gas, as well as the chemical and physical nature of the catalyst. The principal primary reaction of the synthesis on cobalt catalysts may be represented by the equation,

$$2H_2 + CO \longrightarrow 1/n\ (CH_2)_n + H_2O,$$

the usage ratio of hydrogen to carbon monoxide being about 2:1 over a wide range of gas compositions. Previous work demonstrated that bulk-phase cobalt carbide was not formed in the synthesis and that its presence in large amounts seriously decreased the catalytic activity.

A study of the differential reaction rates involved in Fischer-Tropsch synthesis with a cobalt catalyst, was undertaken,[13] in which the synthesis rate and the rates of formation of methane and carbon dioxide were investigated as a function of bed length. Cobalt-thoria-kieselguhr catalyst was operated at atmospheric pressure, with hydrogen to carbon monoxide ratios of 3.5:1, 2:1, and 0.9:1, using argon as an inert reference component. The synthesis rate was found to be high in the initial part of the bed and lower and nearly constant throughout a large portion of the catalyst bed, until the reactants were fairly completely consumed. The data suggest that the rate-controlling step is some process at the catalyst surface and not the rate of adsorption of reactants, which would be strongly dependent upon the partial pressures of the reactants, a characteristic which was not substantiated by the experimental data. A mechanism in which desorption of products is the rate-determining step predicts, qualitatively, many of the characteristics of the synthesis. The reaction is very complicated, because the catalyst is heavily covered with the hydrocarbon products, and differences in the nature of the catalyst surface may be a function of the composition of the gas in contact with catalyst. Over the range of relatively constant rate of synthesis, the gas composition varied considerably.

The ratio of saturated to unsaturated hydrocarbons present at any point in the catalyst bed varied with the ratio of hydrogen to carbon monoxide in the gas at that part of the bed. This indicates that olefins may be a primary product of the reaction, and that these olefins may then be hydrogenated or undergo further reaction to form hydrocarbons of higher molecular weights at the point of formation or in later parts of the catalyst bed.

The processes that form methane appear to be functions of the partial pressures of hydrogen and hydrocarbons, and processes forming carbon dioxide appear to be functions of the partial pressures of water vapor and carbon monoxide. The formation of both methane and carbon dioxide increased with increasing temperature. These reactions occur, at least, to some extent over the entire catalyst bed.

Effect of Reduction on Metal Oxide Catalysts

Changes that occur during reduction may be divided into three classes: (1) Sintering of the unreduced material may occur at the reduction temperature, the "glowing" of oxide gels being an extreme example of this phenomenon; (2) the transition of oxide to metal may be accomplished by a sizeable decrease in volume; and (3) reorientation of the reduced metal into a stable form takes place. If considerable reorientation of the freshly reduced metal does not occur simultaneously with the reduction, the change in volume should make the metal-to-oxide interface accessible

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[13] Anderson, R. B., Krieg, A., Friedel, R. A., and Mason, L. S., Fischer-Tropsch Synthesis - Differential Reaction Rate Studies with Cobalt Catalyst: Ind. Eng. Chem., vol. 41, October 1949, pp. 2189-2197.

to the reducing gas. A study of these changes in the physical structure of precipitated cobalt and iron oxide catalysts on reduction was completed with the electron microscope.[14] To obtain further information about changes occurring during reduction, a number of metal oxides were also studied. A secondary purpose of this research was to determine the usefulness of the electron microscope in studies of granular catalysts.

Consideration of the electron-microscope studies of catalysts indicates that these are useful and informative but that the micrographs are seldom simple to interpret. It is desirable to consider the micrographs, with surface-area results from gas-adsorption studies. From a comparison of particle size and shape with surface area, some estimate of the porosity of the particle can be made. Some precipitated metal oxides have gel structures, with fairly high mechanical strength and high surface areas. Some gels have been postulated to be "felts" of tiny filaments. The electron micrographs support this postulate. Fused-metal oxide catalysts represent an opposite extreme to metal oxide gels. Fused catalysts, such as those used in the ammonia synthesis, are dense solids, with very little internal surface area. On reduction, these catalysts form a porous structure, but their external (unreduced) volumes are not changed. The surface areas of the precipitated oxides and catalysts studied always decreased on reduction.

Separation and Identification of Products of Hydrogenation of Carbon Monoxide

The composition of synthetic liquid fuels obtained from the Fischer-Tropsch process is of interest from the viewpoint of reaction mechanism and in determining the applicability of the products as fuels. Except for general product-distribution studies, little quantitative work on this problem had been reported. A study of the products obtained from laboratory-scale catalyst testing units was therefor undertaken.[15] In the case of iron catalysts, where considerable carbon dioxide is found in the exit gas, studies such as these were facilitated by the development of an improved caustic scrubber.[16]

The product distribution from precipitated cobalt catalysts was obtained by fractional distillation of the liquids and mass spectrometer analysis of the liquids and gases. In connection with the evaluation of distillation columns operating at reduced pressures, the use of a test mixture of a hydrocarbon whose hydrogen atoms are of mass one and the same hydrocarbon containing some deuterium atoms (hydrogen atoms of mass two) is highly desirable. Accordingly, a method of analysis for hydrocarbons containing deuterium was developed,[17][18] which is superior to the analytical

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[14] McCartney, J. T., Seligman, B., Hall, W. K., and Anderson, R. B., An Electron-Microscopic Study of Metal Oxides and Metal Oxide Catalysts: Jour. Phys. Colloid Chem., vol. 54, April 1950, pp. 505-519.

[15] Friedel, R. A., and Anderson, R. B., Composition of Synthetic Liquid Fuels. I. Product Distribution and Analysis of C_5-C_8 Paraffin Isomers from Cobalt Catalyst: Jour. Am. Chem. Soc., vol. 72, March 1950, pp. 1212-1215; correction, May 1950, p. 2307.

[16] Krieg, A., Dudash, A. P., and Anderson, R. B., Improved Caustic Scrubber: Ind. Eng. Chem., vol. 41, July 1949, p. 1508.

[17] Orchin, M., Wender, I., and Friedel, R. A., Determination of Deuterium in Water - Conversion to Methyl Deuteride and Methane and Measurement by Mass Spectrometer: Anal. Chem., vol. 21, September 1949, pp. 1072-1073.

[18] Friedel, R. A., and Sharkey, A. G., Mass Spectrum of Hydrogen Deuteride (HD): Jour. Chem. Phys., vol. 17, June 1949, pp. 584-585.

methods previously used. In a similar manner, the development of a simple method for introducing liquids into a mass spectrometer and the study of negative peaks obtained in mass spectra were extremely valuable.19/20/

The results indicated a high yield of methane, followed by a minimum at C_2, a broad maximum from C_5 to C_{11}, and a gradual decrease to C_{20}. An analysis of paraffin isomers was made in the C_5-C_8 range. Of particular interest was the slight, but progressive, decrease of straight-chain isomers, the major components, with molecular weight. Calculated values for ethyl- and dimethyl-derivatives were very close to zero in all cases, being either slightly positive or negative. This is considered valid evidence for the absence of appreciable concentrations, but the difficulties of carrying out separations on a small scale did not allow precluding the presence of traces. No indication of any compound containing a quaternary carbon atom has been found. These experimental results on isomer distribution of pentane to octane were also deduced from probability considerations.21/22/

Study by means of the infrared spectrometer of the C_6 to C_8 olefins in the product showed the predominance (64-82 percent) of olefins, with the double bond in positions other than the alpha (1,2) position, and that the proportion of these olefins in the product increased with its molecular weight. The product from iron catalysts, on the other hand, contained as high as 90 percent alpha olefins in the C_6 range. Analyses of Diesel oil and wax fractions from iron catalysts in the internally cooled converter have shown, however, that olefins other than alpha predominate in these fractions by about 7 to 1. Since olefins other than alpha are favored thermodynamically, the approach toward equilibrium due to longer residence in the converter, which results from the current practice of recycling the heavy products, would increase the concentration of these olefins at the expense of the alpha olefins.

Homologation of Alcohols

Because of the presence of relatively large amounts of alcohols in the products from the gas synthesis, a study is in progress on possible mechanisms for their formation. The conversion of an alcohol into the primary alcohol containing one carbon atom more than the original is usually a tedious procedure. Recently, a simple one-step homologation reaction was developed,23/ which consists of treating the alcohol with synthesis gas in the presence of a cobalt catalyst under conditions resembling those employed in the Oxo reaction.

The Oxo or hydroformylation reaction employs the conversion, by means of synthesis gas, of an olefin to a mixture of aldehydes containing one carbon atom more than the starting material. The aldehydes are then, customarily, reduced in a separate step, or they may be converted directly to the alcohol by operating the usual Oxo reaction

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19/ Friedel, R. A., Sharkey, A. G., Jr., and Humbert, C. R., Liquid Sampling for Analysis by Mass Spectrometer: Anal. Chem., vol. 21, December 1949, pp. 1572-1573.

20/ Sharkey, A. G., and Friedel, R. A., Anomalous Negative Peaks in Mass Spectra: Jour. Chem. Phys., vol. 17, October 1949, p. 998.

21/ Weller, S., and Friedel, R. A., Isomer Distribution in Hydrocarbons from the Fischer-Tropsch Process: Jour. Chem. Phys., vol. 17, September 1949, pp. 801-803.

22/ Weller, S., and Friedel, R. A., Isomer Distribution in Fischer-Tropsch Hydrocarbons: Jour. Chem. Phys., vol. 18, January 1950, pp. 157-158.

23/ Wender, I., Levine, R., and Orchin, M., Homologation of Alcohols: Jour. Am. Chem. Soc., vol. 71, December 1949, pp. 4160-4161.

at slightly higher temperatures. The homologation reaction, however, need not proceed via an olefin intermediate followed by hydroformylation-hydrogenation, as shown by the fact that β-phenylethyl alcohol is obtained from benzyl alcohol according to the reaction:

$$PhCH_2OH + 2H_2 + CO \longrightarrow PhCH_2CH_2OH.$$

Similarly, isobutyl alcohol was obtained from isopropyl alcohol and isoamyl alcohol from tertiary butyl alcohol.

Study of Fundamental Properties of Fixed and Fluidized Beds

The pressure drop in gases and liquids flowing through beds of granular material is an important factor in the engineering of absorption towers, catalytic reactors, and filters. Recent investigations into the fundamentals of fluidization of solids have shown the intimate relation between pressure drop and fluidization phenomena. It has been established that, if upward flow of a fluid through a bed of granular material is gradually increased, the solids will begin to fluidize when the pressure drop across the bed equals the buoyant weight of the solids. In order to apply this relationship more conveniently to practical examples, a nomogram was developed which facilitates a comparison of pressure drop with fluid velocity, fluid viscosity, size and shape of packing particles, and void space in the bed.[24]

Another problem of fundamental importance is that involving heat flow. Although considerable work has been reported on various types of fixed and moving beds and on a system characterized by clouds of fine particles, little information was available on heat transmission through fluidized beds. A study of this problem[25] resulted in the development of three correlations applicable to particles in the range 0.0015 to 0.0045 inch. With air, carbon dioxide, or helium under various operating conditions, coefficients as low as 0.25 and as high as 8 B.t.u./(hour) (°F.) (square feet) were observed. Within the limits of the experimental data, it was shown that the bed height, the concentration of solids in the bed (state of expansion), the density of the solid phase, and the vessel diameter had no effect upon heat transfer. Effect of slugging upon heat transfer was found to be negligible. Channeling, however, had a very pronounced effect, which was reflected by the erratic temperature distribution through the bed. Absolute particle size and shape are considerably more important than size distribution in inducing channeling.

Pilot-Plant Synthesis of Hydrocarbons from Carbon Monoxide and Hydrogen

Work was continued in pilot-plant units using the internally cooled, moving-bed converter design. In this system, the catalyst is cooled by a flow of circulating oil, which also lifts the catalyst particles and keeps them suspended in a stream of liquid and gas. In most of the pilot plant runs a fused-synthetic-ammonia-type catalyst was reduced with hydrogen at about 450° C. before use in the synthesis. The gases present in the converter were found to be increasingly soluble in the circulating oil in the following order: Hydrogen, carbon monoxide, and carbon dioxide. A conversion about 65 percent as great was obtained by introducing the synthesis gas into the free space at the top of the converter as was obtained by bubbling the gas

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[24] Weintraub, M., and Leva, M., Pressure Drop Through Fixed and Fluidized Beds - A Nomogram: Chem. Eng., vol. 47, January 1950, pp. 110-113.

[25] Leva, M., Weintraub, M., and Grummer, M., Heat Transmission Through Fluidized Beds of Fine Particles: Chem. Eng. Progress, vol. 45, September 1949, pp. 563-572.

up through the converter in the conventional method. This experiment shows that diffusion of the reactants to the catalyst surface is not the rate-controlling step in the process.

Current pilot-plant operations are being directed toward determining the conditions for minimum production of gaseous hydrocarbons. A comparison of Alumina and Cyclocel bauxite catalysts, as a means of removing oxygenated compounds from the overhead vapors, showed both materials to be effective in this respect. Cyclocel bauxite, however, produced appreciable shifting of the double bond, which was not observed with Alumina bauxite.

A stripper column was operated in a manner to simulate actual operation in series with the internally cooled converter. In this column, the dissolved product water was separated from the circulating oil by countercurrent flow of the oil and the feed and recycle gases. As a result, the H_2:CO usage ratio was increased, because the water-gas shift reaction in the converter was depressed.

By scrubbing the carbon dioxide from the recycle gas with potassium carbonate solution during gas-recycle operation, 90-percent conversion of synthesis gas was possible in a single stage by increasing the normal reaction temperature about 10° (to 260° C.). Without removal of carbon dioxide from the recycle gas, not more than 85 percent synthesis-gas conversion could be obtained in a single stage, even at temperatures up to 295° C. and at higher gas-recycle rates.

SYNTHETIC LIQUID FUELS DEMONSTRATION PLANTS

The Coal-Hydrogenation Demonstration Plant went through the necessary break-in procedures. During the year the liquid-phase unit made an exploratory tar oil run of 2 months' duration and three coal runs, during which 150,000 gallons of tar oil was hydrogenated and 1,300 tons of bituminous coal was liquefied.

The Gas-Synthesis Demonstration Plant construction is practically complete. The 1-ton-per-day capacity oxygen-production unit is operating on routine basis. The atmospheric pulverized-coal gasifier made numerous short exploratory runs and an extended run of 2 weeks' duration. The other portions of the plant are ready for trouble shooting and break-in runs.

The emphasis of the engineering work is definitely shifting from the design and construction stages to operational problems, cost estimates, and commercial evaluation of the different processes.

Coal-Hydrogenation Plant

Two papers[26/][27/] described both the process and special features of construction of the demonstration plant.

A successful run was made in the vapor-phase section of the plant.[28/] Light petroleum distillate was hydrogenated to motor gasoline, and a distillate obtained

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[26/] Markovits, J. A., The Coal-Hydrogenation Demonstration Plant at Louisiana, Mo.: Trans. Am. Soc. Mech. Eng., vol. 72, May 1950, pp. 349-355 (Paper 49 - PET-3).

[27/] Hirst, L. L., Complete Coal-to-Oil Demonstration Plants: Ind. and Power, vol. 57, August 1949, pp. 90-93, 116.

[28/] Clarke, E. A., Chaffee, C. C., and Hirst, L. L., Hydrogenation of Petroleum and Lignite Tar Distillates: Bureau of Mines Rept. of Investigations 4676, 1950, 27 pp.

from the tar oil, produced by low-temperature carbonization of North Dakota lignite, was converted to a high cetane number Diesel fuel oil and motor gasoline.

Following a low-pressure (1,600-pounds-per-square-inch) "dummy" run with nitrogen to give the operators experience and to make a break-in test of the equipment, the system was pressured to 10,300-pounds-per-square-inch gage with hydrogen. The temperature was raised slowly to 835° F. and held there to activate the catalyst in the vapor-phase converter. After activation of the catalyst, the temperature was reduced, and a vapor-phase hydrogenation break-in run was made using a 43° A.P.I. gasoline-naphtha-gas-oil blend prepared from an Oklahoma City crude petroleum oil. The charge during the petroleum-hydrogenation run was made up of a blend of 11 percent of 67° A.P.I. straight-run gasoline, 31 percent of 47° A.P.I. naphtha, and 58 percent of 39° A.P.I. gas oil. A total of 29,360 gallons of this blend, with 2,960 M cubic feet of 95 percent hydrogen make-up gas was charged during the 4-day run. Although operating conditions were varied to gain information concerning the operation of controls and the product, the average run conditions were as follows:

Pressure	P.s.i.g.
Converter inlet	9,750
Hydrogen feed	9,975
Recycle compressor suction	9,715
Recycle discharge	10,000

Temperature	°F.
Preheater inlet	395
Preheater outlet	721
Catalyst beds	699
Cold catchpot	92

Hydrogenation at temperatures above 720° F. produced a light product higher than 70° A.P.I. The distillation unit was put on stream to fractionate the hydrogenated product into its components.

The products and yields obtained were:

	°A.P.I.	Gallons	M c.f.
Gasoline	66	15,725	-
Wash oil (naphtha)..	48	5,540	-
Bottoms	43	4,250	-
Off gases	-	-	975

Distillation curves showed that the quantity distilling over 400° F. was increased from 25 percent in the feed to 65 percent in the hydrogenated product, in one pass through the converter. The gasoline cut had an octane number of 63 by the A.S.T.M. D357-48 C.F.R. motor method.

The vapor-phase hydrogenation feed stock then was changed to a lignite-tar distillate oil in order to produce a specification Diesel fuel. This stock was prepared from a 10° A.P.I. North Dakota lignite-tar oil by removing the low-boiling portion and the asphaltic material unsuitable for hydrogenation over the Welheim K-536 catalyst. The portion used represented 70 percent of the raw oil and had a gravity of 12° A.P.I.

The lignite-tar distillate was charged to the unit without any purging of petroleum oil. The total charge, including the petroleum in the system was made up as follows:

Lignite-tar distillate	gallons	12,225
Petroleum oils in system	do.	5,480
Hydrogen make-up gas	M c.f.	2,930

Although temperatures were varied during the run to obtain the optimum reduction of tar acids, the average operating conditions were as follows:

Pressure	P.s.i.g.
Converter inlet	9,725
Hydrogen feed	10,550
Recycle compressor suction	9,690
Recycle compressor discharge	10,100

Temperature	°F.
Preheater inlet	390
Preheater outlet	730
Catalyst beds	727
Cold catchpot	91

The products and yields obtained were:

Gasoline	gallons	8,835
Wash oil (naphtha)	do.	615
Diesel fuel oil	do.	4,520
Slop oil	do.	520
Off gases	M c.f.	955

Distillation curves showed that the quantity distilling over 400° F. was increased in one pass by 25 percent as compared to 40 percent for the petroleum. The gasoline had an octane number of 65 by the motor method. This gasoline has given satisfactory service at this station in all types of motor vehicles formerly using regular motor-grade gasoline.

The bottoms from the product distillation were finished for Diesel fuel oil. Table 16 is a comparison of the properties of the Diesel fuel oil produced with the specifications used by one of the large western railroads for the purchase of Diesel fuel from petroleum:

TABLE 16. - Comparison of Diesel fuel oil produced by coal hydrogenation with railroad specifications

		Louisiana test run	Railroad specifications
Gravity	A.P.I.	36.5	32 - 40
Boiling range:			
10 percent	°F.	440	425 - 500
90 percent	°F.	578	540 - 620
End point	°F.	644	675 max.
Recovery	percent	96	98
Viscosity at 100° F.	S.S.U.	35.5	35 - 45
Flash point (PM)	°F.	194	150 min.
Pour point	°F.	10	5
Ash	percent	0.002	0.1 max.
Cetane number		56	50 min.
Carbon residue (10 percent bottoms)	percent	0.06	0.25 max.

This fuel was used successfully on a 200-mile trial run in a Diesel-electric locomotive hauling a loaded eight-car passenger train.

Four runs have been made on the liquid-phase unit during its first year of operation. The first was an exploratory tar-oil run to test the performance of the equipment and instrumentation. The unit was in continuous operation from October 12 to December 6. During this run, the general reliability of the equipment and its tightness were ascertained. The operators obtained further experience while continuous efforts were being made to have the instruments, particularly control valves, work satisfactorily. The first hydrogenation products appeared at 750° F. The operation was fairly smooth, hydrogenating tar oil at a temperature of 850° F. in the converters, with a 50-50 split between light and heavy oils.

The first coal paste, containing 10 percent of Wyoming Rock Springs coal, was introduced on November 24. The coal content was raised to 25 percent within 20 hours. Operating conditions of pressure, temperature, and flow rates were maintained essentially at the same levels as during the tar-oil operation. The coal-paste phase of the run was continued for 8 days and was terminated when the supply of coal previously obtained to break in the coal-preparation plant was exhausted. The unit was cleaned and inspected, defective instrument tubing was replaced, and improvements in instrumentation were made.

The second liquid-phase hydrogenation run was started on March 31. A light start-up oil blend was used to test the controls and bring the unit up to operating conditions. Coal-tar hydrogenation was started on April 8 and continued until April 12. Operating conditions were fairly smooth.

Coal addition was started to the system on April 12, and the charging rate was rapidly increased to 3,000 pounds per hour, using a paste feed rate of 19 gallons per minute, with gas circulating through the furnace at 840,000 cubic feet per hour. Converter temperatures were held at 840° to 850° F. Shortly after the introduction of coal, instrument failures caused several severe upsets in operations. These conditions were further aggravated by irregular paste injection due to valve troubles. The run was terminated on April 16 because of leaks in the transfer lines between the converters. Inspection of the pumps revealed that all but four of the large injection pumps, as well as several of the smaller pumps, had developed irreparable liquid-end block leaks.

In spite of the pump situation and in view of the delay in obtaining new blocks, it was decided to make a run at 7,500 pounds per square inch to make products, to allow working out more of the mechanical problems, especially those concerned with instruments, and to improve general process control required for smooth operations. After lining out the unit on pasting oil, coal was introduced on May 16. This operation was smooth until converter temperature measurements could not be relied upon. A normal shut-down was made on May 23. Inspection of the equipment showed it to be clean, and for the first time no coke was found in the converters.

For the fourth run, coal-paste injection was started on June 8th and continued for 3 weeks. The operation was the steadiest attained on any run. System pressures, temperatures, rates of flow of liquids and gases, solids removal, uniformity of paste composition, and quality of pasting oil showed a marked improvement, making it possible to charge coal at rates as high as 50 tons per day.

The unit was shut down per schedule on June 27. After cleaning, inspection, and the installation of redesigned fluid-end blocks on the injection pumps, operation will be resumed.

Ammonium molybdate was used as a catalyst when charging coal tar, and stannous oxalate when charging coal.

A total of 140,000 gallons of coal-tar oil and 1,865 tons of coal has been processed during these runs. One hundred sixty-eight thousand gallons of charging stock suitable for further hydrogenation in the vapor phase to make specification motor gasoline have been produced. In addition, there are 90,000 gallons of heavy oils produced during the operations which will require further processing.

A full report of these runs is in preparation and will be published as soon as completed.

Two short runs were made on the steam flash distillation unit for recovering oil from the solids. These operations have shown that:

1. Taking the feed directly from the hot catchpot is not feasible on our unit: (a) It interferes with the operation of the level controls; (b) it is impossible to obtain a steady, measured flow; and (c) the oil-solids mixture is cooled by unavoidable line losses to too low a temperature for flashing.

2. Condensed oil droplets are carried out in the steam exhaust, causing serious hazards.

While these difficulties render the equipment inoperable at present, fortunately, they can be corrected by the installation of a direct-fired heater and a total condenser, which are on order.

Gas-Synthesis Plant

The Gas-Synthesis Demonstration Plant is designed to receive coal by the carload at its gasification unit, to crush and pulverize the coal, and to react it with oxygen and steam to make a synthesis gas. This gas is then to be compressed and purified of sulfur compounds and utilized according to the Fischer-Tropsch synthesis process to make synthetic liquid fuel products, such as gasoline and Diesel oil. The plant capacity is approximately 3,000,000 cubic feet of gas per day, which will be converted into 80 barrels per day of total finished products.

At the end of the fiscal year 1950 construction of this plant is substantially complete. Three of the plant units, namely, the oxygen production and powdered-coal gasification units and the Kerpely gas producer, have been in operation for periods of 2 to 18 months. Two views of the plant are given in figures 30 and 31.

Oxygen Production

The oxygen production unit started initial operations in December 1948 and has been operating since. It produces 24 tons per day of oxygen for use in gasifying coal. Over periods of several months it has been kept in operation 95 percent of the time while producing oxygen of 98-percent purity.

Koppers Powdered-Coal Gasifier

One of the unique units in the demonstration plant is the Koppers-type powdered-coal gasifier. This unit is designed to convert 1 ton of coal per hour to synthesis gas by the use of approximately equivalent quantities of oxygen and steam. The first run at full rates of oxygen, steam, and powdered coal was made in June 1949. Since then 46 runs were made, each run normally of 6 to 10 hours' duration, including 1 run

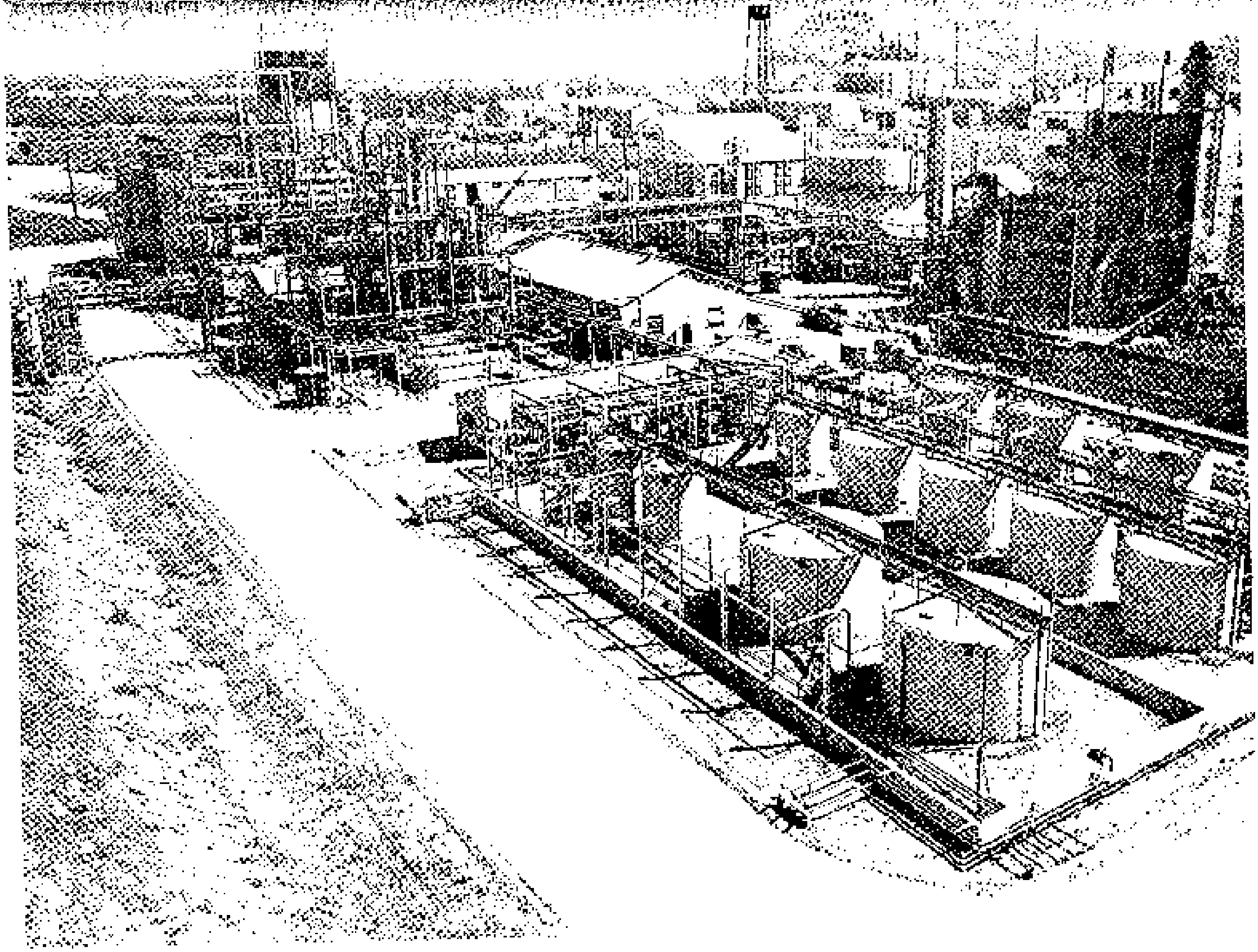

Figure 30. - Gas-Synthesis Demonstration Plant, Louisiana, Mo.

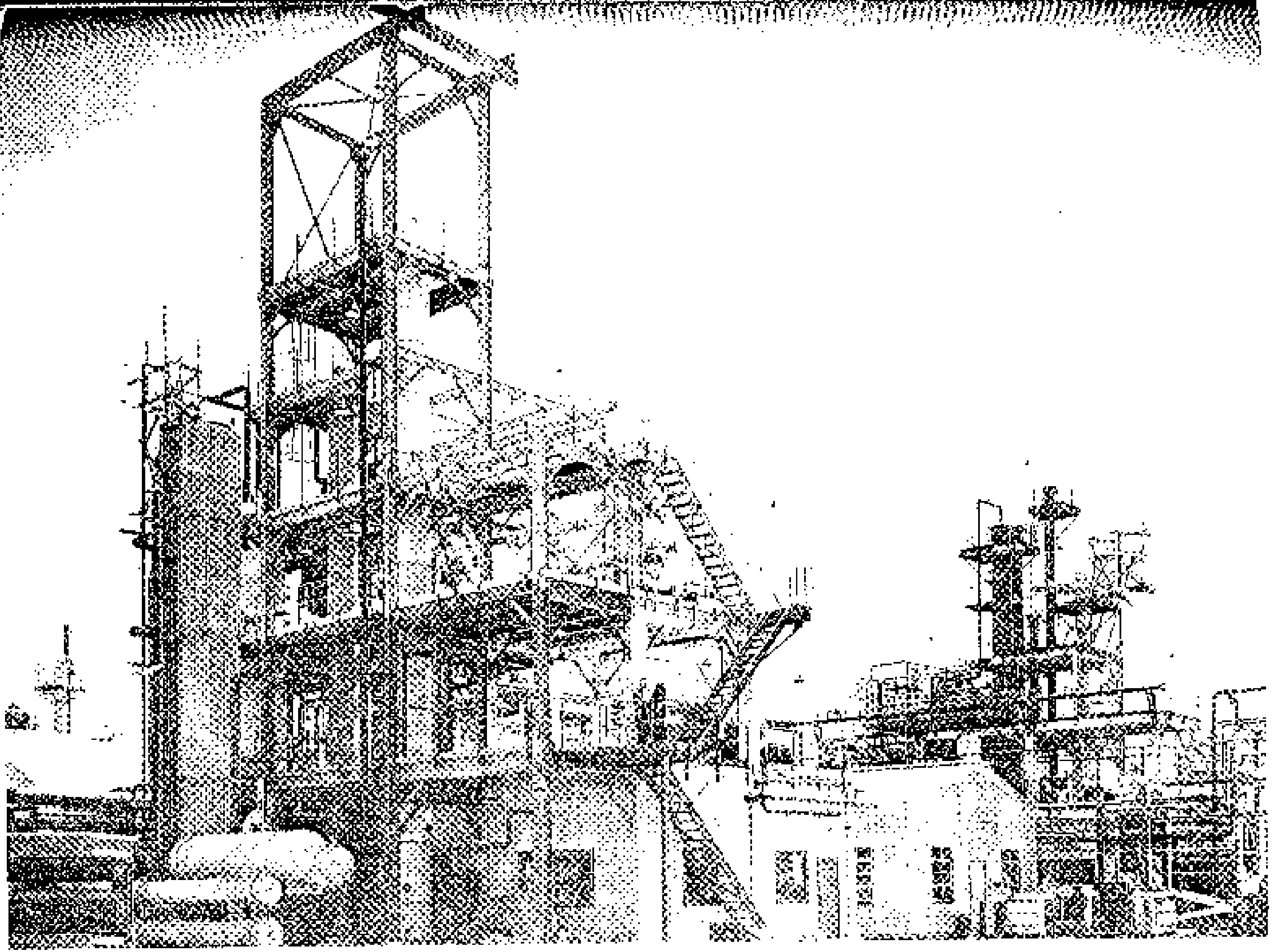

Figure 31. - Gas-Synthesis Demonstration Plant, Louisiana, Mo.