

The difference in concentration of hydrogen sulfide in the rich amine is also reflected in the analysis of acid gas from the reactivator:

| H <sub>2</sub> S<br>per 100 cu.ft.<br>feed gas, grains | H <sub>2</sub> S per gal.<br>rich amine, grains | Acid-gas analysis, percent |                               |                |                  |
|--|---|----------------------------|-------------------------------|----------------|------------------|
|  |   | CO <sub>2</sub>            | CH <sub>4</sub> <sup>1/</sup> | N <sub>2</sub> | H <sub>2</sub> S |
| 600 (for this run)                                     | 196   | 86.6                       | 6.0                           | 2.3            | 5.1              |
| 75 (for run previously reported)                       | 52.3  | 85.0                       | 10.0                          | 4.5            | 0.5              |

<sup>1/</sup> Natural gas is fed into the reactivator reboiler - about 1,000 cu. ft. per hr. to inhibit foaming.

From the daily operating summaries given in table 20, an analysis has been made of the performance of the Girbotol for this operating period. This follows the method set forth in a previous run report (October-November 1951). For this analysis the amine flow divided by the carbon dioxide fed to the Girbotol is plotted against the percent of carbon dioxide absorbed. The curve derived in the earlier report (October-November 1951) and points from performance of this run are shown in figure 35.

Typical flows in the Girbotol system during the run are shown in figure 36.

As in previous runs, the settling tank was in service for continuous decantation of the amine solution.

The excellent scrubbing performance of the absorber tower must be attributed to the cleaning of the Raschig rings and the tower sprays in preparation for the run and for the higher liquid-to-gas ratio at which the absorber was operated. In previous runs, the ratio of amine circulation to feed gas rate usually was about 32 compared with 49 in this run.

## 2. Iron Oxide Towers

These 2 towers were operated in series with gas from the Girbotol unit entering at 105° F. and 340 p.s.i.g., saturated. The gas, which entered the first oxide tower with an average hydrogen sulfide content of 2.5 grains per 100 cu. ft., was free of this impurity at the outlet of the second tower.

## 3. Activated-Carbon Towers

Carbon tower operation in this run period was routine, a tower being steamed every 2 days. The newly steamed tower was put in second position on the line. At all times, even immediately after a newly steamed tower was put back on the line, the total sulfur content at the outlet of the carbon towers was never more than a trace.

## 4. Kinetics of Coal Gasification

The studies on the kinetics of coal gasification have been continued. Twenty separate cases, covering the normal range of operating conditions at atmospheric pressure, were calculated. The calculations have been embodied in a series of multi-parameter matrix charts. Given the coal-feed rate per unit volume, the steam-coal and oxygen-coal ratios, the charts can be used to determine contact time, carbon conversion, synthesis gas yield, oxygen requirement, and product gas temperature and enthalpy.

A preliminary study was made to determine the possibility of using the calculation method derived for reaction at atmospheric pressure for calculation of

results at elevated pressures. The method may be used if certain pressure-sensitive terms are included in the calculation. These studies were embodied in two papers presented before the Division of Gas and Fuel Chemistry of the American Chemical Society in September 1953.<sup>4/</sup>

### Fischer-Tropsch Synthesis

The fourth operating period of the Gas-Synthesis Demonstration Plant began on November 29, 1952, with the starting of the oxygen unit, and continued until January 24, 1953, when all units were shut down. The shutdown was caused by difficulties with the make-gas exhauster. Repairs could have been made and operations resumed, but most of the run objectives had already been achieved.

Operation of the oxygen unit and Kerpely gas producer was routine. A gas containing approximately equal quantities of hydrogen and carbon monoxide was desired for synthesis. Accordingly, a higher steam ratio was used than in previous operations, resulting in a gas containing more carbon dioxide and requiring slightly more oxygen and coke. Almost 100 million cu. ft. of gas was generated.

Difficulties with the purification unit were uncommon. Fouling of the Girbotol absorber spray head and packing resulted in less efficient removal of carbon dioxide than in previous runs, but the content was still low enough to permit excellent sulfur removal and satisfactory synthesis. The absorption system had been cleaned. Removal of sulfur compounds was more uniform and more nearly complete than in preceding operations. About 75 million cu. ft. of purified gas was produced; and, except on rare occasions, the total sulfur content of this gas was less than 0.01 grain per 100 cu. ft.

The synthesis unit was operated for 39 days; data were procured during two periods of high production. At 480 hourly space velocity conversion was 84 percent; at 370 space velocity it was 92 percent. The catalytic activity increased as catalyst decomposed, and finely divided catalyst became suspended in the coolant oil. At a catalyst concentration corresponding to 12 percent ash in the coolant oil, the activity reached a maximum, and further increase in catalyst content had no beneficial effect. Erosion of the circulating pump for cooling oil was severe during the last 10 days of the run, when the concentration of catalyst in the oil was high.

The carbon dioxide scrubber was used throughout the run for substantially complete removal of carbon dioxide from the recycle gas without regard to the efficiency of the scrubbing operation. At no time did the concentration of carbon dioxide in the effluent gas exceed 0.2 percent.

Product was recovered in a conventional oil-absorption system, followed by a debutanizer for separation of LP-gas from gasoline. Results were satisfactory at low rates of gas. At a rate of 5,000 std. cu. ft. per hr., 84 percent of the propylene in the feed may be recovered.

The polymerization unit was in operation from January 1 to 24. Conversion was over 90 percent during 20 of the 24 days. Low conversions resulted from low olefin concentrations in the feed that caused a low heat of reaction and consequently low temperatures in the reactor.

---

<sup>4/</sup> Batchelder, Howard R., Busche, Robert M., The Kinetics of Coal Gasification V. Correlation of Rate Constants and Application of Calculation Method to Pressure Gasification.

Busche, Robert M., Batchelder, Howard R., The Kinetics of Coal Gasification VI. Application to Design of Atmospheric Pressure Gasifiers.

In a typical day during high conversion, 1 million cu. ft. of purified gas was fed to the synthesis unit. The production of this gas (which represented about two-thirds of actual production rate) required 14.8 tons of coke and 11.5 tons of oxygen. From it, 940 gallons of gasoline, 530 gallons of LP-gas, and 110 gallons of heavy oil were produced.

### Gas-Purification Plant

The Girbotol unit was started up on December 4, 1952. The diethanolamine solution (38 percent) was circulated at a rate of 38 g.p.m. Gas flow to the absorber was 72,000 std. cu. ft. per hr. of gas and 19-1/2 percent carbon dioxide. The absorber pressure was 340 p.s.i.g. The reactivator was carried at 12 p.s.i.g., and the reactivator reboiler steam pressure at 30 p.s.i.g. Sixty thousand std. cu. ft. per hr. of purified gas, with a carbon dioxide content of 4 percent and 2 grains of hydrogen sulfide per 100 cu. ft. of gas, were obtained from the Girbotol unit.

Active carbon tower 214 was steamed 10 hours before being put on the line. Carbon tower 215 was steamed next and then put on the line. Tower 219, which had been converted to an active carbon unit in previous operation, was also steamed and placed in standby for emergency service. Iron oxide tower 207 had been recharged with fresh iron oxide before the start-up and was placed second in line (reverse series flow). Iron oxide tower 208, which had been second in line during previous operation and consequently was essentially unexpended, was not recharged and was placed first in line for this run. Its life was 35 percent expended in this operation.

The initial operating conditions were held substantially constant throughout the run except for small variations in flow rate of amine and steam pressure in the reboiler.

Operation of the iron oxide towers was routine to remove all of the hydrogen sulfide remaining in the gas after the Girbotol unit. At all times lead-acetate-paper tests for hydrogen sulfide in the gas leaving the iron oxide towers were negative. Attempts made before the start of the run to revivify the iron oxide with gas containing small concentrations of oxygen are described in another section.

The previous procedure for revivifying the active carbon towers was followed. A given tower was used in second position for 2 days, then in first position for 2 days, after which it was taken out of service and depressured; carbon was regenerated by an upward flow of steam superheated to about 600° F. About 1,500 pounds of steam per hour was used. The treatment was continued until the top temperature of the carbon had been above 400° F. for at least 2 hours, but in no case was the total time less than 10 hours. The tower was then brought to operating pressure and used in second position again. The total sulfur in the purified gas was at all times under the specified maximum of 0.1 grain per 100 cu. ft. The highest sulfur concentrations were obtained just after a freshly steamed tower had been restored to service. In general, these were of the order of 0.01 grain per 100 cu. ft., and after 2 or 3 hours dropped to less than 0.005.

### Synthesis

#### Objectives and Preliminary Preparations

Following runs 2 and 3, which were carried out at a synthesis reactor temperature of 525° and 535° F., respectively, run 4 was set up for a maximum reactor temperature of 515° F. It was hoped that this would prevent the severe and rapid composition of catalyst that had been experienced in run 3.

A second objective was to improve the operation of the oil-circulation pumps which had caused considerable difficulty in earlier runs. The seal-oil piping to the three pumps was modified for this run to provide separate seal-oil supplies for the operating pump and for the standby pumps. In this manner, seal failures could be confined to the particular pump concerned. The method of operation was changed for this run, in that full seal pressure was maintained from the time the pump was placed in service until the pump was ready for overhaul. Previous pump failures apparently had resulted from a change in seal pressure.

The control system in the synthesis unit was modified to improve the performance of the primary recovery system, as well as control of pressure in the synthesis system. In previous runs, pressure in the system was maintained by regulation of tail-gas flow, resulting in surges that caused upsets in the primary recovery system. In this run, the flow of tail gas was controlled directly, and pressure was controlled by providing constant pressure of feed gas.

The gas inlet to the reactor was modified for run 4, both to eliminate erosion experienced in the inlet tee during run 3 and to improve gas distribution in the reactor. The revised inlet consisted of an inverted bell with a serrated edge at the center of the reactor near the bottom. Gas streamed out of the serrations of the 12-inch-diameter bell and was distributed throughout the cross section of the 3-foot-diameter reactor. The oil entered an inlet tee at the bottom of the reactor and was directed upward into the center of the gas distributor bell. The steel balls, previously used in the bottom of the reactor, were confined to the oil-inlet tee during this run.

## History

### Catalyst Charging

Catalyst, stored in drums under Primol "D" oil, was charged on December 14, 1952, as follows:

|                         |              |
|-------------------------|--------------|
| Catalyst volume         | 115 cu. ft.  |
| Catalyst weight         | 13,731 lb.   |
| Oil volume (Primol "D") | 605 gal.     |
| Average reduction       | 88.6 percent |

After coolant oil was added to the system on December 15, oil circulation was begun, followed by heating of the oil in preparation for the stepwise induction of the catalyst.

### Induction Period

The system was maintained at five temperature levels during the induction period:

| <u>Temperature, °F.</u> | <u>Conversion, percent (H<sub>2</sub> + CO)</u> |
|-------------------------|---|
| 440                     | 25  |
| 460                     | 35  |
| 480                     | 50  |
| 500                     | 59  |
| 510                     | 63  |

Each of the conversion levels was maintained for about 24 hours. During this period, and during most of the remainder of the run, the following conditions were maintained:

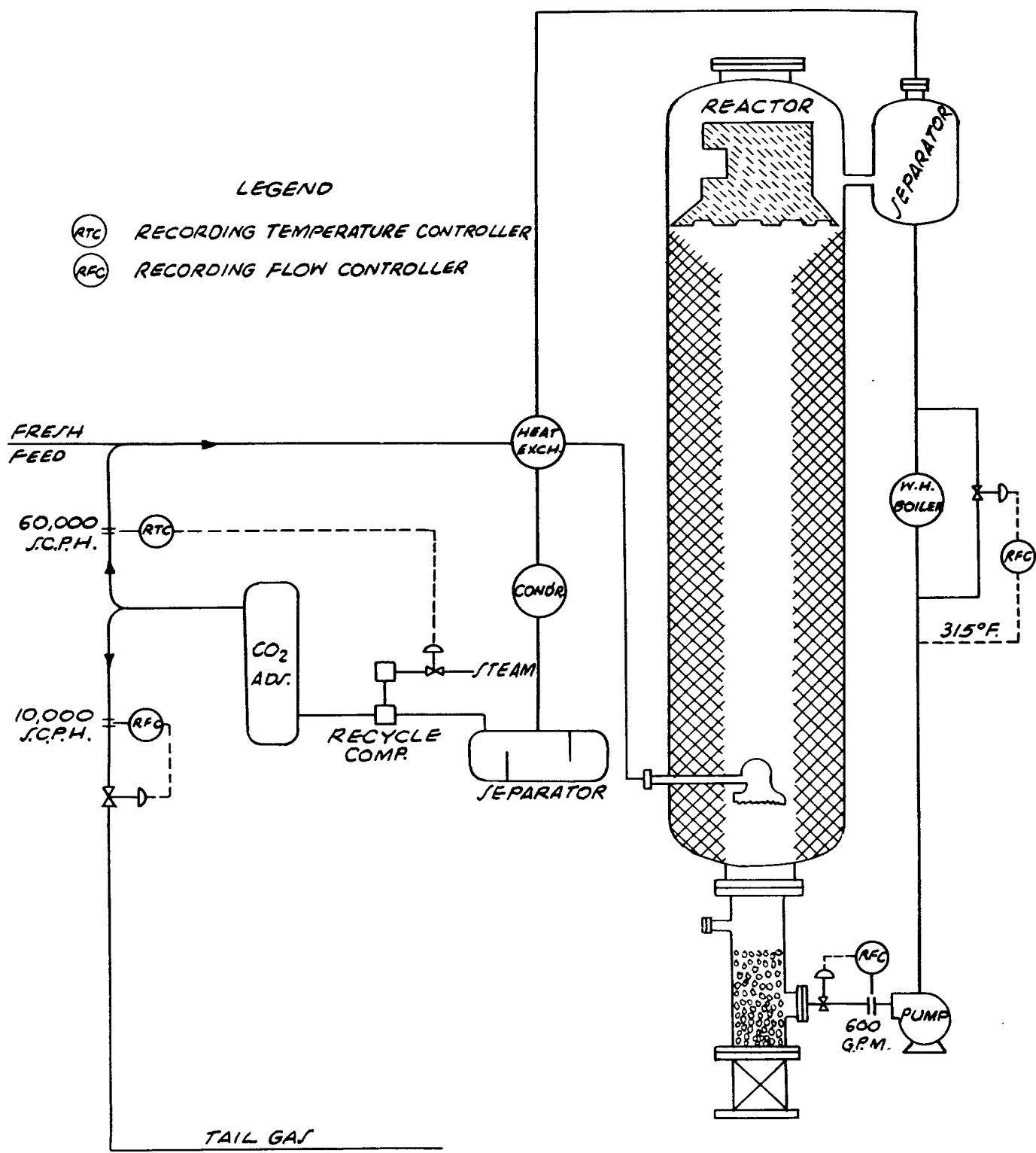


Figure 37. - Simplified flowsheet of synthesis reactor.

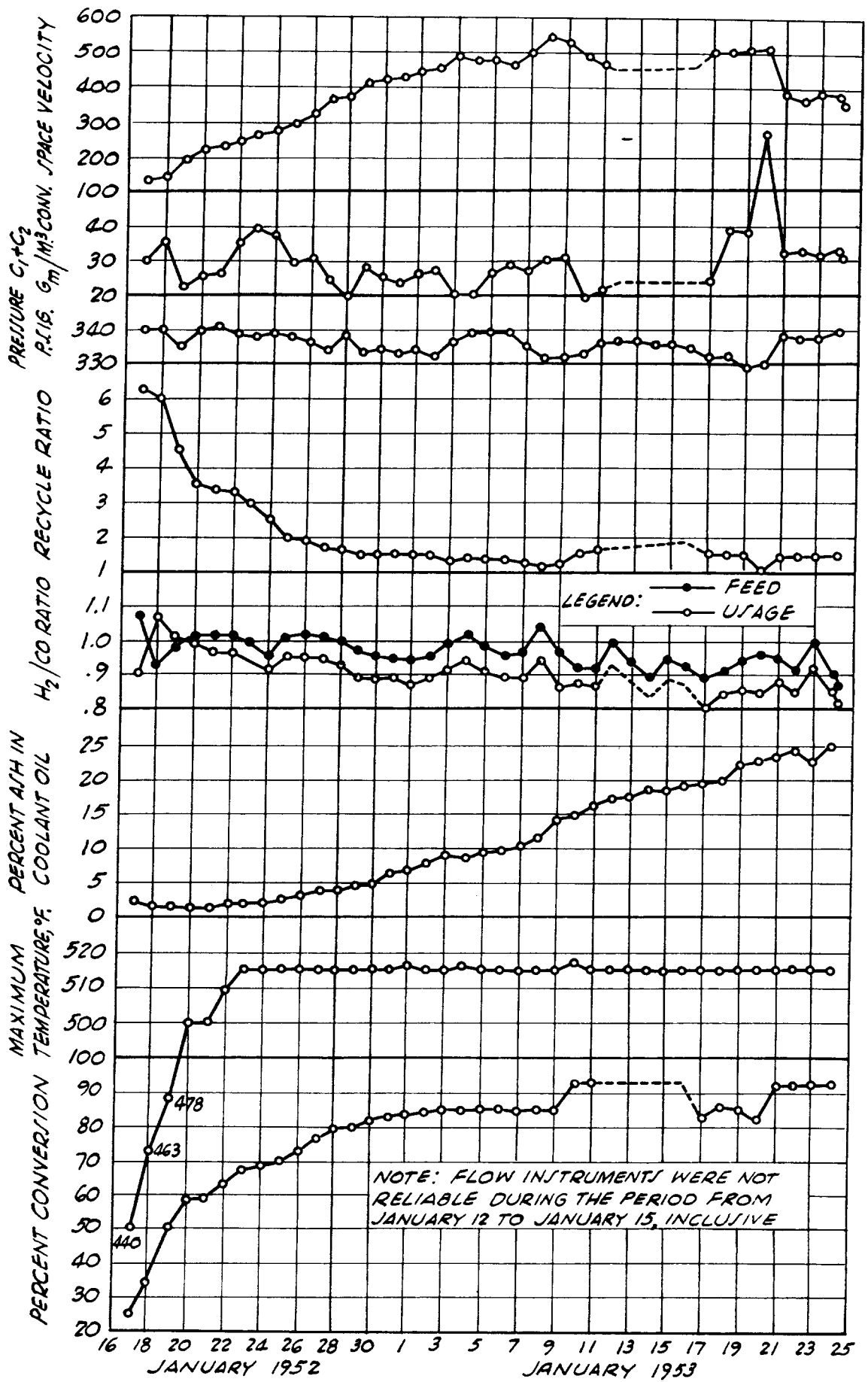


Figure 38. - Daily log, showing variations in synthesis.

| <u>Flow rates</u>            |        |
|------------------------------|--------|
| Tail gas.....std. cu.ft./hr. | 10,000 |
| Recycle gas.....do.          | 60,000 |
| Coolant oil.....gal./min.    | 500    |
| <u>Pressure</u>              |        |
| Reactor inlet.....p.s.i.g.   | 330    |

The method of control is indicated on the simplified flowsheet, figure 37.

#### Normal Operation

Figure 38 shows the daily variations in conditions. On December 23, the temperature was raised to 515° F. and maintained there for the remainder of the run, with only minor variations. Other conditions were maintained as shown above. The flow rate of fresh feed gas was governed only by the activity of the catalyst, as the tail gas rate was held constant. The fresh feed rate, as indicated by the space velocity, increased continuously through January 3. On January 8 and 9, the rate of tail gas was increased to about 12,000 std. cu. ft. per hr. to allow the gasmeter for fresh feed to be checked against a meter in the purification unit. During this period the synthesis unit was being fed with the entire output of gas from the purification unit. On January 10, the flow rate of tail gas was reduced to 5,000 std. cu. ft. per hr. to increase the conversion. The period from January 12 through the 16 was eliminated from consideration as difficulties with the flowmeters made the results inconsistent and unreliable. On January 17, when the difficulties were corrected, maximum flow of gas was repeated (12,000 std. cu. ft. per hr. of tail gas). Another period of using 5,000 std. cu. ft. per hr. of tail-gas was started on the 21st. The unit was shut down on January 24 after 39 days on stream. An interruption in the supply of gas caused a temporary halt on January 9.

#### Discussion of Data

##### Activity and Disintegration of Catalyst

Earlier tests in bench-scale and pilot-plant equipment had shown that several weeks are needed to develop fully the activity of a reduced, fused catalyst. During that period an active, friable layer is formed on the surface of the iron particles. This layer may spall, thus increasing the amount of solid material in the circulating oil and exposing a relatively inactive surface that requires further conditioning. "Induction" of the catalyst before normal synthesis accomplishes this goal best, but conditioning under synthesis conditions is also effective.

Figure 38 shows that these findings also apply to these large-scale experiments. Agitation of the catalyst appears to have been sufficient to remove active material from the surface of the iron. During the first week of synthesis, relatively little spalling occurred, after which the rate of attrition was approximately constant. At the end of the run, about 40 percent of the catalyst was suspended in the circulating oil. The operation, therefore, represents a combination of the moving-bed and slurry processes. The relative contribution of each of these components - large particles and fines - to the overall activity and selectivity can now be estimated.

##### Product Distribution

Table 21 shows the effect of space velocity and possibly also catalyst concentration on product distribution. The distribution of products is given for two periods,

one at 84 percent conversion, the second at 92 percent conversion. The product tends to become lighter at the higher conversion level. Of the butane-and-heavier product, 90.8 percent is in the gasoline boiling range at a conversion level of 92 percent. Comparison of the two periods shows that the yield of propane and butane is greater at the higher conversion level, and yet the yield of polymer gasoline is actually lower. This is the result of a lower percentage of unsaturation in the C<sub>3</sub>-C<sub>4</sub> product.

#### Olefin Concentration

Unsaturation in the C<sub>3</sub> fraction fell very rapidly during the first 3 weeks. That unsaturation at a flow of 5,000 std. cu. ft. per hr. of tail gas was less than that at 12,000 std. cu. ft. per hr. of tail gas confirms earlier observations that unsaturation increases with increasing space velocity and recycling. While the cause of the rapid initial decrease in unsaturation is not known, the effect is real and great. In the later period, one sample showed less than 5 percent unsaturation, and several showed less than 20 percent unsaturation. The results for the C<sub>4</sub> fraction were similar.

#### Mechanical Performance of Equipment

##### Mechanical Seals

The mechanical seals on the coolant-oil-circulation pumps, which caused difficulties in run 3, gave very little trouble in this run. Seal-oil leakage into the system was slight, and the seal failures that took place were due to abrasion of the carbon seal faces after relatively long periods of operation. One pump lasted over 20 days before the seal failed.

The success of the seals during this run probably can be attributed to three factors:

- (1) Careful adjustment and testing in the shop before installation.
- (2) Continuous and constant seal pressure during the life of the seal.
- (3) Independent seal-oil supply for each pump.

##### Coolant-Oil Pumps

Severe erosion of the casings and impellers of the coolant-oil circulation pumps has been observed during periods of high ash content in the coolant oil. Pump 310 was removed from service on January 9 after 24 days of operation, during which time the ash in the coolant oil built up to 13 percent. No appreciable erosion was observed. The pump was put back into service on January 10 and was run until January 21. During this 11-day period, the ash in the coolant oil rose from 14 to 23 percent, and the erosion on the pump was severe enough to necessitate replacement of the impeller and casing. This experience was similar to that encountered in run 3.

##### Control Valves

In run 3, the coolant-oil flow-control-valve body became eroded, and a leak developed, causing a shutdown. This valve was a double-seated, V-port, 6-inch, globe-type valve. For this run, the valve was replaced by a 4-inch, single-seat, linear plug, angle-type valve. Although the erosion, necessitated replacement of the seat and plug, the valve body was not noticeably damaged.



TABLE 21. - Effect of Gas-conversion feed on product distribution

| Period:  | 1-2-53 through 1-7-53                   | 1-21-53 through 1-24-53                                 |                                       |
|--|---|---|---------------------------------------|
| Synthesis data   |   |   |                                       |
| Maximum reactor temperature.....   | 515                                     | 515   |                                       |
| Pressure.....  | 337                                     | 338   |                                       |
| Space velocity.....  | 480                                     | 373   |                                       |
| Recycle ratio.....   | 1.35                                    | 1.43  |                                       |
| Ash (coolant oil), percent   | 7.0                                     | 24.0  |                                       |
| Usage ratio.....   | 0.91                                    | 0.88  |                                       |
| Conversion.....percent   | 84.3                                    | 91.9  |                                       |
| Product distribution - basis: 100,000 std.c.f. (H <sub>2</sub> + CO) converted |   |   |                                       |
| Component  | $\text{O.A.P.I.}$<br>Lb./gal.Vol.(gal.) | $\text{O.A.P.I.}$ Lb./gal.<br>M std. c.f.<br>Vol.(gal.) | M std. c.f.<br>M std. c.f.            |
| CO <sub>2</sub> .....  |   | 20.0  | 21.6                                  |
| Gas(C <sub>1</sub> +C <sub>2</sub> )..   |   | (3.65x10 <sup>6</sup> B.t.u.)<br>2.92                   | 3.57<br>(4.25x10 <sup>6</sup> B.t.u.) |
| Propane.....   | 143.8                                   | 143.8   | 34.9                                  |
| Butane.....  | 110.5                                   | 110.5   | 18.0                                  |
| Polymer  |   |   |                                       |
| gasoline..   | 65.1                                    | 65.1  | 28.3                                  |
| C <sub>5</sub> + 400 e.p.  | 70.0                                    | 70.0  | 65.5                                  |
| Total  | 29.0<br>54.8                            | 65.1<br>70.0  | 111.8<br>157.7                        |
| gasoline..   | 96.5                                    | 48.0  | 8.1                                   |
| Diesel oil..   | 12.6                                    | 32.0  | 3.1                                   |
| Coolant oil.   | 13.3                                    |   |                                       |
| Total.....   | 152.4                                   |   |                                       |
| M std.cu.ft.   | 27.5                                    |   | 26.6                                  |
| (H <sub>2</sub> +CO) conv.   |   |   |                                       |
| Bbl. C <sub>3</sub> +prod.   |   |   |                                       |

## Oil-Inlet Tee

The oil-inlet tee at the bottom of the reactor is fabricated from 8-inch schedule 40 pipe and has a 6-inch welded side outlet. The oil enters the 6-inch side outlet, changes direction and flows upward through the 8-inch pipe, passing through a bed of steel balls used as a catalyst support. The velocity of the oil is sufficient to cause the balls to rattle in the tee, and to exert a continuous hammering effect opposite the 6-inch inlet. Following the shutdown, inspection revealed that the hammering effect of the balls had caused the pipe to bulge approximately 1/2 inch and had decreased the wall thickness at that point from 0.33 inch to 0.13 inch.

## Conclusions

It has been demonstrated that catalyst decomposition is much less rapid at 515° F. than at 535° F. In 39 days of operation at 515° F. the catalyst decomposed to the same extent as in 6 days of operation at 535° F. (run No. 3).

On the other hand, it appears that catalyst decomposition is helpful to maximum development of catalyst activity, as catalyst decomposition was beneficial up to about 12 percent ash in the coolant oil. However, this probably is peculiar to synthesis procedures involving severe mechanical erosion of the catalyst. Tests in small pilot plants with catalysts prepared by activation of steel turnings, and with low linear velocities of cooling oil, show high activity and durability of catalyst with virtually no solids in the coolant oil.

Coolant-oil ash contents of 15 to 25 percent were intolerable in the demonstration plant operations. Ash contents of this magnitude damaged 600-g.p.m. pumps necessitating replacement of internals on two occasions over an 8 to 10-day period.

High percentage conversions can be obtained in a single-stage reaction using recycle gas free of carbon dioxide. Accompanying the high conversion, is a shift in product distribution toward a much lighter product. This is a highly desirable shift as a larger proportion of the product appears in the valuable gasoline range, and the gas yield, at the same time, is not unusually high (28-33 grams per cubic meter converted).

The yield of products for synthesis run No. 4 is given in table 22.

TABLE 22. - Products made in synthesis run 4

|                           | <u>Gallons</u> |
|---------------------------|----------------|
| Polymer gasoline.....     | 5,170          |
| Debutanized gasoline..... | 17,408         |
| Lean oil.....             | 3,619          |
| Coolant oil.....          | <u>2,095</u>   |
| Total.....                | 28,292         |

Following run R-5, the demonstration plant was put in standby condition by midsummer.

RESEARCH AND DEVELOPMENT, Coal-to-Oil Laboratories and Pilot Plants, Bruceton and Pittsburgh, Pa.

1. ANDERSON, R. B. Iron Nitrides as Fischer-Tropsch Catalysts. *Advances in Catalysis*, vol. 5, Academic Press, 1953, pp. 355-384.
2. \_\_\_\_\_. Physical Chemistry of the Fischer-Tropsch Synthesis on Iron Catalysts. Pres. before Petrol. Div. Meeting of American Chemical Society, Los Angeles, Calif. Mar. 15-19, 1953.
3. ANDERSON, R. B., and SHULTZ, J. F. Iron Nitride Catalyst in Carbon Oxide Hydrogenation. U. S. Patent 2,629,728, Feb. 24, 1953.
4. CANTONI, ALDO, and FELDMAN, JULIAN. Vapor-Liquid Equilibria. Binary System Nitromethane-Nitroethane at Subatmospheric Pressure. *Ind. Eng. Chem.* vol. 45, 1953, pp. 2580-2584.
5. COHN, E. M., and HOFER, L. J. E. Some Thermal Reactions of the Higher Iron Carbides. *Jour. Chem. Phys.*, vol. 21, No. 2, February 1953, pp. 354-359.
6. ELLIOTT, MARTIN A., CLARK, EZEKAIL L., and STORCH, HENRY H. Production of Hydrocarbon Synthesis Gas from Coal. U. S. Patent 2,634,286, April 7, 1953.
7. FELDMAN, J., and ORCHIN, M. Composition of Gasoline From Coal Hydrogenation. *Ind. Eng. Chem.*, vol. 44, No. 12, December 1952, pp. 2852-2856.
8. \_\_\_\_\_. Separation of 1- and 2-Methylnaphthalenes by Azeotropic Distillation. *Ind. Eng. Chem.*, vol. 44, No. 12, December 1952, pp. 2909-2914.
9. FELDMAN, J., SVEDI, A., CONNELL, S., and ORCHIN, M. Evaluation of Vacuum Rectification Columns. *Ind. Eng. Chem.*, vol. 45, No. 1, January 1953, pp. 214-215.
10. FRIEDEL, R. A., LOGAR, A. F., and SHULTZ, J. L. Mass Spectrometric Analysis of Paraffin-Naphthene Mixtures. *Appl. Spectroscopy*, vol. 6, No. 5, 1952, pp. 24-28.
11. FRIEDEL, R. A., and PELIPETZ, M. G. Infrared Spectra of Coal and Carbohydrate Chars. *Jour. Opt. Soc. America*, vol. 43, No. 11, 1953, pp. 1051-1052.
12. FRIEDEL, R. A., SHARKEY, A. G., Jr., SHULTZ, J. L., and HUMBERT, C. R. Mass Spectrometric Analysis of Mixtures Containing Nitrogen Dioxide. *Anal. Chem.*, vol. 25, 1953, p. 1314.
13. GOLDEN, P. L., and HEADRICK, A. F. A Sensitive Pressure Controller for High-Pressure Service. *Trans. Am. Soc. Mech. Eng.*, vol. 75, No. 3, April 1953, p. 329.
14. GOLUMBIC, C. Partition Studies. VIII. Silver Complexes of Aromatic Amines. *Jour. Am. Chem. Soc.*, vol. 74, No. 22, 1952, pp. 5777-5778.
15. \_\_\_\_\_. Phenols. A Survey. *Encyclopedia of Chemical Technology*, vol. 10, The Interscience Encyclopedia, Inc., New York, N. Y., pp. 297-320.
16. \_\_\_\_\_. Separation and Analysis of Tar Bases by Countercurrent Distribution. *Anal. Chem.*, vol. 24, No. 11, 1952, pp. 1849-1850.
17. HALL, W. K., DIETER, W. E., HOFER, L. J. E., and ANDERSON, R. B. Preparation and Reactions of Carbonitrides on Iron. *Jour. Am. Chem. Soc.*, vol. 75, No. 6, March 1953, pp. 1442-1447.

18. HINKEL, R. D., and RAYMOND, R. Direct Semimicrodetermination of Oxygen in Organic Substances. *Anal. Chem.*, vol. 25, No. 3, March 1953, pp. 470-479.
19. HITESHUE, R. W., and CLARK, E. L. A Small High-Pressure Pump for Pilot-Plant Service. *Mech. Eng.*, vol. 75, No. 3, March 1953, pp. 205-206.
20. MANES, MILTON. Relationship Between Kinetics and Acoustic Phenomena in Equilibrium Systems. *Jour. Chem. Phys.* vol. 21, No. 10, October 1953, pp. 1791-1796.
21. MANES, M., DAMICK, A. D., MENTSER, M., COHN, E. M., and HOFER, L. J. E. Hexagonal Iron Carbide as an Intermediate in the Carbiding of Iron Fischer-Tropsch Catalysts. *Jour. Am. Chem. Soc.*, vol. 74, No. 24, December 1952, pp. 6207-6209.
22. MANES, M., HOFER, L. J. E., and HINKEL, R. D. Method for the Production of High-Precision Glass Stopcocks. U. S. Patent 2,648,179, Aug. 11, 1953.
23. McCARTNEY, JAMES T., HOFER, L. J. E. SELIGMAN, BERNARD, LECKY, JAMES A., PEEBLES, W. C., and ANDERSON, ROBERT B. Electron and X-Ray Diffraction Studies of Iron Fischer-Tropsch Catalysts. *Jour. Phys. Chem.*, vol. 57, 1953, pp. 730-736.
24. ORCHIN, M., and FELDMAN, J. The Isolation of Coronene From a Coal-Hydrogenation Oil. *Jour. Org. Chem.*, vol. 18, May 1953, pp. 609-613.
25. PANTAGES, P., and FELDMAN, J. Packing Supports for Glass Columns. *Ind. Eng. Chem.*, vol. 44, No. 11, 1952, p. 2783.
26. PELIPETZ, M. G., SALMON, J. R., BAYER, J., and CLARK, E. L. Catalyst-Pressure Relationship in Hydrogenolysis of Coal. *Ind. Eng. Chem.*, vol. 45, No. 4, April 1953, pp. 806-809.
27. STEIN, K. C., and SCHOENEWEIS, F. J. Mercury Piston Gives Close Control of Liquid-Feed Rates. *Chem. Eng.*, vol. 60, 1953, p. 200.
28. SPERNBERG, H. W., WENDER, I., FRIEDEL, R. A., and ORCHIN, M. The Chemistry of Metal Carbonyls, II. Preparation and Properties of Cobalt Hydrocarbonyls. *Jour. Am. Chem. Soc.*, vol. 75, 1953, pp. 2717-2720.
29. \_\_\_\_\_. The Chemistry of Metal Carbonyls, III. The Reaction Between Dicobalt Octacarbonyl and Dimethylamine. *Jour. Am. Chem. Soc.*, vol. 75, No. 13, July 1953, pp. 3148-3152.
30. STORCH, H. H. Book Review of "Advances in Catalysis and Related Subjects. III, IV, and V." *Sci. Monthly*, vol. 76, No. 2, February 1953, pp. 114-115.
31. \_\_\_\_\_. Combustibili Liquidi da Carbone a Scisti Petroliferi (Liquid Fuels From Coal and Oil Shale). *Riv. di Combustib.*, vol. 6, No. 3, 1952, pp. 139-163.
32. \_\_\_\_\_. Problems in the Physical Chemistry of Coal. Proc. 2d Conf. on Origin and Constitution of Coal, Dept. of Mines, Canada.
33. WEINTRAUB, M., and LEVA, M. Fluid Dynamics. *Ind. Eng. Chem.*, vol. 45, No. 1, January 1953, pp. 74-82.
34. WENDER, I., SPERNBERG, H., and ORCHIN, M. Evidence for Cobalt Hydrocarbonyl as the Hydroformylation Catalyst. *Jour. Am. Chem. Soc.*, vol. 75, 1953, pp. 3041-3042.

35. WENDER, I., STERNBERG, H., METLIN, S. J., and ORCHIN, M. Dicobalt Octacarbonyl. Inorganic Syntheses, vol. V (ed., J. C. Bailar), McGraw-Hill, New York.
36. WHITEHOUSE, A. M., GOLDEN, P. L., HITESHUE, R. W., and CLARK, E. L. Bench-Scale Equipment for Reactions at High Pressure. Chem. Eng. Prog., vol. 49, No. 9, September 1953, pp. 491-496.

RESEARCH AND DEVELOPMENT, Synthesis Gas From Coal Pilot Plants, Morgantown, W. Va.

1. BARKLEY, L. W., CORRIGAN, T. E., WAINWRIGHT, H. W., and SANDS, A. E. A Kinetic Study of the Catalytic Reverse Shift Reaction. Pres. before Am. Chem. Soc. New York, N. Y., September 1951; Ind. Eng. Chem., vol. 44, May 1952, p. 1066.
2. STRIMBECK, G. R., CORDINER, J. B., Jr., TAYLOR, H. G., PLANTS, K. D., and SCHMIDT, L. D. Progress Report on Operation of Pressure-Gasification Pilot Plant Utilizing Pulverized Coal and Oxygen. Bureau of Mines Rept. of Investigations 4971, 1953, 27 pp.
3. WAINWRIGHT, H. W., EGLESON, G. C., BROCK, C. M., FISHER, J., and SANDS, A. E. Selective Absorption of Hydrogen Sulfide from Synthesis Gas. Ind. Eng. Chem., vol. 45, June 1953, pp. 1378-1384.

RESEARCH AND DEVELOPMENT, Gorgas Underground Gasification Project, Gorgas, Ala.

1. ELDER, JAMES L., and FIES, MILTON H. Untertagevergasung von Kohle. Pres. before Steinkohlentagung, Aachen, Germany, April 27-28, 1953; pub. Brennstoff-Chemie, vol. 34, No. 13/14, July 1953, pp. 196-199.
2. \_\_\_\_\_. Underground Gasification of Coal - the Gorgas Experiments. Consulting Eng., vol. 2, No. 6, August 1953, pp. 22-25, 62-64.
3. FIES, M. H., and ELDER, JAMES L. Underground Gasification of Coal at Gorgas, Ala. Gasification With Air, Oxygen, and Steam Following Use of Electrolinking-Carbonization for Preparing a Passage in the Coal Bed. Paper 53-A137, Pres. before Am. Soc. Mech. Eng., Fuels Division, New York, N. Y., Nov. 29-Dec. 4, 1953.

SYNTHETIC FUELS DEMONSTRATION PLANT, Louisiana, Mo.

1. BATCHELDER, HOWARD R., BUSCHE, ROBERT M., and ARMSPRONG, WILLARD P. Kinetics of Coal Gasification. Ind. Eng. Chem. vol. 45, No. 9, September 1953, pp. 1856-1878.
2. CHAFFEE, C. C., and HIRST, L. L. Liquid Fuel From Coal. Progress Report on Hydrogenation Demonstration Plant. Ind. Eng. Chem., vol. 45, April 1953, pp. 822-838.
3. CLARKE, EDWIN A., CHAFFEE, C. C., and HIRST, L. L. Early Operations of the Hydrogenation Demonstration Plant Using Rock Springs, Wyo., Coal. Bureau of Mines Rept. of Investigations 4944, 1953, 80 pp.
4. GARDNER, G. D., and DONOVAN, J. T. Corrosion and Erosion in the Synthetic Fuels Demonstration Plant. Trans. Am. Soc. Mech. Eng., vol. 75, No. 4, May 1953, pp. 525-533.

GENERAL - Washington, D. C.

1. BUREAU OF MINES. 1952 Annual Report of the Secretary of the Interior on Synthetic Liquid Fuels. Part I - Oil from Coal. Rept. of Investigations 4942, 1953, 85 pp.

2. BUREAU OF MINES. 1952 Annual Report of the Secretary of the Interior on Synthetic Liquid Fuels. Part II - Oil from Oil Shale. Rept. of Investigations 4943, 1953, 70 pp.
3. STORCH, H. H. Research and Development on the Conversion of Coal to Gas, Liquid Fuels, and Bulk Organic Chemicals. Pres. before Ann. Meeting, Mining Society of Nova Scotia, Sydney, N. S., July 1953. Pub. Trans., vol. 56, 1953, pp. 376-380; Canadian Min. and Met. Bull., vol. 46, No. 499, November 1953, pp. 690-694.
4. \_\_\_\_\_. Synthetic Liquid Fuel Processes. Ind. Eng. Chem., vol. 45, July 1953, pp. 1444-1447.
5. \_\_\_\_\_. Federal Research on Synthetic Fuels by the Bureau of Mines. The Oil Forum (in 3 parts). Part 1, vol. VII, No. 11, November 1953, pp. 420-422, 425; part 2, vol. VII, No. 12, December 1953, pp. 443-444, 453; part 3, vol. VIII, No. 1, January 1954, pp. 30-32.