems to be that of Franz Fischer (see Chap. II, p. 34), who showed that swelling occurred if the decomposition temperature of the oily bitumen obtained by benzol extraction came within the plastic temperature range of the coal. The data[6] in Table CLI for three Pennsylvania gas coals show that swelling seems to occur if the coal has a wide plastic range and if the amount of gas distilled off during the plastic range is large. Swelling, at least with these coals does not seem to be due to the volatile matter nor to the α, β, or γ constituents (see Chap. II).

TABLE CLI

Swelling Characteristics of Coal

Coal	Index of swelling	Plastic range		Temperature at which swelling is most rapid, degrees Fahrenheit	Volume of gas given off during plastic range	Per cent volatile matter	Analysis of coal, per cent constituents		
		Fusion temperature, degrees Fahrenheit	Hardening temperature, degrees Fahrenheit				α	β	γ
Gas coal A	6.25	662	1380	770	201	31.6	72.9	18	9.1
Gas coal B	4.75	662	960	761	62	35.5	70.5	20.6	8.9
Gas coal C	4.75	662	887	752	51	32.9	71.1	18	10.9

The swelling index was determined by the rise in inches of a weighted plunger resting on the coal when the coal was coked under standardized conditions. Many of the strongly coking coals, such as Connellsville, gave a swelling index of less than 1.0. Steaming, probably by causing a more gradual heating of the coal, reduces the hanging in continuous vertical retorts.

Intermittent Vertical Retorts.—The U. G. I. system of intermittent vertical retorts, shown in Fig. 265, is the principal exponent of this type in the United States. The retorts, elliptical in shape, 10 by 22 in. at the top and 18 by 30 in. at the bottom and about 22 ft. long, are set in benches of nine. The retorts are charged by gravity from a traveling larry at the top of the battery of benches. The coke is discharged by gravity into a coke car and is relatively cool, since steam is introduced at the bottom of the retorts during the last 1½ hr. of the coking period.

Benches of these retorts may be provided with a built-in producer, or, as for the setting shown in Fig. 265, may receive

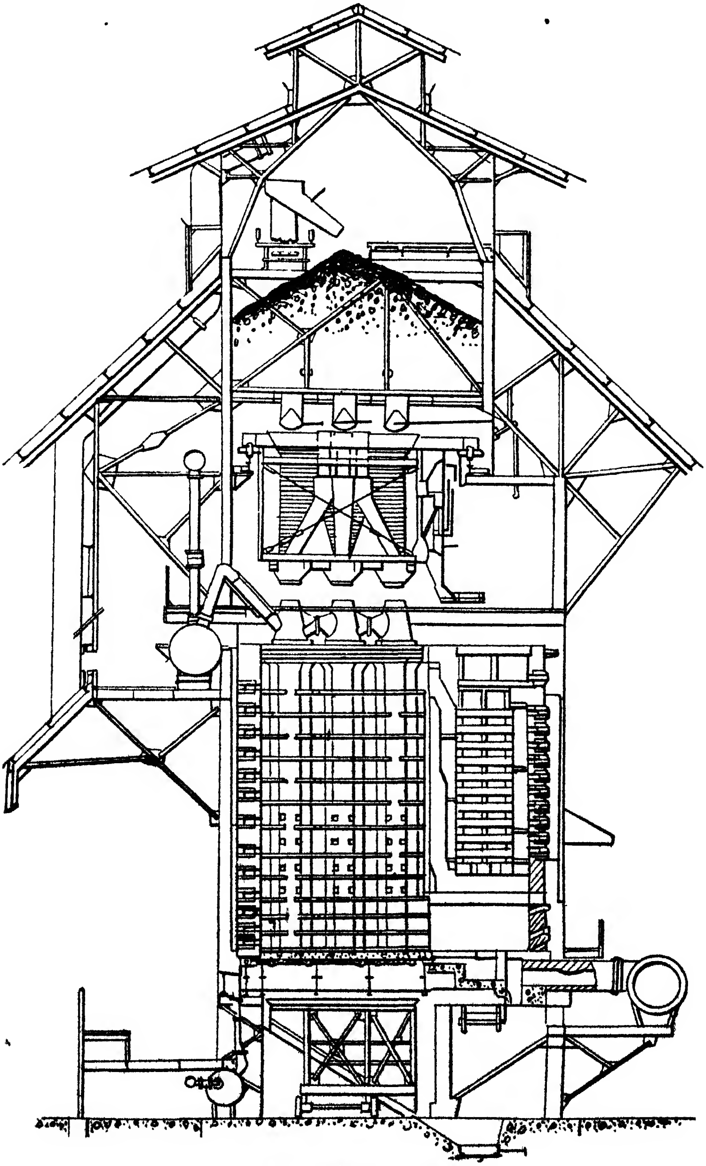

FIG. 265.—U. G. I. intermittent vertical retorts. Central producer type. (*U. G. I. Contracting Company.*)

producer gas from a central producer. The producer gas (at a temperature of about 1000°F.), from which the major part of the dust has been removed by a dust-catching system, enters a small mixing chamber at the base of the setting, where secondary air (preheated to about 1600°F. in the recuperators) is admitted. The burning gases then pass up around the retorts and the products of combustion are drawn down through the recuperators and pass through a waste-heat boiler to the stack. The retort temperatures range from about 2200°F. at the bottom to about 1650°F. at the top, and therefore the top of the charge is liable to be slightly "green." The products of combustion leave the recuperators at about 1100°F. and their temperature is lowered to about 500°F. in passing through the waste-heat boiler.

A heat balance on an installation of these retorts at Rochester, N. Y., showed the following results:[41]

TABLE CLII

HEAT BALANCE ON INTERMITTENT VERTICAL RETORTS

	B.t.u. per lb. coal carbonized	Per cent of total heat input
Heat input (to producers)	1,940	100.0
Heat output		
Heat used in carbonizing	1,078	55.5
Heat utilized in waste-heat boilers	304	15.7
Heat lost in producers	227	11.6
Heat lost in producer flue	20	1.1
Heat lost in stack	312	16.1

The table shows the large amount of heat recovered in the waste heat boilers.

Continuous Vertical Retorts—In continuous vertical retorts coal is fed in by gravity at the top, and coke is continuously extracted by mechanical means at the bottom. Manual labor, therefore, is at a minimum and operating costs can be kept at a low figure. In addition, the gaseous products of carbonization (part of which pass up through the cool central core of the charge) come in contact with the cold entering coal and are not subjected to excessive temperature in the free spaces. This results in a rich gas and in high yields of by-products and tar (since thermal decomposition is at a minimum). The lower portion of the retort can be cooled by preheating the entering secondary air for combustion or by the admission of steam through the charge, and the coke, therefore, can be withdrawn from the retort cool enough to obviate the necessity for quenching. Although developed and

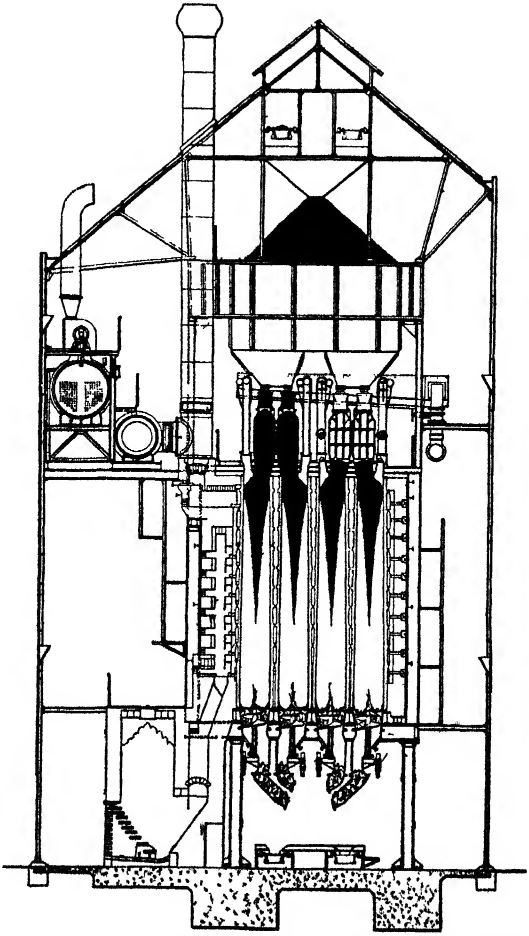

FIG. 266.—Glover-West vertical retorts. (*West Gas Improvement Company of America.*)

principally used in England, this system of carbonization is being adopted in a number of the larger gas plants in the United States.

The Glover-West system of vertical retorts, shown in Fig. 266, uses elliptical retorts (as does the U. G. I. intermittent system) which expand toward the bottom and (in the later installations[42]) flare into a circular chamber where the coke is cooled by the addition of steam. The retorts are set in benches of from four to ten with from two to five retorts in a row Producer gas from an open step-grate producer enters the setting through an uptake, and passes to a series of horizontal flues or combustion chambers around the retorts Here secondary air, preheated by

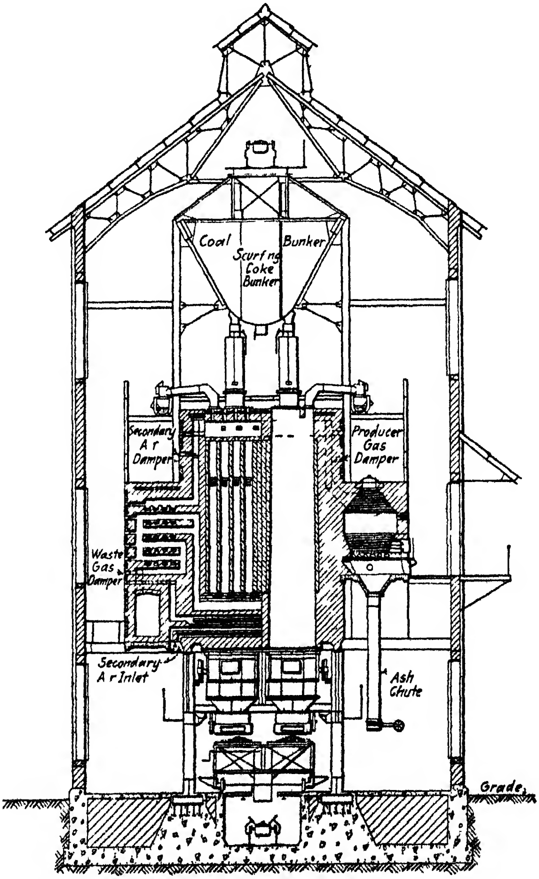

Fig 267 —Woodall Duckham continuous vertical retorts (*Isbell-Porter Co*)

passing across the setting outside the combustion chambers, is admitted, and combustion takes place across the setting. Th waste gases pass up to two horizontal flues at the top of the setting and out through a waste-heat boiler to the stack. Th amount of primary air under the producer grate is controlled b the stack draft, and the admission of secondary air is controlle by adjustable dampers within the setting. A helical coke extractor at the bottom of the retort removes the coke continuously.

The tops of the retorts are maintained at a lower temperature than the bottom, in order to minimize decomposition of the hydrocarbon constituents of the coal gas. The through-put in this system (as built in the United States) is about 5 tons per retort per day.

The Woodall-Duckham system of continuous verticals, shown in Fig. 267, differs markedly from the Glover-West, in that the retorts, instead of being elliptical in shape, are built in the form of rectangular chambers, tapered toward the bottom to assist in the extraction of the finished coke. These retorts, which usually are built in benches of two or four, have a capacity of about 5½ tons of coal per day, though they have been operated at capacities as high as 7½ tons. The through-put is regulated by the speed of a continuous coke extractor at the bottom of the retort, the heat, of course, being regulated in accordance with the rate of carbonization.

The hot gas from the producer passes to the top of the setting, where secondary air, preheated in the recuperator, is admitted. Combustion takes place as the gases travel down through vertical flues parallel to the retort walls, and the products of combustion are collected in a common flue and pass through the recuperator and a waste-heat boiler on their way to the stack. This system gives the highest temperatures at the top of the retorts and therefore would be expected to cause a greater degree of decomposition in the gas than takes place in the Glover-West system, in which the top temperatures are lower. The Woodall-Duckham retorts are larger than the Glover-West, however, and therefore higher temperatures are required to force the fusion zone through the coal. As indicated in Fig. 267, secondary air is admitted through flues near the base of the retorts, and is partially preheated there, thus serving to cool the coke before extraction.

In all gas retorts, whether horizontal, inclined, or vertical, a dense deposit of carbon collects on the retort walls during carbonization. This deposit not only interferes with heat transfer, but also lowers the capacity, and therefore must be removed periodically. The removal (called "scurfing") consists in opening the retort and allowing the carbon to be consumed by the oxygen of the air. In horizontal, inclined, and intermittent vertical retorts, raw coal may be charged as soon as scurfing is complete. This is not practicable with continuous retorts, however, and after scurfing, continuous verticals are first filled with coke (see scurfing coke bunker in Fig. 267) before coal is again charged.

Steaming in Vertical Retorts.— The practice of steaming has been an important factor in the development of continuous vertical retorts for gas-making purposes. In this process, a controlled amount of steam is introduced into the bottom of the retort, reacting with the hot coke to form water gas, and to some extent furthering the devolatilization of the coke. Since the quality and the quantity of the gas, coke, and by-products vary considerably with the amount of steam used, a number of tests have been conducted to determine the most satisfactory operating range. In the most comprehensive of these tests (Gas Investigation Committee of the Institution of Gas Engineers[43]) Lanarkshire coal was carbonized in Glover-West retorts with amounts of steam varying from 0 to 49.9 per cent of the weight of the dry coal. The most important results from these tests are summarized in Tables CLIII, CLIV, and CLV. Table CLIII gives the results expressed on the basis of the coal used, Table CLIV the heat balances on the retorts under the various conditions (also expressed on the basis of a unit amount of coal), and Table CLV the quantity of additional constituents in the gas due to the use of steam.

From these tables it can be seen that steaming to the extent of 50 per cent of the weight of the coal had the following effects:

1. Increase of gas volume from 10,384 to 21,849 cu. ft. per ton, a 110 per cent increase.
2. Decrease in B.t.u. of gas from 544 to 410, a 24.6 per cent decrease.
3. Increase in thermal yield in the form of gas from 5,649,000 to 8,958,000 B.t.u., a 59 per cent increase.
4. Increase in inerts from 9.4 to 17.7 per cent of gas, an 88.5 per cent increase.
5. Increase in ammonium sulfate from 38.2 to 46.1 lb. per ton, a 21 per cent increase.
6. Increase in tar from 14.6 to 23.7 gal. per ton, a 62 per cent increase.
7. Decrease in coke from 1328 to 1152 lb. per ton, a 14 per cent decrease.

TABLE CLIII

RESULTS OF STEAMING IN CONTINUOUS VERTICAL RETORTS

(Per ton coal, dry basis)

	Test No. 1	Test No. 2	Test No. 3	Test No. 4	Test No. 5
Steam as per cent of coal	0	15.0	24.6	36.6	49.9
Cubic feet of gas	10,384	14,215	16,936	19,902	21,849
B.t.u. per cu. ft., gross	544	482	447	436	410
B.t.u. per cu. ft., net	485	431	401	393	368
Thermal value of gas (millions of B.t.u. per ton, gross)	5.649	6.852	7.570	8.677	8.958
Net	5.036	6.127	6.719	7.821	8.040
Dry coke, pounds	1,328	1,295	1,239	1,178	1,152
Tar, gallons	14.6	18.4	18.7	22.8	23.7
Ammonium sulfate equivalent to 24.5 per cent NH_3, pounds	38.2	39.0	40.6	44.6	46.1
Heat to setting, millions of B.t.u.	3.38	3.37	3.50	3.68	3.97
Heat required for steam, millions of B.t.u.		0.54	0.89	1.31	1.79
Analysis of gas					
CO_2	2.2	3.3	3.6	5.1	6.6
C_mH_n	2.3	1.8	1.8	1.8	1.6
O_2	0.4	0.6	0.9	0.9	1.0
CO	10.3	13.8	16.4	17.4	17.9
CH_4	28.5	23.5	19.0	16.8	14.5
H_2	49.5	49.5	48.9	48.8	48.3
N_2	6.8	7.5	9.4	9.2	10.1

TABLE CLIV

Effect of Steaming on Heat Balance of Continuous Vertical Retorts

(Dry basis, heat of combustion of coal carbonized = 100)

	Test No 1		Test No 2		Test No 3		Test No 4		Test No 5	
Heat in steam		Nil		1 31		2 17		3 18		4 35
Potential heat in coke		65 72		61 62		58 38		55 20		53 32
Potential heat in gas		9 26		11 20		11 28		13 76		14 45
Potential heat in tar		20 04		23 69		26 52		30 01		31 20
		95 02		96 51		96 18		98 97		97 97
Less heat in fuel to setting	11 84		11 65		12 27		12 76		13 82	
Less heat in fuel for steam (and total)	Nil	11 84	1 87	13 52	3 12	15 39	4 53	17 29	6 33	20 05
		83 18		82 99		80 79		81 68		77 92
Heat used in process (plus heat lost)		16 82		17 01		19 21		18 32		22 08
		100 0		100 0		100 0		100 0		100 0
Efficiency of gas production, per cent		54 37		58 21		57 99		62 09		58 56
Efficiency of carbonization, per cent		83 18		82 99		80 79		81 68		77 92

TABLE CLV

Effect of Steaming on Volumes of Constituent Gases

(Volumes in cubic feet per ton of dry coal)

	Test No 1, volumes per ton	Test No 2		Test No 3		Test No 4		Test No 5	
		Volumes per ton	Excess over test 1	Volumes per ton	Excess over test 1	Volumes per ton	Excess over test 1	Volumes per ton	Excess over test 1
CO_2	238	469	241	610	382	1,015	787	1,442	1,214
C_mH_n	239	256	17	305	66	358	119	350	111
O_2	42	85	43	152	110	179	137	218	176
CO	1,070	1,962	892	2,778	1,708	3,463	2,393	3,911	2,814
CH_4	2,959	3,341	382	3,218	259	3,344	385	3,168	209
H_2	5,140	7,036	1,896	8,281	3,141	9,712	4,572	10,553	5,413
N_2	706	1,066	360	1,592	886	1,831	1,125	2,207	1,501

8. An increase in the efficiency of gas production from 54.37 to 58.56 per cent.

9. A decrease in the efficiency of carbonization from 83.18 to 77.92 per cent.

It is also important to note the increased formation of methane (CH_4) and hydrocarbons (C_mH_n) which greatly increase the heating value of the gas (over and above the "normal expected" value, namely, that for a mixture of ordinary coal gas and the amount of blue water gas corresponding to the amount of steam used). This shows that the use of steam, by cooling the coke and increasing the velocity of the gases through it, reduces the temperature and time of contact sufficiently to lessen greatly the degree of thermal decomposition of the gas. The increased formation of ammonia is also due to this same effect.

Although the amount of gas made per ton of coal increased with each increase in steam, the actual quantity made per retort decreased when more than 36 per cent steam was used, owing to the excessive cooling of the lower part of the retort, and the resultant slow rate of carbonization. The heating of the lower portion of the charge by conduction from the flues is slow, and with too high steam rates it is impossible to maintain the proper temperatures The low B.t.u. value of the gas and the high percentage of inerts when steaming with over 30 per cent indicate that to go beyond this amount is not advisable with present methods. The point to which steaming should be carried must be determined from an economic balance of the relative costs of producing mixed gas by internal steaming and by the use of external water-gas sets. In current American practice lighter steaming, 10 to 18 per cent, is used in order that the B.t.u. of the coal gas may not be so low that it would be difficult to bring it up to heating-value standards by the admixture of a carburetted water gas. This limitation can be counteracted to a certain extent by the admission of oil to the top of the retort, with consequent enrichment of the coal gas.

The practice of steaming not only gives the increased yields indicated above, but also possesses the following additional advantages:

1. There is greater flexibility in coke production and in the quality and quantity of gas made.

2. Carbon deposition is decreased and the time for scurfing is therefore lessened.

It is interesting to note[44] that all modern installations of Koppers ovens for gas manufacture are being designed to permit steaming during the latter part of the coking period.

Low-temperature Carbonization Equipment.*—The retorts for carbonization at low temperatures may be divided into two main classes, externally heated and internally heated. Each has a number of advantages and each is still in the process of development. The great objection to the use of the ordinary coke oven for low-temperature work is the long coking time, due to the slow rate of heat flow through the coal mass at low-temperature differentials. In low-temperature carbonization, as previously brought out, only small driving temperatures can be employed, and therefore one or more of the following methods are adopted:

1. Mechanical agitation, in order to transfer heat by carrying hot coal into the cool part of the retort and the cool coal into the hot portion.
2. Internal heating with flue gas or other inert gas, thus rendering unnecessary the transfer of heat through the low conductivity coal.
3. The use of thin layers, so that the fusion zone has to travel over a short distance only, and therefore the time required is not excessive.

Certain low-temperature methods, particularly those in which the coal is heated slowly, or is slightly oxidized, destroy the coking properties, and therefore yield a friable coke. Under these circumstances the coke must be briquetted before it is sold for domestic heating.

The subsequent brief summary† indicates the major developments in low-temperature carbonization processes according to the most recent information available.

In the low-temperature process developed by Prof. S. W. Parr at the University of Illinois, two stages are used: The coal is first treated in a revolving drum heater, raised to a temperature of about 500°F. (*i.e.*, just below the fusion point), and then quickly transferred (in order to conserve the sensible heat and prevent excessive oxidation) to an externally heated vertical

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* It is neither possible nor advisable in a work of this kind to go into the details of the various types of retorts and processes for low-temperature carbonization. Changes in the various processes and the development of new types would make such discussion obsolete in a very short time. The various processes are described in references 16, 35, 45, and 46, of the Bibliography.

† Mainly based on statements made by PORTER[35] (March, 1925) and FIELDNER,[46] July, 1925

retort in which coking is completed. The maximum temperature employed is about 1300°F. With the high-oxygen Illinois coals which Parr uses, exothermic reactions in the coal itself supply part of the necessary heat. Since no agitation is used in the final stage of the process, a strong coke of excellent quality is obtained, with a volatile-matter content of from 4 to 7 per cent.

The Greene-Laucks process is being tried out on a commercial scale in Illinois. In this process the coal is propelled upward on a worm conveyor in a vertical retort in which heat is supplied both on the outside and within a hollow shaft. The primary objective in this case is to make a good solid fuel with recovery of tar oils, and no attempt is made to reduce the volatile matter below 15 to 17 per cent. It is reported that retorts operating with this process have carbonized, over considerable periods, as high as 24 tons of coal per day. The fuel ignites readily and burns with a smokeless flame.

The Carbo-coal process, probably the most widely known American low-temperature development, distills the coal in a stationary retort with internal agitation. Although the plant is not operating completely on a commercial scale at the present time (Sept., 1925), development work is being carried on and a new type of metal-lined coking chamber has been designed. Porter[28] reports that this retort, with a length of setting of 16 ft., has carbonized as high as 50 tons of Pittsburgh coal per day. With the retort shell at 1100 to 1200°F., the semi-coke contained 12 to 14 per cent volatile matter, about 3000 cu. ft. of 875 B.t.u. gas were made, and 23 to 30 gal. of tar oils were produced. The semi-coke,[47] being granular, is briquetted with pitch, and again carbonized, this time at a higher temperature with further recovery of by-products.

An installation of a 400-ton per day plant using the Piron-Caracristi process is being attempted[28, 48] at the Ford Motor Company plant at Ford, Ont. In this process the coal is distilled in a thin layer on a traveling metallic belt, which floats in, and is heated by a molten-lead bath. This process is designed to yield a pebbly coke suitable for use in powdered form as a furnace fuel, or to be used for mixing with high-volatile coals for coking in by-product ovens.

A new development[28, 49] in low-temperature processes is being tried out on a commercial scale by the Economic Carbonization Company at Petersburg, Va. This process (designed along the lines of the principles given by Warner, see p. 654) withdraws the volatile constituents from the center of the coal charge in order to keep thermal decomposition at a minimum. Metallic vertical retorts are used and the volatile constituents are drawn off through the hollow central shaft of a worm conveyor which propels the coal upward. The retorts are heated to relatively high temperatures at the top and are designed to yield a low-volatile coke.

The MacLaurin low temperature process,[46] the Laing and Nielsen sensible heat process in England, and the Stinnes double rotary retort in Germany, have all attracted wide attention in Europe. The MacLaurin carbonizing plant consists essentially of a large, by-product recovery gas producer, in which part of the semicoke is transformed into producer gas, the sensible heat of which carbonizes the descending coal charge. The producers are tall and of square cross-section, the experimental producer at Grangemouth, Scotland, being about 45 ft high and 8 ft. wide at the widest part. The air blast is introduced about 12 ft. above the discharging doors, through a large number of ports in opposite side walls and through similar ports in a central dividing wall at the same level. Steam is injected at the discharging doors, cooling the descending coke, and forming blue water gas in the upper part of the producer. The mixed producer gas and water gas passes through an annular condenser (for the separation of oils and tars), which surrounds the upper part of the charge, and is then led through coolers and scrubbers. This process yields a low B.t.u. gas, and the coke obtained is small and somewhat friable. Trouble is experienced with strongly coking coals, but the producers have the advantage of great flexibility in that they can be run either for the complete gasification of coal, or for the production of smokeless fuel, depending on the amount of air admitted. A battery of five of these producers has been installed at the gas plant of the Glasgow Gas Corporation.

The Laing and Nielsen sensible heat process utilizes a rotary kiln which feeds directly to a gas producer. Coal enters the rotary kiln, is carbonized by the sensible heat of the ascending producer gas, and the resulting coke is fed directly into the gas producer. The gases leaving the kiln are cooled in order to condense the valuable light oils. In its latest form, the kiln is arranged so that producer gas, hot products of combustion, or steam may be used for carbonizing the coal.

The Stinnes inclined, double, rotary retort installed by the Mathias Stinnes Colliery at Karnap, near Essen, Germany, utilizes the middle space of a large cylinder for a concentric drier and preheater for the incoming coal charge, thereby increasing the throughput. The retort is heated externally with producer gas, provision being made for diluting the hot combustion gas with flue gas in order to provide uniform and closely regulated temperature. The inner retort is never heated above the temperature at which the steel has maximum tensile strength (482°F. or 250°C.), and therefore is used to support the entire weight of both retorts. The outer retort may be heated to as high as 1112 to 1292°F. (600 to 700°C.). The slow heating of the coal in the inner retort changes the coking characteristics, and absence of sticking is reported even with moderately coking coals. The path of the evolved gases is countercurrent to the flow of coal and, therefore, little or no thermal decomposition of the vapors takes place, the yields of tar and oils being exceptionally high. Steam is injected into the outer shell in order to sweep out the high B.t.u. gases and to prevent the coal from sticking. In August 1925 this retort was producing over 70 tons a day of domestic and gas producer coke from bituminous dust.

Complete Gasification Processes.*—Complete gasification processes have for their object the utilization of the sensible heat in producer gas or water gas to effect a partial carbonization of coal in the upper part of a generator, thus making possible the recovery of ammonia and part of the liquid by-products. The partially coked coal is then completely gasified in the lower part of the same generator. This process yields a gas of from 180 to 350 B.t.u. per cu. ft., and since the entire process is carried out in a single shell a certain heat economy is possible over separate machines for low-temperature carbonization and producer-gas or water-gas generation. One such process has already been described, namely, the Wollaston producer, discussed in Chap. XVIII. The reactions taking place in the fuel bed of the generator have been treated for gas producers in Chap. XVIII and for water gas in Chap. XIX.

In addition to the complete gasification process already described, the Doherty and the Tully should also be mentioned.

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* For a bibliography of complete gasification processes see reference 50.

In the Doherty process, shown[50] in Fig 268, the distillation zone is above a water-gas generator. Coal and coke are fed in the top, coming in contact with gases at successively higher temperatures during passage through the furnace. When the air blast is on, no gas flows through the distillation zone, but all goes through the regenerators Water is admitted to the regenerators and the blue water gas formed by the action of steam on the coke passes through the distillation zone, volatilizing the lighter oils and ammonia which may then be recovered. It must be noted that in this process the volatile products never come in contact with material at a higher temperature than that at which they distill, so that no cracking takes place. The unburned coke from the base of the shaft is returned as fuel, after most of the ash has been separated mechanically.

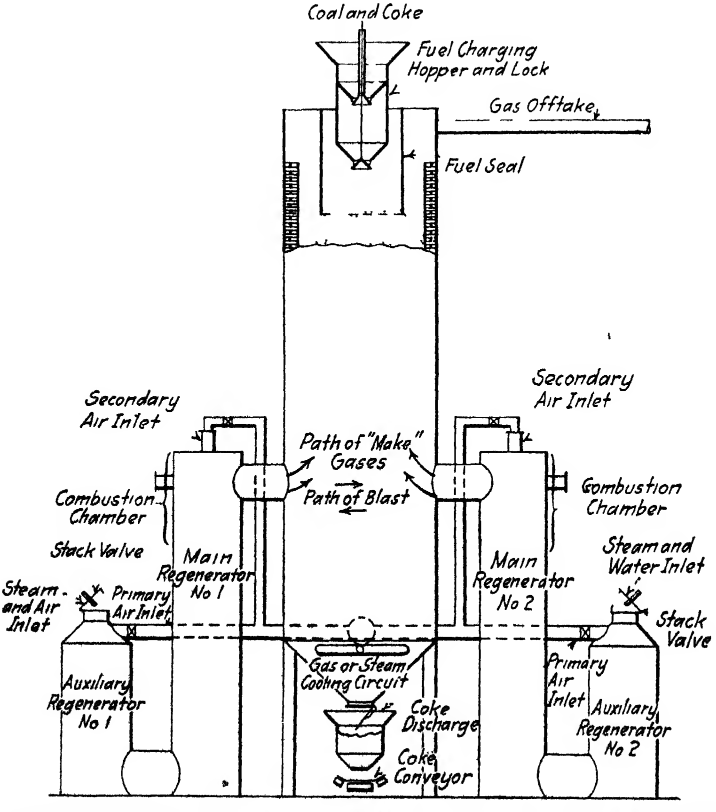

Fig 268—Doherty process of complete gasification (*Courtesy of the American Gas Association*)

The Tully process,[51] shown in Fig. 269, has been quite widely adopted in a number of small English gas works on account of its flexibility and low labor requirements. The apparatus consists essentially of a tapered retort for low-temperature carbonization set on a water-gas machine. Coal, or a mixture of coal and coke, is fed in through a side opening near the top, is distilled and partly coked in passing through the retort, and then is completely gasified in the water-gas generator at the bottom. The method of operation is somewhat as follows: An air blast is blown up through the coke fuel bed in the lower part of the machine, and the resultant producer gas passes through a series of nostrils at the base of the carbonizing chamber. A blast of secondary air is admitted and the gas burns up through an annular space *D* filled with checkerbrick. The brick is heated rapidly and the heat for partial carbonization is transmitted to the coal charge through the walls of the retort. At the conclusion of the air blow, the

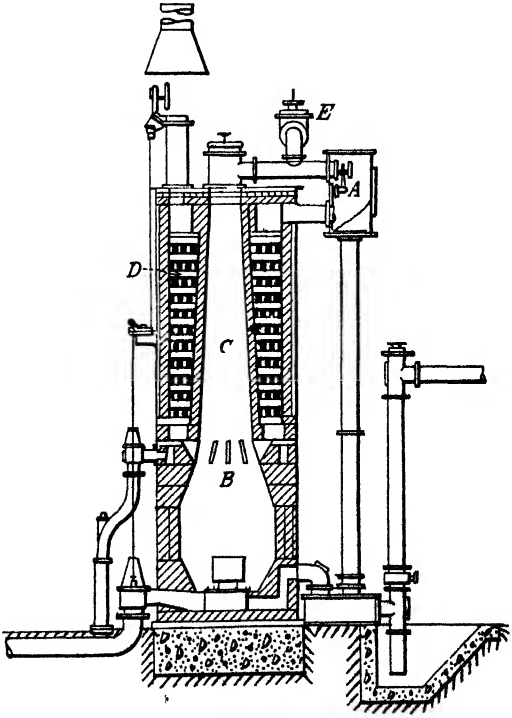

FIG. 269.—Tully complete gasification process.

stack damper is closed, steam is forced through the fuel bed, and the hot water gas passes up through the retort, sweeping out the products of carbonization. Following this make period, the bed is blown with air again, and then steam is admitted at the top of the checkers, is superheated by them, and passes down through the fuel bed, the water gas now being withdrawn from the bottom of the generator. According to Meade,[35] the highest output obtained from the Tully plant is a make of 63,000 cu. ft. of 360 to 370 B t.u gas and 8 to 9 gal of tar per ton of coal.

Complete gasification processes have not been adopted on a commercial scale in America, since the gas made is so far below the present heating-value standards. In England, on the other hand, gas is sold by the therm (a therm equals 100,000 B.t u.) and the low heating value does not offer such a disadvantage.

Methods of Heating Gas Retorts.—In the metallurgical coke industry, as already pointed out, the heat for carbonization is obtained by the combustion of part of the charge in beehive ovens, or, in the more efficient and modern by-product ovens, by the combustion of part of the coke-oven gas. In the application of by-product ovens to gas manufacture, the ovens are heated wholly or in part either with producer gas (which requires preheating of both gas and air) or with blue water gas. Early D-type gas retorts were direct fired, coal or coke being burned on grates immediately below the retorts. Direct firing is now practically obsolete, however, and, as illustrated by the preceding description of individual retort types, modern gas retorts are almost invariably fired with producer gas. The producers may be: (1) built-in producers, forming an integral part of the setting beneath the retorts; (2) semi-built-in producers, the producers still forming part of the setting, but built out in front of the retorts; and (3) external producers, in which producer gas may be generated in a central plant and distributed to a number of benches. In all these types it is extremely important that the producer gas be relatively free from dust and dirt in order to prevent clogging of the flues and combustion spaces.

Built-in producers, although retained by some of the older and smaller plants, are being discarded on account of difficulties in maintaining efficient regulation and inaccessibility for repairs. Semi-built-in producers can be charged with hot coke from the retorts, are easier to regulate and repair than the built-in type, and find quite a wide application in horizontal and some vertical-retort installations. External or central producers have a higher first cost than the two preceding types, and are used mainly in the larger plants. These producers may be run with recovery of by-products if coal is used as fuel. Other advantages of external producers are the lower depreciation and maintenance charges, the greater ease of proper regulation, and the better chance for cleaning the gas before it enters the retort setting. There is, however, a heat loss from the distributing main.

THE PRODUCTS OF CARBONIZATION

The properties of coke and gas, the major products of carbonization, and the methods of by-product recovery are highly important. The remainder of this chapter takes up these subjects, and includes a discussion of the essential properties of coke, a summary of the effect of various factors on the type of coal gas obtained, and an outline of the methods employed in the recovery of ammonia, benzol, and tar.

The Properties of Coke.—Coke produced from the distillation of coal for use as a blast-furnace fuel must have certain fairly definite characteristics. Of these the chemical properties, except chemical reactivity, which is probably due to physical structure, are of minor importance, since little can be accomplished in the carbonization process to remedy them. The limits within which the important chemical constituents of the coke should fall are well recognized and are as follows:

	Per Cent
Moisture	Not over 5
Ash	Not over 12
Sulfur	Not over 1
Phosphorus	Not over 0.02

Of these, the sulfur content is the most important, since a large percentage lengthens the time of working in the steel furnace. A low-sulfur coke is extremely important from the point of view of blast-furnace operation. The removal of sulfur is not a function of the operation of the coke oven, but rather of the initial qualities of the coal used. This is also true to a large extent of the other constituents in the coal, and since a relatively large quantity of coals having the required analyses are available, the production of coke having the requisite chemical properties is not difficult.

The physical properties of coke are of great importance, and a number of tests have been designed to determine them The physical tests* usually carried out are as follows:

1. The determination of the apparent specific gravity. This is a displacement test, the apparent specific gravity being

$$\frac{\text{Weight of Dry Coke}}{\text{Weight of Water Displaced by Coke} + (\text{Weight of Wet Coke} - \text{Weight of Dry Coke})}$$

2. The determination of the true specific gravity. This consists in determining the weight of a Hogarth flask containing a sample of coke plus a known weight of water. The true specific gravity is:

$$\text{True specific gravity} = \frac{W}{W - (W^1 - P)},$$

where W = weight in grams of dry coke, W^1 = weight in grams of bottle, dry coke, and water to fill bottle, and P = weight in grams of bottle plus water required to fill bottle.

3. The determination of porosity or per cent of volume occupied by cell space.

$$\text{Porosity} = 100 - \frac{\text{Apparent Specific Gravity}}{\text{True Specific Gravity}} \times 100.$$

4. The resistance of the coke to breakage. This is determined by the "shatter test," in which a 50-lb coke sample (of such size that it will not in any position pass through a 2-in. square mesh screen) is dropped four times onto an iron plate from a 6-ft. height, and the amount of coke which stays on a 2-in. screen noted.

5. The resistance of the coke to abrasion as determined by[52] the "tumbler test." A weighed coke sample is tumbled in a barrel for a definite length of time and a subsequent screen analysis gives a figure supposed to represent the friability of the coke. This test has no significance when applied to widely different cokes.

The variation in results obtained in running these tests on American blast-furnace cokes is given in the following table.

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* See SPERR,[14] CHRISTOPHER and BYROM,[31] and KINNEY and PERROT[64] for complete details of testing methods. See also Am. Soc. Test. Materials, Standards, pp. 1025 to 1032 (1924).

TABLE CLVI

RESULTS OF PHYSICAL TESTS ON COKE FOR BLAST FURNACES

TEST	RANGE
Apparent specific gravity (water = 1.0) . .	0.80–1.1
True specific gravity (water = 1.0)	1.8–2.0
Porosity, per cent of coke volume occupied by cell space	41–55
Shatter test, per cent through 2-in. screen	13–30
Shatter test, per cent retained on 2-in. screen	87–70

In addition to fulfilling the above requirements, the coke must be strong in order to stand the burden in the blast furnace, and should be carefully sized. American practice in this latter respect is to discard coke of less than ¾-in. size, and in some plants the screening may run as high as 1 in.

The structure of the coke may be controlled in the by-product oven by varying (among other factors) the coal mixture, temperature, coking time, oven width, and size of coal, and the regulation of the quality of resultant coke is thus to some extent in the hands of the oven operator. Coke made in by-product ovens is one-half the width of the oven in length and varies in thickness from 1 to 3 in., the average size being about 2 in. Beehive coke is considerably larger, since the coking zone travels the whole depth of the fuel bed. Another characteristic difference is in the appearance of the coke surface, the beehive coke having a more lustrous gray sheen than the by-product variety.

Combustibility of Coke.—The combustibility of coke has been given separate mention, not only on account of the importance of the subject, but also on account of the large amount of discussion which has appeared in the literature[53] in recent years. Combustibility of coke usually means the ease with which coke reacts either with air or with carbon dioxide, and inasmuch as combustibility deals with the rate of reaction, it would be more appropriate to speak of the reactivity of coke. The three main fields in which coke is used are:

1. The use of metallurgical coke in the blast furnace.
2. The use of coke for domestic heating.
3. The use of coke in the production of water gas.

In all these fields the coke is consumed by the reactions

$$\text{Oxidation} \qquad C + O_2 = CO_2 \qquad (1)$$

$$\text{Reduction} \qquad C + CO_2 = 2CO \qquad (2)$$

and in the water-gas process by the additional reactions

$$C + H_2O = CO + H_2 \quad (3)$$
$$C + 2H_2O = CO_2 + 2H_2 \quad (4)$$

The structure of the coke, its porosity, and the character of the cell walls (that is, whether or not they are covered by graphitic carbon) all are important factors in controlling its combustibility or reactivity.

A number of tests have been proposed for measuring the combustibility of coke for use in blast furnaces. Some of the more important of these tests and the limitations restricting their use are as follows:

1. Oxygen is passed over a weighed sample of crushed coke[54] for a given length of time at a definite temperature, the combustibility or the reactivity of the coke being measured by the loss in weight of the sample. This test measures the rate of oxidation of the coke, and, as was seen on page 338, oxidation of carbon, or reaction (1), depends on the rate of diffusion of the oxygen to the coke. Therefore, the combustibility of coke as measured by this test depends upon the velocity of the oxygen and upon the area of the surface of the coke. As would be expected from Chap. XII, all cokes show practically the same combustibility by this test. Another trouble with this test is that a small sample of the coke is ground up for a laboratory test and therefore the surface presented for the action of oxygen differs very greatly in character from the original surface of the large pieces.

2. In the Salamander test a sample of the coke is burned in a cylindrical iron shell by a natural draft for a given length of time and the combustibility is measured by the loss in weight. It has already been seen that with a fuel bed of this character the rate of combustion depends upon the rate at which air flows through the bed. The rate of flow of air depends upon the resistance of the fuel bed and this is governed largely by the size and shape of the coke and the way in which it is packed. Therefore, this test does not truly measure the activity of the coke surface.

3. Koppers[55] proposed the same type of test as the Salamander test, but substituted the character of the flame above the fuel bed for the loss in weight as the measure of combustibility. In this test, the longer the flame the more combustible the coke. This

test, in distinction to those previously mentioned, measures the rate of reduction of carbon dioxide by the coke in accordance with reaction (2). This reaction is a time reaction (see p. 338) and therefore the faster the rate of air flow the less the time of reduction, and, consequently, a bed of coke that has a high resistance to the flow of air will show low combustibility compared to the same coke in a loose, free fuel bed. Another objection to this test is that the personal factor in judging the height of the flame cannot be avoided. Koppers[55] also proposes to blow a measured volume of air through the fuel bed, and, while this plan is much better, it still depends upon the eye as the measure of combustibility.

Bunte and Fitz[59] define the reactivity of coke as its ignition temperature. Heating the coke to a high temperature increased its ignition temperature. For instance, with one sample, heating increased the ignition temperature from 622 to 1153°F. However, Bunte and Fitz believe that the temperature at which the coke is formed has even more influence on the reactivity (ignition temperature) than the temperature to which the coke is afterwards heated.

Thau[60] points out that the cells in coke that are visible to the eye have no effect on reactivity and that high reactivity is assured only if the tiny cells are open. If these cells are closed, as, for example, by graphite or ash, the coke will be less combustible. As a result, Thau considers that the density of coke is not important and that a very dense coke may have a high combustibility, provided the coal from which it is made is free from impurities and that the cell structure of the resulting coke is open. Low-temperature coke is more combustible, in his opinion, not on account of the volatile matter it contains, but because as the coke is burned the volatile matter is driven off slowly, leaving an abundance of small cells free from graphitic carbon. In this connection it may be mentioned that a number of investigators[59,61] consider that the permeability of the coke to hydrogen and to carbon dioxide indicates an open type of cell structure and therefore a combustible coke. An important review of these and other views is given by Sherman and Kinney.[62]

Sutcliffe and Evans[63] also believe that a fuel with interconnected pores is more combustible than a fuel with large, discontinuous pores.

To summarize the various ideas, combustibility (depending on point of view) may mean oxidation of coke with oxygen, reduction of carbon dioxide by coke, or ignition temperature of coke. Cokes have practically the same rate of oxidation, since this is controlled more by the velocity of the air than by the character of the coke. The rate of reduction of carbon dioxide by coke, which is a reaction with a rate proportional to the active surface, depends on the size of the coke, its pores (that is, whether or not the pores are open and interconnected), and the temperature at which the coke is formed, since volatile matter left in the coke creates open pores on burning Ignition temperature depends on the size of the coke, on its previous heat treatment, *i.e.*, the temperature at which it was formed, on its subsequent heating, and on the character of the coal substance from which the coke was produced. Whether rate of reduction with carbon dioxide or ignition temperature is the more important factor is an open question.

Coke as a Fuel for Domestic Heating.—When coke is used for domestic heating, the most important factors are:

1. A low percentage of non-clinkering ash.
2. A high apparent or bulk density.
3. Low water content.
4. An easily ignited coke.
5. A coke that burns efficiently without constant draft regulation.
6. A coke that will stay ignited for long intervals between firing.

For home consumption it is obvious that a coke with a low water content and a low percentage of non-clinkering ash is desirable. High bulk density (pounds per cu. ft.) is required, because this characteristic permits longer intervals between firings—a very important consideration for a domestic fuel. For example, the bulk density of anthracite coal is 50 to 55 lb. per cu. ft., whereas the usual domestic coke weighs only 30 to 35 lb. per cu. ft. Coke should be capable of easy ignition so that the fire will pick up quickly when fuel is thrown on a low fire. A coke that burns efficiently without constant draft regulation is one that burns almost wholly to carbon dioxide in the fuel bed, *i.e.*, the gases coming off from the top of the fuel bed should be low in carbon monoxide. Under these conditions, the amount of fuel being burned and the efficiency of combustion (for coke

21 per cent CO_2 and no CO indicates 100 per cent efficient *combustion*) may be controlled by the setting of the stack damper or the ash-pit door only. Further, if the coke has high reactivity with CO_2, the gases shortly after firing (due to the deep fuel bed) will contain a large amount of CO, thus lower the efficiency of combustion and require the damper in the firing door to be open. As the fuel burns out, the bed becomes thinner, less CO has to be burned in the combustion space over the fuel bed, and the damper in the firing door should be closed. In the average home, such a constant regulation of dampers or the firing of fuel at frequent intervals is not feasible.

It has been found that the time required to ignite coke by a blast or hot air (1300°F.) varies 300 per cent with different domestic fuels. Also, the ratio of carbon as CO to the total carbon ($CO_2 + CO$) in the gases coming off the top of a fuel bed of an ordinary hot-water boiler varies from 0.4 to 0.2, the coke having a ratio of $\frac{CO}{CO + CO_2} = 0.4$ being the more difficult to burn efficiently in the home. The difference between the cokes is due to variations in the coal used and the temperatures of carbonization.

The ability of a coke fire to stand long periods of banking depends on the heat capacity, conductivity, and density, and, to some extent, on its ability to withstand combustion with small amounts of air leaking up through the ash-pit door.

Coke for Water-gas Production.—In the water-gas process a coke is desired that will have a high reactivity with steam and a low CO loss during the air blow. On account of the similarity of the reactions of coke with steam,

$$C + H_2O = CO + H_2$$

$$C + 2H_2O = CO_2 + 2H_2$$

and the reaction of coke with CO_2 (formed in the lower portion of the fuel bed during the air blow)

$$C + CO_2 = 2CO,$$

it is probable that a coke with a high reactivity with steam will also have a higher CO loss during the air blow. This loss, however, can be cut down by a high-velocity air blow, since the amount of reduction of CO_2 by carbon depends on the time of contact. Therefore, for water-gas production it would seem as if the coke with the highest reactivity with H_2O (and, therefore, probably CO_2, although this is not desired) would be most suitable, leaving the elimination of the CO loss to the use of a high-velocity blast. As shown earlier, the coke with highest reactivity with CO_2 and, therefore, with probably the highest reactivity with steam, is a low-temperature, small-size coke, with interconnected pores. A non-graphitic coke, such as that made in a steamed, vertical retort, would seem to be very suitable. The fusion point of the ash must be high in order to lessen trouble from clinker formation (see p. 609)

Coal Gas.—The previous discussion of the mechanism of carbonization, and the comparison of different types of retorts have brought out the fact that the quality and quantity of coal gas made depends on:

1. The coal used.
2. The temperature employed.
3. The kind of retort in which carbonization is carried out.

The richer the coal used (in general) the higher the gas yield will be under a given set of conditions.* It was shown on page 669 that the coals used for gas making are chosen to a large extent on this qualification, namely, as high gas yields as are compatible with the production of a coke suitable for the particular local market. It has also been pointed out that high temperatures cause thermal decomposition and a resultant decrease in the heating value of the gas. A comparison of different types of carbonizing equipment (Table CXLVIII, p. 673) has shown the temperatures, times of contact in the free space, etc and affords an indication of the influence of the kind of retort on the gas made.

The properties of gases have been taken up in Chap. XI, and the composition of a representative coal gas and coke-oven gas given in Table LXVI, page 282.

The Recovery of Ammonia.—The recovery of ammonia from the gases evolved during carbonization is an important factor in the economic success of the process. The gross value of this product may be as high as 6 to 10 per cent of the cost of the coal used. Recovery became important as the available supply of nitrogen compounds decreased and those countries which had most difficulty in obtaining nitrogen salts were the foremost in developing processes. The perfection of methods for the

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* This statement is not strictly true in all cases, since better gas yields may be obtained from Westmoreland coal, for example (32 to 34 per cent volatile matter) than from some higher-volatile coals.

fixation of atmospheric nitrogen, however, probably will make ammonia recovery from coal carbonization much less profitable. Progress in the recovery of ammonia has been rapid during recent years and in all countries the maximum amount is obtained from by-product and gas ovens.

The ammonia present in the gas depends to a large extent on the nitrogen present in the original coal. Other factors of importance are the distillation temperature, time of distillation, and exposure of gases to heat and decomposition.

Processes for ammonia recovery (in which the ammonia is recovered as sulfate) may be divided into two main classes—direct and indirect. The latter was the original method and is still retained to some extent. In the indirect process the gases from the coke or gas oven are first cooled, the tar extracted, and the ammonia dissolved in water in suitable scrubbing towers. This liquor must then be distilled and the ammonia dissolved in sulfuric acid In the direct process, the absorption in water is not necessary, and all the gases are passed through a sulfuric acid bath to precipitate the ammonium sulfate directly. This process originated with Brunck in 1903, but failed because the tar was not entirely removed from the gas and dirty sulfate resulted More efficient tar removal has eliminated the difficulty and the direct process is used almost exclusively in the larger by-product plants today.

In the indirect process, ammonia liquor from the cooling, scrubbing, and absorbing towers is preheated and enters the top of the still at a temperature of 80 to 85°C. Free ammonia is distilled in the upper section and lime is then pumped into a lower section in order to liberate the fixed ammonia. The gases leaving the still pass into a water separator (to condense and remove water vapor) and then enter a saturator containing strong sulfuric acid, where ammonium sulfate is crystallized. The remaining gas and uncondensed steam may be filtered through iron oxide purifiers or may be led to the oven stack. The sulfate is withdrawn continuously from the saturator, and dried centrifugally. This process may also be used for the production of concentrated ammonia liquor by using water as the absorbing medium, after passing the gases through a dephlegmating tower.* At certain times liquor may be of more

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* See Walker, Lewis, and McAdams[64] for a discussion of the individual steps.

value than sulfate and many plants are equipped for recovery of both.

The Koppers direct process is the one most widely used in by-product coke-oven plants in the United States at the present time. The foul gas from the ovens passes through primary coolers and tar extractors (as in the indirect process). After tar removal, the gases (containing about 75 per cent of the ammonia originally present in the gas) are preheated to about 150°F. and pass into a saturator containing sulfuric acid. The tar and ammonia liquor removed in the coolers and extractors are separated (by allowing the mixture to stand) and the ammonia is distilled off and added to the gases just before they enter the saturators. The ammonium sulfate is handled as in the indirect process.

Benzol Recovery.—The recovery of benzol or light oil from coke-oven gas consists of absorption in mineral oil and subsequent steam distillation of the benzol (and toluol, etc) from this mixture. The mineral oil is recirculated through the scrubbing towers and the light oil further purified. The general scheme may be subdivided into the following processes.

1. Extraction of a complex mixture of aromatic hydrocarbons by treating the gas with solvents of higher boiler point, such as straw oil, cresols, etc.
2. Expulsion of the mixture from the solvent by regulated heat treatment.
3. Preliminary fractionation into groups of compounds according to the respective boiling points.
4. Chemical treatment of the various fractions.
5. Redistillation and careful fractionation of the chemically treated products.

The efficiency of such a process depends on the proper heat recovery and heat transfer from the hot products to the feed entering the stills.* The actual arrangement of the heating system varies in different types of plant.

The main products obtained are benzol, toluol, and solvent naphtha, with a certain amount of other hydrocarbons of this series. A large number of impurities present in the gas accumulate in the benzol plant, the most important of which are carbon bisulfide, thiophene, phenols, pyridine, and naphthalene.

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* See Walker, Lewis, and McAdams[64] for a discussion of the individual steps.

The recovery process consists in separating the various products by the methods outlined above.

Tar.—Tar is deposited from the gas at certain places in the system, the major portion separating on the initial cooling. All this is collected into one large tank. It may then be shipped directly or, in some plants, distilled to separate various products.

The constituents of tar may be grouped as:

1. Hydrocarbons:
   (*a*) Paraffin.
   (*b*) Aromatic.
2. Tar acids—oxygen-containing bodies of phenolic type (higher phenols, cresols, etc.).
3. Nitrogen-containing bodies—(ammonia, pyridine, carbazole).
4. Sulfur-containing bodies.
5. Pitch.
6. Free carbon.

The proportion of these constituents varies considerably with temperature, the yield of hydrocarbons being greater at low temperatures and the amount of pitch and free carbon increasing as the temperature (and thermal decomposition) is increased. The yield of tar decreases and its specific gravity increases with increasing retort temperatures.

As previously mentioned, the tar is the source of many derivatives, is used as a road and roofing material, and is becoming of increasing importance as a fuel in steel and other mills which manufacture their own coke (see Chap. XVI).

BIBLIOGRAPHY

1. U. S. Geological Survey.
2. U. S. Bureau of Statistics, Dept. of Commerce.
3. "Coal Catalog," Keystone Consolidated Publishing Co. (1925).
4. Campredon: *Compt. rend.* (1895).
   Pollard: Memoirs of the Geological Survey (Eng.), p. 11 (1908).
   Dunn: *J. Soc. Chem. Ind.*, **32**, 397 (1913).
   Sinnatt and Grounds: *Ibid.*, **39**, 83T (1920).
   Weighhill: *Ibid.*, **41**, 12T (1922).
   Gray: *Fuel*, **2**, 42 (1923).
   Badarau and Tideswell: *Ibid*, **2**, 62 (1923).
5. Parr and Olin: Univ. of Illinois Eng. Expt. Sta., *Bull.* 60 (1910); and Parr and Olin: *Ibid.*, *Bull.* 79 (1915).
6. Schoenfeld and Loring, Mass. Inst. Tech. *S. B. Thesis* (1924); Keller, Mass. Inst. Tech., *S. B. Thesis* (1924).

7. A. W. Warner: *Proc.* Am. Gas Assoc., **5,** 928 (1923).
8. Von Hilgenstock: *J. Gasbel.*, **45,** 617 (1902).
9. Von Schiewind: *Trans.* Am. Inst. Mining Eng., **33,** 776.
10. Simmersbach: "Grundlagen der Koks-Chemie," Julius Springer, Berlin (1914).
11. Cooper: *J.* Iron Steel Inst., **11** (1914).
12. Ruehl: *Blast Furnace and Steel Plant* (February, 1923).
13. Ramsburg and Sperr: *J.* Franklin Inst., **183,** 391 (1917).
14. Sperr: "Technology of Coke," in Bacon and Hamor: "American Fuels," McGraw-Hill Book Co., Inc. (1922).
15. O. B. Evans: *Proc.* Am. Gas Inst. (1913).
16. Porter: "Coal Carbonization," Am. Chem. Soc. *Monograph.*, Chemical Catalog Co. (1924).
17. E. V. Evans: Cantor Lectures. *J.* Roy. Soc. of Arts, **72,** 767, 785, 805 (1924).
18. D. W. Wilson: *Proc.* Am. Gas Assoc., **5,** 952 (1923).
19. Ryan: "Rate of Travel of Fusion Zone in Coke Oven," *Proc.* Am. Gas Assoc. (preprint, 1925), p. 29.
20. Foxwell: *Fuel,* **3,** 122, 174, 206, 227, 276, 315, 371 (1924).
21. Biddulph-Smith: *Fuel,* **1,** 68 (1922).
22. W. A. Bone: "Coal and its Scientific Uses," Longmans, Green & Co. (1922).
23. Davidson: *Trans.* Inst. Gas Eng., p. 269 (1910).
24. Jones: *J.* Franklin Inst., **177,** 531 (1914).
25. Paper by Joseph Becker: "Modern By-product Coking," printed by the Koppers Co.
26. Davis: *Proc.* Am. Gas Assoc., Tech. Session, **2,** 105 (1920) and *Ibid.*, **4,** 65 (1922).
27. Southern: *Fuels and Furnaces,* **3,** 475 (1925).
28. Porter: *J.* Franklin Inst., **199,** 381 (1925).
29. Pratt: *Chem. Met. Eng.*, p. 27 (1917).
30. Porter: *Trans.* Am. Inst. Min. Met. Eng., **63,** 902 (1920).
31. Christopher and Byrom: "Modern Coking Practice," D. Van Nostrand Co. (1921).
32. G. S. Cooper: "By-product Coking," Benn Bros., Ltd. (1922).
33. Meisner: *Trans.* Am. Iron Steel Inst., p. 118 (1913).
34. Davidson: "Gas Manufacture," Longmans, Green & Co. (1923).
35. Meade: "Modern Gas Works Practice," Benn Bros., Ltd. (1921).
36. Ward: *J.* Soc. Chem. Ind., **39,** 198T, (1920).
37. Mott: Series of articles in *Fuel,* **4** (1925).
38. Wilson, Forrest, and Herty: *Ind. Eng. Chem.*, **15,** 251 (1923).
39. Munster: *Blast Furnace and Steel Plant,* **11,** 389 (1923).
40. Davis and Place: *Ind. Eng. Chem.*, **16,** 589 (1924).
41. Earle: *Proc.* Am. Gas Assoc., Tech. Session, **2,** 201 (1920).
42. R. S. Carter: "Vertical-retort Carbonizing Plant of the Malden & Melrose Gas Light Co." (May 9, 1925).
43. *Gas World,* **72,** 492 (1920).
44. Sperr: *J.* Western Soc. Eng., **29,** 327 (1924); *cf.* also Sperr and Powell: *Mech Eng.* (June–July, 1924).

45. *Proc.* Am. Gas. Assoc., Tech. Session, **2,** 94, 105 (1920); *Ibid.*, **3** (1921); *Ibid.*, **4,** 76 (1922).
46. Series of articles by Brownlie: *Combustion* (1924 and 1925); cf. also Fieldner, Publication No. 25–72, National Electric Light Association, p. 34 (1925).
47. Curtis *et al.*: *Chem. Met. Eng.*, **28,** 11, 60, 118, 171 (1923).
48. Caracristi: *Power*, **57,** 831 (1923); *Chem. Age*, **31,** 361 (1923).
49. *Proc.* Am. Gas Assoc., **5,** 982 (1923).
50. R. G. Porter: *Proc.* Am. Gas Assoc., Tech. Session, **4,** 374 (1922).
51. *Gas J.* (Dec. 22, 1920; May 11 and Oct. 26, 1921; Jan. 11, Feb. 8, and June 14, 1922; Jan. 23, 1924).
52. Kinney and Perrot: *Ind. Eng. Chem.*, **14,** 926 (1922).
53. Haslam and Thiele: *Ind. Eng. Chem.*, **16,** 749 (1924).
54. Sherman and Blizard: *Trans.* Am. Inst. Min. Met. Eng., **1217S** (1923).
55. Koppers: *Fuel*, **2,** 160 (1923); Fischer, *et al.*, *Brennstoff-Chem.*, **4,** 33 (1922).
56. Sperr and Jacobson: *Blast Furnace and Steel Plant*, **11,** 314 (1923).
57. Broche: *Brennstoff-Chem.*, **4,** 343 (1923).
58. Bahr: *Stahl u. Eisen*, **1,** 39 (1924).
59. Bunte and Fitz: *Gas J.*, pp. 167 and 513 (1924).
60. Thau: *Chem. Met. Eng.*, **30,** 34 (1924).
61. Bardenheuer and Thanheiser: *Mitt. Inst. Eisenforschung.*
62. Sherman and Kinney: *Iron Age*, **111,** 1839 (1923).
63. Sutcliffe and Evans: *J.* Soc. Chem. Ind., **41,** 196 (1922).
64. Walker, Lewis, and McAdams: "Principles of Chemical Engineering," McGraw-Hill Book Co., Inc. (1923).

APPENDIX I

FLOW OF LIQUIDS AND GASES*

The flow of fluids, both liquids and gases, is of great importance in the field of combustion, since liquids, such as oil and water, frequently must be transported and the problem of moving gases is encountered in all furnaces. In considering the flow of liquids, such as water and oil, the pressure drop through the system, and the head necessary to maintain a definite rate of flow, are the main factors of interest. The flow of gases, on the other hand, involves only slight pressure drops and, although the resistance to flow is important, the distribution of the gases is an added factor of equal or greater importance. Liquids flow at high velocities through pipes of small cross-section, so that perfect mixing usually takes place. Furnace gases, however, often flow at low velocities with little turbulence through large flues and over large hearths, a condition which prevents thorough mixing. Consequently, a discussion of fluid flow may be divided into two sections: first, friction losses or resistance to flow; and, second, distribution of gases.

FRICTION DROP

The solution of all problems in flow of fluids is based on one of two equations, both known under the name of Bernoullis' theorem. Both are based upon consideration of a given section of the conduit through which the flow is taking place. The first is a mechanical energy balance, based on the fact that the mechanical energy can be increased within the section itself only by one of two mechanisms, either by doing external work on the fluid (*e.g.* by means of a pump), or by the mechanical energy of expansion of the fluid upon itself. The mechanical energy can be dissipated only through friction. The second equation is based on the law of conservation of energy, and represents an overall energy balance applied to the section. The two equations are:

$$\begin{aligned} x_1 + p_1v_1 + \frac{u_1^2}{2g} + R + \int pdv &= \\ x_2 + p_2v_2 + \frac{u_2^2}{2g} + F & \end{aligned} \tag{1}$$

and

$$\begin{aligned} E_1 + x_1 + p_1v_1 + \frac{u_2^2}{2g} + R + H &= \\ E_2 + x_2 + p_2v_2 + \frac{u_2^2}{2g} & \end{aligned} \tag{1a}$$

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* For a more complete discussion of this subject see WALKER, LEWIS and MCADAMS, "Principles of Chemical Engineering."

where R equals the mechanical work done upon the fluid in the section by external means (zero unless a pump, etc. is used), $\int pdv$ is the work done by the expansion of the fluid upon itself, F is friction loss, *i.e.*, mechanical energy converted into heat, and H equals the heat added from external sources, expressed in mechanical units.

E represents the intrinsic energy of the fluid, x the height above a datum plane, p the pressure on the fluid, v its volume, u its velocity, and w any external work done on the fluid. The subscripts 1 and 2 refer to the value of these terms at two cross-sections of the path of flow. The equations are based on a unit weight of fluid flowing, or, in English units, 1 lb., so that the terms are energy terms, *i e.*, foot-pounds.

Which of these equations to use depends on the specific problem, but in almost all cases the first equation is the one required. The second is, however, the more useful Furthermore, pumping devices usually are treated separately, so that in the following discussion they are not considered and thus $R = 0$. For convenience $\int pdv$ is represented by w. The terms x and $\frac{u^2}{2g}$ represent, respectively, the potential energy of the fluid, due to its height above the datum plane, and the kinetic energy due to its velocity. The product pv is the mechanical work done in forcing the fluid into, or out of, the section.

Fluid Heads.—Since 1 lb. of the material is assumed in Eq. (1), the units of the energy terms must be feet in order that the resulting units will be foot-pounds of work. The feet of fluid may be considered as a "head" and Eq. (1), when there is no friction loss, then may be rewritten

$$h_p' + h_s' + h_v' + w = h_p'' + h_s'' + h_v'', \tag{2}$$

where h_p, h_s, and h_v are the potential, static or pressure, and velocity heads respectively, in feet of the fluid. The value of the potential head h_p is the distance above the datum line, which is usually taken as the lowest point in the system to avoid negative values. *Static or pressure head* is the fluid head exerted by the material on a plane parallel to its direction of flow. *Velocity head* is the fluid head exerted on a plane perpendicular to the direction of flow, minus the pressure head. Velocity head may be measured in this way, since the kinetic energy is converted entirely to pressure head at the moment of impact. Figure 270 illustrates the method of determining the values of each head. Velocity head may be measured directly by the differential manometer at the right, one arm determining pressure head and the other determining the sum of pressure and velocity heads, so that the reading gives the velocity head by difference. The head may be measured in pounds per sq. in., inches of water, or inches of mercury, and the value in feet of the fluid flowing calculated from the density. The density of water is 62.3 lb. per cu. ft. and a column of water 1 ft. high exerts a pressure of 62.3 lb. per sq. ft., or 0.432 lb. per sq. in. The head in feet of air at 60°F. may be determined from a measured head in inches of water by multiplying the ratio of the density of water to that of air, and by dividing by 12, namely, $\frac{62.3}{0.075 \times 12}$, or 69.2.

Equation (2) indicates that any kind of fluid head may be converted into any other without outside work being done on the system. The sum of the three heads at one section must be equal to their sum at any other section provided no external work is done or no energy is lost by friction. If at any point the velocity head is increased, then either the potential or the static head must be decreased by the same amount. This fact is the basis of the design of many meters for measuring the rate of fluid flow, which operate under such conditions that there is a simple relationship between the changes in static and velocity heads.

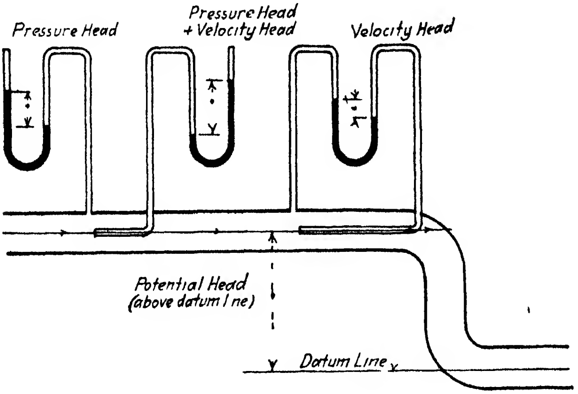
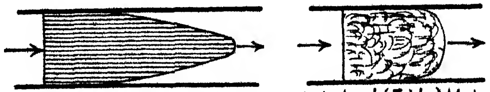


FIG. 270.—Measurement of fluid heads.

Cause of Flow.—The flow of any fluid starting from rest is caused by differences in head, either pressure head or potential head. Thus water will flow from a higher level to a lower level by gravity, or by difference in potential head between the two points. It also will flow through a horizontal conduit with no difference in potential head, provided there is a difference in pressure head. In addition, flow will take place if there is a difference in the sum of the pressure and potential heads at the two points. Velocity head serves to continue the flow after the pressure and potential heads have been eliminated.



Straight Line (Viscous) Flow   Turbulent (Eddy) Motion

Abs = actual net velocity at a given point

FIG. 271.—Velocity distribution in circular pipe.

Types of Motion.—A fluid, flowing through a conduit under constant conditions (*i.e.*, constant pressure drop), may adopt either of two types of motion, straight-line or turbulent. In straight-line motion each particle flows in a path parallel to the conduit and no transverse or convection currents are set up[1] (see Fig. 271). As the velocity increases there is a sudden transition at a definite point from this type of flow to turbulent motion, which is characterized by the presence of a multitude of eddy currents (see right side of Fig. 271).

Straight-line Motion.—The theoretical law of Poiseuille, which, for straight line motion, has been experimentally confirmed for circular pipes up to 12 in. in diameter, is

$$\Delta p = \frac{32\mu L u}{g d^2}, \tag{3}$$

where Δp = pressure drop (pounds per sq. ft.).
L = length of straight pipe (feet) plus equivalent length due to bends, etc.
u = average velocity (feet per sec.).
g = a constant (32.2).
d = inside diameter (feet).
μ = absolute viscosity (lb. per sec. per ft.), or
μ = $0.000672z$, where z is the viscosity relative to water at 68°F.

Turbulent Motion.—The equation governing the pressure drop when the motion of the fluid is turbulent is given by Fanning's equation:

$$\Delta p = \frac{f\rho L u^2}{2gm} \tag{4}$$

where Δp = pressure drop (pounds per sq. ft)
f = friction factor.
p = density (pounds per cu. ft.).
L = length of straight pipe in feet plus equivalent length due to bends, etc. (Table CLVII).
u = velocity (feet per sec.).
m = hydraulic radius in feet, which equals the cross-sectional area divided by the perimeter of all surfaces exposed or wetted.

For circular pipes or square flues where $m = \frac{d}{4}$ this equation becomes

$$\Delta p = \frac{2f\rho L u^2}{gd}. \tag{5}$$

TABLE CLVII

Equivalent Lengths,* Expressed as Diameters of Additional Straight Pipe (Turbulent Motion)

	Number of Diameters
90-deg. elbows:	
1–2½ in.	30
3–6 in.	40
7–10 in.	50
Globe valves:	
1–2½ in.	45
3–6 in.	60
7–10 in.	75
Tees:	
1–4 in.	60

* Walker, Lewis, and McAdams, Principles of Chemical Engineering."

Critical Velocity.—At the critical velocity u_c, or the velocity where the change from straight-line to turbulent motion occurs, the pressure drops in both types of motion must be equal, and from Eq. (3) and (5) we have,

$$\Delta p = \frac{32\mu L u_c}{gd^2} = \frac{2f_c\rho L u_c^2}{gd}, \tag{6}$$

whence

$$u_c = \frac{16\mu}{f_c\rho d} \tag{7}$$

As will be shown later, the coefficient of friction at the critical velocity f_c is 0.0176 for all fluids, so that the equation on critical velocity becomes

$$u_c = \frac{942\mu}{\rho d}, \tag{8}$$

or substituting viscosity relative to water, z ($\mu = 0.000672z$), specific gravity relative to water, s, and diameter in inches, D, for μ, ρ, and d respectively, Eq. (8) reduces to

$$u_c = \frac{0.122z}{Ds}. \tag{9}$$

The following table shows the values of u_c for various fluids flowing inside a 2-in. pipe calculated from Eq. (8). The critical velocity for air at 1150°F. flowing through a 6-ft. main determined from this equation is 0.157 ft. per sec. From a consideration of these values it is evident that the critical velocities for air and water are so low as almost never to be found in practice. It is only when dealing with the viscous oils that straight-line flow is encountered.

TABLE CLVIII

CALCULATED CRITICAL VELOCITIES IN 2-INCH PIPE

Fluid	Temperature, degrees Fahrenheit	Absolute pressure, atmospheres	Density pounds per cu. ft.	Absolute viscosity English units	Critical velocity, feet per sec
Hydrogen	70	1	0.00522	0.0000062	6.5
Air	70	1	0.0750	0.0000125	0.91
Air	70	10	0.750	0.0000125	0.091
Steam	212	1	0.0373	0.0000081	1.19
Steam	357	10	0.326	0.0000097	0.16
Water	68	Any	62.3	0.000672	0.059
Refined oil[a]	68	Any	56.0	0.0067	0.65
California crude oil	68	Any	60.0	2.32	211

Friction Factor.—Equation (4), governing the case of turbulent motion, is thus the most important equation for the determination of friction losses. Before this can be applied, however, it is necessary that the values of the friction factor be known. A multitude of empirical relationships expressing f as some function of the pipe size and the velocity, density, and viscosity of the fluid have been developed. Each of these relationships is applicable to the particular fluid in question and under a limited range of conditions.

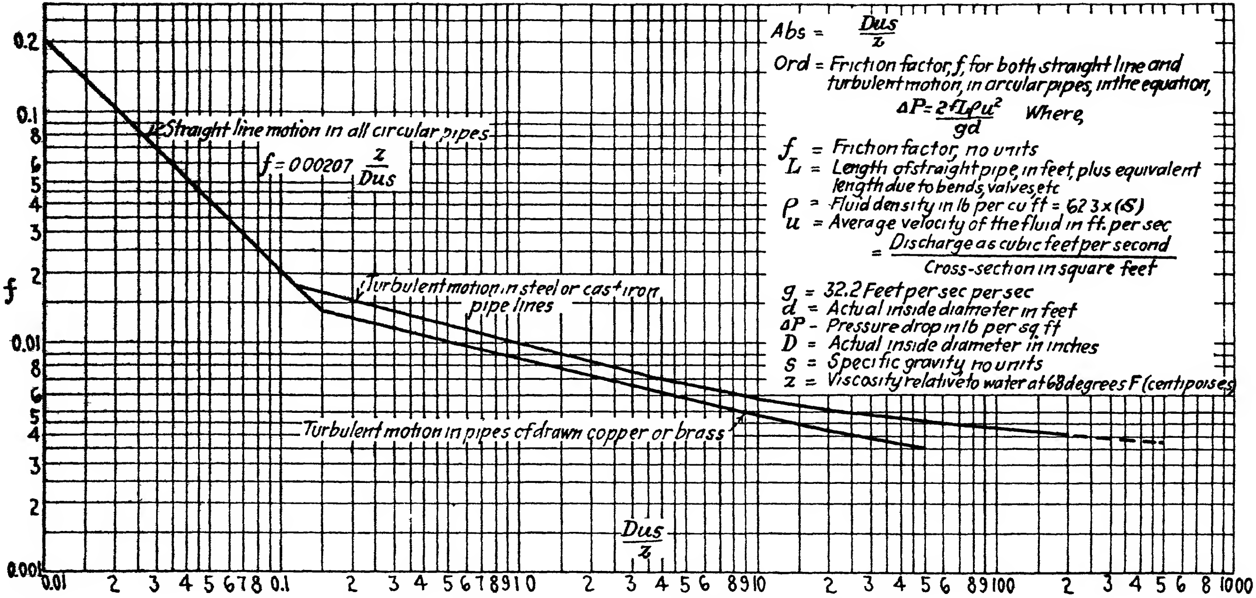


FIG. 272 —Friction factors for all fluids in circular pipes.

The friction factor, however, is a function of all four of these variables and, since it is a coefficient without units, it must vary as some function of $\frac{Dus}{z}$, since this is the only combination of D, u, s, and z which has no units. The value of this function has been determined[2] by correlation of the most reliable data available on the flow of gases, water, and oils. It may be represented best by the chart[1] (Fig. 272), in which the friction factor is plotted against the value of the modulus $\frac{Dus}{z}$. For the solution of any problem in the turbulent-flow range it is simply necessary to calculate $\frac{Dus}{z}$, look up the friction factor f from the chart, and substitute in the equation

$$\Delta p = \frac{fL\rho u^2}{2gm}. \tag{4}$$

It should be noted that Eq. (3), $\Delta p = \frac{32\mu Lu}{gd^2}$, for straight-line motion may be changed to the form

$\Delta p = \frac{2fL\rho u^2}{gd}$ by substituting f for $\frac{16}{du\rho}$, or f equals $0.00207\,\frac{z}{Dus}$, so that by drawing the line $f = 0.00207\,\frac{z}{Dus}$ to intersect with the line giving the values of f in turbulent motion a chart is obtained from which the value of f may be determined at any velocity for use with equation

$$\Delta p = \frac{2fL\rho u^2}{gd} \tag{5}$$

regardless of whether the flow is turbulent or straight line. The intersection of these lines gives the value of f at the critical velocity equal to 0.0176. The value of the modulus $\frac{Dus}{z}$ at this point is 0.122 for fluids flowing in cast-iron or steel pipes.

Flow of Liquids in Pipes.—The method of determining friction loss through a pipe line carrying liquids may be divided into three steps: First, calculation of $\frac{Dus}{z}$ from the size of pipe or flue and the characteristics of the liquid flowing; second, determination of f from Fig. 272; and, third, calculation of the pressure drop in pounds per sq. ft. by substitution in Eq. (5). The external work in foot-pounds (550 ft.-lb. per sec. = 1 hp.) necessary to overcome the friction is the product of the pressure drop times the cubic feet of fluid flowing. The total work necessary to move the fluid is that necessary to overcome friction plus that to compensate for differences in potential, static, and velocity heads at the two ends of the system. In many commercial problems in flow of liquids, differences in velocity and static head are negligible and external work is supplied chiefly to overcome friction and potential heads.

Flow of Gases in Flues.—For the solution of problems in the flow of liquids the above method is most satisfactory. In combustion practice, however, problems in *liquid* flow are not often met with, and the flow of *gases* is of greater importance. The equations above are equally applicable to the flow of gases, however, but for brick flues the values of the friction factor are higher. The chart (Fig. 273) shows the values of f calculated from data given by Trinks[3] for the friction factor in flues with walls having various degrees of roughness. In this plot and in all subsequent discussion, for convenience in handling problems on the flow of gases, the unit of pressure drop, or friction loss, is inches of water. The value of f for iron pipe (Fig. 272) must be divided by 5.2 to conform to these units, and the results are plotted as the dotted line in Fig. 273. The actual determination of the value of f under various conditions is difficult, and few accurate data are available. Menzin[4] gives as the equation for pressure drop in chimneys

$$p = 0.008\frac{L}{D}\frac{V^2}{t},$$

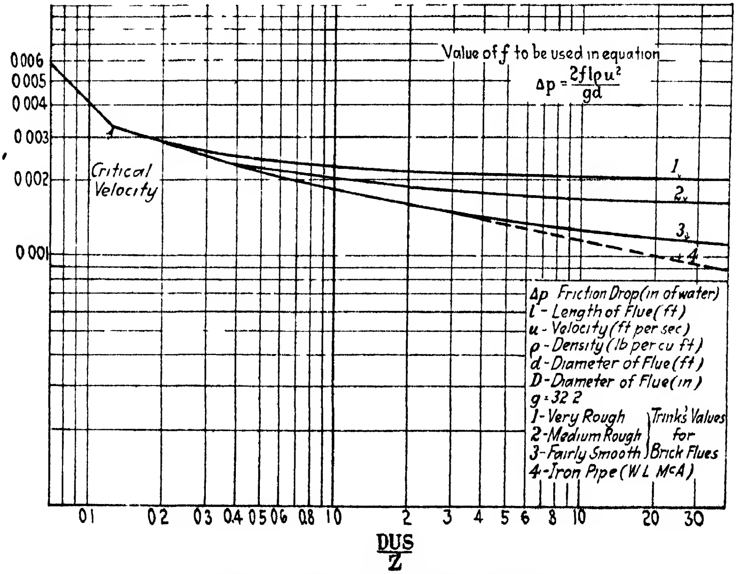


FIG. 273.—Friction factors for gas flow.

which, in the units and form developed previously, is

$$\Delta p = \frac{2(0.00325)Lu^2}{gd},$$

or a friction factor of 0.00325. This would correspond to a value of $\frac{Dus}{z}$ of 0.13 and is probably much lower than the usual prevailing conditions. The values given in Fig. 273 represent the best figures obtainable.

The usual range of values of $\frac{Dus}{z}$ in large flues is from 1 to 30, and in most cases the value is between 12 and 30. In this range (from 1 to 30) the value of the friction factor for very smooth walls falls from 0.00185 to 0.00093 and for very rough walls from 0.0023 to 0.0021. Evidently, with rough walls f is more nearly constant and varies much more with the degree of

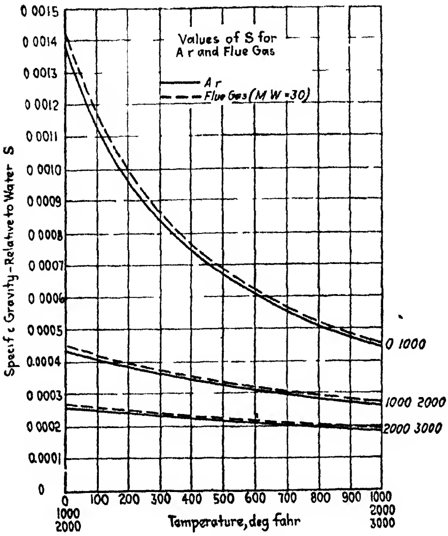


Fig. 274.—Specific gravities of air and flue gas at different temperatures.

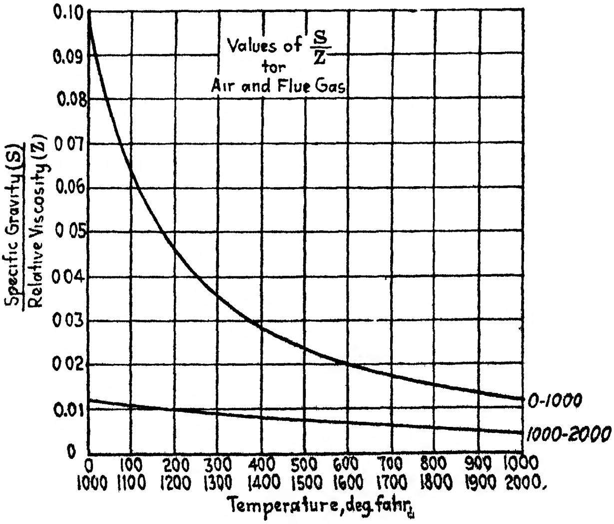


Fig. 275.—Values of $\frac{\text{Specific gravity (S)}}{\text{Relative viscosity (Z)}}$ for air and flue gas at different temperatures.

roughness than with $\frac{Dus}{z}$. For this reason many investigators have used a constant value for f over the range in which they were interested. The use of such a value is apt to lead to erroneous conclusions, especially when attempting to extrapolate outside of the range in which the value was obtained. It is, therefore, recommended that the value of $\frac{Dus}{z}$ always be calculated and the friction factor obtained from the chart (Fig. 273).

Trinks defines "fairly smooth" walls as brickwork which has been evenly laid and brush-coated with cement; "medium rough" as brickwork evenly laid but not so coated; and "very rough" as uneven brickwork, or brickwork which has been covered with slag. It will be noted that the effect of roughness at low values of $\frac{Dus}{z}$ (low velocity) is negligible, and the difference is accentuated as the velocity increases.

The values of s and $\frac{s}{z}$ for air and flue gas at atmospheric pressure, as functions of the temperature in degrees Fahrenheit, are given in Figs. 274 and 275, respectively. The method of using the charts may be illustrated by the following problem.

Example 1.—What will be the pressure drop in inches of water if flue gas at 800°F. is flowing at a velocity of 20 ft. per sec. through a 5-ft.-diameter flue 300 ft. in length, having fairly smooth walls

Step 1. Determine $\frac{Dus}{z}$

where D = 5 ft. 60 in.
u = 20 ft. per sec.
$\frac{s}{z}$ = from Fig. 275 = 0.016.

$$\therefore \frac{Dus}{z} = 60 \times 20 \times 0.016 = 19.2.$$

Step 2. Find friction factor from Fig. 273. Friction factor corresponding to $\frac{Dus}{z} = 19.2$ for fairly smooth walls equals 0.00119.

Step 3. Substitute the proper values of f, L_ρ, u, and d in Eq. (5). $\Delta p = \frac{2f\rho Lu^2}{gd}$

$$\Delta p = \frac{2\overbrace{(0.00119)}^{f} \times \overbrace{(0.00051 \times 62.3)}^{\rho} \times \overbrace{(300)}^{L} \times \overbrace{(20)^2}^{u^2}}{32.2 \times 5}$$

= 0.0564 in. of water.

If the walls were very rough, the pressure drop would be

$$\Delta p = \frac{2\overbrace{(0.0020)}^{f} \times \overbrace{(0.00051 \times 62.3)}^{\rho} \times \overbrace{(300)}^{L} \times \overbrace{(20)^2}^{u^2}}{(32.2)\ (5)}$$

= 0.0948 in. of water,

or almost twice as great for rough walls as for smooth walls.

Loss of Velocity Head Due to Bends.—As indicated by Table CLVII, the pressure drop through 90 deg. bends is equivalent to that through a straight length of flue equal to 30 to 50 diameters of the flue. Many authorities on the flow of gases consider the friction at a right-angle bend to be equivalent to a loss of the total velocity head. Thus, in changing the direction of flow the velocity decreases to zero, with a loss of the entire velocity head. The velocity then increases to its former value at the expense of a loss in pressure head, so that the net result is a loss in pressure head equal to the velocity head at the bend. The method of calculating for this case is given below under Velocity Heads (Eq. (11)).

Other Sources of Pressure Drop.—In considering the flow of liquids, the loss through pipes and bends is the chief item. This is due to the small sizes of pipe with high pressures and, consequently, high pressure drop. Gases, however, are transported at low pressures, and since the volume is extremely large, only small pressure drops are allowable. In the example cited above, the friction drop was only 0.1 inch of water at high velocity and in rough flues. In considering the flow of gases, there are a number of other sources of pressure drop which often are of greater importance than the drop through flues. Among these are (1) pressure drops due to friction through fuel beds, boiler tubes, economizers, regenerators, and ports; and (2) pressure drops due to the change of pressure head to velocity head caused by changing the velocity of the gases and by expansion and contraction losses.

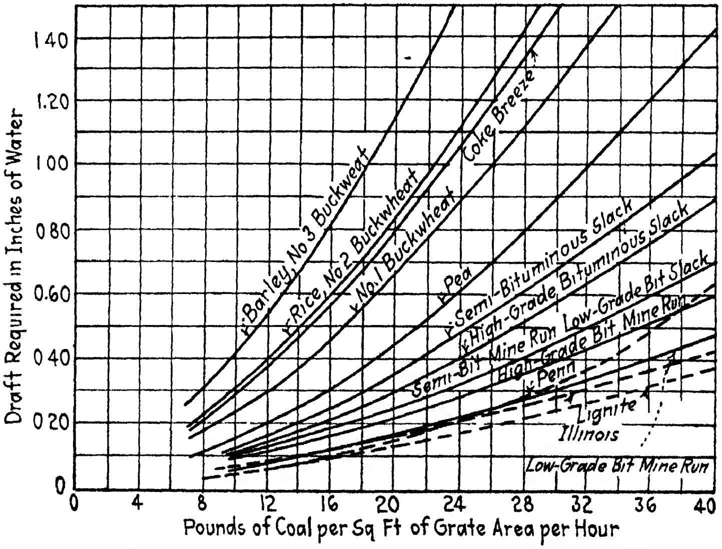


FIG. 276.—Draft loss through fuel beds with different fuels.

Pressure Drop in Fuel Bed.—Friction drop through the fuel bed of a coal-fired furnace cannot be calculated directly from any equations and must be experimentally determined. Figure 276 shows the average values of the draft required in the furnace chamber for burning various fuels at different ratings as given by Cotton.[5] The curves are for reasonably clean fires with hand firing and with average conditions using stokers. Although these represent average values, it is advisable to obtain the data required on any particular stoker from the manufacturer.

Pressure Drop Past Boiler Tubes.—The pressure drop through boiler tubes may be calculated from Eq. (4), where the mean hydraulic radius, or ratio of the cross-sectional area of gas passage to the perimeter of all surfaces touched by the gas, is used instead of a pipe diameter. The results obtained are not of great accuracy, however, since the equation does not hold rigidly under these conditions. The pressure loss depends to a large extent on the type of setting and on the volume of gases flowing. It is advisable to obtain data on the particular setting to be used from the boiler manufacturer.

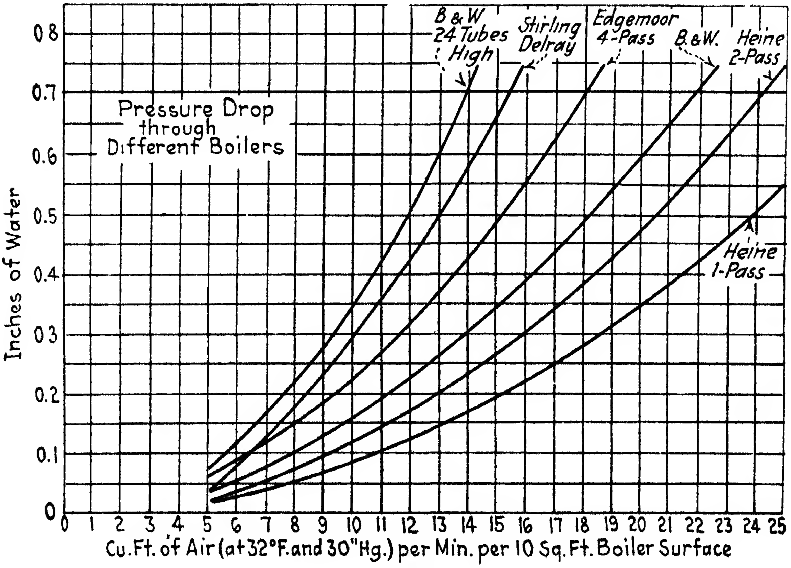


FIG. 277.—Draft losses through boilers.

Figure 277 shows[7] the draft loss through several types of boilers as a function of the air used. Any arrangement whereby the draft loss through the boiler tubes is decreased means a decrease in heating efficiency. Both the rate of heat transfer and the friction drop increase with gas velocity, so that, in general, a balance must be maintained between efficiency of heating and cost of draft. Although in a horizontal water-tube boiler vertical baffles offer more resistance to flow of gases than horizontal baffles, they also give higher rates of heating.

Pressure Drop through Economizers.—Menzin[4] gives the draft required by an economizer as

$$\Delta p = \frac{6.61F^2NT}{10^{12}}, \tag{10}$$

where Δp is the pressure drop (inches of water).

F is the flow of gases (pounds per hr. per lineal foot of pipe in each economizer section).

N is the number of sections.

T is the mean temperature of the gases (degrees Fahrenheit absolute).

The use of this equation is illustrated by the following example:

What will be the pressure drop in an economizer consisting of 36 vertical sections with 10 tubes 10 ft long in each section, when 80,000 lb. of gas per hr. at an average temperature of 400°F. are passing through it?

1. Calculate F:

$$10 \times 10 = 100 \text{ ft. of pipe in each section.}$$

$$F = \frac{80,000}{100} = 800 \text{ lb. per lineal foot per hr.}$$

2. Substitute values in Eq. (10).

$$N = 36,\ T = 460 + 400 = 860\text{°F. abs.}$$

$$\Delta p = \frac{6.61(800)^2(36)(860)}{10^{12}} = 0.131 \text{ in. of water.}$$

Pressure Drop Due to Velocity Head.—The head necessary to impress a velocity on a fluid is

$$h_v = \frac{u^2}{2g}, \tag{11}$$

where h_v is the head (feet of fluid).

u is the velocity (feet per sec.).

g is a constant $= 32.2$.

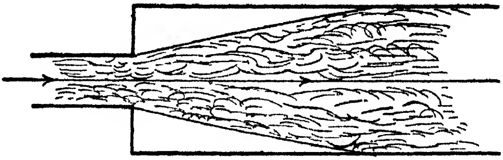


FIG. 278.—Expansion of a fluid from a small into a large section.

Or, expressing h_v in inches of water,

$$h_v = \frac{u^2}{2g}(s)(12) = 0.186u^2s, \tag{12}$$

where s is the specific gravity of the fluid relative to water. This equation applies to all cases where the change in velocity head is from 0 to u. For a change in velocity from u_1 to u_2 the equation is

$$\Delta p = 0.186s(u_2^2 - u_1^2). \tag{12a}$$

The velocity head must be taken into consideration in all cases where fluids are flowing. Where the velocity is low and the other resistances large, the effect of velocity head may be negligible, but in many cases it is a large proportion of the total. The velocity head of a stream of air at 100°F. flowing at the rate of 100 ft. per sec., for example, is

$$\Delta p = 0.186(100)^2(0.00114) = 2.12 \text{ in. of water.}$$

Losses Due to Expansion.—When a fluid expands from a smaller section to a larger, as in Fig. 278, a portion of the velocity head is converted to static head in the larger section, but not all the decrease in velocity head is so converted. The loss in total head when such expansion takes place may be determined from the equation

$$\Delta h = \frac{(u_1 - u_2)^2}{(2g)},$$

when Δh is in feet of the fluid, or

$$\Delta p = 0.186\ (u_1 - u_2)^2 s \tag{13}$$

where Δp is in inches of water and s is the specific gravity relative to water.

Losses Due to Contraction.—When the fluid passage is suddenly contracted, a further loss takes place, since a considerable proportion of the static pressure in the larger section is lost during the conversion to velocity pressure. This loss, in inches of water, may be expressed by the equation

$$\Delta p = 0.186K(u)^2 s, \tag{14}$$

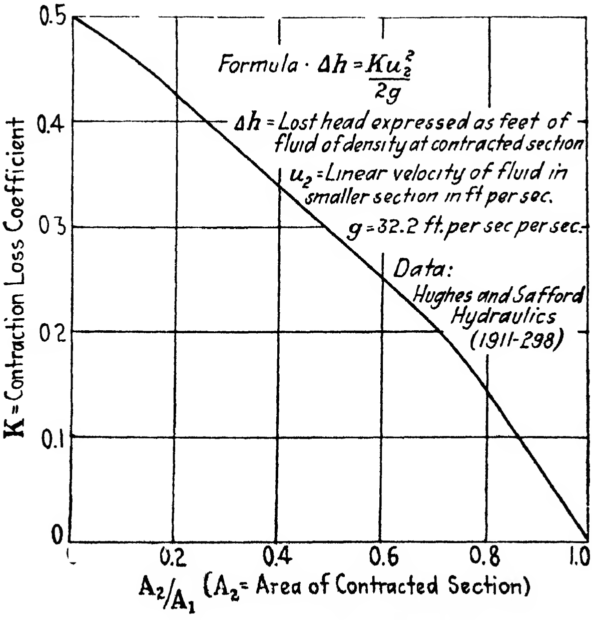


FIG. 279.—Values of contraction loss coefficient K.

where u is the velocity in ft. per sec. in the smaller section and K is the coefficient of contraction, varying with the ratio of areas in the two sections. The value of K as a function of this ratio is given in Fig. 279.

The importance of expansion and contraction losses is great when the friction drop is small, *i.e.*, when the flues are short, but for small changes in size they often may be neglected. The contraction loss in an air stream on entering a flue is 0.186 (0.5) $(u_2)s$, or $0.093u_2s$, since the ratio of $\frac{A_2}{A}$ is 0 and K is 0.5 (Fig. 279). The *total* pressure necessary to give a quantity of air a definite velocity u is the sum of the contraction and velocity heads, or for this last case

$$\Delta p = 0.186u^2s + 0.093u^2s = 0.279u^2s \tag{15}$$

Pressure Drop through Ports.—The friction drop through ports, which are in reality short flues, is usually negligible. However, there may be considerable drop in pressure due to the conversion of pressure head to velocity head. This loss should be considered as a contraction loss and determined from Eq. 14, using the value of K as determined from Fig. 279.

*Recuperators and Regenerators.**—Friction drops through recuperators may be calculated from Eq. (5) as for flues. The friction drop in regenerators may be calculated from Eq. (5), by using the equivalent diameter calculated from the mean hydraulic radius m. The mean hydraulic radius m is the cross-sectional free area for gas flow divided by the perimeter of all surfaces past which the gas is flowing. This value of m may be multiplied by 4 to give the *equivalent diameter* for substitution in Fig. 273 or in Eq. (5). A considerable proportion of the friction loss in regenerators is caused by expansion and contraction losses due to the varying size of gas passage, and should be calculated from Eqs. (12) and (13). In the majority of cases (on account of the relatively low velocities) the drop through the regenerator is small compared to other losses. The "stack effect" in the regenerators, or difference in density between cold air and hot gases, usually is the sole means of bringing air into the furnace and is used largely in giving velocity head to the air. The friction drop through the regenerators must be low for this reason and also to prevent deposition of dust, etc. on the brick from checking the air flow to any appreciable extent.

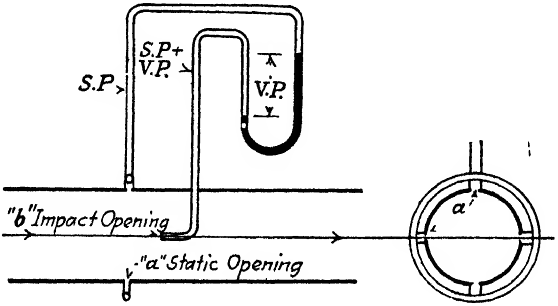


FIG. 280.—Open tube and piezometer ring.

Methods of Measuring the Flow of Fluids.—As previously outlined, velocity head may be expressed by the equation

$$h_v = \frac{u^2}{2g}, \tag{11}$$

from which

$$u = \sqrt{2gh_v}, \tag{16}$$

where u is the velocity in feet per sec. and h_v is in feet of gas flowing. This relation between velocity and the velocity head forms the basis of most instruments for measuring the rate of fluid flow.

Pitot Tube.—The Pitot tube is an instrument for measuring the velocity head directly. Two tubes are inserted into a gas stream, one of which, a

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* For further discussion on recuperators and regenerators see Chap. XI, p. 293.

(Fig. 280), is perpendicular to the gas stream and measures static head only. The second, *b*, is parallel to the gas stream and measures static head plus velocity head. By connecting both tubes to a manometer, the difference, or the velocity head, may be read directly and the velocity calculated. The commercial Pitot tube usually has both tubes integral as in Fig. 281, the static pressure measured by the outside tube with the small holes. This arrangement insures measuring the static and velocity pressures at nearly the same point.

Pitot tubes are used chiefly for exploration to determine the variation of velocity throughout a flue. The velocity in a pipe varies from the sides to the center, and for turbulent motion the *average* value is 0.8 of the *maximum* value (which is usually at the center of the pipe). For accurate work the pipe should be explored and the measurements averaged. Pitot tubes cannot be successfully used with furnace gases due to the relatively low velocities encountered. For example the velocity head for air flowing at 10 ft. per sec. at 70°F. is only 0.0224 in. of water. One advantage of the Pitot tube is that its use causes no appreciable pressure drop in the gas stream.

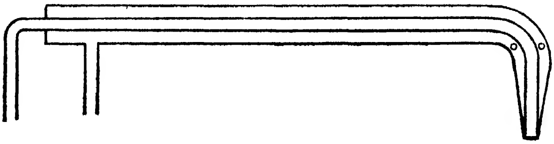

FIG. 281.—Pitot tube.

Orifice.—The most general method of measuring fluid flow is by means of an orifice. A sudden restriction of the gas passage causes the conversion of a portion of the static head to velocity head. The decrease in static head, Δh_s, in feet of gas may be determined from the equation.

$$\frac{u_2^2}{2g} - \frac{u_1^2}{2g} = C'\Delta h_s$$

or

$$\sqrt{u_2^2 - u_1^2} = C\sqrt{2g\,\Delta h_s}, \tag{17}$$

where u_2 and u_1 are the final and initial velocities, C is a coefficient dependent on the type of orifice, and Δh_s is the decrease in static head. If the diameter of the pipe is at least five times the diameter of the orifice, u_1 is $\frac{1}{25}u_2$, u_1^2 is only $\frac{1}{625}u_2^2$, and thus may be neglected in the term $\sqrt{u_2^2 - u_1^2}$. The equation then becomes

$$u = C\sqrt{2g\Delta h_s}. \tag{18}$$

The value of C when a thin plate orifice of type shown in Fig. 282 is used, and when the chamber diameter is at least five times that of the orifice, is 0.61. For a rounded orifice in a thick plate (see Fig. 282) the value is 0.98. The value of C for a short-tube or channel orifice varies greatly with conditions between 0.7 and 0.9 and this type should not be used unless calibrated for the specific conditions under which it is to be used. The thin-plate orifice usually is the cheapest measuring device, is easiest to construct accurately, and is, therefore, in most general use. For the purpose a steel plate, not over $\frac{3}{16}$ in. thick, may be drilled and bolted between flanges. The downstream manometer should be attached a distance of $0.4D$ below the orifice and the upstream manometer a distance $0.8D$ above the orifice, where D is the diameter of the pipe in which the orifice is placed. If the manometers are placed at these points the maximum difference in head is obtained, and it is only under these conditions that the coefficient of discharge is 0.61.

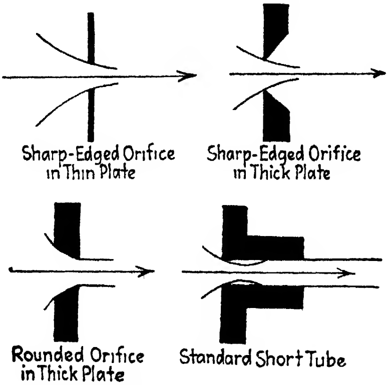

FIG. 282.—Types of orifices.

Venturi Meter.—The major disadvantage of the orifice is the large pressure drop through it. The pressure drop measured is a direct loss and must be supplied by a pump and so represents a continuous cost of operation. This objection is overcome by the *venturi meter,* which operates in the same manner as the orifice but is constructed differently. By making a gradual decrease in the size of the pipe to a minimum and a still more gradual increase back to the original size, the velocity head may be reconverted almost quantitatively to pressure head. The construction is shown in Fig. 283 with the proper angles of divergence on the up- and downstream sides. The throat diameter in this type of instrument is usually about one-fourth the upstream diameter and the coefficient of discharge C is 0.97 to 0.99 in the equation

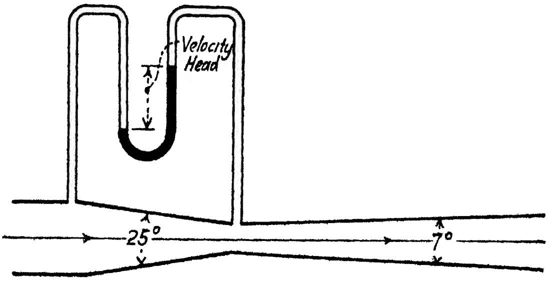

FIG. 283.—Venturi meter.

$$u = C\sqrt{2g\Delta h_s} \tag{18}$$

Since the coefficient depends on accurate machining, each venturi usually is calibrated before being used.

Chimney Design.—In order fully to understand the flow of gases and the design of chimneys, it is necessary to have a knowledge of "stack effect." By this is meant the effect of hot gases rising and being displaced by colder gases, caused by the difference in density between gases at different temperatures. Since the volume of a gas is proportional to its absolute temperature and the density is inversely proportional to the volume, the density of a gas is inversely proportional to the absolute temperature.

The densities of air, and of flue gas having on the average a molecular weight of 30.5, are shown in Fig. 274, as functions of the temperature in degrees Fahrenheit.

Density.—Circulation of gases caused by "stack effect" may be illustrated by a diagram (Fig. 284), in which A represents a column of air at 60°F. and B a column of air of the same height at 600°F. Since the density of air at 60°F. is 0.076 and at 600°F. is 0.038 lb. per cu. ft., there will be an unbalanced force of 0.076 − 0.038, or 0.038 lb. per cu. ft, acting to force the cold gas down column A into B. Under normal conditions column A is not enclosed but is represented by the outside atmosphere. The net "stack effect" is then the difference in densities of the hot gas in the column B and of the outside air. This "stack effect" is present in all places where hot gases rise, including chimneys, regenerators, and vertical furnaces, and must be considered in determining pressure losses. Its theoretical value may be obtained from the difference in densities or, since the density is a function of the absolute temperature, it may be calculated from the equation

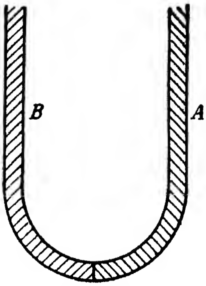

FIG. 284.—Diagram illustrating stack effect.

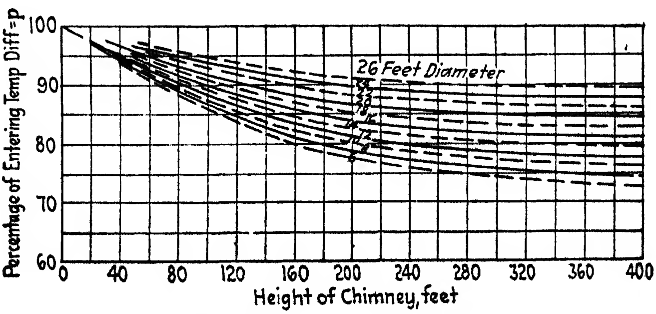

FIG. 285.—Factors for determining the mean temperature of the gases in chimneys.

$$\Delta p = 0.00881P\left(\frac{29}{T_1} - \frac{M_2}{T_2}\right)H, \tag{19}$$

when Δp = draft in inches of water.

P = barometric pressure (inches of mercury).

M_2 = molecular weight of hot gas (29 for air; 30 to 31 for flue gases).

T_1 and T_2 = temperature of air and hot gas (degrees Fahrenheit absolute).

H = height in feet.

Mean Temperature.—The temperature T_2 to be used in this equation is the mean temperature (degrees Fahrenheit absolute) of the column of hot gas. In the design of chimneys, this temperature cannot be taken as the temperature at the base of the chimney. Cotton[5] gives the following method for determining the mean value, using Fig. 285.

t_1 = temperature of atmosphere (degrees Fahrenheit).

t_2 = temperature of hot gases at base of stack (degrees Fahrenheit).

P = factor appropriate to diameter and height from Fig. 285.

t_m = mean temperature of gases in chimney (degrees Fahrenheit).

$$t_m = (t_2 - t_1)P + t_1 \qquad (20)$$

For most cases this may be reduced to

$$t_m = Pt_2 \qquad (20a)$$

without introducing any serious error in calculation.

Effective Draft.—The effective draft for a given chimney is the draft (Eq. (19)) minus the friction loss (Eq. (5)). This effective draft must be equal to the sum of all pressure drops throughout the system, including dampers serving to regulate the amount of draft. In the design of such a system, a balance thus exists between the size of flues and the size of chimney. The smaller the flues the greater the chimney height to maintain the necessary draft. The minimum height of a stack often is fixed by the necessity of discharging the gases at a sufficiently high level to protect surrounding property.

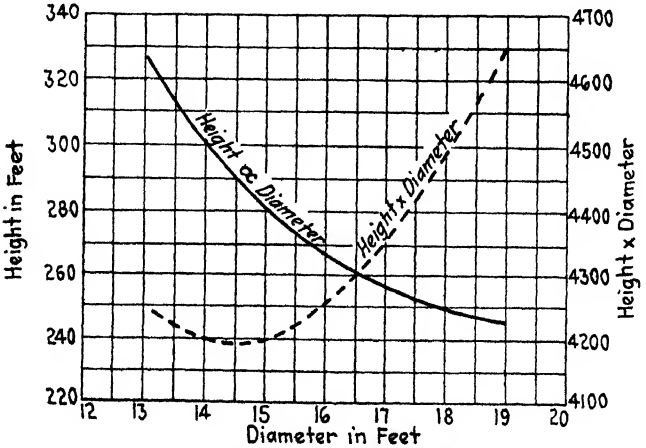

FIG. 286.—Chimney size, giving combinations of height and diameter. Height vs. diameter, full line. Height times diameter, dotted line.

Stack Diameter.—The question of stack diameter is one of friction drop entirely. The stack should be sufficiently large so that at high loads the friction will not be great enough to reduce the available draft materially. The design of stacks to take care of the combustion products from furnaces is essentially that of determining minimum height. The minimum diameter of a brick chimney, due to bricklaying difficulties, is 27 in. and the smallest size usually built is 30 to 36 in. This diameter is sufficient for most smaller furnaces, so that determination of diameter is important only on large installations. For small furnaces, steel stacks may be used and the diameter made only large enough to give the required effective draft.

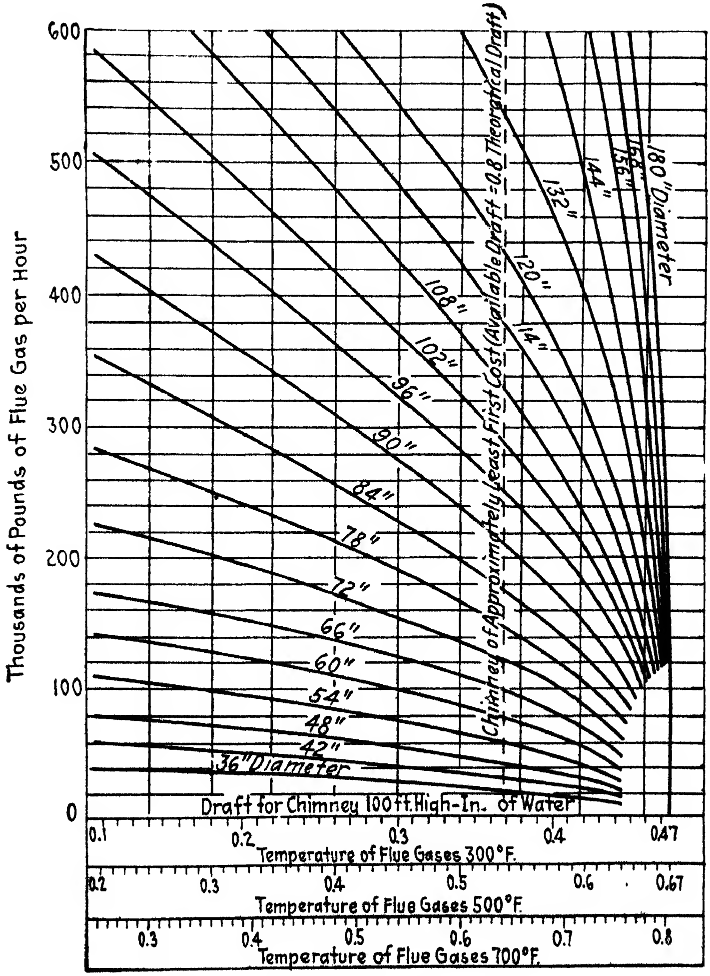

FIG. 287.—Available draft at chimney base for various rates of gas flow and various chimney diameters.

From a rigid friction-drop standpoint, there may be an infinite number of solutions to any problem in the design of flues and stack to operate in a given system. If the flues are made smaller, the stack must be higher, while large flues do not need such a high stack. Practical limitations of sizes, economic consideration of construction costs, and required rates of flow in various sections confine the sizes within narrower limits. For example, the cost of a chimney is approximately proportional to the product of the height and the diameter. The proper chimney to build is the one which gives the lowest value of this product. In determining the size, the diameter first may be assumed and the height calculated. The values of the product of height and diameter will fall on a curve[5] such as Fig. 286 and the cheapest chimney in the case illustrated is one which has a diameter of 14.5 and a height of 238 ft. According to Sibley and Delaney,[6] the chimney of lowest first cost is one in which the available draft is four-fifths the theoretical draft. Figure 287 gives[6] the available draft for various rates of gas flow and chimney diameters per 100 ft. of chimney height. The dotted line in Fig. 287 indicates the chimney of least first cost.

The following example illustrates the method for determining chimney size. A new chimney is to be built for two 500-hp. B. & W. boilers, to operate at normal rating, with maximum loads of 150 per cent rating, using high-grade run-of-mine bituminous coal with a heating value of 13,800 B.t.u. per lb. The grate area is 80 sq. ft. and the boiler tube area 5000 sq. ft. in each boiler. No economizer is used and the temperature of the gases entering the stack is 625°F. The breeching is very short and large in area and the pressure drop through it may be neglected. The size chimney which should be built is calculated as follows:

Step 1. Calculate amounts of coal and of flue gases (assume for purposes of illustration 10.0 lb. air theoretically required per lb. coal, 50 per cent excess air, and 60 per cent efficiency, *i.e.*, a design for poor conditions).

Hp. rated	Max. cap. / rated cap.	B.t.u. per hp.	B.t.u. in coal per B.t.u. in steam	B.t.u. per lb. coal	
500	× 1.50	× 33,479	$\times \frac{1}{0.60}$	$\times \frac{1}{13,800}$	= 3030 lb. of coal per hr. per boiler

Lb. coal per hr.	Lb. air per lb. coal	
3030	× 10.0 × 1.50	= 45,500 lb. air per hr.

If coal has 8 per cent cinder, then the weight of coal to flue gas = (100 − 8)(3030) = 2800 lb. per hr. or total weight of flue gas = 48,300 lb. per hr.

Step 2. Determine loss through fuel bed.

Lb. coal	Sq. ft. grate area	
3030	$\times \frac{1}{80}$	= 38.3 lb. coal per sq. ft. of grate area.

From Fig. 276 pressure drop for bituminous run-of-mine coal (high-grade) at 38.3 lb. of coal per sq. ft. is 0.56 in. of water.

Step 3. Determine pressure drop through boiler.

	Lb. per hr.	Min. per hr.	Cu. ft. (32°) per lb.	Sq. ft. heating surface	
Air used =	46,500		359		= 1.88 cu. ft. per sq. ft.
		60	29	5000	

From Fig. 277 pressure drop through a B. & W. boiler at 18.8 cu. ft. of air per 10 sq. ft. of boiler surface is 0.53 in.

Step 4. Assume or determine a suitable gas velocity and calculate chimney height and size.

Assume velocity of 30 ft. per sec.

	Lb. per hr. per boiler	No. boilers	Cu. ft. per lb. at 32°F.	Temperature correction	
Volume of gas =	48,300	x 2	x $\frac{359}{30}$	x $\frac{1160}{492}$	= 2,725,000 cu. ft. per hr.

$$\frac{2{,}725{,}000}{3600} = 757 \text{ cu. ft. per sec.}$$

$$\frac{757}{30} = 25.3 \text{ sq. ft. area of stack, or, diam.} = 5.68 \text{ ft.} = 68.2 \text{ in.}$$

Velocity head $= 0.186\, u^2 s = 0.186 \times \overline{30}^2 \times 0.00057 = 0.09$ in. water. Calculating the pressure drop in chimney from Eq. (5), where, $D = 68.2$ in., $u = 30$ ft. per sec., $s/z = 0.018$,

and $$\therefore \frac{Dus}{z} = 68.2 \times 30 \times 0.018 = 36.8.$$

From Fig. 273 friction factor (very rough) = 0.00209.

Then $$\Delta p = \frac{2(0.002)(100)(30^2)(0.00057)(62.3)}{(32.2)(5.68)} = 0.0740 \text{ in. water per 100 ft.}$$

Find the mean temperature of stack gases; assume the height to be about 150 ft. The correction factor (for a 6-ft. diameter stack) from Fig. 285 is 0.80 and the mean temperature is, therefore, approximately 500°F. The total draft produced per 100 ft. (Eq. (19)) when the flue gas is 500°F. and the air at 70°F. is

$$\Delta p = 0.00881(30)\left(\frac{29}{530} - \frac{30}{960}\right)100 = 0.621 \text{ in. water.}$$

The available draft (total − friction) is 0.621 − 0.075 = 0.546 in. water 30°C.

The draft necessary is 0.56 + 0.53 + 0.09 = 1.18 in. water

The chimney height must then be

$$(100)\frac{1.18}{0.546} = 216 \text{ ft.}$$

Similar solutions of the problem at different assumed gas velocities are possible. From the combination of heights and diameters, and a knowledge of building costs, the most economical size may be determined. From Fig. 287, the size having the least initial cost may be determined as follows:

Determine weight of gases: 48,300 × 2 = 96,600 lb. per hr.

From Fig. 287 the chimney of lowest first cost for this weight of flue gases has a diameter of 66 in. and an available draft (at 500°F.) of 0.538 in. per 100 ft. necessitating a chimney 219 ft. high to provide 1.18 in. draft.

Distribution of Gases in Furnaces.—Aside from the question of friction, or pressure drop, there is a second factor of great importance in furnace design—the *distribution of the gases*. For large furnaces, such as open hearths and reheating furnaces, it may have greater influence on the working of the furnace than actual pressure drops. Whereas a consideration of friction losses determines *size* of gas passages, distribution is concerned with the *shape* and the *location* of the passages. Friction drop is most important in dealing with ports, flues, and chimneys, while in the working part of the furnace it is usually negligible, or at most, very small, and thus of little significance. On the other hand, the distribution of the gases often determines the effectiveness of the furnace design.

Groume-Grjimailo[8] compares the flow of gases in furnaces to the flow of a light fluid, such as kerosene, through a heavier medium, such as water. Kerosene represents the hot gases and water the surrounding air. The difference in density of hot gases and cold air is very great and the actual ratio of the densities may be as great as that of water to molten iron. Thus at 70°F. the specific gravity of air is 0.0012 and at 3000°F. the specific gravity of flue gas is 0.00019 and the ratio of the density of flue gas to that of air is $\frac{0.00019}{0.0012} = 0.158$. The specific gravity of molten iron is approximately 6.7 and the ratio of the density of water to that of iron is $\frac{1}{6.7} = 0.149$ or in other words the density of hot flue gas to that of air is relatively the same as water to liquid iron. In this conception of gas flow, density is the predominant factor. The great variation in the density of gases with temperature may be noted from Fig. 274. Water may be made to flow *downhill* in a definite path by enclosing the *bottom* and two sides, as in a river. Gases lighter than air may be made to flow *uphill* by enclosing the *top* and two sides. Thus, hydrogen will flow from one bottle to another through the inverted channel. Hot gases in furnaces do likewise and thus may be said to flow in a *reverse* manner to water. The hot gases will first fill the top section of a furnace and, as the quantity increases, they more nearly fill the entire furnace completely. By the use of models of various types of furnaces in which kerosene and water were utilized, Groume-Grjimailo[8] studied the action of several factors on the flow of gases. This method of study gives a good indication of the efficacy of furnace design, but does not take into consideration the effects of mixing and diffusion of the gases. These factors become more important at high velocities. Considering the flow of a *liquid* in a pipe, unless the discharge is at a higher level than the top of the pipe there will be a velocity below which the pipe will not be full. If *hot gases* are flowing through a pipe at a low velocity, they will, according to Groume-Grjimailo,[8] fill the pipe to a level at a definite distance from the top. From this basic conception, the flow of gases may be treated in the same manner as the flow of water and the various formulæ of hydraulics applied to furnace design.

Hot gases tend to rise, due to the difference in density between the gases and the surrounding air, and the greater this difference the greater is the force acting to cause the gases to rise. Low density of a hot gas is equivalent to a pump forcing the gases upward. For example, in regenerators on an open-hearth furnace, the only force acting on the entering air used for combustion is the difference in the density of the hot air in the regenerator and that of the cold air surrounding the furnace. The pressure in the furnace itself is about atmospheric at the furnace level and the pressure at the air intake to the regenerators is atmospheric at the air-intake level, but the pressure at the top of the regenerators is slightly higher than atmospheric pressure. In other words, the force due to the difference in density between the hot gases and cold air for the height of the regenerators is sufficient to overcome the friction losses and to supply the necessary velocity head to the air. It should be noted here that the difference in atmospheric pressure between two points separated vertically by 20 ft. is about 0.31 in. of water.

Since the force acting is greater at higher temperatures, it is *impossible to obtain uniform heating of portions of a divided furnace where the flow of the hot gas is upward.* As the temperature becomes higher in any section of a regenerator, for example, more of the gas flows through that section, since the energy available to cause flow becomes greater. In the cooler section the energy becomes less and it is conceivable that circulation of gases may take place within the regenerator, hot gases flowing up through the hottest parts and down through the coldest parts.

Conversely, it is *impossible to obtain uniform distribution of gases being heated when they are passing down* through a regenerator. The force acting is greatest in those places where the gas flowing is the coldest and least where the gas is the hottest. The result is not only non-uniformity, but a large amount of ineffective furnace area, and a consequent lower temperature of the heated gas. Thus, for uniformity, when gases are giving up heat to surroundings, the flow should be downward and, conversely, when gases are picking up heat from the surroundings the flow should be upward.

Density of the gases is also a very important factor in the operation of kilns. In an up-draft kiln the hot gases rise to the top rapidly and pass out. If the opening at the top is large enough, the stream of hot gases may rise rapidly through the center of the kiln, with cool gases on the sides. The outlet may be reduced so that the time of contact in the kiln is increased and the hot gases forced to fill the entire volume. Even under these conditions, however, the gases which leave the top of the kiln are the very hottest gases. If the kiln is built on the down-draft principle, then the gases leaving are certain to be the coldest gases in the kiln, and will be at a lower temperature with greater economy of operation than in the up-draft kiln. For greatest uniformity of heat distribution, and for maximum economy, the gases should be removed from the kiln, or furnace, at the point of lowest temperature, *i.e.*, the lowest point in the furnace.

These points may be summarized in the statement that, for greatest uniformity, maximum efficiency, and maximum capacity, gases giving up heat must flow downwards and gases receiving heat must flow upwards.

From the analogy of the flow of water and the flow of gas, several of the formulæ developed for flow of water may be directly applied to the flow of gas. Several of these computation methods as advanced by Groume-Grjimailo[8] are given in the following pages.

Flow under Thick-crested Weir.—The formula for the flow of water over a thick-crested weir may be expressed by the equation

$$Q = Kb\sqrt{2gH^3},$$

where Q is the quantity of water.

K is a constant.

b is the length of the weir.

H is the height of water in the reservoir above the weir level.

When h represents the height of water above the weir,

$$h = \tfrac{2}{3}H$$

and

$$h = \sqrt[3]{\frac{Q^2}{K^2b^2}}.$$

The depth of the water flowing *over* the weir may be determined from the rate of flow if the constant K is known. This same equation may be applied to the flow of gases *under* a weir, since the gases may be considered as "water upside down." Yessman[9] gives the equation as

$$h = A\sqrt[3]{\frac{Q^2}{B^2\left(\frac{t-32}{1.8}\right)}}, \tag{21}$$

where h is the thickness of the gas stream in feet, t is the temperature of the gas in degrees Fahrenheit, Q is cubic feet of gas at t°F., B is the width of the furnace in feet, and the value of A varies with h and B, as shown in the accompanying table.

TABLE CLIX

VALUES[8] OF A TO BE USED IN EQ. (21)

Values of h, ft.	Values of B, ft.				
	3 ft.	6 ft.	9 ft.	12 ft.	16 ft.
1.0	2.28	2.36	2.40	2.42	2.43
1.5	2.22	2.32	2.38	2.40	2.41
2.0	2.15	2.29	2.35	2.38	2.39
2.5	2.08	2.25	2.32	2.36	2.38
3.0	2.01	2.20	2.30	2.35	2.38

Reverberatory Furnaces.—A reverberatory furnace may be considered as an inverted weir and the depth of gases flowing through it calculated from the above equation. For efficient working it is necessary that the depth of the gas stream h be as great as the height of the furnace, otherwise the hearth will not be bathed effectively in heated gases.

This method of calculating furnace heights does not take into account the diffusion, eddying, and expansion of the gases, and for this reason may lead to erroneous conclusions. Groume-Grjimailo, however, cites examples to show that the calculation does give accurate indications of the effectiveness of furnace design, and the formula is worthy of wider use.

When h is less than the height of the furnace, the distribution of the heating gases will be poor and the rate and uniformity of heating will be decreased. In order to obtain the required heating effectiveness of such a furnace, it is necessary to increase the value of Q, or, in other words, the fuel consumption, and the economy of a furnace, are thus dependent on the height. If Q is not increased, the time necessary to heat the charge will be greatly increased on account of the poor contact between the hot gases and the product, and both economy and uniformity of heating will be sacrificed. In many furnaces the height of the furnace is determined by combustion space required and is much larger than that necessary for heating. The method outlined here assumes that combustion is complete before the gases enter the furnace The producer-gas-fired open hearth is the most flagrant example of poor gas distribution from the point of view of direct heating. The height of the furnace is far greater than would correspond to the equation above, but the furnace must be large, since it serves as a combustion chamber and, in addition, large-size materials are used often in charging.

Weir and Reservoir.—In certain cases the furnace may be so designed that it is not a thick-crested weir as above, but is a weir with a reservoir. That is, the crest of the weir may be cut to form two smaller crests with a depression between. When a reservoir of this type is used for the flow of water, the level of the water remains the same as with a thick-crested weir, but the velocity in the cistern is decreased. This relationship also holds with hot gases. The actual height h of the weir should be equal to the distance from the top of the outlet port to the bottom of the furnace. This type of furnace permits the heating of large materials while retaining the effective gas distribution of the thick-crested weir.

Inclined Ports.—In order to overcome the tendency of the gases to flow along the top of the open-hearth-type furnace, it is necessary to incline the ports and utilize the velocity head of the gases to cause them to impinge on the hearth. The hot gases naturally will rise, due to the difference in density between them and the relatively cooler gases in the furnace. The force acting in a volume V is $V(P_1 - P_2)$, where P_1 and P_2 are the densities of the stationary and moving gases. The mass of the moving gas is $\frac{V \cdot P_2}{g}$ and the acceleration a is, therefore,

$$\frac{V \cdot (P_1 - P_2)}{\frac{VP_2}{g}} = g\left(\frac{P_1 - P_2}{P_2}\right). \tag{22}$$

The curve showing the path of the gas is given in Fig. 288, where H represents the maximum fall of gases. This fall must be at least as great as the height of the port above the hearth in order to insure proper heating. The depth to which the gas will descend is

$$H = \frac{V^2 \sin^2 \alpha}{2g} \left(\frac{P_2}{P_1 - P_2} \right), \tag{23}$$

and since density is a function of temperature and may be expressed

$$P_2 = P_1 \frac{T_1}{T_2}$$

The equation becomes

$$H = \frac{V^2 \sin^2 \alpha}{2g} \left(\frac{T_1}{T_2 - T_1} \right), \tag{24}$$

where T_1 and T_2 are the temperatures of the stationary and moving gases respectively in degrees Fahrenheit absolute.

Thus if the angle of the center line of a port is 30 deg. ($\sin 30° = 0.5$), the temperature of the hot gases 3300°F., the temperature of the cold gases 600°F., and the velocity of the combustion products 60 ft. per sec., the value of H is

$$H = \frac{(60)^2(0.5)^2}{64.4} \times \frac{600 + 460}{3300 - 600} = 5.68 \text{ ft.}$$

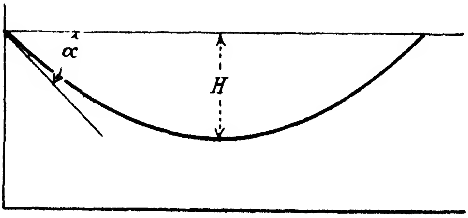

FIG. 288.—Path of gas in a jet.

The value of H, according to Eq. (24), increases as the temperature of the stationary gases increases, until it reaches infinity, when the temperatures of the gases are the same. The minimum value is obtained when the gases in the furnace are cold, and the port should be so designed that the flame will impinge on the hearth when T_1 is a minimum, and thus facilitate the sintering of the furnace bottom. The minimum allowable angle of inclination when the temperature of the stationary gases is 40°F. ($T_1 = 500$) may be determined from the equation

$$\sin \alpha = \frac{1}{V} \sqrt{2gH \frac{(T_2 - 500)}{500}}, \tag{25}$$

where H is the difference in level between the port sill and the furnace bottom in feet. If H is 3 ft., V is 60 ft. per sec., and T_2 is 3300°,

$$\sin \alpha = \sqrt{\frac{64.4(3)\frac{(3300 - 500)}{500}}{60}} = 0.548$$

and $\alpha = 33.3°$.

This method of determining proper inclination is not rigid, since numerous factors come into play which are not accounted for by it. Among these are the combustion taking place in the jet and diffusion and turbulence in the gaseous streams. The value of V and α when the air and gas ports have different inclinations must be determined from the resultant of the two.

Summary on Distribution of Gases in Furnaces.—The foregoing discussion of gas distribution is not intended to form an accurate quantitative basis for furnace design because of the small amount of published data quantitatively comparing actual results with those calculated. The qualitative relationships derived are useful, however, in considering the rational design of new furnaces and in improving the operation of existing furnaces. Consideration of the density difference leads to the following general conclusions:

1. The furnace should be so constructed that the gases withdrawn to the stack are the coldest, and therefore are taken from the *lowest* section of the furnace. Heat distribution in up-draft furnaces, or where the outlet is near the top, will be much poorer than in the down-draft type.

2. The height of the furnace above the hearth should be less than the depth of the gaseous stream calculated for an inverted weir (Eq. (21)).

BIBLIOGRAPHY

1. Walker, Lewis, and McAdams: "Principles of Chemical Engineering," McGraw-Hill Book Coompany, Inc. (1923). The major portion of the part of the chapter dealing with friction losses has been condensed from this book.
2. Wilson, McAdams, and Seltzer: *J. Ind. Eng. Chem.*, **14** (No. 2), 105–119 (1922).
3. W. Trinks: "Industrial Furnaces," vol. 1, John Wiley & Sons (1923).
4. A. L. Menzin: *Trans.* A. S. M. E., **37**, 1065 (1915).
5. Alfred Cotton: Paper on "Chimney Sizes," presented at Spring Meeting of the A.S.M.E. (1923).
6. Sibley and Delaney: "Fuel Oil and Steam Engineering," McGraw-Hill Book Coompany, Inc. (1921).
7. H. S. B.W. Cochrane Corporation: "Finding and Stopping Waste in Modern Boiler Rooms."
8. Groume-Grjimailo (translated by Williams): "Flow of Gases in Furnaces," John Wiley & Sons (1923).
9. Yessman: *Rev. Soc. Russe metál.*, pp. 319-343 (1910).

APPENDIX II

FLOW OF HEAT*

The previous chapters have discussed the development of heat from the combustion of the various fuels. The mechanism of the transfer of heat thus developed to the material being heated is also of great importance.

MECHANISM OF HEAT TRANSFER

The flow of heat may take place by three distinct mechanisms, conduction, convection, and radiation. These may be defined as follows:

1. *Conduction* is the transfer of heat from one part of a body to another part of the same body, or from one body to another in physical contact with it, without appreciable displacement of the particles of the body.
2. *Convection* is the transfer of heat from one place to another within a fluid (liquid or gas) by the mixing of one portion of the fluid with another.
3. *Radiation* is the transfer of heat in the form of radiant energy from one body to another. A hot body gives off radiant energy which, on striking another body, may be partially reflected, partially transmitted, and partially absorbed. The amount absorbed is quantitatively transformed into heat except in very rare cases where photochemical reactions or other energy-absorbing effects occur.

Heat Transfer by Conduction.—The basic principle governing the flow of heat by conduction is *Newton's Law,* which, in general, states that the quantity of heat dQ flowing past a given section at right angles to the path along which the heat is flowing is directly proportional to the area of the section A, to the difference in temperature between the two faces of the section dt, to the time $d\theta$ during which the heat is flowing, and is inversely proportional to the thickness of the section dL. This law may be expressed mathematically as

$$\frac{dQ}{d\theta} = \frac{KAdt}{dL} \tag{1}$$

or, since this discussion deals only with constant rate of flow through any given section,

$$\frac{Q}{\theta} = \frac{KAdt}{dL}, \tag{2}$$

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* Mainly condensed from WALKER, LEWIS, and MCADAMS, "Principles of Chemical Engineering."

where $\frac{Q}{\theta}$ the heat transferred between two faces (in B.t.u. per hr.).
A = the area of cross-section through which the heat is flowing (in square feet.)
dt = differential temperature between the two faces (in degrees Fahrenheit).
dL = differential thickness of the section (in feet).
K = a proportionality constant (B.t.u. per sq. ft. per °F. temperature difference per ft. of length).

In other words, the equation states that the amount of heat which will flow through a body in a given time is directly proportional to the area and to the rate of change of temperature with the length (or the change of temperature per unit of length), and inversely proportional to the thickness.

Conductivity.—The proportionality constant K is called the coefficient of heat conductivity and is expressed as B.t.u. per hr. per sq. ft. per °F. temperature difference, per ft. of length. The value of K varies widely for different materials, and for any given material varies with temperature. Table CLX gives the values of K for a number of different materials. In many problems the introduction of a temperature coefficient is an unnecessary correction, since its value is less than that of the possible error in the value of K itself. This is not true, however, if the values for the particular material used have been determined accurately.

TABLE CLX*

Coefficient of Heat Conductivity of Solids†

(K = B.t.u. per hr. per sq. ft. per °F. difference per ft. thickness.)

Material	Experimental temperature range, degrees Fahrenheit	K	Reference number[d]
(*a*) Metals			
Aluminum	32–212	85	1
Brass, yellow	32–212	55	1
Copper, pure	At 68	238	1
Iron:			
Pure	At 76	37	1
Wrought	32–527	40	1
Cast, 3.5 per cent carbon	At 86	36	1
Lead	At 59	20	1
Nickel	68–392	32	1
Platinum	64–212	41	1
Silver	At 64	243	1
Steel	At 82	25[a]	1
Tin	At 32	35	1
Zinc	At 64	64	1

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* Walker, Lewis, and McAdams: "Principles of Chemical Engineering," p. 181.

† The presence of impurities, especially in the metals, may cause variation in the coefficient of conductivity of almost 100 per cent.

[a] This is the lowest value usually quoted.

TABLE CLX.—(*Continued*).

(*b*) Non-metals

Asbestos	100–1000	0.12	2
Brick:			
Carborundum	At 1800	5.6	2
Magnesia	At 1800	1.8	2
Fire	At 1800	0.75[b]	
Building	At 1800	0.8	2
Silica	At 1800	0.5	2
Kieselguhr	At 1800	0.4	2
Sil-O-Cel	Ar 1800	0.03	2
Brick and mortar wall	......	0.4[b]	
Portland cement, neat	At 95	0.5	2
Cork	122–392	0.03	2
Electrode carbon	212–1700	0.32[c]	2
Glass, flint	50–59	0.3	2
Infusorial earth, 12.5 lb per cu. ft	At 122	0.05	2
Leather	72– 97	0.1	1
Mineral wool	70–350	0.03	2
Magnesia	68–310	0.04	2
Rubber	At 220	0.1	1
Sand	68–310	0.2	2
Wood, pine:			
Parallel to fibers	......	0.07	1
Perpendicular to fibers	......	0.02	1
Petroleum oil	At 80	0.087	1

(*c*) Gases

	K at 32°F.	*K* at 212°F.	Reference number
Air	0.0137	0.0174	3
Carbon monoxide	0.0131	......	3
Carbon dioxide	0.00804	0.0120	3
Oxygen	0.0138	0.0180	3
Nitrogen	0.0138	0.0174	3
Hydrogen	0.101	0.0121	3
Methane	0.0174	......	3
Steam	0.0095	0.0129	4

[b] Average of results from several sources.

[c] The coefficient of conductivity of this material is given as only 10 per cent of this value by other experimenters.

[d] Numbers refer to following works:

1. "Landolt-Bornstein Tabellen" (1905 ed.).
2. Carl Hering: *Met. and Chem. Eng.*, **9**, 652 (1911).
3. "Smithsonian Tables," Smithsonian Inst., Washington, D. C., p. 217 (1920 ed.).
4. Marks: "Mechanical Engineers' Handbook," McGraw-Hill Book Company, Inc., p. 306 (1916 ed.).

Path of Constant Cross-section.—The integration of Eq. (2) for each individual problem may be avoided in most cases by determining its integrated form for the two most general cases met in practice. For the flow of heat across a path of constant cross-section the integrated form of the equation is

$$\frac{Q}{\theta} = \frac{KA(t_1 - t_2)}{L_2 - L_1} = \frac{KA\Delta t}{L}, \tag{3}$$

where Δt is the temperature drop through the section and L is the length of the path of heat flow (in feet).

Example.—The heat loss through a 9-in brick and mortar wall 10 ft. high and 6 ft. wide (when the inner and outer temperatures are 330 and 130°F. respectively) is

$$\frac{Q}{\theta_q} = \frac{KA\Delta t}{L} = \overset{K}{0.4} \times \overset{A}{10 \times 6} \times \overset{\Delta t}{(330 - 130)} \times \overset{1/L}{\frac{1}{\frac{9}{12}}} = 6400 \text{ B.t.u. per hr.}$$

Cross-section Proportional to Length of Path.—Equation (3) is the simplest integrated form of Newton's law and is frequently considered as the law itself. To use this form for cases where the area of the path varies, average values of the area must be employed. The most important case is where the cross-section of the path is proportional to the distance through which the heat is flowing, *i.e.*, through cylindrical objects, and most generally through hollow cylinders (lagged pipes). The integrated form of the equation for this case is

$$\frac{Q}{\theta} = \frac{KbN(t_1 - t_2)}{2.3 \log \left(\frac{L_2}{L_1}\right)}, \tag{4}$$

where N = length of the cylinder (in feet).

b = a constant depending on the shape of the section through which heat is flowing ($b = 2\pi$ if section is circular).

Logarithmic Mean.—Equation (3) may be used to replace Eq. (4) if the proper value of the area of the path of heat flow is used. An average value must be employed because of the varying cross-section. The equation is, then,

$$\frac{Q}{\theta} = \frac{KA_{av}\Delta T}{L}, \tag{3a}$$

where A_{av} is an average area of the path of heat flow. The correct value of A_{av} may be determined from the equation,

$$A_{av} = \frac{A_2 - A_1}{2.3 \log \frac{A_2}{A_1}}, \tag{5}$$

where A_2 and A_1 are the areas of the path at the two limits of the path.

The average of two quantities determined in this way is called the logarithmic mean and is a value frequently used in problems of heat flow. For cases where A_2 is less than twice A_1, the use of the arithmetic mean

$$\frac{A_2 + A_1}{2} \tag{6}$$

instead of the logarithmic mean introduces an error of less than 4 per cent and may be used. When A_2 is more than twice A_1 the logarithmic mean (Eq. 5) should be used. The alignment chart* (Fig. 289) may be used to determine the logarithmic mean graphically. A line connecting two points on the outer curves intersects the middle line at the logarithmic mean of the two outer points. Thus, the logarithmic mean of 30 and 90, as shown by the line on the chart, is 54.8.

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* By E. H. M. Lehnhardt.

Example.—Calculate the heat loss per 24 hr. from 100 ft. of 1-in. pipe carrying steam at 150 lb. gage pressure if the pipe is lagged with 2 in. of magnesia pipe covering ($K = 0.04$). The temperature is 90°F.

Diameter of outside of 1-in pipe = 1 315 in.

Diameter of outside of lagging = 5.315 in.

Temperature of steam at 150-lb. pressure = 366°F.

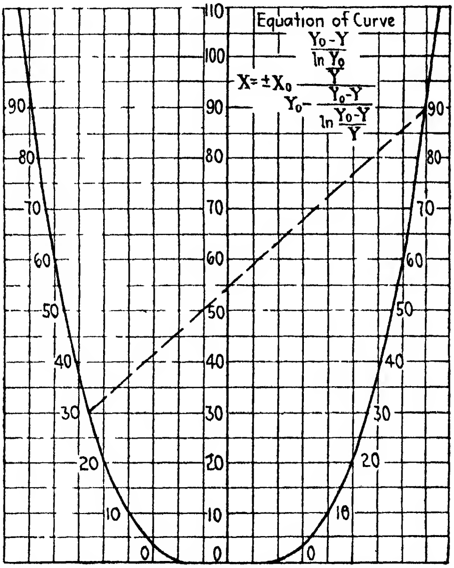

FIG. 289.—Alignment chart for finding logarithmic mean.

This may be assumed to be the temperature of the inner edge of the pipe lagging.

Step 1. Obtaining the logarithmic mean area (Eq. (5)),

$$A_2 = 3.14 \;\Big|\; \frac{5.315}{12} \;\Big|^{\text{Sq. ft.}}\; 100 \;\Big|^{\text{L ft.}} = 139.2$$

$$A_1 = 3.14 \times \frac{1.315}{12} \times 100 = 34.4$$

$$A_{av} = \frac{139.2 - 34.4}{2.3 \log \dfrac{139.2}{34.4}} = 75 \text{ sq. ft.}$$

Step 2. By Eq. (3*a*),

$$Q = 24 \;\Big|^{\theta}\; 0.04 \;\Big|^{K}\; 75 \;\Big|^{A_{av}}\; 366\text{-}90 \;\Big|^{\Delta t}\; \frac{\frac{1}{2}}{12} \;\Big|^{1/L} = 119{,}000 \text{ B.t.u.}$$

The optimum thickness of pipe covering to be used on a steam line may be determined from an economic balance of cost of steam against cost of lagging. Figure 290 shows the heat loss for various thicknesses of lagging, with the cost of lagging and the value of steam saved. The curve of net savings is a maximum, *under these conditions*, at 2.8 in.

Potential Concept.—The quantity of energy or matter transferred can be conceived as controlled by two factors: one a potential and the other a resistance factor. The familiar use of this concept is in electricity, the basic equation for flow of current being $\frac{E}{R} = I$, where E is the potential factor (volts), R the resistance factor (ohms), and I the quantity factor (amperes). The rate of flow of current is here proportional to the potential difference and in the same manner the rate of flow of heat is proportional to the temperature difference. Equation (3) may be written in the same form as Ohm's law:

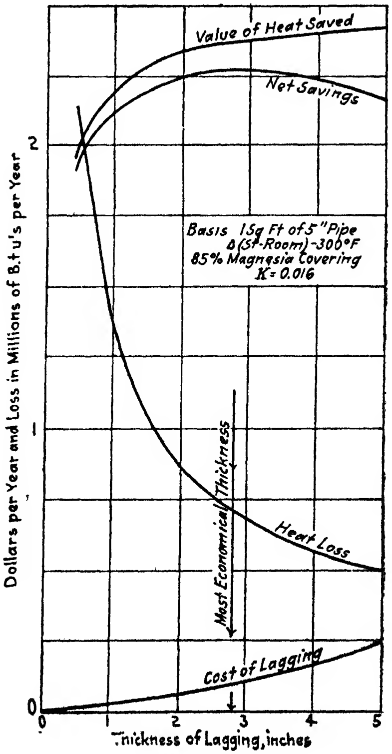

Fig. 290.—Determination of optimum thickness of pipe covering.

$$\frac{Q}{\theta} = \frac{\Delta t}{\frac{L}{KA}}$$

The resistance factor in this case is $\frac{L}{KA}$, the potential factor is Δt, and the quantity factor is $\frac{Q}{\theta}$. The transfer of heat thus may be considered as analogous to the flow of electricity. The amount of heat flowing through a number of resistances in *series* with one another may be determined by dividing the potential factor by the sum of the resistance factors, or

$$\frac{Q}{\theta} = \frac{\Delta t}{\Sigma R} = \frac{\Delta t}{\frac{L_1}{K_1A_1} + \frac{L_2}{K_2A_2} + \frac{L_3}{K_3A_3} + \frac{L_n}{K_nA_n}}, \tag{7}$$

where L_1, L_2, L_3, etc. are the lengths of the path of heat flow through the various resistances, and K_1, K_2, K_3, etc. and A_1, A_2, A_3, etc. are the conductivities and areas of the corresponding resistances.

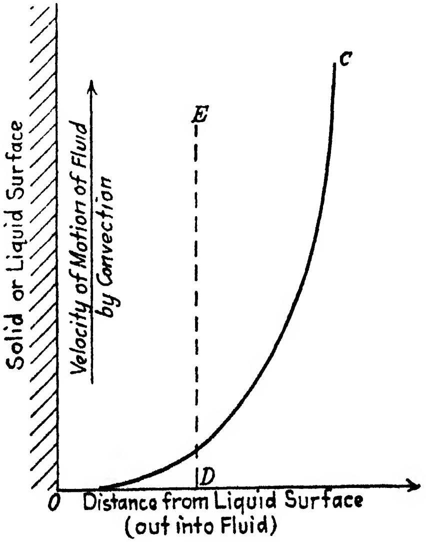

Fig. 291.—Velocity in a fluid film—diagrammatic.

When a number of resistances are in *parallel* the rate of flow is determined by multiplying the sum of the reciprocals of the resistances by the temperature drop, or

$$\frac{Q}{\theta} = \left(\frac{1}{\frac{L_1}{K_1A_1}} + \frac{1}{\frac{L_2}{K_2A_2}} + \frac{1}{\frac{L_n}{K_nA_n}}\right)\Delta t$$

or

$$\frac{Q}{\theta} = \left(\frac{KA}{L_1} + \frac{K_2A_2}{L_2} + \frac{K_nA_n}{L_n}\right)\Delta t. \tag{8}$$

Heat Transfer by Convection.—When a liquid or gas is in contact with a solid, or when a liquid or gas moves over a solid surface, or when a gas moves over a liquid surface, there is a film of the moving substance upon the surface of the stationary one. This film has little motion, and acts toward the transfer of heat, vapor, or matter as though it were a separate material. Gases or liquids mix intimately in large spaces, due to the small frictional resistance, but at the surface of a solid the motion of the fluid is so small as to be almost negligible and the film may be considered as stationary. A plot of fluid velocity as determined by distance from a solid wall would be of the type shown in Fig. 291. Langmuir[2] experimentally showed the existence of this film by measuring air temperatures while gradually approaching a heated silver plate. At a point approximately 0.1 in from the edge of the silver plate the temperature began to rise rapidly until it reached that obtained by touching the plate. This film acts to retard the flow of heat energy or material and is of great importance in absorption, corrosion, and heat transfer.

Film Coefficient.—Although heat transfer by convection in a fluid is extremely rapid, due to low frictional resistance in the fluid and consequent intimate mixing, the transfer of heat from a solid wall to a liquid is comparatively slow. Heat can be transferred through the stationary film by *conduction* only, but, having passed the film boundary, may rapidly distribute itself by *convection* through the main body of the fluid. Figure 292 illustrates the temperature gradient in a heat interchanger, where heat is being transferred from one fluid to another fluid through a solid wall. The temperature throughout each fluid is essentially constant except near the boundary. The maximum temperature drops are (t_1-t_2) and (t_3-t_4) through the stationary films at the solid surface.

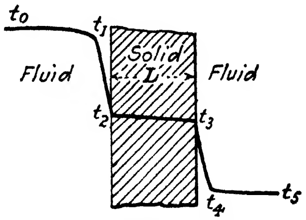

FIG. 292.—Diagram illustrating film concept.

Since the thickness of the fluid film is not susceptible to accurate measurement, the conduction Eq. (3) is expressed for this case as

$$\frac{Q}{\theta} = hA(\Delta t), \tag{9}$$

where h replaces the term $\frac{K(\text{conductivity of film})}{L(\text{thickness of film})}$ in Eq. (3). The coefficient h is called the "film coefficient" or surface coefficient and is expressed as B.t.u. per hr. per sq. ft. per °F. temperature difference. The value of Δt is the temperature drop from the surface of the solid to the edge of the film, or t_3-t_4 in Fig. 292 Since t_4 is essentially equal to t_5, the value used is t_3-t_5.

Although the film coefficient h is probably the most important of any in the consideration of heat flow, it cannot be evaluated as simply as can the conductivity coefficient K. The variation of h under many conditions cannot be estimated except qualitatively. Its value depends on the physical characteristics of the fluid, the nature and the shape of the solid surface, and the velocity of the fluid past the solid boundary. Any mechanism which tends to disrupt or to decrease the thickness of fluid film increases the rate of heat transfer or the value of h.

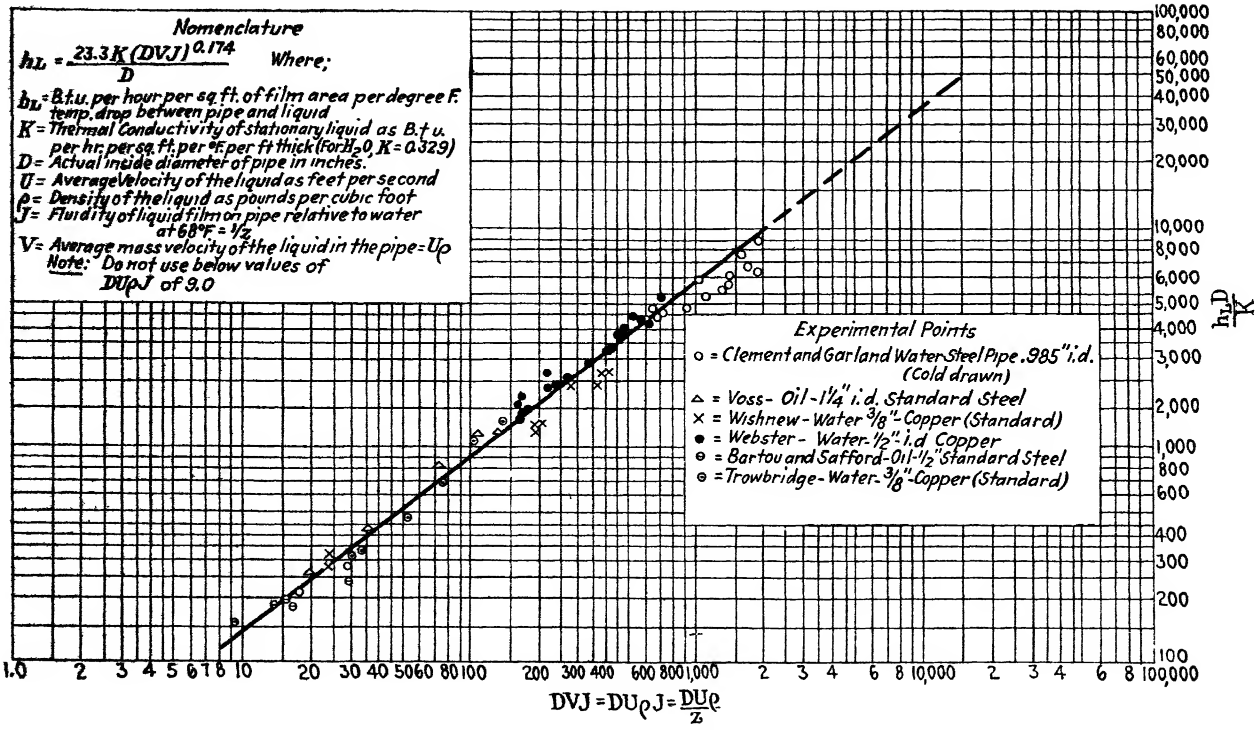

FIG. 293.—Film coefficient for liquids flowing through pipes in turbulent motion

Overall Coefficient.—The rate of heat flow in an interchanger as in Fig. 292 may be determined from the concept of series flow, where the resistances are the two liquid films and the intervening wall, $\frac{1}{h_1A_1}$, $\frac{L}{K_2A_2}$, and $\frac{1}{h_3A_3}$, respectively. The sum of these resistances may be replaced by an "overall" resistance $\frac{1}{H}$ and the resultant equation becomes

$$\frac{Q}{\theta} = \frac{\Delta t}{\frac{1}{h_1A_1} + \frac{L}{K_2A_2} + \frac{1}{h_3A_3}} = HA_{av}\Delta t, \tag{10}$$

where H = overall coefficient of heat transfer.
A_{av} = average cross-sectional area of path.

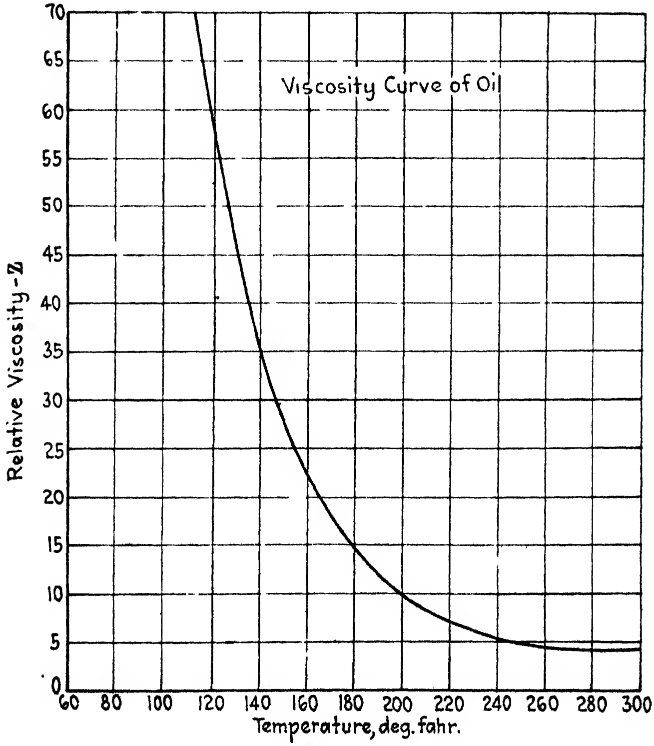

FIG. 294.—Viscosity curve of oil.

Early investigators used the "overall coefficient," but since any factor which influences any one of the resistances involved does not affect the overall coefficient to the same degree, it is impossible to calculate or to forecast the value of *H* directly. The logical and simplest method is to determine each resistance separately, then combine and use them in Eq. (10). The calculation of the film resistances is taken up below.

Forced Convection.—One of the most important cases of the film coefficients for liquids is that for a liquid flowing rapidly under pressure inside a pipe. As a result of a critical survey of the results of many investigators, McAdams and Frost[3] found that such a film coefficient varies directly as fluid conductivity, inversely as the diameter of the pipe, and is a function of the modulus $\frac{DU\rho}{Z}$, where D is the pipe diameter, U the linear velocity, ρ the density, and Z the viscosity of the fluid relative to water. Figure 293 gives the values of $\frac{h_L D}{K}$ for various values of $\frac{DU\rho}{Z}$, and since D is known and K may be determined from tables similar to Table CLX, the value of h_L to be used in Eq (9) may be calculated. This same plot may be used for liquids flowing in the circular space between the pipes, if the diameter D is taken as that of a pipe having the same hydraulic radius as the circular space. The equation of this line (Fig. 293) is

$$h = 23.3 \frac{K\left(\frac{DU\rho}{Z}\right)^{0.8}}{D} = \frac{23.3KU^{0.8}\rho^{0.8}}{D^{0.2}Z^{0.8}}. \tag{11}$$

The following example indicates the method of using the above equations.

Example.—An oil, having a specific gravity of 0.936 and a viscosity curve as shown* in Fig. 294, enters a double pipe heater at 85°F. Live steam at 100 lb. per sq. in. pressure is inside a 1¼-in. pipe and the oil is in the circular space between the 1¼ and a 2-in. pipe. The heater† consists of seven pieces of pipe, each 17 ft. long, connected by double bends. What will be the exit temperature of the oil when flowing at a velocity of 6 ft. per sec.

Step 1. Determine $\frac{DU\rho}{Z}$

Outside diameter of 1¼-in. pipe = 1.66 in.
Inside diameter of 2-in. pipe = 2.07 in.

For annular space mean hydraulic radius m = area of annular space divided by the wetted perimeter =

$$\frac{\frac{\pi(d_2^2 - d_1^2)}{d}}{\pi(d_2 + d_1)} = \frac{d_2 - d_1}{4},$$

where d_1 is the outside diameter of the small pipe and d_2 the inside diameter of the large pipe. Since the equivalent diameter of a section is $4m$, the value for this case is $d_2 - d_1$, or 2.07 − 1.66 = 0.41 in.

To determine the average value of Z, assume the exit temperature to be about 200°F.

Temperature of steam, 338°F.
Temperature entering oil, 85°F.
Temperature exit oil (assumed), 200°F.

Assume temperature drop through pipe negligible, then average temperature of the oil film at entrance is $\frac{338 + 85}{2} = 212$°F. and at exit it is

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* See oil p in Fig. 34, Chap. VI.

† See Fig. 182, Chap. XVI.

$\frac{338 + 200}{2} = 269°\text{F.}$ Viscosity at entrance (Fig. 294) = 8.0 and at exit = 4.2. Average viscosity $= \frac{8.0 + 4.2}{2} = 6.1$

$$\frac{DU\rho}{Z} = \frac{0.41 \times 6 \times 0.936 \times 62.3}{6.1} = 23.5$$

Step 2. Determine h_L

From Fig. 293, $\frac{h_L D}{K} = 290.$

From Table CLX, $K = 0.087.$

$$h_L = \frac{290(0.087)}{0.41} = 61.5.$$

From Eq. (11), $h = \frac{23.3(0.087)(23.5)^{0.8}}{0.41} = 61.7$

Step 3. Determine area of heating surface.

Outside area of 1¼-in pipe =

$$17 \times 7 \times \frac{1.66 \times \pi}{12} = 51.7 \text{ sq. ft.}$$

Step 4. Determine $\frac{Q}{\theta}$.

Specific heat of oil = 0.50 (Table LV, Chap. X).

Area of oil space $= \frac{\pi}{4 \times (144)} (2.07^2 - 1.66^2)$

$= 0.00835$ sq. ft.

Amount of oil per hr. =

$$\overset{\text{Sq. ft.}}{0.00835} \times \overset{\text{Ft. per sec.}}{6} \times \overset{\text{Sec. per hr.}}{3600} \times 62.3 \times \overset{\text{Lb. per cu. ft.}}{0.936} = 10{,}500 \text{ lb. per hr. or 1350 gal. per hr.}$$

$\frac{Q}{\theta} = 10{,}500 \times 0.5 \times (t - 85)$, where t is exit temperature of the oil.

Step 5. Calculate Δt.

Steam at entrance = 368°F.; oil temperature = 85; $\Delta t = 283$.

Steam temperature at exit = 368°F.; oil temperature = t; $\Delta t = 368 - t$.

$$\text{log mean } \Delta t = \frac{283 - (368 - t)}{2.3 \log \frac{283}{368 - t}} = \frac{t - 85}{2.3 \log \frac{283}{368 - t}}.$$

Step 6. Substitute values in Eq. (9).

$$10{,}500 \times 0.5 \times (t - 85) = 61.7 \times 51.7 \times \frac{(t - 85)}{2.3 \log \frac{283}{368 - t}}$$

$$\log \frac{283}{368 - t} = 0.265$$

$$283 = 1.84(368 - t)$$

$$t = 214°\text{F.}$$

The value of Z, determined from assumption of 200°F. exit temperature, may be recalculated. When $t = 214$, temperature of the film is $\frac{338 + 214}{2} = 276°$ and $Z = 4.1$. The average value of Z is $\frac{8 + 4.1}{2} = 6.05$, which is so close to 6.1, as calculated from the assumed temperature, that the correction is negligible.

The temperatures of the oil, calculated in this manner for various rates of oil flow, are shown in Fig. 183 of Chap. XVI.

Natural Convection.—Where only natural convection takes place two other types of liquid film occur in practice, first, when the liquid is boiling and, second, where the liquid is not boiling. When the liquid is not boiling, values of h from 50 to 300 may be encountered with water, while for more viscous fluids the value may be as low as 10 to 20. Stose and Whittemore[4] give the following empirical relationship governing the value of h in sugar solutions.

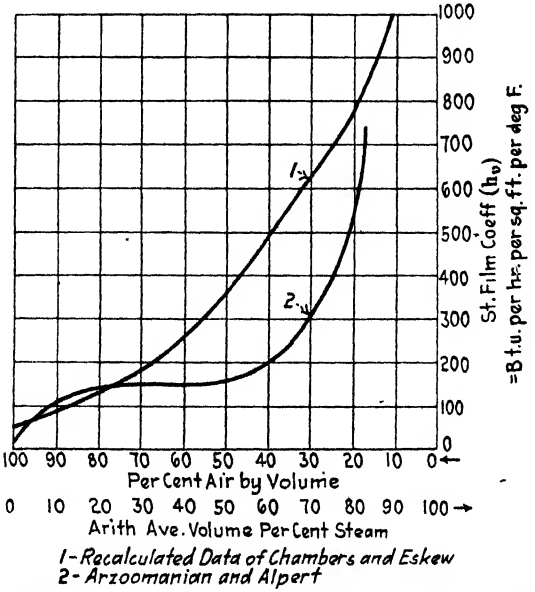

FIG. 295.—Effect of air on the film coefficient from condensing steam to pipe.

$$h = \frac{33.6(\Delta t)^{0.412}}{Z^{0.314}} \tag{12}$$

For boiling liquids the value of h is much higher, due to the disruption of the liquid film. No values of the film coefficient under these circumstances are available, but values of the overall coefficient from condensing steam to boiling liquid, ranging from 130 to 510, have been determined in evaporators in sugar factories.[5]

Condensing Vapor Films.—Extensive experiments have been carried out to determine the value of h from *condensing* steam to a solid. The values vary from 1400 to 3400, and since the effect of all factors is not known, an average value of 2000 is recommended. For other condensing vapors the equation

$$h_v = 2200\frac{K}{Z}, \tag{13}$$

where K equals the conductivity of the condensing liquid and Z equals its viscosity relative to water, may be used as an approximation. The effect of non-condensible gases is extremely important in connection with heat transfer through condensing vapor films. Figure 295 shows the results of two investigators[6,7] on the effect of air in steam on the value of h_v.

Gas Films.—The resistance to heat flow offered by gas films is always much greater than that offered by liquid films. The film coefficient of a gas depends on the thickness of the film and the conductivity of the gas, which, in turn, depend on the film temperature. The value of the gas film coefficient h_G is not accurately known for natural convection conditions. Langmuir[2] from experiments with a silver disc gives the equation

$$h_G = 0.28(\Delta t)^{0.25}, \tag{14}$$

where h_G is the coefficient from a vertical surface. For a horizontal disc Langmuir[2] found the coefficient to be 10 per cent greater on the upper side and 50 per cent less on the lower side as compared with the sides of vertical discs. The equation[1] recommended for estimating the convection from vertical walls is

$$h_G = 0.7 + \frac{\Delta t}{375}, \tag{15}$$

where Δt is the difference in temperature between the wall surface and the surrounding air.

With higher velocities, caused by forced convection, inside pipes or flues the value of h_G becomes higher as the film becomes thinner. Although a large amount of experimental work has been carried out to determine value of h_G, most of it has been for special conditions, and empirical formulas have been developed for special cases. Weber[8] considered all available data and derived the following equation, which is recommended as the most general and reliable one to be used when gases flow through pipes and flues:

$$h_G = \frac{0.88(V^{0.8})(T^{0.6})(S^{0.2})(C_p)}{M^{0.3}}, \tag{16}$$

where V is the mass velocity of the gas (pounds per sec. per sq. ft.).

T is the temperature of the gas (degrees Fahrenheit absolute).

S is the surface factor, *i.e.*, area of surface (square feet) divided by volume of open space through which gas is flowing (cubic feet). For circular pipes $S = \frac{4}{d}$.

C_p is the average specific heat of the gas at constant pressure.

M is the average molecular weight of the gas.

The data for this equation were obtained in pipes from ½ to 2 in. in diameter, for temperatures from 70 to 2000°F., and gas velocities of 1 to 20 lb. per sq. ft. of free area per sec. These limitations must be considered in applying the equation to larger pipes. For ordinary calculations with air inside pipes of 1- to 2-in. diameter the equation becomes approximately

$$h_G = cV^{0.8}, \tag{17}$$

where c varies between 3.7 and 4.5. For air at 200°F. in a 1-in. pipe the value of c is 4.2 and the value changes as indicated by Eq. (16).

The equation governing the values of h_G outside pipes was developed by Carrier[9] from tests on various heaters in which air flows at right angles to the length of the pipes. The equation is

$$h_G = \frac{22.3V}{V + 1.42}, \tag{18}$$

where V is the mass velocity in pounds per sec. per sq. ft. of free area at the line of centers in any row.

Optimum Velocity.—Anything that increases the pressure drop through a system also increases the coefficient of heat transfer. Greater friction and disruption of the gas film decrease the film resistance to the transfer of heat.

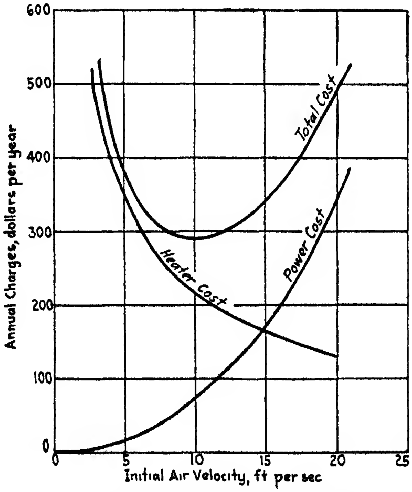

FIG. 296.—Optimum velocity in an air heater (by an economic balance).

In the design of air heaters there is an optimum velocity where the cost of the heater plus the power cost is a minimum. In Fig. 296 the fixed charges and power cost for heating 10,000 cu. ft. of air per min. from 70 to 170°F. have been plotted versus the gas velocity. The minimum yearly cost from this plot occurs at a gas velocity of 10 ft. per sec. The actual velocity to be used in any case depends on the rate for fixed charges and the cost of power and should be determined by an economic balance of this type.

Logarithmic Mean Temperatures in Heater Calculations.—In all types of heaters there is a variable temperature difference between the heating material and the material being heated. For substitution in any of the equations an average value, Δt_{av}, must be used. If equilibrium has been established and each point within the system is maintained at a constant temperature, and the resistance to heat flow is constant throughout the system, the logarithmic mean of the values of Δt at each end of the system gives the true average.

$$\Delta t_{av} = \frac{\Delta t_2 - \Delta t_1}{2.3 \log \frac{\Delta t_2}{\Delta t_1}}. \tag{19}$$

Where Δt_1 is over 50 per cent of Δt_2 the arithmetic mean may be used without serious error. The arithmetic mean usually gives a value which is quite high. For example, consider a countercurrent heater in which oil is being cooled from 150 to 25°F. by water which enters at 20 and leaves at 70°F. The temperature differences at the two ends are 80 and 5° and the logarithmic mean value is

$$\Delta t_{av} = \frac{80 - 5}{2.3 \log \frac{80}{5}} = 27.1^\circ\text{F.}$$

The arithmetic mean is $\frac{80 + 5}{2} = 42.5^\circ$F., or 57 per cent higher than the correct value. The value as determined from Fig. 289 is 27.1.

Heat Transfer by Radiation.—All materials emit and receive energy in the form of radiant heat. Part of the heat thus received is absorbed and the ratio of the amount absorbed to the total amount received is essentially constant for a given material. A substance which absorbs all the heat received, and for which the above ratio is unity, is called a "black body." The ratio of the heat absorbed to the heat received is the black-body coefficient p.

The total energy radiated per unit area for a "black" body is independent of the material and is proportional to the fourth power of the temperature of the body. Mathematically, this law may be expressed as

$$\frac{Q}{\theta} = bAT^4 \tag{20}$$

where b is a constant, A is the area (square feet), and T is the temperature (degrees Fahrenheit absolute).

Not all bodies may be considered as "black," however, and the equation, corrected for relative blackness, becomes

$$\frac{Q}{\theta} = bApT^4, \tag{21}$$

where p is the black-body coefficient. The values of p to be used in this equation are given in Table CLXI.

The important factor is not the total radiation from a body, but the difference in the heat radiated by two bodies at different temperatures. This *net* radiation may be determined by substituting in Eq. (21) the values of T^4 corresponding to the temperatures and subtracting the smaller from the larger.

The *net radiation* between two parallel planes, indefinite in extent, and each at a constant temperature, may be expressed by the equation

$$\frac{Q}{\theta} = \frac{A\left[\left(\frac{T_1}{100}\right)^4 - \left(\frac{T_2}{100}\right)^4\right]}{\frac{1}{p_1C} + \frac{1}{p_2C} - \frac{1}{C}} \tag{22}$$

where T_1 and T_2 are the temperature of the two planes in degrees Fahrenheit absolute, p_1 and p_2 are the blackness coefficients of the two planes, and C is a constant (0.162 is used throughout this chapter, although the value 0.172 is now more generally accepted).

TABLE CLXI[*]

RELATIVE RADIATING POWER OF SOLIDS BELOW 1600°C.

(Ideal black body, 1.0.)

	Average value (= p)	Reference number[b]
Iron:		
Cast and bright	0.22	1
Oxidized	0.62	1
Steel, polished	0.20	2
Steel, highly oxidized	0.90[a]	
Copper:		
Polished	0.10[a]	
Calorized	0.26	1
Oxidized	0.72	1
Aluminum paint	0.50	1
Gold enamel	0.37	1
Monel metal	0.43	1
Silver	0.03	1
Gold, polished	0.05[a]	
Brass:		
Bright	0.07[a]	
Dull	0.23	2
Zinc:		
Bright	0.10[a]	
Dull	0.22	2
Tin, polished	0.10[a]	
Nickel	0.35[a]	
Mercury	0.23[a]	
Platinum	0.24[a]	

[*] Walker, Lewis, and McAdams: "Principles of Chemical Engineering," p. 184.
[a] Average of values given by various authorities.
[b] Numbers refer to following works:
1. LANGMUIR: *Trans.* Am. Electrochem. Soc., **23** (1913).
2. WAMSLER: *Mitt. Forschungsarbeiten*, **98** (1911).

This equation also applies to spheres of large, and nearly equal, diameters, and may also be used as an approximation in the calculation of heat transfer by radiation in recuperators and concentric pipes.

It may be shown mathematically that in a space entirely surrounded by a solid body, whatever the shape or blackness of the inner walls may be, the total radiation and reflection per unit area is the same as though the walls were black. Furthermore, any small body inside the enclosure receives as much energy as if the walls formed a true black body. The radiation emitted from it is dependent on its coefficient of blackness, so that the net radiation under these circumstances is

$$\frac{Q}{\theta} = ApC\left[\left(\frac{T_1}{100}\right)^4 - \left(\frac{T_2}{100}\right)^4\right], \tag{23}$$

where A is the area of the smaller body (square feet).

C is a constant = (0.162).

p is the coefficient of blackness of the small body.

T_1 and T_2 are the temperatures of the walls and the small body respectively.

*Example.**—A furnace at 1500°F. has an opening 2 sq. ft. in area exposed to surroundings at 70°F. Determine the heat lost by radiation from the opening.

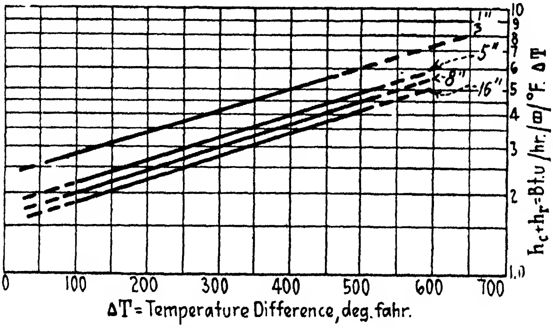

FIG. 297.—Coefficient of heat transfer by radiation plus convection $(h_c + h_r)$ from bare pipes to a room at 70°F.

Data: Black-body coefficient of surroundings . . 0.9
Black-body coefficient of furnace. 0.9

$$\frac{Q}{\theta} = \overset{C}{0.162}\;\overset{p}{0.9}\;\overset{A}{2}\left[\left(\frac{1500+460}{100}\right)^4 - \left(\frac{70+460}{100}\right)^4\right]$$

$$= 0.291 \times (148{,}000 - 800)$$
$$= 42{,}800 \text{ B.t.u. per hr.}$$

For convenience in determining the heat loss from a hot body the loss by radiation and convection sometimes may be added together. Over short ranges the fourth power equation may be represented, approximately, as proportional to the temperature difference. Since this is a case of parallel flow, the equation becomes

$$\frac{Q}{\theta} = (h_c + h_r)\Delta t$$

---

* For further application of this equation see following chapter.

where h_c is the film coefficient for convection and h_r is the simplified radiation coefficient.

When the enclosing walls are at 70°F. the values of h_r from 0 to 350°F. and 350 to 700°F. become

$$h_r = p\left(1 + \frac{\Delta t}{300}\right) \text{ (0 to 350°F.).} \tag{24}$$

$$h_r = p\frac{\Delta t}{145}\text{(350 to 700°F.).} \tag{25}$$

Values of $h_c + h_r$ from bare pipes to a room at 70°F. are given in Fig. 297 for pipes ½ to 16 in. in diameter. Values of p for metals are given in Table CLXI. For most other materials, such as furnace walls, a black-body coefficient of 0.9 may be used without serious error.

BIBLIOGRPAHY

1. Walker, Lewis, and McAdams: "Principles of Chemical Engineering," McGraw-Hill Book Company, Inc. (1923).
2. Langmuir: *Trans.* Am. Electrochem. Soc., **23,** (1913).
3. McAdams and Frost: *Ind. Eng. Chem.,* **14,** 1, 12 (1922).
4. Stose and Whittemore: Mass. Inst. Tech., *Thesis* (1922).
5. Kerr: *Trans* A.S.M.E., **38,** 67 (1912).
6. Chambers and Eskew: Mass. Inst. Tech., *Thesis* (1921).
7. Arzoomanian and Alpert: Mass Inst. Tech., *Thesis* (1922).
8. Weber: Mass. Inst. Tech., *Thesis* (1919).
9. Carrier: *Trans.* A.S.M.E., **33,** (1911).

APPENDIX III

RATE OF HEATING

In the previous chapter heat flow under equilibrium (constant) conditions only was discussed. Many heat-flow problems fall into that classification; that is, the process is continuous and, a short time after starting, temperatures throughout the apparatus become constant. In furnace work a considerable number of problems involving the flow of heat come under another class, namely, the process is not continuous and the operation consists in heating a material to a definite, final temperature. Heat-treating and reheating furnaces come under this classification and for the solution of such problems an entirely different method is necessary.

Mathematical Development.—The basic equation expressing Newton's law mathematically is founded on the conservation of energy. The heat remaining in a differential cube of material, that is, the difference between input and output, may be expressed as

$$K\left(\frac{\partial^2 t}{\partial x^2}+\frac{\partial^2 t}{\partial y^2}+\frac{\partial^2 t}{\partial z^2}\right)dx\ dy\ dz\ d\theta, \tag{1}$$

where K is the conductivity; $\frac{\partial^2 t}{\partial x^2}$, $\frac{\partial^2 t}{\partial y^2}$, and $\frac{\partial^2 t}{\partial x^2}$ are the differences in the rates of change of temperature between the two opposite faces of the cube in each direction; dx, dy, dz is the volume; and $d\theta$ is the time.

This may be equated to the increase of heat in the cube

$$SC\ dx\ dy\ dz\ \frac{\partial t}{\partial \theta}d\theta, \tag{2}$$

where S is the density, C the specific heat, and $\frac{\partial t}{\partial \theta}$ the rate of change of temperature. Equations (1) and (2) give the general form of Newton's law

$$\frac{K}{SC}\left(\frac{\partial^2 t}{\partial x^2}+\frac{\partial^2 t}{\partial y^2}+\frac{\partial^2 t}{\partial z^2}\right)=\frac{\partial t}{\partial \theta}. \tag{3}$$

Considering the case where equilibrium (constant) conditions are maintained, with flow in one direction only, the solution of this differential equation gives the usual form of Newton's law as expressed by Eq. (3) in the preceding chapter. Considering unsteady conditions with flow in one direction only gives the basic equation

$$\frac{K}{SC}\left(\frac{\partial^2 t}{\partial x^2}\right)=\frac{\partial t}{\partial \theta}, \tag{4}$$

where $\frac{K}{SC}$ is the diffusivity and is usually designated as (α^2).

Considering a slab of material with heat flow in one direction only, the heat flow through a unit area at the surface may be expressed as

$$\frac{dQ}{d\theta} = -K\frac{dt}{dx}, \tag{5}$$

and the flow from the surface into the surrounding atmosphere is

$$\frac{dQ}{d\theta} = h(t - t_o), \tag{6}$$

where h is the coefficient of heat transfer from the surface, t is the temperature of the surface, and t_o is the temperature of the atmosphere.

Equating Eqs. (5) and (6) gives the differential equation

$$\frac{dt}{dx} - \frac{h}{K}(t - t_o) = 0. \tag{7}$$

Solution of Equations.—Assuming h and K to be constant over the temperature range in which the equation is to be used, Eq. (7) may be solved and from the solution the temperature distribution throughout the slab at any time may be determined. Similar equations and solutions have been set up for materials of different shape. The solution of such equations is not extremely difficult but is tedious on account of their unwieldy nature. A graphical method of solution, therefore, is advisable.

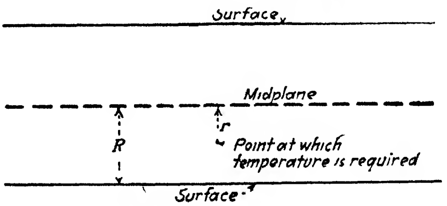

FIG. 298.—Diagram showing rate of heating symbols.

Variables.—The final temperature at any point in the material to be heated is a function of nine independent variables, in addition to the shape. These variables are:

t_o = initial uniform temperature of material (degrees Fahrenheit).
t_s = temperature of surroundings (degrees Fahrenheit).
C = specific heat.
S = density (pounds per cu. ft.).
θ = time (hours).
r = distance from center of material to the point at which the temperature is required* (in feet).
R = one-half thickness of slab or radius of sphere or cylinder* (in feet).
h_{c+r} = surface coefficient of heat transfer between the material and its surroundings (B.t.u. per hr. per sq. ft. per °F.).†
K = conductivity of the material (B.t.u. per hr. per sq. ft. per °F. per ft. thickness).

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* See Fig. 298.

† h_{c+r} is the symbol used to denote the sum of the coefficient of heat transfer by convection h_c plus the coefficient of heat transfer by radiation h_r; *i.e.* $h_{c+r} = h_c + h_r$.

Combination of Variables.—Although it is impossible to present the effect of each of the variables independently and to make the tabulation comprehensive, it is possible to combine them so as to obtain four non-dimensional groups of variables.[1] The variables are so connected that the rate of heating is affected not by a change in one variable, but by a change in certain ratios. An advantage of non-dimensional variables is that either English or French units may be used in the original variables without affecting the numerical value of the final variables. The non-dimensional variables used are Δ, m, ρ, and τ.

$$\Delta = \frac{t_s - t}{t_s - t_0}. \tag{8}$$

$$m = \frac{K}{h_{c+r}R}. \tag{9}$$

$$\rho = \frac{r}{R}. \tag{10}$$

$$\tau = \frac{\alpha^2\theta}{R^2} = \frac{K\theta}{SCR^2}. \tag{11}$$

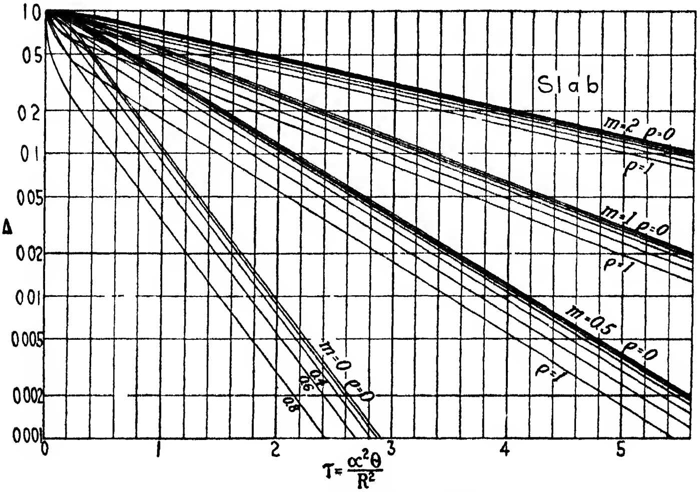

FIG. 299.—Δ versus τ for slab

The first variable, Δ, is the ratio of unaccomplished temperature change to total possible temperature change, or the temperature difference between the heating or cooling medium t_s and the temperature of the point under consideration, t, divided by the difference between the heating or cooling medium temperature t_s and the original temperature of material to be heated, t_0.

The ratio of the conductance of heat *through the material* $\frac{K}{R}$ to the conductance of heat *to the surface of the material*, h_{c+r} is expressed by the second variable m.

The third variable, ρ, is the ratio of the distance from the midplane or center line to any point at which it is desired to know the temperature to the total distance from the midplane or center line to the outer surface.

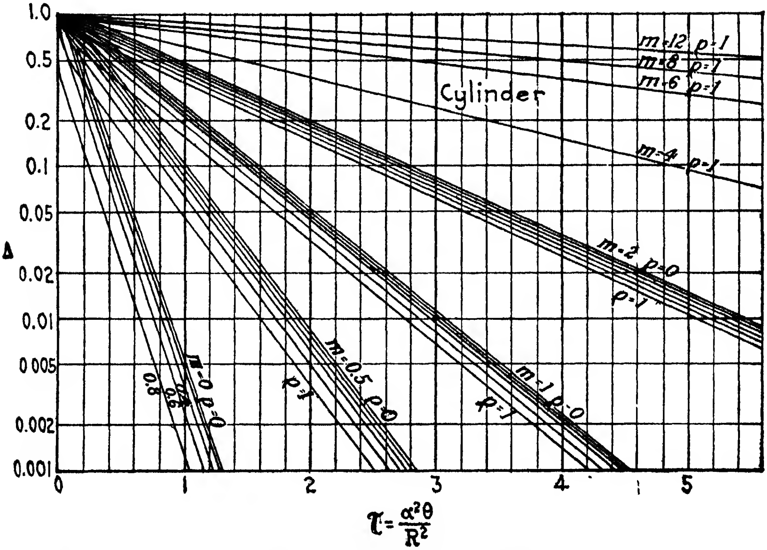

FIG. 300.—Δ versus τ for cylinders.

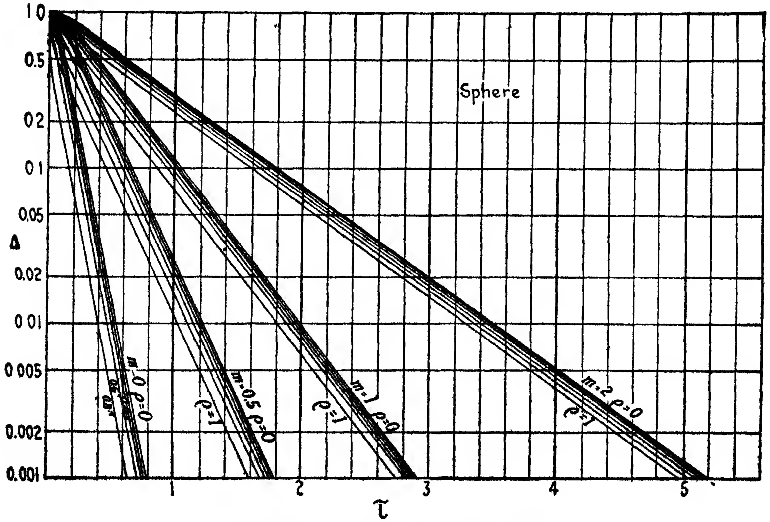

FIG. 301.—Δ versus τ for spheres.

The fourth variable, τ, or "relative time," is the product of diffusivity, α^2, and time, θ, divided by the square of the distance from surface to midplane or center line R^2 (see also p. 212).

Physical Constants.—The effect of these variables is shown* in Figs. 299, 300 and 301, where Δ is plotted against τ at various values of ρ and m for slabs, cylinders, and spheres. It will be noted that these figures are semi-logarithmic plots. Before using these plots, the physical constants for the materials to be heated must be evaluated. Of these, t_o, t_s, θ, r, and R depend on the special conditions under which the operation is carried out. Values of K, S, and C for various materials[1,2,3,4,5] and the corresponding diffusivities, α^2, are shown in Table CLXII. See also p. 212.

TABLE CLXII[1,2,3,4,5]

PROPERTIES OF MATERIALS AFFECTING RATE OF HEATING

Material	Density S in pounds per cu. ft.	Specific heat C in B.t.u. per lb. per °F.	Heat conductivity K in B.t.u. per hr. per sq. ft. per °F. per ft. thick	Heat diffusivity α^2 in sq. ft. per hr.
Silver	656	0.055	242	6.71
Copper	554	0.091	223	4.43
Aluminum	168	0.215	116	3.21
Zinc	444	0.093	65.4	1.58
Red brass	514	0.090	61.0	1.31
Yellow brass	515	0.088	49.5	1.09
Pure iron	490	0.114	39.0	0.70
Cast iron	490	0.110	35.0	0.65
Wrought iron	490	0.106	34.0	0.65
Medium steel	490	0.165	34.0	0.42
Steel (1 per cent carbon)	490	0.150	26.0	0.35
Hardened steel	490	0.140	25.0	0.36
Nickel	550	0.093	34.0	0.66
Lead	707	0.031	19.4	0.88
Monel metal	562	0.127	15.0	0.21
Graphite	140	0.160	9.0	0.40
Limestone	158	0.22	1.21	0.0349
Fire brick	107	0.22	0.34	0.0145
Water	62.4	1.0	0.33	0.053
Asbestos (paper)	26.0	0.196	0.104	0.0020
Petroleum oil	56	0.46	0.087	0.0034
Rubber	73	0.30	0.085	0.00387
Wood (kiln dried)	34.0	0.60	0.073	0.0036

Classification of Problems.—Two classifications may be made of problems in heating: first, where the material is of low diffusivity or large thickness or both, and, second, where the material is of high diffusivity or small thickness or both. In the first case, heat diffusion through the material is the

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*These figures are adapted from the calculation of H. P. GURNEY and J. LURIE.[1]

controlling factor, while in the second case heat transfer to the material is the controlling factor. Figure 302 shows diagrammatically the temperature gradients for these two cases. The line 1, 2, 5 represents the gradient in the first case and the line 1, 4, 5 represents the second case. Obviously, it is possible to have such conditions that both heat transfer to the material, and diffusion of heat through it, are equal in importance and the gradient in this case is shown by line 1, 3, 5. The heating of large pieces of materials, such as rubber, to relatively low temperatures (300°F.) falls in the first classification, while the heating of small pieces of metals to high temperatures (1500°F.) falls in the second class. The heating of large pieces of metals by gases at high temperatures may fall in the first class, but usually is between the two classes.

The criterion determining the classification of the heating process is the value of m, or the ratio $\frac{K}{h_{c+r}R}$. With high conductivity and low rate of heat transfer to the material, the transfer of heat into the interior may be considered instantaneous, while when the conditions are reversed, the surface material may be assumed to reach the furnace temperature instantaneously. In Fig. 300 curves are given for high values of m. As m increases, the difference between the temperatures at the surface ($\rho = 1$) and at the center ($\rho = 0$) becomes smaller. Above values of $m = 12$ the difference is negligible. Since K for steel is 34, then when m equals 12, and h_{c+r} is 17 or less (see p. 763), the diameter of a steel rod throughout which there will be no difference in temperature is 4 in. This value is obtained from the equation

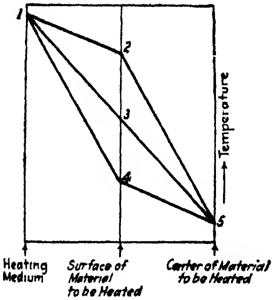

FIG. 302.—Temperature gradients through a material being heated.

$$m = \frac{K}{hR}\left(\text{or } R = \frac{K}{mh}\right).$$

The value of R under these circumstances is $\frac{34}{17 \times 12} = \frac{1}{6}$ ft., or 2 in., and the diameter ($2R$) is 4 in. In the same way the maximum value of R for a copper rod, in which the temperature will be uniform (when $h_{c+r} = 17$ or less), is $\frac{220}{17 \times 12} = 1.08$ ft., or 13 in., and the corresponding diameter is 26 in. Thus, when iron or steel rods less than 4 in. in diameter or copper rods less than 26 in. in diameter are heated in a furnace, where $h_{c+r} < 17$ the temperature throughout the metal is constant and the controlling factor is the rate of heat transfer to the material.

Value of h_{c+r}.—The variable of most importance in determining rate of heating is h_{c+r} the coefficient of heat transfer to the material. In most problems, values of K, the conductivity of the material itself, may be determined from tables (as Table CLXII) or may be readily determined experimentally with sufficient accuracy for the purpose at hand. Values of h_{c+r} are more difficult to ascertain in the majority of cases. When the conductivity is low, and diffusion of heat is the controlling factor, a large error in h_{c+r} will have no great effect on the rate of heating. It is in such problems, however, that h_{c+r} is most accurately known, since such problems are usually low-temperature conditions and the heat is transferred by convection and conduction.

Low-temperature Heating.—When heat transfer is by convection and conduction only (*i.e.*, low temperatures), the value of h_{c+r} may be approximated from equations in Appendix II and the value of m, Δ, τ, and ρ may readily be determined.

For example, a slab of rubber is to be vulcanized in a steam-heated mold by heating from both sides at 200°F. and the temperature distribution at the end of 1 hr. is required.

Data:

Thickness of slab = 6 in. ($\therefore R = \frac{1}{4}$ ft.).
$h_{c+r} = 3$; $\alpha^2 = 0.00387$; $K = 0.085$.
Initial temperature, 70°F.
Mold temperature, 200°F.

Step 1. Determine m.

$$m = \frac{K}{h_{c+r}R} = \frac{0.085}{3 \times \frac{1}{4}} = 0.11.$$

or m may be considered as 0 in Fig. 299.

Step 2. Determine τ.

$$\tau = \frac{\alpha^2\theta}{R^2} = \frac{0.00387 \times 1}{(\frac{1}{4})^2} = 0.620.$$

Step 3. From Fig 299 determine values of Δ for values of ρ between 0 and 1, when $m = 0$ and $\tau = 0.620$.

ρ	1	0.8	0.6	0.4	0.2	0.0
Δ	0	0.088	0.164	0.230	0.275	0.295

Step 4. Calculate temperature corresponding to these values of Δ.

$$\Delta = \frac{t_s - t}{t_s - t_0} = \frac{200 - t}{200 - 70}$$

$\therefore t = 200 - 130\Delta$

ρ	1.0	0.8	0.6	0.4	0.2	0.0
t	200	189	176	170	164	162

These values of t may be plotted against ρ as in Fig. 303 to show the temperature distribution throughout the slab. Since the slab was heated from both sides, the temperature at corresponding points in each half will be the same. If a 3-in. slab were heated from one side only, the distribution would be the same as in one-half of the curve shown.

Transfer of Heat at High Temperatures.—In heating metals in a furnace the heat may be transferred directly from the heating medium (or flame) to the charge, or it may be transferred to the furnace walls and then re-transferred to the charge. The heat is transferred from the gases to the walls or charge by convection and radiation, and from the walls to the charge by thermal radiation. The amount of heat transferred from a flame by radiation varies widely with the degree of luminosity. A non-luminous flame radiates only a small amount of heat (10 to 15 per cent of the total), while a very luminous flame may radiate a much larger proportion. Since heat transfer by radiation is much more rapid than transfer by convection at high temperatures, the luminosity of the flame is an important factor in determining the value of h.

Heating of Small Charge.—The simplest case of high-temperature heating is one where a small charge is heated in a large furnace. Under such conditions the heat received by the charge is small compared with the heat capacity of the furnace, and the furnace-wall temperature remains practically constant during the heating period. The amount of heat received by the charge from the flame is small and most of the heat is radiated to the charge from the furnace walls. The black-body coefficient of the charge is of major importance in determining the quantity of heat received by radiation.

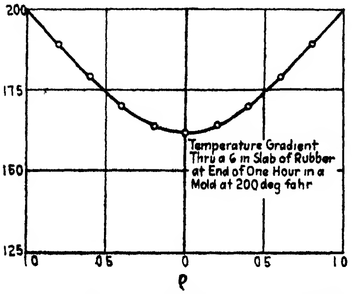

FIG. 303.—Temperature gradient through a 6-in. slab of rubber at end of one hour in a mold at 200°F.

Black-body Coefficient of Metals. The results of experimental work[6] with small charges of copper placed in furnaces of varying, but large, volume prove that under these circumstances the controlling factor is radiation of heat from the walls of the furnace to the charge. Even when oil-fired reverberatory furnaces or electric furnaces were used, the rate of heating followed the curve which would be expected from the radiation equation:

$$\frac{dQ}{d\theta} = 0.162\,pA\left(\frac{(T_w^4)}{100} - \frac{(T_p^4)}{100}\right), \tag{12}$$

where T_w represents the degrees Fahrenheit absolute of the walls and T_p represents the degrees F. absolute of the product to be heated. Under these conditions the black-body coefficient for copper was found to be 0.5. Comparative tests on iron, brass and copper in smaller furnaces indicated the same value for each of these metals, so that under these circumstances a value of 0.5 may be considered an average value for each. Other values of black-body coefficients may be found in Table CLXI.

Graphical Method.—Solution of the radiation Eq. (12) for each case of heating by radiation is difficult, and in many cases impossible. The equation may be converted* to an equation of the same type as the convection equation

$$\frac{Q}{\theta} = h_rA(T_w - T_p) \tag{13}$$

* The radiation Eq. (12) may be rewritten as

$$\frac{Q}{\theta} = pcA(T_w - T)(T_w^3 + T_w^2T_p + T_p^2 + T_p^3) \tag{14}$$

or, for any given values of T_w and T_p,

$$\frac{Q}{\theta} = h_rA(T_w - T_p),$$

where h_r includes p, c, and the term $(T_w^3 + T_w^2T_p + T_wT_p^2 + T_p^3)$.

by substituting the proper value of h_r, but the magnitude of h_r will vary with the temperature of the wall and product. The values of the coefficient h_r to be used in Eq. (13) may be calculated for any given wall and product temperatures from the term $pc(T_w^3 + T_w^2T_p + T_p^2T_w + T_p^3)$. Figure 304 shows the variation of h_r with wall and product temperatures, calculated in this way for a black-body coefficient of 0.5. The values of h_r determined from this plot may be used in conjunction with Figs. 299, 300, and 301 to determine the time necessary to heat metal charges in furnaces.

The value of h_r varies during the heating period and an average value must be used The experiments cited above (with a small charge) indicated that the true mean value corresponded to the arithmetic average within 5 per cent, and therefore the arithmetic average of the initial and final values should be used. When the black-body coefficient of the material to be heated is not 0.5 the heat-transfer coefficient h_r must be multiplied by the ratio of the actual black-body coefficient p to 0.5.

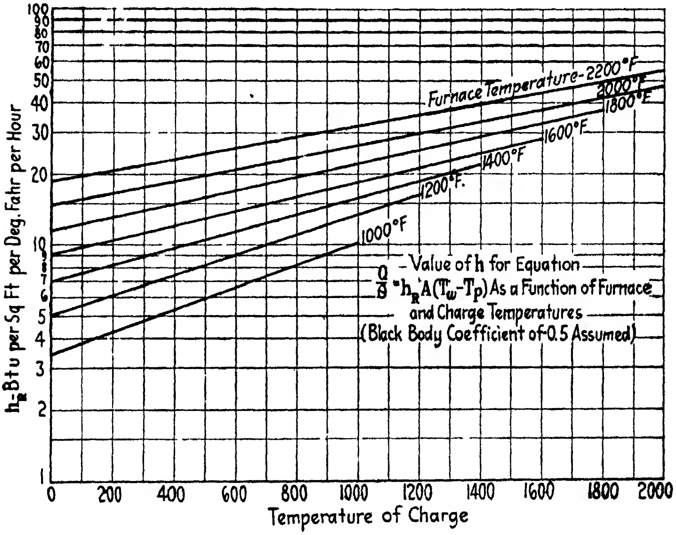

FIG. 304.—Chart for determining the rate of heating. Valus of h_r for equation $\frac{Q}{\theta} = h_rA(T_w - T_p)$.

Example.—A cylinder of medium steel 2 in. in diameter is to be heated from 100 to 1400°F. at the center in a large furnace at 1600°F. How long will it take to heat?

$$\text{Data;}\quad K = 34;\quad \alpha^2 = 0.42;\ R = 1/12 = 0.0833\ \text{ft.}$$

h_r(on 1600°F. line) at 1400°F. = 24
h_r(on 1600°F. line) at 100°F. = 9.8
Average h_r = 16.9

Since the black-body coefficient in this case is 0.5, the values of h_r are correct as read directly from the chart.

$$m = \frac{K}{h_r R} = \frac{34}{16.9 \times 0.0833} = 12$$

$$\Delta = \frac{1600 - 1400}{1600 - 100} = \frac{200}{1500} = 0.133.$$

From Fig. 300, where $\Delta = 0.133$ and $m = 12$, $\rho = 1.0$; $\tau = 15.8$

$$\tau = \frac{\alpha^2\theta}{R^2} = \frac{0.42\theta}{\left(\frac{1}{12}\right)^2} = 15.8.$$

Whence $\theta = 0.26$ hr., or 15.6 min., is the time required for heating.

Factors in Heating.—Cases of the above type are not encountered very frequently in practice, except in heat-treating or hardening furnaces where small objects are heated. In such cases the above method of solution may be used with considerable accuracy. Such furnaces have a high ratio of area of furnace walls to area of charge, and therefore are operating much below their maximum capacity, and usually at relatively low efficiency. Uniformity of heating is the main factor.

In the majority of large industrial furnaces the wall surface is utilized to its utmost and the calculation of heating rate becomes complicated by the importance of a number of other factors. Among these are: (1) the type of heating medium (temperature and luminosity of flame), and (2) the ratio of furnace-wall surface to surface of the product. A luminous flame transfers far more heat by radiation than a non-luminous flame at the same temperature. The effective ratio of wall surface to product surface is important, since when it is large most of the heat is transferred from the flame to the walls and retransferred to the charge. Obviously, the heating rate will differ widely in various furnaces.

In order to determine the rate of heating it is necessary to know the rate of heat transfer from flame to charge and to walls by radiation and convection and from walls to charge by radiation. The latter has been discussed under the previous case of small charges. The transfer of heat from the flame is more difficult to determine and accurate generalizations are impossible; the whole question of heat transfer at high temperatures is subject to errors in calculation or measurement of temperatures. An error of only 50° in flame temperature may cause an error of 10 per cent in the calculated rate of heating.

Heat Transfer from Flames.—Heat transfer from flames may be by radiation and convection. The amount transferred by either mechanism varies widely in different applications. The coefficients of heat transfer for each case may be determined, and the sum of the coefficients used in the convection equation.

Influence of Convection.—The convection coefficient h_c is approximately 3 under the conditions of most commercial operations, and may be determined from Eq. (16) of Appendix II. If the velocity of the gases is low, the value of h_c may become as low as 1.0 to 2.0. The major factors affecting the coefficient of heat transfer by convection, as shown by the equation referred to above, are the temperature and the velocity of the gases. Since heat transfer by convection is essentially a film proposition, any mechanism which tends to decrease the thickness of film will also increase the heat-transfer rate. Causing the hot gases to impinge on the surface to be heated is the most common method of doing this. In the majority of cases, however, the increase in heating rate to be gained in this manner is almost negligible for high temperature furnaces. The transfer of heat directly to the charge by convection is small and the convection coefficient of heat transfer for this reason may be considered constant under normal conditions at approximately 2 to 3 B t.u. per sq. ft. per °F. per hr.

Influence of Radiation.—Heat transfer by radiation is proportional to the difference of the fourth powers of the absolute temperatures of the bodies emitting and receiving the radiation. For this reason radiation becomes more important as the temperatures increase. Figure 304 shows the values of the coefficient in the simplified radiation equation which correspond to the convection coefficient under the same conditions. While the convection coefficient h_c remains essentially constant at 2 to 3, the corresponding radiation coefficient h_r for a *50 per cent black body* increases from 3.5 when the furnace temperature is 1000°F. and the product temperature is 0°F. to 55 when the respective temperatures are 2200 and 2000°F. For a true black body these values would be 7.0 and 110 respectively, and they show the tremendous importance of radiation at high temperatures. It is only when the furnace temperature is below 1000°F. that convection becomes of any great importance in heat transfer. Turbulence, is of course, very important from the standpoint of rapidity of combustion, which controls the rapidity with which heat is *generated*. In this manner turbulence also affects heat transfer.

Radiation from the *walls* of a furnace to a metallic charge has been shown to accord with a black-body coefficient of 0.5, and the rate of heat transferred in this manner may be calculated readily. Radiation from the *flame** to the walls and to the charge is not susceptible to the same methods. Neither the area of the flame nor its black-body coefficient is accurately known. In addition, the temperatures of charge and furnace walls are usually different, especially during the earlier stages of heating. For these reasons any method of calculating rate of heat transfer to walls and charge must, of necessity, be approximate.

Luminosity of a flame determines, to a large extent, its radiating power. Heat transfer from a strictly non-luminous flame would be chiefly by convection. All flames carry a certain amount of dust particles and thus emit considerable radiation, while luminous flames transfer much of their heat in this manner. Table CLXIII shows the effect of luminosity on the rate of transfer of heat from flames to furnace walls as given by Trinks.[7] The values given in this table are for certain conditions only, but serve to illustrate the great increase in heat transfer from luminous flames at high temperatures.

Consideration of the fact that heat must be transferred from the flame to the walls of the furnace and to the charge, and that heat may then be transferred from the walls to the charge, shows the necessity of knowing flame, furnace, and charge temperature, and coefficients of heat transfer between flame and walls, flame and charge, and walls and charge. If the furnace is large and the charge is small, the controlling factor is radia-

*See Chap. XI.

TABLE CLXIII

Effect of Luminosity on Heat Transfer from Flames

(Flame temperature 200°F. above final temperature of product)
h_{c+r} = B.t.u. per sq. ft. of product per hr. per °F.

	h_{c+r}, when temperature to which product is to be heated is		
	1200°F.	1600°F.	2000°F.
Non-luminous flame	6.0	6.5	7.5
Semi-luminous flame	8.0	11.5	17.0
Luminous flame	11.0	17.5	27.0

tion from the walls to the charge and the rate of heating may be calculated. If the charge is large, the furnace temperature will be lowered and the rate of heat transfer from the flame to the walls and charge becomes the controlling factor. Calculations of heating rate in this case become more complicated.

Wall Temperature.—Assuming no losses by conduction of heat through the furnace walls, the heat transferred to the walls at equilibrium (constant) conditions will be equal to that transferred to the charge from the walls, or

$$\frac{Q}{\theta} = h_w A_w (t_f - t_w) = h_p A_c (t_w - t_p), \tag{15}$$

where t_f is flame temperature, degrees Fahrenheit.

t_w is wall temperature, degrees Fahrenheit.

t_p is charge temperature, degrees Fahrenheit.

h_w is coefficient of heat transfer from the flame to the walls (B.t.u. per hr. per sq. ft. wall surface per °F.).

h_p is coefficient of heat transfer from walls to product (B.t.u. per hr. per sq. ft. product surface per °F.).

A_w is wall area, in square feet.

A_p is product area, in square feet.

If the area of the charge, A_p, is small compared with the wall area, A_w, and if the coefficient of heat transfer from the flame to the walls, h_w, is of the same magnitude as the coefficient of heat transfer from the walls to the charge, then the difference between the flame and wall temperatures $T_f - T_w$ is small compared with the difference between the wall and charge temperatures $T_w - T_p$, and the flame and wall temperatures are nearly the same. For example, if $\frac{A_w}{A_p} = 200$; $T_w - T_p = 800$; and $\frac{h_p}{h_w} = 4$; then, by substituting in Eq. (15), the value of $T_f - T_w$ is found to be only 16°F. That is, in a large furnace where the charge is small, the wall temperature approaches the flame temperature. On the other hand, if $\frac{A_w}{A_p} = 2$; $T_w - T_p = 400$;

and $\frac{h_p}{h_w} = 4$; then $T_f - T_w = 800°F.$ That is, when a very large charge is heated in a furnace the wall temperature may be far below the flame temperature.

Available data on the values of the coefficient of heat transfer from the flame, h_w, are meager and mostly inaccurate. Until such data are presented, values of the coefficient of blackness of the flame may be assumed as 0.1 for non-luminous flames,* 0.3 for semi-luminous flames, and 0.4 for luminous flames. Flame temperatures under operating conditions do not approach theoretical values due to incomplete mixing, excess or deficiency of air, and radiation before combustion is entirely complete. For these reasons, a flame temperature at least 15 per cent lower than the theoretical temperature (after allowing for dissociation) should be used in calculations. By making these assumptions, coefficients of heat transfer may be calculated, using Fig. 304 to determine the value of h_r for a black-body coefficient of 0.5 and multiplying the values of h_r thus obtained by the ratio of the actual black-body coefficient to 0.5. The coefficient of heat transfer from the flame to the metal may be determined in the same manner. Average values must be used to avoid tedious point-by-point determinations. For this purpose the arithmetic average is recommended because of its simplicity and because it is within the accuracy of other assumptions.

Knowing the relative rates of heat transfer from flame to wall and from wall to charge, the effective temperature of the wall may be calculated from the equation

$$ah_wA_w(t_f - t_w) = h_pA_p(t_w - t_p),$$

where a is the fraction of heat received by the walls which is transferred to the charge. This factor varies from 0.5 to 0.9, depending on the furnace, and may usually be assumed to be 0.8. The solution of the above equation involves a trial-and-error method, since both h_p and h_w vary with the wall temperature t_w. In other words, an approximate wall temperature must be assumed and the values of each side of the equation calculated. If they agree, the assumption is correct; if not, other assumptions must be made until the correct solution is found. Two substitutions will usually determine the temperature with sufficient accuracy. In the majority of cases, where the wall area is greater than 10 times the charge area, the temperature of the furnace may be assumed to be within 100°F. of the flame temperature.

Heating Rate.—The heat received by the charge is the sum of that radiated to it from the walls, and that transferred by convection and radiation from the flame. When the wall temperature has been determined as outlined above, the values of the coefficient of heat transfer by radiation from the wall to the charge h_w and from the flame to the charge h_f may be determined from Fig. 304. To the latter, the convection coefficient h_c† must be added to give the overall coefficient h_{c+r}.

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* Although the radiation equation is used to determine the amount of heat radiated from a non-luminous flame, this does not mean that the mechanism of the heat transfer is by thermal radiation, but that this equation may be *used* for determining the amount.

† Usually h_c varies from 1 to 3.

The use of these values in calculating the time necessary to heat a large charge in the furnace is difficult. If the wall and charge temperatures are nearly alike, the heating may be considered to be controlled by the rate of heat transfer from the wall. The method is the same as that used for small charges. When the wall temperature is considerably lower than the flame temperature, any method of calculating with required accuracy is tedious and usually impossible. The values of the coefficients as calculated in the above manner are indications of what the heating rates should be.

BIBLIOGRAPHY

1. H. P. Gurney and J. Lurie: *J. Ind. Eng. Chem.*, **15**, 1170 (1923).
2. "Landolt-Bornstein Roth Tabellen" (1923 ed.).
3. Walker, Lewis, and McAdams: "Principles of Chemical Engineering," McGraw-Hill Book Company, Inc. (1923).
4. Marks: "Mechanical Engineers' Handbook," McGraw-Hill Book Company, Inc. (1916).
5. Fowle: "Smithsonian Physical Tables," *Smithsonian Misc. Pub.*, **79** (No. 1), 7th rev. ed. (1920).
6. H. C. Wagner: Mass. Inst. of Tech. *Thesis*, unpublished (1922).
7. W. Trinks: "Industrial Furnaces," vol. 1, John Wiley & Sons (1923).

Thermometer Scales.—Let F and C denote the readings on the Fahrenheit and Centigrade (or Celsius) scales, respectively, for the same temperature; then

$$C = \tfrac{5}{9}(F - 32), \qquad F = \tfrac{9}{5}C + 32$$

The table below gives corresponding readings on the two scales.

CONVERSION OF THERMOMETER READINGS

Degrees Centigrade to Degrees Fahrenheit

C	F	C	F	C	F	C	F	C	F	C	F
−40	−40.0	+5	+41.0	+40	+104.0	+175	+347	+350	+662	+750	+1382
−38	−36.4	6	42.8	41	105.8	180	356	355	671	800	1472
−36	−32.8	7	44.6	42	107.6	185	365	360	680	850	1562
−34	−29.2	8	46.4	43	109.4	190	374	365	689	900	1652
−32	−25.6	9	48.2	44	111.2	195	383	370	698	950	1742
−30	−22.0	10	50.0	45	113.0	200	392	375	707	1000	1832
−28	−18.4	11	51.8	46	114.8	205	401	380	716	1050	1922
−26	−14.8	12	53.6	47	116.6	210	410	385	725	1100	2012
−24	−11.2	13	55.4	48	118.4	215	419	390	734	1150	2102
−22	− 7.6	14	57.2	49	120.2	220	428	395	743	1200	2192
−20	− 4.0	15	59.0	50	122.0	225	437	400	752	1250	2282
−19	− 2.2	16	60.8	55	131.0	230	446	405	761	1300	2372
−18	− 0.4	17	62.6	60	140.0	235	455	410	770	1350	2462
−17	+ 1.4	18	64.4	65	149.0	240	464	415	779	1400	2552
−16	3.2	19	66.2	70	158.0	245	473	420	788	1450	2642
−15	5.0	20	68.0	75	167.0	250	482	425	797	1500	2732
−14	6.8	21	69.8	80	176.0	255	491	430	806	1550	2822
−13	8.6	22	71.6	85	185.0	260	500	435	815	1600	2912
−12	10.4	23	73.4	90	194.0	265	509	440	824	1650	3002
−11	12.2	24	75.2	95	203.0	270	518	445	833	1700	3092
−10	14.0	25	77.0	100	212.0	275	527	450	842	1750	3182
− 9	15.8	26	78.8	105	221.0	280	536	455	851	1800	3272
− 8	17.6	27	80.6	110	230.0	285	545	460	860	1850	3362
− 7	19.4	28	82.4	115	239.0	290	554	465	869	1900	3452
− 6	21.2	29	84.2	120	248.0	295	563	470	878	1950	3542
− 5	23.0	30	86.0	125	257.0	300	572	475	887	2000	3632
− 4	24.8	31	87.8	130	266.0	305	581	480	896	2050	3722
− 3	26.6	32	89.6	135	275.0	310	590	485	905	2100	3812
− 2	28.4	33	91.4	140	284.0	315	599	490	914	2150	3902
− 1	30.2	34	93.2	145	293.0	320	608	495	923	2200	3992
0	32.0	35	95.0	150	302.0	325	617	500	932	2250	4082
+ 1	33.8	36	96.8	155	311.0	330	626	550	1022	2300	4172
2	35.6	37	98.6	160	320.0	335	635	600	1112	2350	4262
3	37.4	38	100.4	165	329.0	340	644	650	1202	2400	4352
4	39.2	39	102.2	170	338.0	345	653	700	1292	2450	4442

TABLE OF VALUES FOR INTERPOLATION IN THE ABOVE TABLE

Degrees Centigrade	1	2	3	4	5	6	7	8	9
Degrees Fahrenheit	1.8	3.6	5.4	7.2	9.0	10.8	12.6	14.4	16.2

Fixed Temperatures.—The following fixed temperatures, adopted by the U. S. Bureau of Standards, are useful in the calibration of pyrometers:

	DEG. FAHR.	DEG CENT.
Liquid tin solidifies at	449	232
Liquid lead solidifies at	621	327
Liquid zinc solidifies at	787	419.4
Liquid sulfur boils at	832.5	444.7
Liquid antimony solidifies at	1167	630.5
Liquid aluminum (97.7 per cent pure) solidifies at	1216	658
Solid gold melts at	1947	1064
Liquid copper solidifies at	1983	1084
Solid nickel melts at	2615	1435
Solid palladium melts at	2815	1546
Solid platinum melts at	3187	1753

High Temperatures and Color.—High temperatures may be judged approximately by color, though the estimate of an experienced observer is likely to be 100°Fahrenheit from the true value. The following table associating color and temperature of iron or steel is due to White and Taylor.

	DEG. FAHR.		DEG. FAHR.
Dark blood red, black red	990	Orange, free scaling heat	1650
Dark red, blood red, low red	1050	Light orange	1725
Dark cherry red	1175	Yellow	1825
Medium cherry red	1250	Light yellow	1975
Cherry, full red	1375	White	2200
Light cherry, light red	1550		

CONVERSION FACTORS

MULTIPLY	BY	TO OBTAIN
B.t.u. per lb.	0.55556	calories per kg.
B.t.u. per cu. ft.	0.0089	calories per l.
B.t.u. per cu. ft.	8.899	calories per cu. m.
Calories per kg.	1.8	B.t.u. per lb.
Calories per l.	112.37	B.t.u. per cu. ft.
Calories per cu. m.	0.11237	B.t.u. per cu. ft.

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