FURTHER STUDIES OF THE FISCHER-TROPSCH SYNTHESIS USING GAS RECYCLE COOLING (HOT-GAS-RECYCLE PROCESS)

By J. H. Field, D. Bienstock, A. J. Forney,
and R. J. Demski

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FURTHER STUDIES OF THE FISCHER-TROPSCH SYNTHESIS USING GAS RECYCLE COOLING (HOT-GAS-RECYCLE PROCESS)[1]

by

J. H. Field,[2] D. Bienstock,[3] A. J. Forney,[3]
and R. J. Demski[4]

SUMMARY

Synthesis of hydrocarbons by the Fischer-Tropsch reaction in a hot-gas-recycle system was demonstrated to be technically feasible. This project was undertaken, by the Bureau of Mines, to determine if the system was operable and to ascertain the optimum conditions for the production of gasoline.

Carbon steel turnings packed with a void volume from 88 to 90 pct. were satisfactory catalysts, but steel catalysts with greater void volume, either turnings or steel wool, were not as active.

Good temperature control with a pressure drop of less than 0.5 pounds per square inch (p.s.i.) per foot of catalyst bed of turnings was obtained by operating with a recycle to fresh feed ratio of 20 and by splitting the total gas flow and injecting portions into the reactor at different entry ports.

At an hourly space velocity of 1,000, 90 pct. of the feed gas was converted at 300° to 320° C. At a space velocity of 1,500, 88 pct. was converted The proportion of gasoline in the hydrocarbon product ranged from 55 to 66 pct. The finished gasoline with 3 cc. of tetraethyl lead had research octane ratings from 90 to 96. Synthesis of gasoline in a hot-gas-recycle process is preferable to processes using oil as coolant. At the higher temperatures possible in the hot-gas system, a larger percentage of the hydrocarbon product is gasoline.

INTRODUCTION

The Bureau's program on the synthesis of liquid fuels from coal includes the development and evaluation of reactor systems for the catalytic

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[1] Work on manuscript completed March 1961.

[2] Project Coordinator, Pittsburgh Coal Research Center, Bureau of Mines, Pittsburgh, Pa.

[3] Supervisory chemical engineer, Pittsburgh Coal Research Center, Bureau of Mines, Pittsburgh, Pa.

[4] Engineering technician, Pittsburgh Coal Research Center, Bureau of Mines, Pittsburgh, Pa.

hydrogenation of carbon monoxide. Methods for conducting the synthesis differ principally in the technique of removing the exothermic heat of reaction and in the type and form of catalyst. Heat is removed directly by a cooling medium contacting the catalyst surface in the oil circulation,[5] slurry,[6] and hot-gas-recycle[7] processes, and indirectly in fluidized bed[8] and fixed bed[9] systems.

In the hot-gas recycle process the exothermic heat of reaction (about 70 B.t.u./cu.ft. of synthesis gas converted) is absorbed by the sensible heating of large volumes of recycle gas circulating through the reactor in direct contact with the catalyst. The heat absorbed by the recycle gas is then removed externally in a waste heat boiler by generating steam. Because of the large volumes of gas and the high cost of compression, it is necessary to have a low pressure drop through the catalyst bed. The development by the Bureau of an active lathe turning catalyst having a considerably lower resistance to gas flow and better heat-transfer characteristics than granular catalysts has overcome the major difficulties experienced by the Germans in 1938.[10]

Assuming a constant specific heat of the gas and a constant rate of heat transfer from the catalyst to the gas, the amount of recycle required depends on the temperature increase which can be tolerated across the catalyst bed. An excessive temperature rise would inactivate the catalyst and promote the deposition of carbon. A temperature rise of 40° gave no operating difficulties in the present studies; this is 4 times the temperature differential that could be tolerated in the early German studies.[11]

The purpose of these experiments was to demonstrate the feasibility of the hot-gas-recycle process with an active catalyst that offers a low resistance to gas flow and to determine the operating variables which affect the production and quality of gasoline and the life and activity of the catalyst. A previous report[12] discussed the use of a simulated hot-gas-recycle system in

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[5] Benson, H. E., Field, J. H., Bienstock, D., Nagel, R. R., Brunn, L. W., Hawk, C. O., Crowell, J. H., and Storch, H. H., Development of the Fischer-Tropsch Oil-Recycle Process: Bureau of Mines Bull. 568, 1957, 72 pp.

[6] Schlesinger, M. D., Crowell, J. H., Leva, M., and Storch, H. H., Fischer-Tropsch Synthesis in Slurry Phase: Ind. Eng. Chem., vol. 43, 1951, pp. 1474-1479.

[7] Bienstock, D., Jimeson, R. M., Field, J. H., and Benson, H. E., The Fischer-Tropsch Synthesis Using Gas Recycle Cooling (Simulated Hot-Gas-Recycle Process): Bureau of Mines Rept. of Investigations 5655, 1960, 25 pp.

[8] Demeter, J. J., and Schlesinger, M. D., Fischer-Tropsch Synthesis in a Fluidized-Catalyst Reactor with a Nitrided, Fused-Iron Catalyst: Bureau of Mines Rept. of Investigations 5456, 1959, 16 pp.

[9] Storch, H. H., Golumbic, N., and Anderson, R. B., The Fischer-Tropsch and Related Syntheses: John Wiley and Sons, Inc., New York, N.Y., 1951, 610 pp.

[10] Work cited in footnote 9.

[11] Work cited in footnote 9.

[12] Work cited in footnote 7.

which the recycle gas was cooled, compressed, and reheated to reactor temperature. Cooling the recycle gas below 25° C. results in condensing and removing from the system water and higher-boiling hydrocarbons. This simulated system was used until a compressor capable of circulating hot gases was installed in the pilot plant.

EXPERIMENTAL PROCEDURES

Raw Materials

Synthesis Gas

Synthesis gas was produced for the hot-gas-recycle pilot plant by reforming natural gas with steam and carbon dioxide over a nickel catalyst at atmospheric pressure in a small commercial unit. Hydrogen-to-carbon monoxide ratios were varied from 1.0 to 3.0. Impurities in the synthesis gas, methane, carbon dioxide and nitrogen amounted to less than 1 pct. The sulfur content was kept below 0.1 grain (gr.) per 100 cu.ft. by passing the gas through activated carbon. Usually the sulfur content was less than 0.02 gr. per 100 cu.ft.

Catalyst

Steel lathe turnings and steel wool were catalysts in these experiments. The turnings, made of 1018 carbon steel, were cut on a lathe, and the cutting depth, angle, and speed could be adjusted to produce turnings with void volumes from 88 to 96 pct.[13] No. 6 commercial grade steel wool was used in two experiments and was packed in the reactor to a void volume of 97 pct.

Apparatus and Method of Operation

The pilot plant consisted of a reactor, recycle compressor, carbon dioxide scrubbing system, product recovery system, and heat exchangers. The catalyst was oxidized in a separate unit. A simplified flowsheet of the hot-gas-recycle process is shown in figure 1. A more detailed view of the reactor indicating gas ports, pressure taps, catalyst sampler, and thermocouples is illustrated in figure 2. The reactor is a 3-in. schedule 80 pipe made of ASTM type A-106 steel, grade B. A maximum catalyst bed height of 10 ft. could be used, but all of the experiments discussed in this report were made with a 6-ft. bed equal to a volume of 0.275 cu.ft.

As shown in figure 1, the total feed gas, consisting of fresh and recycle gases, enters the top of the reactor and flows downward through the catalyst bed at a superficial linear velocity of about 4 to 5 ft./sec. The gas leaving the reactor enters a gas-to-gas heat exchanger where it is cooled to 200°-250° C. The effluent gas stream from the reactor is divided in the following manner: The product gas stream is cooled, metered, and sampled; the remainder is returned to the system by the hot-gas-recycle compressor. Five to 15 pct. of the hot recycle gas is cooled to condense the water vapor and oil and treated with monoethanolamine to absorb the carbon dioxide. Thus the carbon dioxide

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[13] Work cited in footnote 7, p. 2.

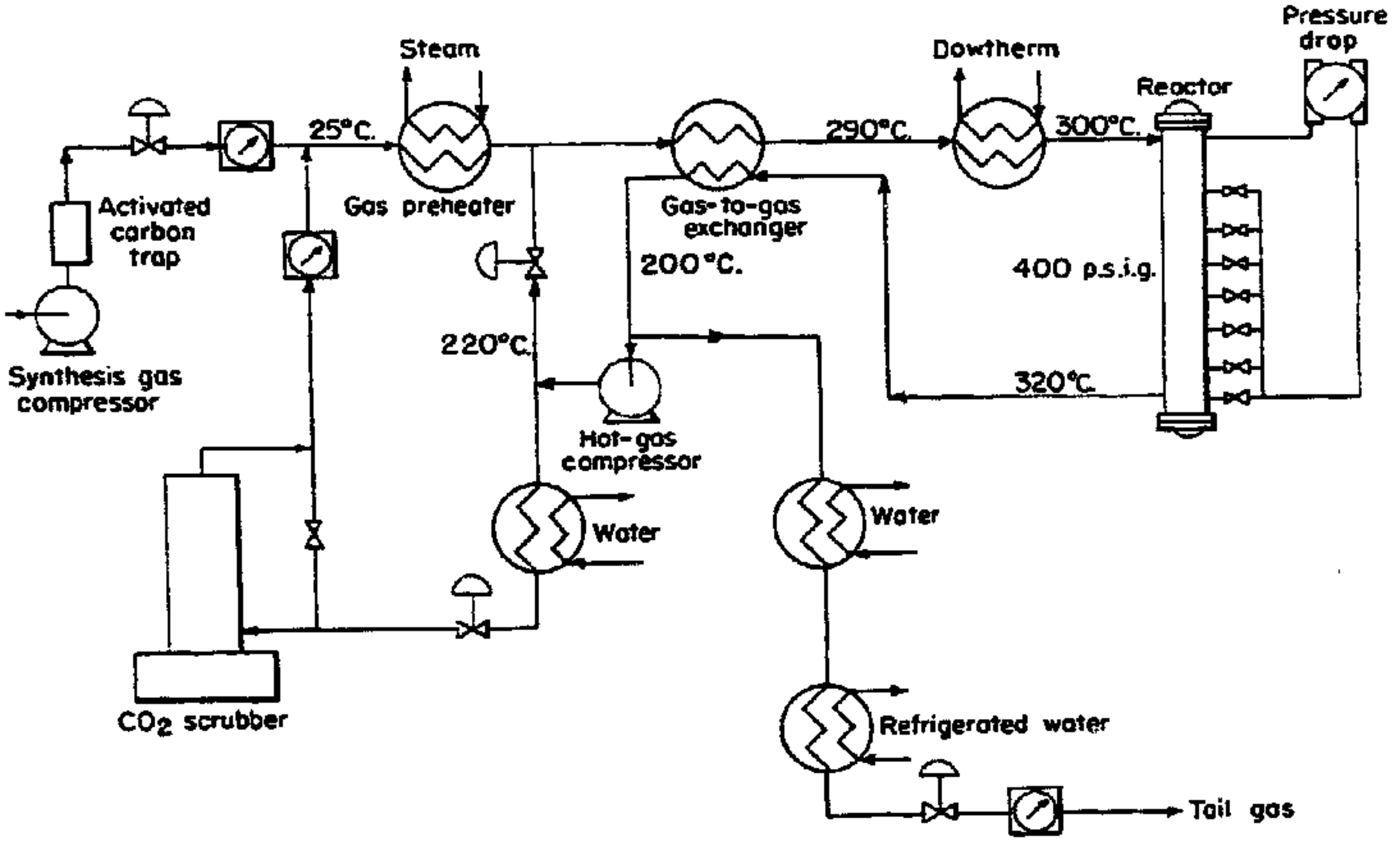

FIGURE 1. - Schematic Flowsheet of Hot-Gas-Recycle Process.

and water vapor content of the recycle stream are both maintained below 10 pct. This cold recycle then recombines with the remainder of the hot recycle gas and with the fresh feed gas, and flows to the reactor through the gas-to-gas heat exchanger and a Dowtherm-jacketed[14] gas heater. The recycle booster compressor is shown in detail in figure 3; it is a single-cylinder, double-acting reciprocating type. Figure 4 shows the compressor. The gas is compressed from 400 to 500 pounds per square inch gage (p.s.i.g.), and can be operated at temperatures as high as 325° C. The capacity at these conditions is 27,000 standard cubic feet per hour (std. cu.ft./hr.).

Carbon dioxide is removed from the cold recycle gas by a 20-pct. monoethanolamine solution that is fed into the top of the absorber through a distributor and flows countercurrent to the gas. The absorber column is 6 in. in diameter, packed to a height of 8 ft. with 3/4-in. ceramic Raschig rings. The stripper column is also 6 in. in diameter, containing 6 ft. of the same type packing. Any part of the cooled recycle gas could be sent to the absorber. This method of operation permitted the water vapor content and the carbon dioxide content in the recycle gas to be controlled independently.

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[14] Reference to brand names is made to facilitate understanding and does not imply endorsement of such brands by the Bureau of Mines.

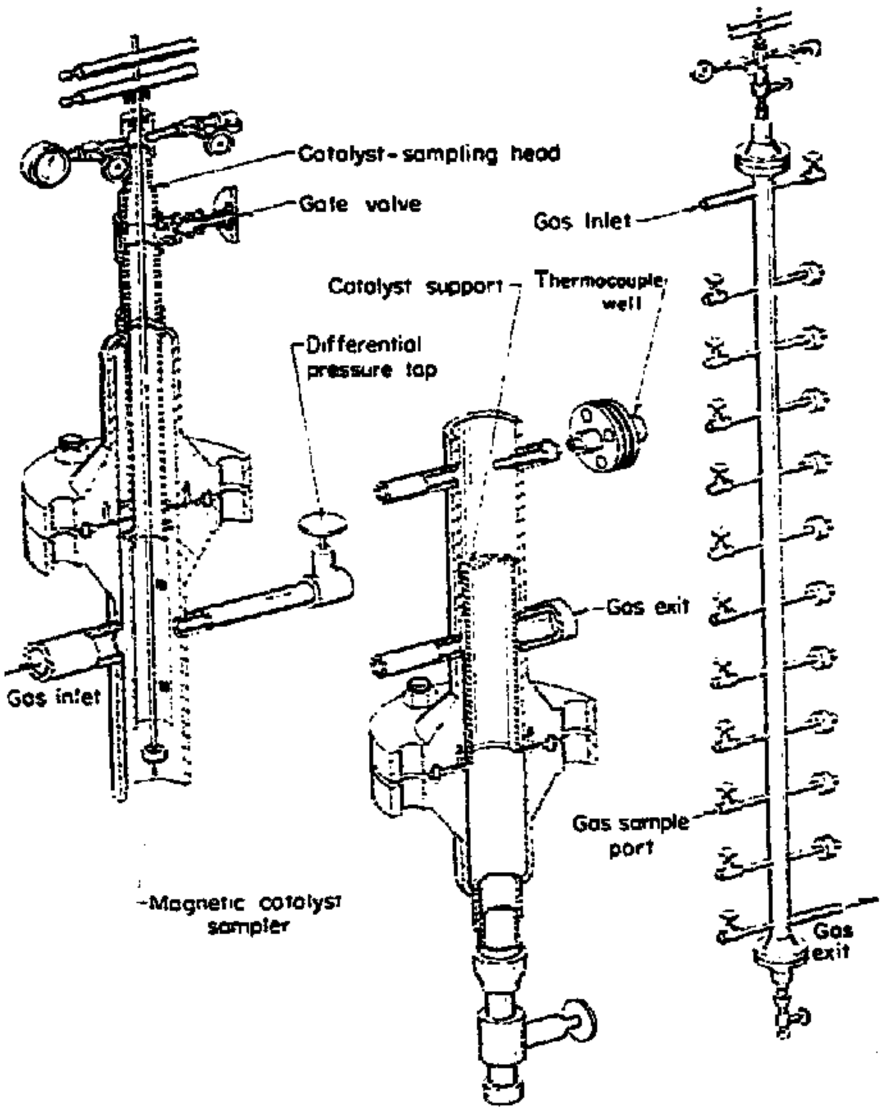

FIGURE 2. - Hot-Gas-Recycle Reactor.

Figure 5 shows schematically the unit used for oxidizing massive iron catalysts such as turnings and steel wool. The oxidizer is a 6-in. pipe made of 304-stainless steel. Its capacity is 0.3 cu.ft. Distilled water is converted to steam which is then superheated by passing through a tube coil wound around the oxidizer. The steam reacts with the iron to form Fe_3O_4 and hydrogen. The unreacted steam is condensed and the hydrogen flows through a drying tower and is metered. The yield of hydrogen indicates the degree of oxidation.

Operations

Oxidation

Oxidation followed by reduction of the turnings makes the steel porous and increases the surface area, resulting in a catalytically active material. The oxidations were made at 600° C. and atmospheric pressure using a steam flow of about 400 hourly space velocity. The steel was oxidized until 20 pct. of the iron was converted to Fe_3O_4. The degree of oxidation is checked by comparing the weight gain of the catalyst with the amount of hydrogen evolved. Oxidation in this manner produces a layer of iron oxide with the bulk of the steel in the interior unoxidized.

Impregnation

Storch has shown that alkali impregnation of massive iron catalysts shifts the product distribution toward the higher boiling hydrocarbons and minimizes the yield of gaseous hydrocarbons.[15] The catalysts used in these experiments were impregnated with an aqueous solution of potassium carbonate, and then dried. The strength of the solution and length of time the material

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[15] Work cited in footnote 9, p. 2.

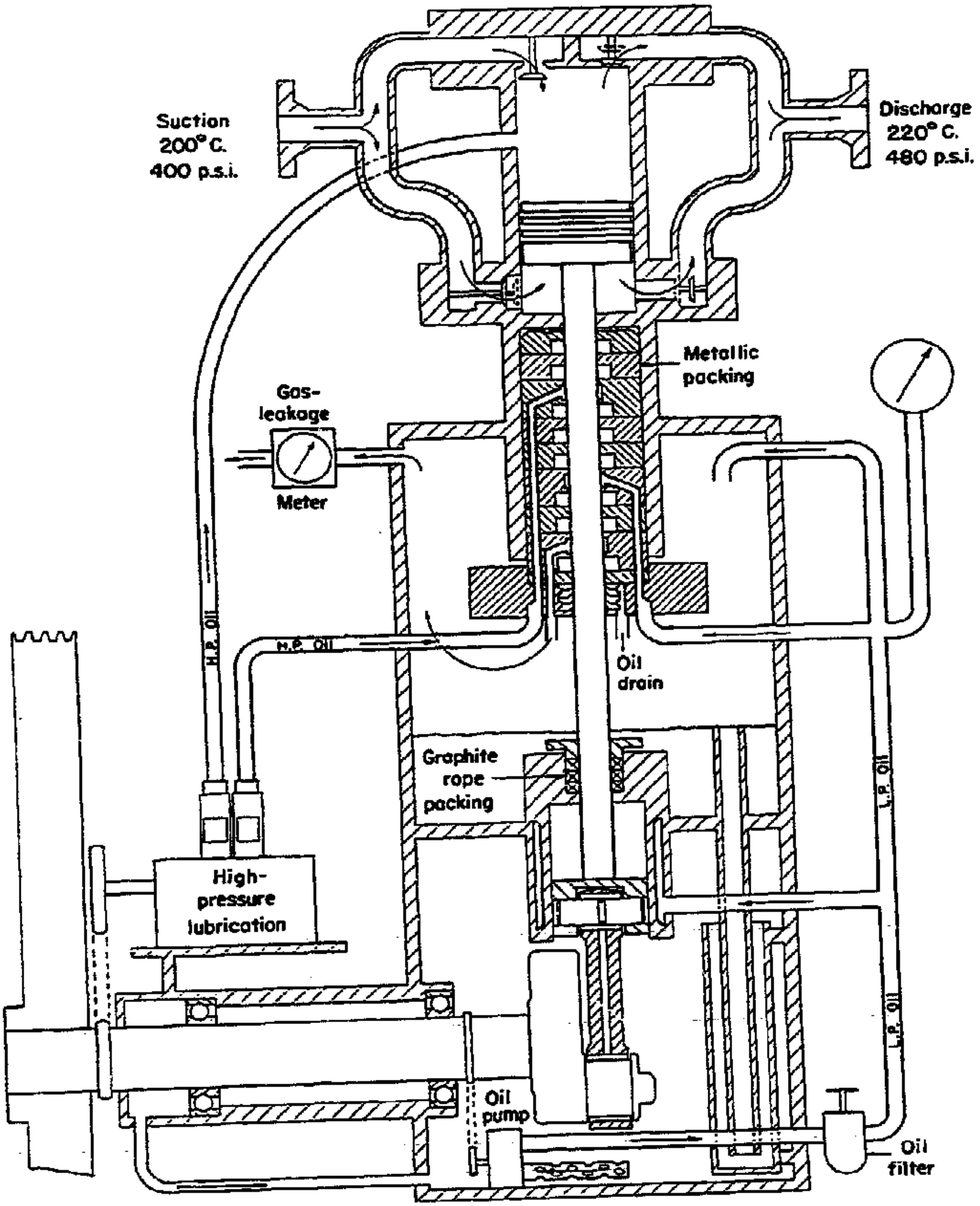

FIGURE 3. - Hot-Gas-Recycle Compressor System.

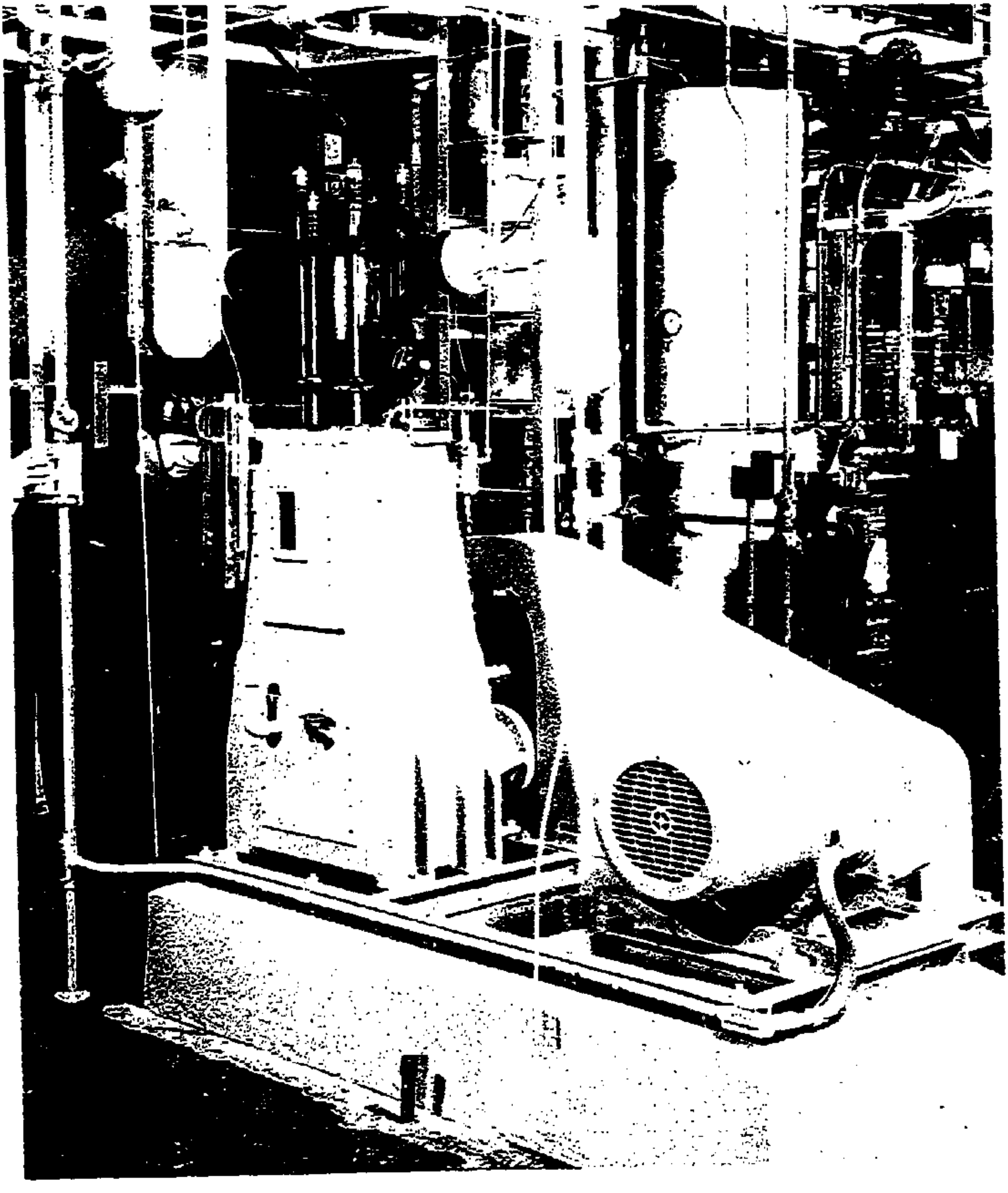

FIGURE 4. - Hot-Gas-Recycle Compressor.

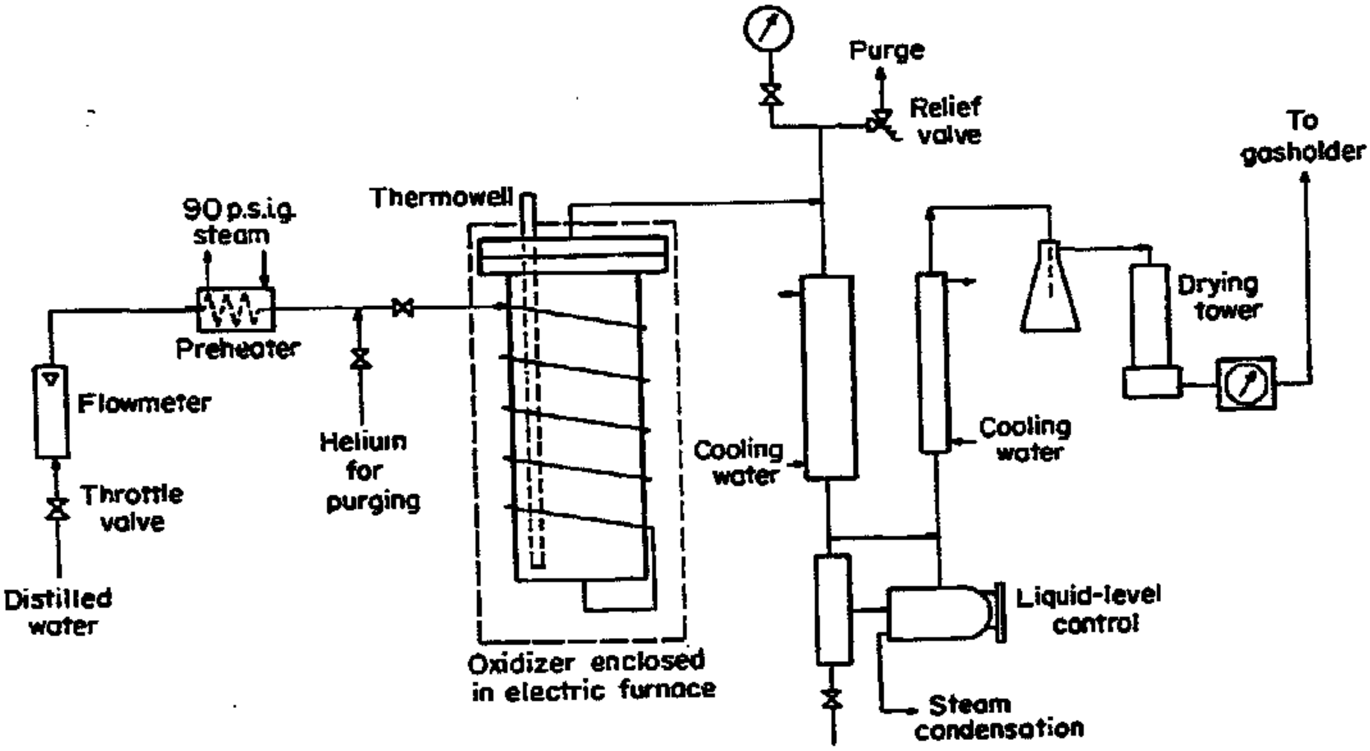

FIGURE 5. - Oxidation Unit for Preparing the Steel Catalyst for the Hot-Gas-Recycle Process.

was immersed in the solution determined the potassium content of the catalyst. The potassium content of catalysts, expressed as potassium oxide, varied from 0.04 to 0.36 pct.

Reduction

The catalysts were reduced after oxidation and impregnation. The reductions were conducted in the reactor of the pilot plant. Usual conditions were: Thirty cubic feet per hour of fresh hydrogen, 2,000 cu.ft./hr of recycle hydrogen, 100 p.s.i.g. reactor pressure, and 400° C. average temperature. The recycle stream was cooled to remove water and then reheated. The temperature was limited by the heat transfer fluid (dowtherm) used in the preheater. All reductions were 40 to 60 hours in duration. The phase present on the surface of the catalyst after reduction was shown by X-ray diffraction to be predominantly alpha-Fe.

Synthesis

After reduction the catalysts were usually inducted in the reactor to bring the system to operating conditions. Induction was accomplished by operating 10 days at the following conditions: An hourly space velocity of 200 to 250, an average bed temperature of 260°-270° C., a reactor pressure of 400 p.s.i.g., and a hot-gas-recycle rate sufficient to maintain the temperature differential over the catalyst bed at 10° C. The feed gas had a hydrogen-carbon monoxide ratio of 1.0. After the induction period, the hydrogen-carbon monoxide ratio of the fresh gas was changed to 1.3-1.5, the hourly space velocity was increased while maintaining a CO_2-free contraction of 65 to 70 pct., and the average bed temperature was raised to 300°-320° C. The recycle rate was adjusted to give a differential across the bed of 20° C. The cold recycle stream was generally 5 pct. of the total recycle. Samples of gaseous and liquid products were taken at periodic intervals after steady-state conditions were obtained. Weight balances were also made during these periods.

Catalyst Analysis

Catalyst samples were analyzed by X-ray diffraction for the determination of the solid phases present. These samples were taken from the top of the catalyst bed during synthesis by using the sampler shown in figure 2. The apparatus was attached to the top of the reactor and the magnetic sampler was lowered to recover sufficient turnings for X-ray analysis. When steel wool was used as a catalyst, similarly treated turnings were placed above the bed of steel wool to supply representative catalyst samples. The catalyst was sampled after reduction, activation, and during synthesis.

Product Evaluation

Liquid product was drained periodically from the unit during steady-state periods, and samples were taken of these products for analysis. Oil samples were analyzed for bromine numbers, specific gravities, ASTM distillations, and other pertinent information. Usually the liquid was distilled into 50° C. cuts which were then analyzed by infrared spectrometry for determination of the oxygenate and olefin content. Water samples from the decanted liquid were also analyzed by mass spectrometry for their oxygenate content--alcohols, ketones, and acids.

Proportionated product gas and feed gas samples were collected in gas-holders over a 24-hour period. Gas from these holders was analyzed by mass spectrometry. The carbon dioxide content of the tail gas was determined directly by hourly Orsat analysis.

Chromatographic analysis was used occasionally to determine the branched hydrocarbon content of the gaseous products.

DISCUSSION OF RESULTS

Operating Variables

The operating variables investigated in these experiments are those affecting product distribution, pressure drop through the catalyst bed, and catalyst activity. To standardize the experiments, an equal volume (0.275 cu.ft.) of catalyst was used. All catalysts had been oxidized before reduction with hydrogen. Table 1 shows a summary of experiments 7 to 17. This table lists the purpose of the experiments, type of catalyst used, induction techniques, and the important conditions of synthesis.

TABLE 1. - Summary of experiments 7 to 17

Experiment No.	Main purpose	Catalyst[1]				Catalyst induction	Synthesis gas			
		Type[3]	Voids, pct.	Weight of charge, g.	K_2O content, pct.		Fresh[2] gas space velocity vol./vol.-hr.	Fresh gas ratio, H_2:CO	Hot recycle: Fresh feed ratio	Cold recycle: Fresh feed ratio
7	First experiment with hot-gas compressor.	1018 steel	89.5	6,433	0.05	None. Temperature raised gradually to 305° C. in 76 hr.	600	1.5 and 1.3	60 and 30	2.5
8	To obtain better activity by induction technique.	1018 steel	88.5	6,990	.04	148 hr. at 270° C. at 200 S.V.H. with $1H_2$:1CO feed gas.	200 to 1,000	1.3	60	1.4 and 3.0
9	To decrease methane yield by increasing K_2O content of catalyst.	1018 steel	88.5	6,990	.16	280 hr. at 270° C. at 200 S.V.H. with $1H_2$:1CO feed gas.	200 to 1,000	1.3	60	2.5
10	Test activity of steel wool catalyst.	No. 6 steel wool	97.0	1,958	.16	None, because of low activity.	200	1.0 and 1.4	100	0 and 1
11	Test activity of steel wool catalyst.	No. 6 steel wool	97.0	1,962	.12	None, because of low activity.	200 to 265	1.0 and 1.4	160 and 80	3
12	Compare use of high void turning to steel wool. Use of bauxite to up-grade products.	1018 steel	96.0	2,469	.36	138 hr. at 275° C. at 200 S.V.H. with $1H_2$:1CO feed gas.	200 to 300	1.0 and 1.3	60	2.5
13	Check four process variables: (1) Reforming gaseous products with alumina; (2) 1.4:1 synthesis gas for induction; (3) Improve olefin yield using activated carbon; (4) Operate with hot recycle at 250° C.	1018 steel	88.0	7,177	.13	144 hr. at 270° C. at 200 S.V.H. with $1.4H_2$:1CO feed gas.	200 to 800	1.4	40 and 20	2.0 and 0.6
14	Induction studies and alumina reforming.	1018 steel	88.0	7,200	.17	240 hr. at 274° C. at 300 S.V.H. with $1H_2$:1CO feed gas.	200 to 1,000	1, 1.4 and 2	100 and 40	1.9 and 0.9
15	Induction studies and alumina reforming, repeat of experiment 14.	1018 steel	88.0	7,200	.17	Same as Experiment 14.	250 to 1,000	1, 1.4 and 2	77 and 45	2.1 and 3.5
16	Induction studies, and lower recycle ratios.	1018 steel	88.0	7,200	.17	240 hr. at 270° C. at 200-250 S.V.H. with $1.4H_2$:1CO feed gas.	200 to 1,000	1.4 and 3	93, 44, 16	2.0 and 4.5
17	No induction, multiple feed.	1018 steel	88.0	7,200	.18	None. Temperature raised gradually to 320° C. in 43 hr.	1,000 and 1,500	1.4	43 and 21	2.4

[1] Catalyst volume equals 0.275 cu.ft. [2] Volume gas per volume catalyst-hour. [3] Steel turnings except as noted.

TABLE 1. - Summary of experiments 7 to 17 (Con.)

Synthesis (Con.)												
Initial conditions					Final conditions					Duration, hr.	General comments	Experiment No.
Reactor temperature, ° C.			H_2+CO conversion, pct.	Usage ratio	Reactor temperature, ° C.			H_2+CO conversion, pct.	Usage ratio			
In gas	Out gas	Average			In gas	Out gas	Average					
300	316	305	75.3	0.97	311	329	315	90.7	1.12	682	Activity lower than experiments with cold recycle compressor.	7
298	318	305	89.4	1.15	321	341	329	90.1	1.15	979	Good activity; methane yield too high.	8
320	340	330	93.9	1.22	310	330	318	90.6	1.15	2,265	Lowered methane yield by increasing K_2O content of catalyst.	9
300	303	301	(1)	(1)	393	399	396	(1)	(1)	191	Unable to get good activity.	10
349	359	353	71.5	0.96	379	401	390	89.7	1.22	366	Low activity. Shut down due to carbon build-up.	11
310	330	318	87.0	1.11	-	-	-	-	-	330	Catalyst poisoned by bauxite trap operation.	12
310	330	319	85.3	1.13	311	329	318	51.1	1.85	1,145	Catalyst poisoned.	13
310	331	317	93.4	1.44	310	330	317	79.9	1.52	393	Catalyst poisoned.	14
310	330	318	93.6	1.29	310	330	321	91.2	1.78	740	Gas yields higher than normal. The effect of alumina reforming was negligible.	15
310	331	319	92.1	1.28	310	350	334	91.3	2.67	920	Unit operated satisfactorily with lower recycle ratios.	16
310	330	321	88.0	1.17	300	340	317	87.5	1.17	711	Unit operated satisfactorily with multiple feed.	17

[1] Not calculated.

As shown in table 1, steel turnings with a void volume of from 88 to 90 pct. were used in experiments 7, 8, 9, and 13 through 17. In experiments 10 and 11 steel wool with 97 pct. voids and in experiment 12 steel turnings with 96 pct. voids were used.

Product Distribution

Tables 7 to 12 (in the Appendix) contain the operating and yield data for the steady-state periods of experiments 7 to 17. Included in the tables are the temperature and temperature differential, pressure and pressure differential, gas conversions, analyses of the product gas, yields and distribution of products, and weight balances. The product distribution groups are C_1 plus C_2 gases, C_3 gas (propane), gasoline (propylene to 204° C.), diesel oil (204°-316° C.), and fuel oil and wax (>316° C.). The oxygenated hydrocarbons and gaseous olefins are also reported.

Table 2 shows the oxygenates found in the product obtained from experiment 17. The products are typical of these experiments. Ethyl alcohol is the major component. Methanol and higher hydrocarbons are also produced. In addition to the alcohols, acetone, methyl-ethyl ketone, acetic acid, and propionic acid are found in the water phase. Esters, ketones, and higher molecular weight alcohols and acids are found in the oil phase.

Effect of Variables

The yield of oxygenates in the products from the hot gas recycle experiments is low and operating variables do not affect the yield appreciably. Usually less than 9 g./m.^{3} of H_2+CO converted was obtained, about the same as in the oil circulation process. These yields may be compared with that of 52 g. obtained with a nitrided catalyst[16] which is a specific catalyst for the production of oxygenates.

The effect of variables on product distribution is shown in table 3, which is a condensation of tables 7 through 12. It lists the variables that produce appreciable changes in product distribution. These changes are also shown in graph form in figure 6.

The operating variables include temperature, the H_2:CO ratio of the fresh gas, the fresh gas rates (hourly space velocity), and water and carbon dioxide content of the recycle gas. The effects of chloride inhibition and potassium oxide impregnation are also discussed. Pressure was held constant at 400 p.s.i.g. in these experiments. The effect of pressure on the Fischer-Tropsch synthesis has been discussed by Benson.[17]

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[16] Bienstock, D., Field, J. H., Forney, A. J., Myers, J. G., and Benson, H. E., The Fischer-Tropsch Synthesis in the Oil-Circulation Process: Experiments with a Nitrided Fused-Iron Catalyst: Bureau of Mines Rept. of Investigations 5603, 1960, 32 pp.

[17] Work cited in footnote 5, p. 2.

TABLE 2. - Oxygenates and olefins found in experiment 17, products in aqueous phase

Period.........	A	B	C	D	E	F	G
Product water:							
Specific gravity.....	0.978	0.991	0.981	0.982	0.982	0.981	0.985
Acid number..	17.5	14.0	6.9	8.3	11.4	13.7	22.3
Refractive index.......	1.3422	1.3412	1.3409	1.3410	1.3417	1.3420	1.3413
Product yield, g./m.^{3} H_2+CO converted:							
Water........	114.2	124.0	129.8	121.3	122.8	123.9	122.8
C_1OH.........	0.9	0.8	0.8	1.0	0.8	0.9	0.8
C_2OH.........	9.2	9.2	8.1	9.9	10.6	12.2	8.4
C_3OH.........	2.3	2.9	2.6	2.9	2.8	2.6	2.1
C_4OH.........	0.1	0.3	0.3	0.4	0.6	0.4	0.4
C_5OH.........	0.3	0	0	0.3	0.3	0.2	0.1
C_6OH.........	0.1	0.3	0.3	0.1	0	0.1	0.1
Acetone......	0.5	0.7	0.5	0.4	0.4	0.4	0.3
Methyl ethyl ketone......	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Acetic acid..	1.6	1.0	0.5	0.6	0.7	0.7	1.6
Propionic acid........	0.3	0.3	0.1	0.1	0	0.1	0.3
Total yield	129.6	139.6	143.1	137.1	139.1	141.6	137.0

Products in oil phase, period D
(g./m.^{3} H_2+CO converted)

	Product, net grams	Acid	Ester	Ketone	Terminal olefin	Internal olefin	Branched olefin
F.D. - 100° C.........	26.9	0.09	0.10	0.12	13.28	4.24	0.87
100° - 150° C.........	15.2	.07	.10	.36	11.62	2.44	.44
150° - 200° C.........	10.2	.07	.10	.26	6.47	1.54	.35
200° - 250° C.........	4.6	.01	.03	.10	2.39	0.83	.15
250° - 300° C.........	5.0	0	.06	.13	2.22	1.11	.10
> 300° C.........	12.3	0	.27	.28	3.01	2.53	.65
Total.............	74.2	0.24	0.66	1.25	38.99	10.30	2.47

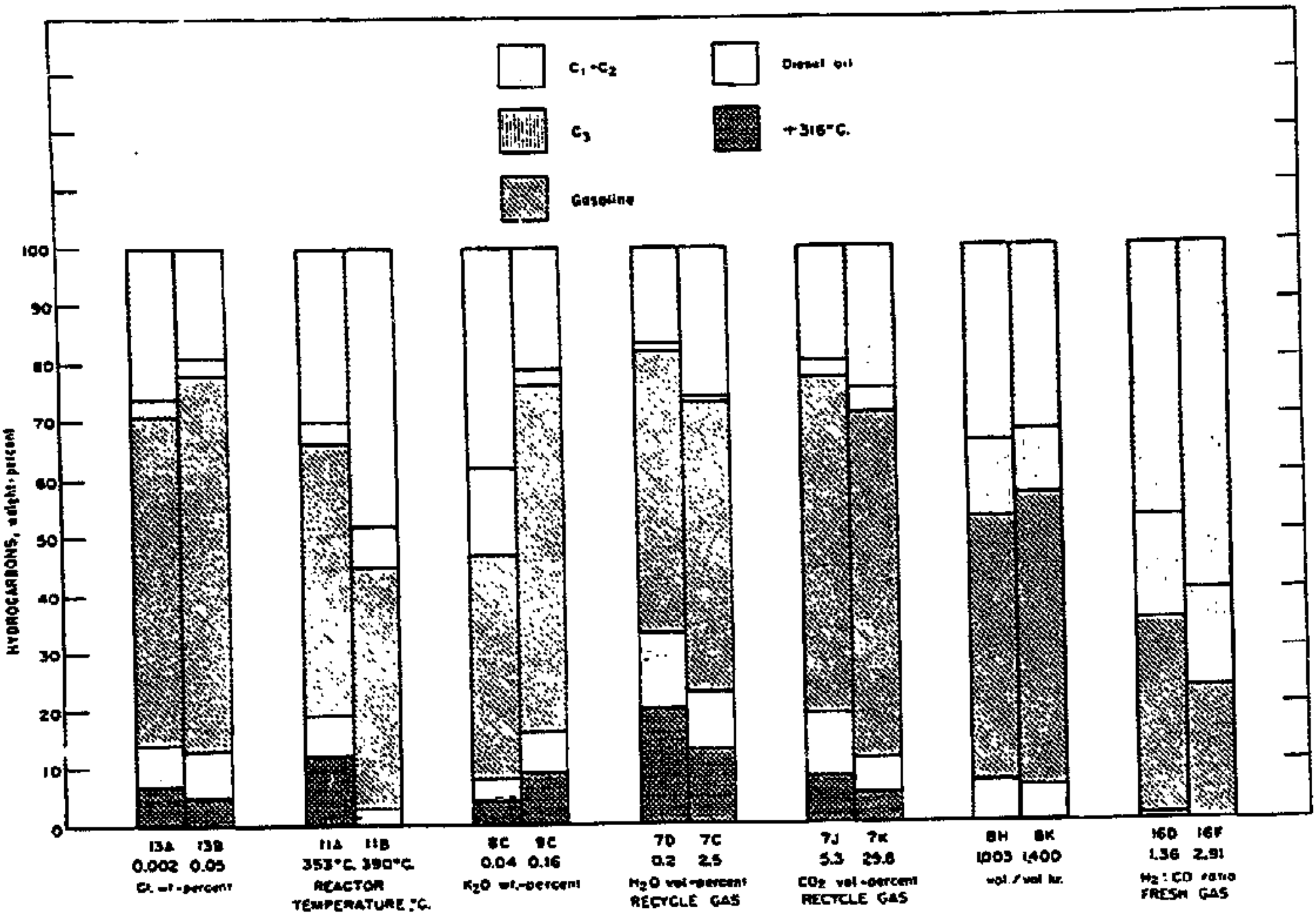

FIGURE 6. - Effect of Variables on Product Distribution.

Reactor Temperature

When the reactor temperature was raised from 353° to 390° C. in experiment 11, the yield of gaseous hydrocarbons increased and the heavier hydrocarbons decreased. As shown in table 3, the C_1 plus C_2 yield increased from 30.1 to 48.3 pct., while the gasoline fraction decreased from 46.4 to 42.1 pct. The fraction boiling over 316° C. decreased from 12.2 to 0.3 pct. Increasing the reactor temperature had little effect on the yields of oxygenates and olefins. In the oil circulation process[18] more gaseous hydrocarbons and gasoline are made as the temperature is increased. However, the temperature range in the oil circulation unit was 235° to 270° C., and at these lower temperatures the yield of hydrocarbons boiling above the gasoline range is higher.

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[18] Benson, H. E., Field, J. H., Bienstock, D., and Storch, H. H., Oil Circulation Process for Fischer-Tropsch Synthesis: Ind. Eng. Chem., vol. 46, 1954, p. 2284.

TABLE 3. - Effect of variables on product distribution[1]

	Reactor temperature change		Fresh gas H_2:CO ratio		Hourly space velocity		Percent water in recycle gas		Percent CO_2 in recycle gas		Chlorine deposition on catalyst		Potassium impregnation	
Experiment and period	11-A	11-B	16-D	16-F	8-H	8-K	7-D	7-C	7-J	7-K	13-A	13-B	8-C	9-C
Recycle to fresh feed ratio:														
Total	162	81	17	16	60	60	67	65	60	60	44	35	105	108
Cold	2.6	2.8	2.6	4.5	2.8	2.9	67	2.5	2.5	0	2.2	2.2	3.0	2.4
Average reactor temperature ° C.	353	390	305	313	300	329	305	310	319	319	319	319	303	315
CO_2 in recycle gas pct.	4.8	5.0	5.0	2.4	10.1	10.1	6.4	4.9	5.3	29.8	5.3	1.6	4.6	5.3
Water in recycle gas pct.	-	-	4.3	4.9	4.1	4.0	0.2	2.5	5.0	7.3	5.1	7.3	3.4	4.8
Space velocity vol./vol.-hr.	265	244	1,001	1,000	1,003	1,400	400	400	600	600	799	800	400	399
H_2:CO ratio:														
Fresh gas	1.41	1.23	1.36	2.91	1.29	1.30	1.28	1.31	1.29	1.29	1.40	1.46	1.28	1.33
Usage	0.96	1.02	1.26	2.61	1.16	1.15	1.21	0.97	1.09	0.73	1.13	1.73	1.16	1.18
Recycle	4.32	3.58	10.0	22.4	6.51	4.03	1.50	3.21	8.04	7.68	8.31	0.88	4.33	3.87
Weight-percent of hydrocarbons:														
C_1+C_2	30.1	48.3	47.3	59.5	33.9	32.1	17.3	25.7	19.5	24.9	25.6	18.8	38.0	21.4
C_3	4.2	6.8	16.5	17.5	13.0	11.2	0.5	1.3	3.1	3.7	3.3	2.5	15.2	3.1
Gasoline (C_3= - 204° C.)	46.4	42.1	35.5	23.0	45.8	50.6	48.9	49.5	58.1	59.8	57.7	66.2	39.4	60.3
Diesel oil (204°-316° C.)	7.1	2.5	0.7	0	6.8	5.8	13.0	10.2	11.3	6.7	6.7	7.7	3.6	10.4
Fuel oil + wax (316° C. +)	12.2	0.3	0	0	0.4	0.3	20.3	13.3	7.4	4.9	6.7	4.8	3.8	4.8
Specific yield, g./m.^{3} H_2+CO converted:														
Oxygenates	8.8	7.5	2.5	3.8	5.9	5.4	10.8	8.1	9.6	6.2	9.0	2.6	3.3	8.9
C_2-C_6 olefins	41.3	37.5	10.7	7.5	18.1	31.6	28.5	34.6	39.0	74.4	41.3	51.5	16.9	44.3
Potassium oxide content on catalyst wt.-pct.	-	-	-	-	-	-	-	-	-	-	-	-	0.04	0.16
Chlorine content on catalyst wt.-pct.	-	-	-	-	-	-	-	-	-	-	[2]0.002	[3]0.05	-	-

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[1] Underlined figures indicate change in variable.

[2] Chlorine content of oxidized and impregnated turnings similar to that used in experiment 13.

[3] Chlorine content of the discharged catalyst.

H_2:CO Ratio of the Fresh Gas

An increase in the H_2:CO ratio of the fresh gas caused a decrease in gasoline production. In experiment 16 the H_2:CO ratio was increased from 1.36 to 2.91 and the gasoline fraction decreased from 35.5 to 23.0 pct. Simultaneously C_1 plus C_2 gases increased from 47.3 to 59.5 pct. As shown in figure 6, the product distribution shifted to the lighter hydrocarbons. The same result is also shown in table 11 in periods D and E of experiment 15 where the H_2:CO ratio was changed from 1.41 to 2.01.

Throughput of Fresh Gas

As shown in table 3, tests 8H and 8K, an increase in the throughput of the fresh gas had very little effect on the gasoline production or the oxygenate yield, but increased the yield of gaseous olefins. An increase in the fresh-gas flow from an hourly space velocity of 1,003 to 1,400 increased the yield of C_2-C_6 olefins from 18.1 to 31.6 g./m.^{3} of H_2+CO converted.

Water Vapor and Carbon Dioxide Content of the Recycle Gas

An increase in the water vapor or the carbon dioxide content of the recycle gas shifted the product distribution to lighter hydrocarbons with only a slight increase in the gasoline yield. An increase in either component decreases the oxygenate yield but increases the olefin yield.

The carbon dioxide content of the total recycle stream is usually maintained at 5 pct. by scrubbing the cold recycle stream with monoethanolamine. When this scrubbing operation was suspended the carbon dioxide content of the total recycle stream increased from 5.3 to 29.8 pct. The C_2-C_6 olefin yield increased from 39.0 to 74.4 g./m.^{3} of H_2+CO converted. Although the water vapor content also increased 2.3 pct. at this time, it is believed that the large change in olefin content is due primarily to the carbon dioxide increase. Experiment 7 (table 3) supports this belief. Here the water vapor content of the total recycle stream changed from 0.2 to 2.5 pct. (the carbon dioxide decreased from 6.4 to 4.9 pct.); the C_2-C_6 olefin content of the tail gas only increased from 28.5 to 34.6 g./m.^{3} of H_2+CO converted.

Because of possible oxidation of the catalyst at this high concentration of carbon dioxide, resulting in decreased activity, it is undesirable to operate in this manner. A slight increase of the carbon dioxide content from 5 to 10 pct., shown in table 8 for periods F and G of experiment 8, increased the gaseous olefin yield by only 2 g./m.^{3} of H_2+CO converted.

Chloride Deposition on the Catalyst

The catalyst of experiment 13 was presumed to have been exposed to hydrogen chloride that was transferred by the recycle gas from an alumina-containing trap. (The purpose of the alumina trap was to convert the terminal olefins in the recycle gas to internal olefins and to decrease the oxygenate yield. The effect of the alumina, however, was negligible.) The chlorine content of the catalyst was 0.05 pct. after discharge. No analysis was made of the catalyst when charged to the reactor, but an unused catalyst similar to that used in experiment 13 had a chlorine content of 0.002 pct. The gasoline fraction increased from 57.7 to 66.2 pct. of the total hydrocarbon yield after the catalyst was poisoned. This was the largest percentage of gasoline in the product obtained in any of the experiments. Unfortunately, the chloride decreased the activity of the catalyst.

The presence of chloride on the catalyst increased the yield of C_3-C_6 olefins from 41.3 to 51.5 g./m.^{3} and decreased the oxygenate yield from 9.0 to 2.6 g./m.^{3} The effects of halogen compounds on Fischer-Tropsch catalysts are described by Davis and Wilson.[19] They state that these compounds on unalkalized catalysts cause the C_2-C_4 olefinic fraction to increase, the overall conversion to decrease, and the methane and carbon dioxide yields to decrease. These experiments corroborate their findings.

Potassium Impregnation

The addition of potassium to a catalyst increases the yield of liquid and solid hydrocarbons, and the yield of oxygenates and olefins. An increase in potassium content from 0.04 pct. in experiment 8C to 0.16 pct. in experiment 9C increased the gasoline yield from 39 to 60 pct., as shown in table 3. Figure 6 illustrates how the product distribution was shifted toward higher boiling compounds. As the percentage of C_1 plus C_2 and C_3 decrease, the yield of liquid and solid products increases. This shift is also shown in table 12 in experiment 17. When potassium hydroxide dissolved in alcohol was added to the catalyst in periods E and F, the yield of liquid and solids increased.

As the catalyst ages, the active coating on the surface of the catalyst flakes off. As a result of this loss of alkali, the yield of gaseous hydrocarbons increases with age of catalyst. This effect is illustrated in table 9, experiment 9. After 612 hours of synthesis the percentage of the product above C_3 was 75 pct.; after 1,859 hours it was 65 pct.

Table 3 also shows that when the potassium oxide content was increased from 0.04 to 0.16, the oxygenate and C_2-C_6 olefin yield increased. The olefin yield increased from 16.9 to 44.3 g./m.^{3} of H_2+CO converted, and the oxygenate yield from 3.3 to 8.9 g./m.^{3} of H_2+CO. The effect of alkali on Fischer-Tropsch catalyst has been discussed by Anderson and others.[20]

In summation, the highest yields of C_1 plus C_2 gases were obtained in experiment 11C (75.4 pct.) operating at a very high temperature of 390° C., and in experiment 16 (61.5 pct.) operating with a very high (2.94) H_2:CO ratio. The highest yield of gasoline was obtained in experiment 13B (66.2 pct.) after a chloride had been deposited on the catalyst.

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[19] Davis, H. G., and Wilson, T. P., U.S. Patent 2,717,259, Sept. 6, 1955.

[20] Anderson, R. B., Seligman, B., Shultz, J. F., Kelly, R., Elliott, M. A., Fischer-Tropsch Synthesis--Some Important Variables of the Synthesis on Iron Catalysts: Ind. Eng. Chem., vol. 44, 1952, pp. 391-397.

Pressure Drop

The pressure drop through the catalyst bed is affected by the shape of the catalyst, the linear velocity, and the density of the gas. A comparison of steel turnings with 88 pct. voids and steel wool with 97 pct. voids[21] shows that the pressure drop is about 50 pct. less with steel wool. However, turnings of 88 pct. voids were employed as catalyst because turnings and wool of greater void volume were less active, both on a weight and a volume basis.

As compression costs are a function of both the recycle flow and pressure drop across the bed, any reduction in either factor would be desirable. However, the flow must be sufficient to maintain stable temperature control. In experiment 16, shown in table 12, as the total recycle-to-fresh feed ratio was reduced from 44 in period C to 17 in period D, the pressure drop per foot of catalyst bed height decreased from 46 to 17 in. of water, a decrease of 63 pct. The temperature differential increased from 20° to 40° C., but there was no difficulty in maintaining control of the temperatures.

Further reduction in pressure drop may be achieved by splitting the recycle stream and proportionating the flow along the reactor length. As only part of the gas traverses the entire length of the bed, a reduction in the overall pressure drop across the bed occurs. In experiment 17 (table 12) when the gas feed was changed from single entry at the top of the reactor in period A to multiple feed (with the gas flow divided into 3 equal parts) in period B, the pressure drop decreased from 26 to 19 in. of water although the total recycle flow was constant at about 40 to 1. Continuing the multiple feed system in period C, the recycle ratio was decreased from 40 to 21 and the pressure drop decreased to 8 in. of water per foot of catalyst bed. Thus a combination of lower recycle and multiple feed along the length of the reactor bed results in a very low pressure drop while still maintaining a high gas conversion and the same product distribution.

Another factor affecting the pressure drop is the gradual accumulation of fines during synthesis as a result of carbon deposition and the flaking of the catalyst surface. The fines collect in the lower section of the catalyst bed and cause an increased pressure drop. This factor is illustrated in figure 7.

The pressure drop increased from 16 in. of water per foot of catalyst bed at 300 hours of operation to 70 in. after 550 hours--a fourfold increase in 250 hours. The calculated pressure drop[22] across the turnings, (neglecting the presence of fines) remained approximately the same, 12 to 15 in., during this period. This problem of pressure drop increasing with catalyst aging was present in almost all of these experiments: The pressure drop in experiment 7 increased 55 pct. in 682 hours of operation, in experiment 9, 36 pct. in 1,007 hours, and in experiment 16, 120 pct. in 416 hours. The comparisons of pressure drop are made at similar conditions of recycle and fresh feed flows.

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[21] Jimeson, R. M., and Decker, W. A., Pressure-Drop Measurements Through Beds of Granular Iron, Lathe Turnings, Steel Wool, and Parallel-Plate Assemblies: Bureau of Mines Rept. of Investigations 5541, 1959, 51 pp.

[22] Work cited in footnote 20.

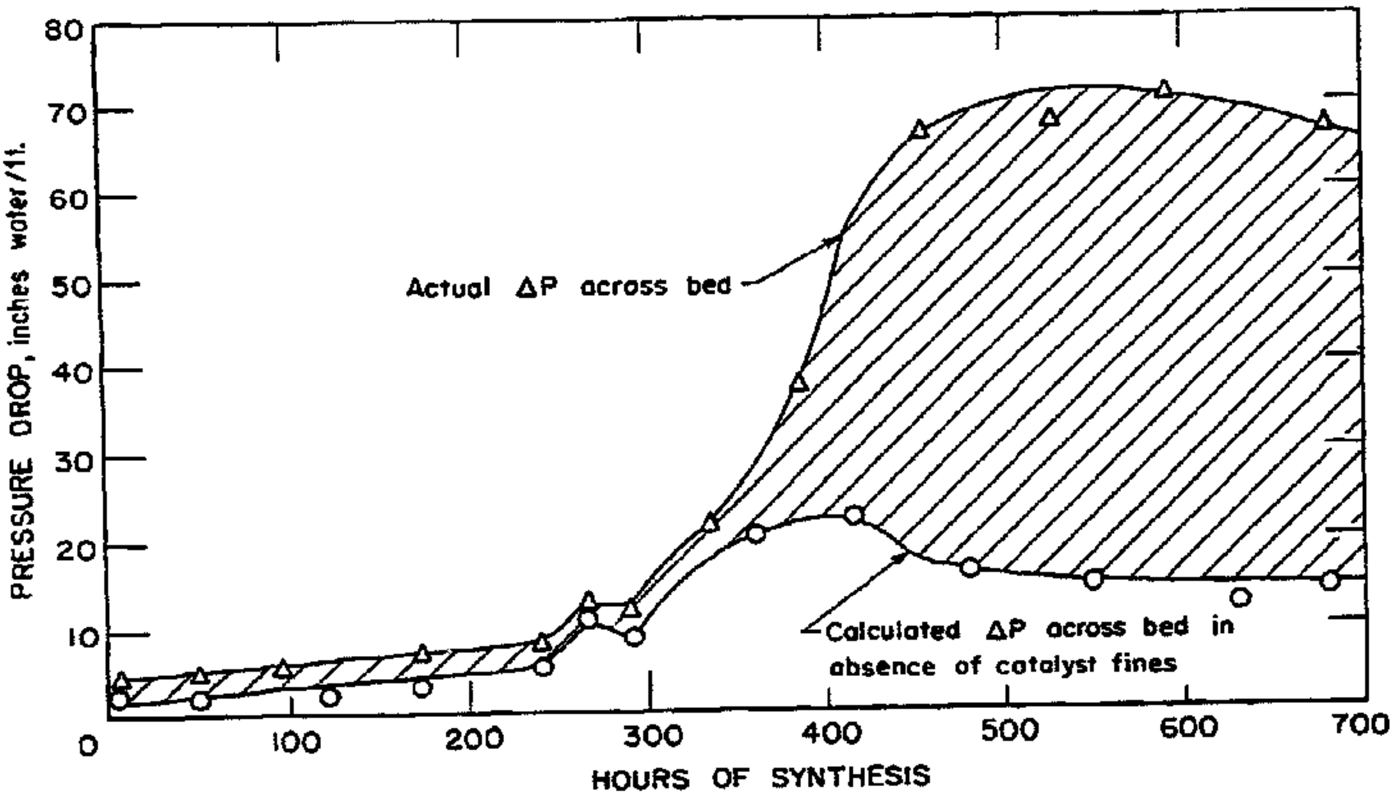

FIGURE 7. - Pressure Drop Across a Bed of Steel Lathe Turnings, Experiment 15.

Although there is sufficient evidence to indicate that the pressure drop increase is due to carbon deposition and catalyst flaking, this difficulty could not be avoided during these experiments. In later experiments, particularly experiment 21, high water vapor concentration in the recycle gas was found to prevent this condition.[23]

However, with the use of turnings, recycle-to-fresh feed flows of 20 to 1 and multiple gas entry to the reactor, a pressure drop of less than 0.5 p.s.i./ft. of catalyst height has been obtained. This is sufficiently low so that it contributes a minor percentage to the cost of gasoline made by this process.

Catalyst Activity

A method of indicating catalyst activity is to calculate activity coefficients A_v in terms of volumes of feed gas (H_2+CO) converted per hour per volume of catalyst referred to a standard reference temperature.[24] These coefficients A_v (activity per unit volume of catalyst) and A_{Fe} (activity per gram of iron) are shown in table 4. Another method of assessing activity at the

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[23] Bienstock, D., Field, J. H., Forney, A. J., and Demski, R. J., Pilot Plant Development of the Hot-Gas-Recycle Process for the Synthesis of High-B.t.u. Gas: Bureau of Mines Rept. of Investigations 5841, 1961, 27 pp.
[24] Work cited in footnote 19, p. 17.

conditions used in these experiments is to consider the conversions that can be achieved when operating with maximum space velocity at the same temperature. By either of these methods the most active catalysts were those of experiments 8, 9, 15, 16, and 17. These five experiments had values of A_V greater than 140, and A_{Fe} values greater than 154, and had conversions of feed gas greater than 90 pct. when operating at an hourly space velocity of 1,000 at an average temperature of 319° C. or lower. The catalysts of the other experiments had A_V values less than 87, and none could be operated at an hourly space velocity of 1,000 with 90 pct. conversions.

The more active catalysts on both A_{Fe} and A_V bases were those which had reactor charges of 15 and 16 lb. compared with 4 to 5 lb. used in experiments 10, 11, and 12. The catalysts used in experiments 13 and 14 exhibited low activity because of poisoning by chlorides. No reason for the low activity of the catalyst in experiment 7 is known. While sulfur poisoning was originally suspected, the possibility was discounted because the yield of C_2-C_8 olefins was high, 29 to 50 g./$m.^3$ Davis[25] states the effect of sulfur on iron-based catalysts is to decrease formation on unsaturated gaseous hydrocarbons, so it may be concluded that in this experiment sulfur poisoning was not a factor.

The effect of alkali impregnation in the range from 0.04 to 0.36 pct. as potassium oxide upon catalyst activity is not significant in the temperature range used in these experiments. The catalyst in experiment 8 with a potassium oxide content of 0.04 pct., had as high an activity as the catalyst in experiment 9 with 0.16 pct. In experiment 17 (table 12) potassium as KOH was added to the catalyst during periods E and F. The H_2+CO conversion was about the same (91.9 pct. compared with 91.2 pct.) before and after this addition, but at these high conversions little increase can be expected.

In most of these experiments an induction period was used after the reduction to get the system on stream. The temperature of the reaction was raised gradually to 270° C. and maintained at that level for 10 days, then increased to the desired operating temperature. The feed gas during the induction period was usually $1H_2$+1CO. In experiment 15 the feed gas used during the induction period was $1.4H_2$+1CO, the gas used during the main part of the experiments. In experiment 16 the hydrogen-rich gas was again used and the temperatures were raised as fast as practical to the desired operating range. The catalysts used in these two experiments had activities (A_V) greater than 140, indicating that the method of induction of the catalyst has little effect on its activity.

Hägg Fe_2C was the major form of iron on the surface of the catalysts shown in table 4. The difference in activity observed in these tests was not related to the carbide content of the catalyst.

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[25] Davis, H. G., and Wilson, T. P., U.S. Patent 2,717,260, Sept. 6, 1955.

TABLE 4. - Activity data for experiments 7 through 17

Experimental period	Space velocity, vol./vol.-hr.	Average catalyst temperature, ° C.	Differential temperature, ° C.	H_2+CO conversion, pct.	Operation, hr.	Voids, pct.	Catalyst charge, lb.	Alkali content, wt.-pct. K_2O	H_2:CO ratio of fresh gas	A_v[1]	A_{Fe}[2]	Phases, present[3]
7-L	600	315	18	90.7	674-682	89.5	14.2	0.05	1.3	81	99	Hägg Fe_2C.
8-F	1,003	300	20	92.2	451-547	88.5	15.4	.04	1.30	201	224	Do.
9-K	1,001	319	20	93.6	852-972	88.5	15.5	.16	1.35	141	157	Do.
10[4]	(b)	(5)	-	-	-	97	4.3	.16	1.4	-	-	-
11[4]	206	305	12	78.6	72-96	97	4.3	.12	1.6	24	104	-
12-A	300	318	20	87.0	186-258	96	5.4	.36	1.4	32	102	Hägg Fe_2C, Fe, Fe_3C.
13-A	799	319	20	85.3	392-464	88	15.8	.13	1.40	78	83	Hägg Fe_2C.
14-B	1,000	317	20	79.9	359-393	88	15.9	.17	1.84	86	92	Hägg Fe_2C, Fe.
15-B	1,002	318	20	93.6	396-468	88	15.9	.17	1.40	145	155	Do.
16-E	1,000	295	20	92.9	712-760	88	15.9	.17	1.4	265	284	Hägg Fe_2C.
17-D	1,002	309	20	91.9	423-471	88	15.9	.18	1.4	171	190	Do.
17-G	1,501	317	40	87.5	663-711	88	15.9	.18	1.4	168	179	Do.
13-B	802	319	20	76.0	528-557	88	15.8	.13	1.46	59	63	-

[1] Activity coefficient, activity per unit volume.

[2] Activity coefficient, activity per gram of iron.

[3] Determined by X-ray diffraction.

[4] Steel wool catalysts; others are lathe turnings.

[b] No data.

Chlorides reduce catalyst activity. The catalyst used in experiment 13 was poisoned during period B by chlorides present in the alumina used for treating the recycle gas stream. Table 10 shows the H_2+CO conversion decreased from 85.3 pct. before treating the recycle gas in period A to 76.0 pct. after contamination with the chloride (presumably hydrogen chloride). This poisoning effect of halides has been discussed by Davis and Wilson.[28]

Gasoline Processing

A finished gasoline was prepared from period P of experiment 9 in the manner depicted in figure 8. Operating conditions for this experiment are shown in table 9. The heavy and light oil streams were combined and distilled, initially at atmospheric pressure and finally at a reduced pressure of 5 mm. of mercury to prevent cracking of the heavier oil. The fraction boiling below 204° C. was combined with the charcoal spirits (a fraction collected by passing the tail gas from the reactor through activated carbon and recovering the hydrocarbons by regenerating the carbon with steam) stabilized at room temperature, and then reformed over bauxite.

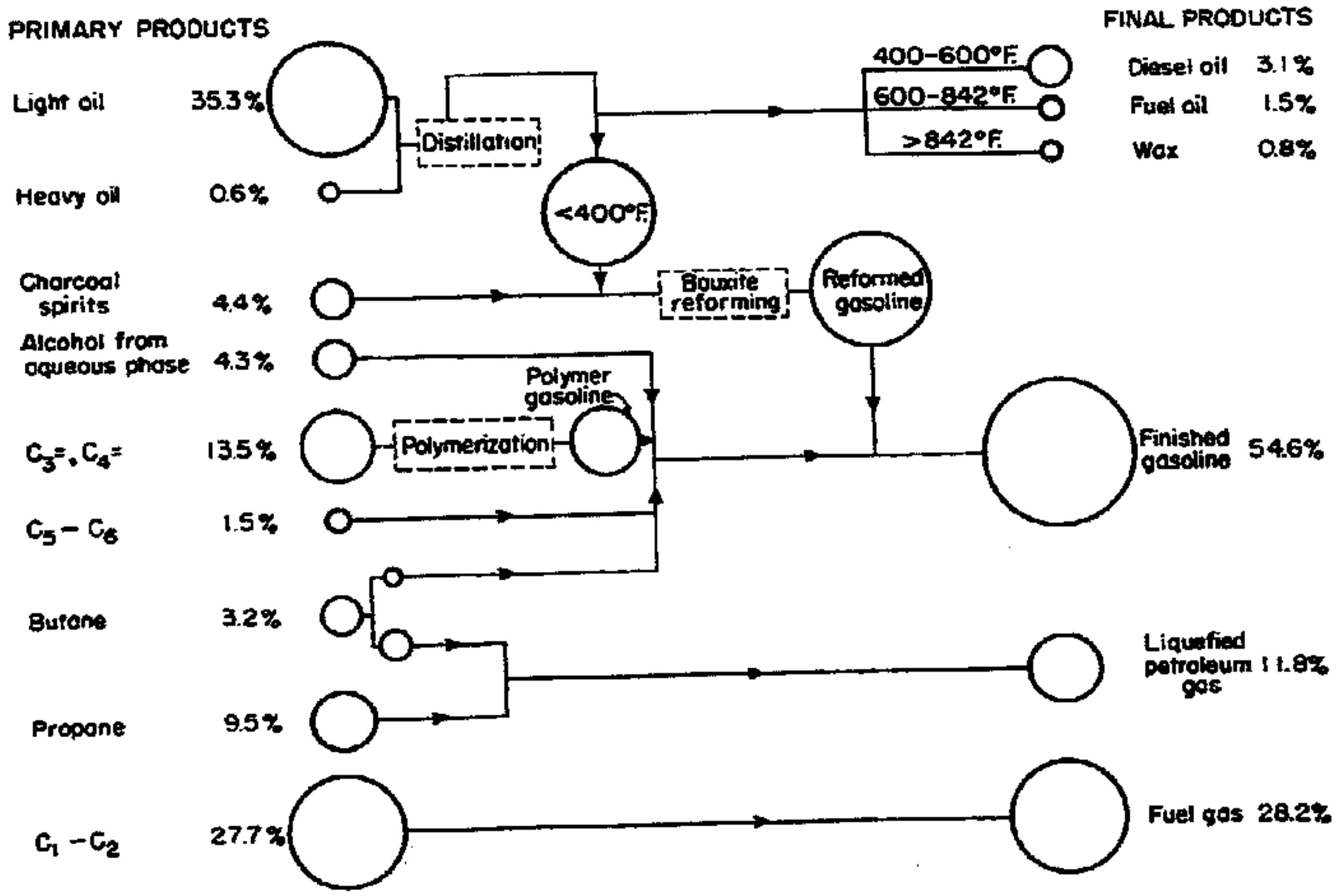

FIGURE 8. - Processing of Synthesis Products to Produce Gasoline, Experiment 9.

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[28] Work cited in footnote 18, p. 14.

The purpose of the bauxite treatment is to upgrade the product by dehydrating the oxygenated hydrocarbons and shifting the double bond in the terminal olefins to an internal position. The conditions for bauxite treatment were as follows: A temperature of 353° C., a pressure of 25 p.s.i.g., and a liquid hourly space velocity of 1.5 vol. of liquid per volume of catalyst per hour. The catalyst used in this treatment was activated bauxite from 4- to 12-mesh. Industrially the C_3 and C_4 olefins would be sent to a polymerization unit for further processing. To simplify the blending procedure, a polymer gasoline, obtained commercially, was added to the reformed gasoline in an amount equivalent to that which would be obtained by polymerizing the C_3 and C_4 olefins. The finished gasoline contained the reformed and polymer gasoline and sufficient butane to adjust the Reid vapor pressure to about 10 p.s.i. To another gasoline sample, prepared in the same manner, the water-soluble alcohols, methyl, ethyl and n-propyl, were added in the amounts produced in the product water. The research octane number of these gasolines was determined as such, and after the addition of 1 and 3 cc. of tetraethyl lead. The results of these tests are shown in table 5.

TABLE 5. - Octane number of the finished gasoline, experiment 9

	Reid vapor pressure, p.s.i.	Research octane number		
		Clear	1 cc. T.E.L.[1]	3 cc. T.E.L.[1]
Polymer gasoline.............	-	95.1	-	-
Reformed + polymer + butane...	9.9	77.1	85.3	90.2
Reformed + polymer + alcohol[2] + butane....................	10.6	81.2	88.5	92.4

[1] Tetraethyl lead.
[2] Alcohol, 8.3 pct. of total blend.

The addition of alcohol, equivalent to 8.3 pct. of the total blend raised the octane number from 77 to 81. Addition of 3 cc. of tetraethyl lead further increased th octane rating to 92. As the use of alkylate is supplanting polymer for upgrading gasoline, the gasoline from experiment 15 was blended with both alkylate and polymer gasoline. As shown in table 6, the blend containing alkylate gasoline imparted a higher octane number than the one with polymer. The alkylate gasoline with 3 cc. of tetraethyl lead had an octane number of 96.

CONCLUSIONS

The hot-gas-recycle pilot plant has been operated successfully with carbon-steel lathe turnings as catalyst. Turnings packed with 88 to 90 pct. voids were the most effective, being of high activity and low pressure drop. Turnings with higher void volumes and steel wool had lower activity.

Good temperature control was achieved operating with a recycle-to-fresh feed ratio as low as 20. At this recycle flow and with use of multiple injection of the gas into the catalyst bed a pressure drop of less than 0.5

p.s.i./ft. of catalyst bed was obtained. The cost of recompression of the recycle gas therefore is small, amounting to a few tenths of a cent per gallon in a commercial sized reactor.

Hourly space velocities of fresh gas as high as 1,500 and conversions as high as 95 pct. were achieved. Successful pilot plant operations of 2,265 hours (over 3 months) duration were obtained. The optimum hydrocarbon production consisted of 66 pct. gasoline, 10 pct. diesel oil, 2 pct. fuel oil and wax, and the balance of 22 pct. gases. This was achieved when operating with a reactor temperature of 310°-330° C. The gasoline containing alkylate had a research octane number of 86; with addition of 3 cc. of tetraethyl lead the octane number was raised to 96.

TABLE 6. - Octane numbers of the finished gasoline, experiment 15, period D

	Reid vapor pressure, p.s.i.	Research octane number		
		Clear	1 cc. T.E.L.	3 cc. T.E.L.
Polymer gasoline	-	96.8	-	101.1
Alkylate gasoline	-	94.2	-	103.8
Reformed + polymer + butane	8.9	84.5	91.4	94.1
Reformed + alkylate + butane	8.4	85.6	93.5	96.4

Gases containing traces of hydrogen chloride reduced the activity of catalysts; however, the product contained a greater percentage of gasoline. An alkali promoter, K_2O, on the catalyst caused a shift in the product distribution, making more gasoline and less gaseous products. The recommended quantity of K_2O to produce the maximum gasoline yield is from 0.13 to 0.18 pct.

The only operating difficulty encountered was the tendency of the catalyst to spall after prolonged synthesis. The fines produced by spalling tended to collect in the lower section of the bed together with carbon formed in the reaction. These deposits decreased the void volume and caused an increased pressure drop. After 2 to 3 months of synthesis the catalyst activity declined due to the spalling of the active layer. As a result of the loss of alkali the production of C_1 plus C_2 gases increased.

Future objectives for the pilot plant are the development of a durable catalyst that will be active for 6 months with little or no change in the product yield or the pressure drop.

APPENDIX

Tables 7 through 12 are included in the appendix.

TABLE 7. - Operating conditions and results, experiment 7

Period	A	B	C	D	E	F	G	H	I	J	K	L
Hours of synthesis	72-120	120-263	340-388	423-431	431-487	503-527	527-551	554-575	611-622	626-646	646-670	674-682
Fresh gas space velocity vol./vol.-hr.	601	603	401	400	400	400	500	600	400	600	600	600
Recycle:fresh gas ratios:												
Total	65	65	65	67	65	28	28	28	65	58	60	58
Cold	2.2	2.5	2.5	67.0	2.9	2.5	2.6	2.3	2.9	2.5	0	2.4
CO_2 scrubbed	2.2	2.5	2.5	2.0	2.9	2.5	2.6	2.3	2.9	2.5	0	2.4
Water vapor in recycle gas, vol.-pct.	3.8	3.5	2.5	0.2	2.5	4.3	4.1	5.0	1.7	5.0	7.3	6.0
Cold recycle, pct. of total	3	4	4	100	4	9	9	9	4	4	0	4
H_2:CO ratios:												
Fresh gas	1.44	1.49	1.31	1.28	1.28	1.17	1.36	1.36	1.38	1.29	1.29	1.31
Recycle gas	11.82	9.91	3.22	1.50	2.71	3.28	5.26	8.73	5.00	8.04	7.68	13.36
Usage	0.97	1.07	0.97	1.21	0.98	1.06	1.13	1.20	1.02	1.09	0.73	1.12
Reactor pressure, p.s.i.g.	403	400	402	400	400	400	400	600	400	400	400	400
Catalyst pressure drop, in. H_2O/ft.	9	8	7	7	4	2	2	3	7	12	13	14
Catalyst temperature, ° C.:												
Top (inlet)	300	300	305	300	303	302	306	308	305	314	314	311
Bottom (outlet)	316	316	321	317	319	342	347	348	322	334	333	329
Differential temperature	16	16	16	17	16	40	41	40	17	20	19	18
Average	305	305	311	305	309	315	318	318	309	319	319	315
H_2 conversion, pct.	62.0	67.8	63.1	73.8	63.7	74.9	77.6	77.9	67.8	81.7	51.7	84.8
CO conversion, pct.	95.4	95.2	85.1	77.9	82.9	90.2	94.3	97.5	91.3	97.1	91.8	98.5
H_2+CO conversion, pct.	75.3	78.8	72.6	75.6	71.9	81.3	84.2	89.6	77.7	88.4	69.2	90.7
Exit gas analysis (dry basis, vol.-pct.):												
H_2	60.2	66.4	62.2	48.1	59.1	57.0	55.3	52.4	56.5	56.3	49.9	54.7
CO	5.0	6.7	19.3	32.0	21.8	17.4	10.6	6.0	11.3	7.0	6.5	4.1
N_2	2.9	0.7	0.8	0.6	0.4	0.5	3.9	1.1	2.4	2.0	1.0	1.9
CO_2	5.0	4.6	4.9	6.4	4.5	5.1	5.1	5.5	5.4	5.3	29.8	5.1
C_1	10.4	10.6	7.6	6.6	7.1	8.6	12.6	18.2	7.6	14.1	6.0	21.5
$C_2^=$	1.7	1.5	1.3	1.1	1.1	1.4	1.5	1.5	1.7	1.5	1.1	0.9
C_2	2.0	2.7	1.5	1.4	1.6	2.3	3.0	4.6	2.0	3.8	1.3	4.3
$C_3^=$	3.0	3.4	1.6	2.0	2.0	3.1	3.3	4.3	4.8	5.1	2.0	3.1
C_3	0.8	0.9	0.2	0.1	0.5	0.9	1.0	1.6	1.3	1.5	0.5	1.6
$C_4^=$	1.2	1.5	0.9	1.1	1.0	1.9	1.9	2.4	2.9	2.3	1.2	1.1
C_4	0.3	0.4	0.1	0.2	0.3	0.5	0.5	0.7	1.0	0.7	0.2	0.4
$C_5^=$	0.3	0.4	0.0	0.3	0.4	0.8	0.8	1.0	1.5	0.2	0.4	0.2
C_5	0.0	0.1	0.0	0.1	0.1	0.2	0.2	0.2	0.4	0.2	0.1	0.1
$C_6^=$	0.0	0.1	0.0	0.0	0.1	0.3	0.3	0.4	0.7	0.0	0.0	0.0
C_6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.0	0.0	0.0	0.0
Yield, g./m.³ H_2+CO converted:												
C_1	32.6	26.5	25.4	11.8	21.9	17.0	22.0	23.7	19.4	18.3	29.8	24.3
$C_2^=$	10.2	7.2	8.2	4.5	7.3	5.8	5.5	4.1	9.7	4.3	10.8	2.2
C_2	17.0	14.0	10.2	6.1	11.4	10.3	11.9	13.1	12.1	11.8	13.9	13.5
$C_3^=$	26.9	23.7	15.9	12.2	19.3	19.0	18.0	16.4	29.1	20.9	30.2	11.5
C_3	7.0	6.8	2.3	0.6	5.3	5.8	5.7	6.3	11.5	6.5	8.0	5.5
$C_4^=$	14.2	14.2	8.5	8.9	13.1	15.0	13.6	12.6	31.1	12.5	23.5	4.7
C_4	2.8	5.6	3.4	2.4	5.6	6.9	5.5	6.7	13.0	5.2	4.2	3.3
$C_5^=$	5.2	4.7	2.0	2.9	6.2	8.4	6.7	6.6	20.4	1.3	4.9	1.0
C_5	0.5	1.8	0.0	0.9	1.3	2.5	2.2	2.4	6.1	1.8	2.6	1.5
$C_6^=$	0.0	2.6	0.0	0.0	0.7	3.4	2.9	3.6	11.1	0.0	0.0	0.0
C_6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.8	0.0	0.0	0.0	0.0
Oil	79.0	63.9	86.6	68.7	90.6	83.4	81.1	64.1	74.4	83.8	79.7	33.0
Aqueous phase	117.9	102.6	81.1	129.7	90.0	117.5	108.8	116.0	60.2	122.2	45.9	137.6
C_1-C_5OH	9.7	8.9	5.9	5.8	6.8	(2)	9.9	8.5	(2)	8.0	4.9	(2)
Other oxygenates[1]	2.7	2.0	2.2	5.0	2.7	(2)	2.9	1.8	(2)	1.6	1.3	(2)
Water	105.5	91.7	73.0	118.9	80.5	(2)	94.0	105.7	(2)	112.6	39.7	(2)
CO_2	346.9	343.2	383.1	166.3	379.2	367.9	345.4	322.6	400.9	291.2	458.5	327.1
Hydrocarbon recovery	207.8	180.9	170.6	129.8	191.8	(2)	184.9	175.7	(2)	176.0	218.8	(2)
Theoretical hydrocarbon recovery	200.8	201.7	202.8	205.5	203.3	203.9	203.0	204.7	200.1	203.7	201.1	202.2
Overall weight balance, pct.	103.4	95.2	93.6	92.1	95.6	101.9	100.4	97.5	102.5	95.9	95.1	84.7
Hydrocarbons recovered, lb./1,000 ft.³ fresh gas	9.7	8.9	7.7	8.0	8.6	(2)	9.7	9.8	(2)	9.7	9.4	(2)
Weight-percent of hydrocarbons recovered:												
C_1-C_3	28.7	26.4	35.7	17.3	21.2	(2)	21.3	23.3	(2)	19.5	24.9	(2)
C_4	3.4	3.8	1.3	0.5	2.8	(2)	3.1	3.6	(2)	4.7	3.7	(2)
Gasoline (C_5-204° C.)	52.6	57.2	40.5	48.9	49.2	(2)	55.8	55.6	(2)	58.1	59.8	(2)
Diesel oil (204°-316° C.)	6.9	6.8	10.2	13.0	8.8	(2)	9.1	9.8	(2)	11.3	6.7	(2)
Fuel oil (316°-450° C.)	2.7	1.8	5.2	2.4	5.0	(2)	3.7	3.2	(2)	2.6	1.9	(2)
Wax (> 450° C.)	5.7	4.0	8.1	17.9	12.8	(2)	7.0	4.5	(2)	2.8	3.0	(2)

[1] Calculated as hydrocarbons.

[2] Liquids not analyzed.

TABLE 8. - Operating conditions and results, experiment 8

Period	A	B	C	D	E	F	G	H	J	K
Hours of synthesis	52-148	208-256	280-304	316-328	354-396	451-567	555-571	579-619	699-907	931-979
Fresh gas space velocity...vol./vol.-hr.	200	400	400	801	1,001	1,003	1,004	1,003	1,202	1,400
Recycle:fresh gas ratios:										
Total	100	105	105	60	60	60	60	60	60	60
Cold	1.8	2.9	3.0	2.3	2.4	2.9	1.4	2.8	2.9	2.9
CO_2 scrubbed	1.8	2.9	3.0	2.3	2.4	2.9	1.4	1.4	1.5	1.5
Water vapor in recycle gas.....vol.-pct.	3	3	4	4	4	4	6	4	4	4
Cold recycle, pct. of total	7	3	3	4	4	5	2	5	5	5
H_2:CO ratios:										
Fresh gas	1.00	0.97	1.28	1.31	1.30	1.31	1.31	1.29	1.31	1.28
Recycle gas	1.11	1.32	4.33	4.51	4.80	8.10	9.37	6.42	3.65	4.06
Usage	0.97	0.93	1.16	1.14	1.15	1.15	1.12	1.16	1.19	1.15
Reactor pressure...p.s.i.g.	400	406	405	407	411	409	409	410	414	408
Catalyst pressure drop, in. H_2O/ft. catalyst	5	16	13	18	33	32	41	43	61	64
Catalyst temperature, ° C.:										
Top (inlet)	266	299	298	299	298	293	298	293	302	321
Bottom (outlet)	278	310	310	319	318	313	318	313	322	341
Differential temperature	12	11	12	20	20	20	20	20	20	20
Average	271	303	303	306	305	300	305	300	309	329
H_2 conversion...pct.	57.0	85.7	87.7	82.7	84.6	86.3	83.6	87.2	86.5	85.9
CO conversion...pct.	58.8	89.5	96.4	95.0	95.8	97.8	97.7	97.5	95.3	95.6
H_2-CO conversion...pct.	57.9	87.6	91.5	88.0	89.4	92.2	89.7	91.7	89.9	90.1
Exit gas analysis (dry basis, vol.-pct.):										
H_2	46.7	36.1	38.5	43.7	41.8	41.3	43.1	38.5	37.2	38.2
CO	42.1	27.4	8.9	9.7	8.7	5.1	4.6	6.0	10.2	9.4
N_2	0.9	1.3	1.5	0.5	0.5	1.1	1.0	0.8	0.5	0.4
CO_2	4.8	5.5	4.6	5.6	5.5	5.1	9.9	10.1	10.4	10.4
C_1	2.2	16.8	29.9	23.6	26.1	27.6	23.6	25.5	23.5	23.7
C_2=	0.6	0.6	0.3	0.3	0.2	0.2	0.2	0.2	0.5	1.2
C_2	0.4	5.0	9.3	7.3	7.8	8.7	7.7	8.1	7.3	5.8
C_3=	1.0	2.8	1.1	1.5	1.5	1.2	1.3	1.6	2.0	2.9
C_3	0.2	2.5	5.3	4.6	5.0	5.8	5.2	5.6	4.9	4.5
C_4=	0.6	1.4	0.3	1.2	1.2	1.3	1.3	1.3	1.5	1.9
C_4	0.1	0.6	0.2	1.2	1.3	1.5	1.3	1.4	1.2	1.0
C_5=	0.3	0.0	0.1	0.4	0.3	0.4	0.4	0.5	0.5	0.5
C_5	0.1	0.0	0.0	0.3	0.2	0.4	0.3	0.2	0.2	0.1
C_6=	0.0	0.0	0.0	0.1	0.0	0.2	0.1	0.2	0.0	0.0
C_6	0.0	0.0	0.0	0.0	0.0	0.1	0.0	0.0	0.0	0.0
Product yield, g./m.^{3} H_2-CO converted:										
C_1	10.9	25.9	41.5	42.1	42.5	39.2	40.0	36.9	36.1	37.9
C_2=	7.2	1.7	0.8	1.0	0.6	0.5	0.1	0.6	1.5	3.6
C_2	5.0	16.3	26.1	26.3	25.0	24.9	26.5	23.2	22.0	18.9
C_3=	18.1	14.2	5.7	7.9	6.8	4.9	6.3	6.3	8.9	13.1
C_3	3.8	12.8	27.4	24.2	24.1	24.1	25.5	23.3	21.2	11.1
C_4=	13.1	10.1	7.2	7.6	7.1	6.9	8.2	6.8	8.8	11.3
C_4	4.1	6.1	9.3	9.2	8.6	8.8	8.8	8.2	7.3	6.6
C_5=	6.4	1.3	2.8	3.2	2.3	2.8	3.0	2.9	3.0	3.6
C_5	1.8	0.3	1.6	2.5	1.2	2.6	2.4	1.7	1.4	0.8
C_6=	1.0	0.0	0.4	1.0	0.0	1.4	0.9	1.5	0.4	0.0
C_6	0.0	0.0	0.0	0.0	0.0	0.7	0.0	0.2	0.1	0.0
Oil	116.5	91.8	54.0	49.4	52.3	57.4	49.0	61.6	69.1	65.8
Aqueous phase	93.3	82.5	98.8	98.5	101.9	102.6	93.9	110.8	113.9	115.5
C_1-C_3OH[1]	3.3	4.3	2.2	4.7	4.8	5.3	4.4	5.3	5.5	4.5
Other oxygenates[1]	1.8	0.9	1.1	0.6	0.8	0.7	0.6	0.6	0.9	0.9
H_2O	88.2	77.3	95.5	93.2	96.3	96.6	88.9	104.9	107.5	110.1
CO_2	345.6	414.3	340.1	349.2	352.4	338.4	359.9	334.0	330.6	334.6
Total hydrocarbon recovery	193.0	185.7	180.1	179.7	176.1	180.2	175.7	179.1	186.2	188.1
Theoretical recovery	205.4	201.4	197.0	197.2	197.2	198.6	197.2	199.1	198.9	198.9
Overall weight balance...pct.	95.4	98.7	99.2	98.9	98.5	98.7	98.8	97.7	99.0	100.3
Hydrocarbons recovered, lb./1,000 ft.^{3} fresh gas	7.0	10.1	10.3	9.9	9.8	10.4	9.8	10.2	10.4	10.6
Weight-percent of hydrocarbons recovered:										
C_1-C_3	11.9	23.6	38.0	38.6	38.7	35.8	37.9	33.9	32.0	32.1
C_4	2.0	6.9	15.2	13.5	13.7	13.4	14.5	13.0	11.4	11.2
Gasoline (C_5= - 204° C.)	43.7	53.1	39.4	41.3	41.6	43.4	42.2	45.8	49.0	50.6
Diesel fuel (204°-316° C.)	10.4	10.5	3.6	5.3	5.0	6.2	5.0	6.8	7.1	5.8
Fuel oil (316°-450° C.)	9.3	4.4	0.8	0.2	0.2	0.2	0.1	0.1	0.1	0.0
Wax (> 450° C.)	22.7	1.5	3.0	1.1	0.8	1.0	0.3	0.4	0.4	0.3

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[1] Calculated as hydrocarbons.

TABLE 9. - Operating conditions and results, experiment 9

Period	A	B	C	D	E	F	G	H	J	K	L	M	N	O	P	Q	R
Hours of synthesis	180-252	324-348	348-396	420-468	540-564	588-612	612-660	660-732	756-828	852-972	972-996	996-1,068	1,092-1,188	1,188-1.260	1,260-1,368	1,408-1,649	1,667-1,859
Fresh gas space velocity...vol./vol.-hr.	202	312	399	501	600	801	801	801	998	1,001	1,002	1,002	1,001	1,001	1,001	1,001	1,001
Recycle:fresh gas ratios:																	
Total	114	114	108	60	60	60	60	60	60	60	60	60	60	60	60	60	60
Cold	0.8	2.7	2.4	2.5	2.4	2.5	2.5	2.3	2.2	2.3	2.3	2.3	2.2	2.2	2.1	2.2	2.2
CO_2 scrubbed	0.8	2.7	2.4	2.5	2.4	2.5	2.5	2.3	2.2	2.3	2.3	2.3	1.1	1.1	1.1	1.2	1.2
Water vapor in recycle gas...vol.-pct.	3.8	4.3	4.8	4.4	6.5	5.7	-	6.3	6.9	6.1	6.1	5.9	6.8	6.5	6.4	5.6	4.5
Cold recycle, pct. of total	1	2	2	4	4	4	4	4	4	4	4	4	4	4	4	4	4
H_2:CO ratios:																	
Fresh gas	0.99	1.31	1.33	1.33	1.31	1.30	1.31	1.36	1.35	1.38	1.37	1.37	1.37	1.34	1.36	1.33	1.30
Recycle gas	1.36	2.94	3.87	3.75	18.64	6.79	4.15	12.87	13.53	18.57	20.04	14.47	10.47	9.07	9.76	8.62	6.23
Usage	0.88	1.20	1.18	1.15	1.21	1.17	1.13	1.22	1.22	1.24	1.22	1.23	1.25	1.21	1.23	1.21	1.15
Reactor pressure...p.s.i.g.	402	401	402	400	401	403	403	403	405	406	406	407	408	409	404	406	407
Pressure differential across catalyst...in. H_2O/ft.	4	7	8	7	8	14	16	17	27	28	31	31	37	39	43	42	38
Catalyst temperature, ° C.:																	
Top (inlet)	270	309	310	303	310	310	315	320	320	310	310	310	310	309	310	310	310
Bottom (outlet)	281	320	322	323	330	330	335	340	340	330	330	330	330	330	330	330	330
Differential temperature	11	11	12	20	20	20	20	20	20	20	20	20	20	21	20	20	20
Average	275	314	315	311	320	320	325	330	330	319	319	319	319	319	319	318	318
H_2 conversion...pct.	67.8	86.9	83.0	80.9	92.2	87.7	81.0	88.5	90.2	89.6	88.2	88.4	89.8	89.0	89.5	89.0	85.9
CO conversion...pct.	76.6	94.5	96.1	93.3	99.5	97.7	94.1	98.7	98.9	98.9	99.2	98.8	98.6	98.4	98.5	98.2	96.9
H_2+CO conversion...pct.	72.2	90.2	87.7	86.3	95.3	92.0	86.6	92.5	93.9	93.6	92.8	92.8	93.5	93.1	93.3	92.9	90.6
Exit gas analysis (dry basis, vol.-pct.:																	
H_2	48.3	51.5	51.1	51.1	41.0	47.5	53.5	46.9	40.2	40.7	46.1	43.7	39.2	39.8	38.7	43.3	46.6
CO	35.5	17.5	13.5	13.7	2.2	7.0	12.9	3.9	3.1	2.6	2.3	3.3	4.1	4.4	3.8	5.1	7.7
N_2	0.2	1.1	1.9	1.8	2.5	1.4	0.8	1.0	0.8	1.6	0.0	0.6	1.8	1.4	1.6	1.1	1.0
CO_2	10.6	5.7	5.3	5.0	5.5	5.2	5.0	5.3	5.6	5.4	5.3	5.5	10.2	10.2	10.2	10.3	9.9
C_1	2.8	10.6	13.3	14.7	20.9	18.5	14.4	22.3	25.6	25.4	25.4	24.5	22.2	22.4	24.5	20.7	16.5
$C_2^=$	0.5	1.8	1.8	1.8	1.0	1.9	1.0	1.2	0.9	0.9	0.8	0.6	0.7	0.5	0.6	0.6	0.5
C_2	0.6	2.6	3.6	3.8	7.1	5.9	4.3	7.1	9.2	9.5	9.7	9.3	9.0	8.8	8.8	7.5	6.0
$C_3^=$	0.8	4.2	4.6	4.1	8.7	5.8	3.6	4.6	5.2	4.3	3.8	3.3	4.2	4.1	3.3	3.2	3.0
C_3	0.0	1.0	1.2	1.1	3.2	2.2	1.5	3.3	4.4	4.8	2.9	5.1	4.3	4.2	4.1	3.7	3.1
$C_4^=$	0.5	2.2	2.2	1.7	4.5	2.6	1.4	2.3	2.7	2.4	2.1	2.0	2.3	2.1	2.1	2.2	1.9
C_4	0.1	0.3	0.4	0.4	1.2	0.7	0.4	1.0	1.0	1.2	1.1	1.1	1.0	1.1	1.1	1.1	1.0
$C_5^=$	0.2	0.9	0.8	0.6	1.5	0.8	0.1	0.7	0.8	0.7	0.3	0.6	0.7	0.6	0.6	0.7	0.5
C_5	-	0.2	0.1	0.1	0.3	0.2	0.4	0.2	0.2	0.2	0.1	0.2	0.2	0.2	0.3	0.3	0.2
$C_6^=$	0.1	0.2	0.2	0.1	0.4	0.3	0.1	0.2	0.1	0.2	0.1	0.1	0.1	0.2	0.2	0.2	0.1
C_6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.0	0.1	0.0	0.0	0.1	0.0	0.0
Product yield, g./m.^{3} H_2+CO converted:																	
C_1	8.5	9.0	17.7	22.5	16.9	20.8	23.6	22.5	26.2	27.1	28.6	28.5	25.1	26.6	28.2	22.5	23.6
$C_2^=$	3.6	4.6	5.7	6.2	1.8	4.0	3.2	2.4	1.7	1.8	1.8	1.3	1.3	1.1	1.0	1.3	1.3
C_2	4.3	6.4	11.3	12.8	12.1	13.3	13.8	13.2	19.8	20.9	21.8	21.7	20.0	20.7	21.1	17.0	15.6
$C_3^=$	8.6	15.0	19.9	20.0	19.6	17.5	16.3	13.6	15.0	13.2	11.6	10.8	13.3	13.7	12.9	10.1	10.7
C_3	0.9	3.7	5.0	5.4	7.2	7.6	7.2	9.9	13.3	15.2	9.4	17.2	14.0	14.5	15.2	11.6	11.4
$C_4^=$	6.5	9.8	11.7	10.9	12.7	10.6	8.6	8.8	10.3	9.7	8.9	8.6	9.5	9.2	9.6	8.8	8.9
C_4	2.0	4.2	3.5	4.3	7.3	4.2	4.3	5.3	5.0	5.6	6.1	5.8	4.5	5.3	5.7	5.4	5.4
$C_5^=$	3.7	4.5	5.4	4.8	5.2	4.0	2.8	3.4	3.7	3.7	1.6	3.4	3.3	3.2	2.5	3.3	3.1
C_5	0.0	0.8	1.0	0.8	1.0	1.4	1.2	1.3	1.5	1.4	0.9	1.7	1.1	1.5	1.1	1.9	1.4
$C_6^=$	1.2	1.0	1.6	1.3	1.7	1.9	0.2	0.9	0.8	1.3	0.6	0.7	0.9	1.3	0.6	1.1	1.2
C_6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.2	0.0	0.5	0.0	0.1	0.0	0.1	0.0	0.1	0.6
Oil	74.5	57.9	70.1	78.9	77.9	87.9		78.3	72.4	75.8	81.2	72.1	78.7	72.6	70.6	65.2	58.5
Aqueous phase	49.7	107.7	107.3	111.2	134.8	131.0		136.8	134.7	130.6	128.9	126.3	134.5	128.9	126.9	113.4	97.1
C_1-C_5OH^1	1.9	7.3	6.7	7.2	7.7	6.9		7.1	6.0	5.8	6.2	5.7	5.9	6.7	4.8	5.7	5.1
Other oxygenates^2	1.3	2.2	2.2	2.5	2.5	2.4		2.0	1.7	1.1	1.1	1.1	1.2	1.3	0.9	1.2	0.9
H_2O	46.5	98.2	98.4	101.5	124.6	121.7		127.7	127.0	123.7	121.6	119.5	127.4	120.9	121.2	106.5	91.1
CO_2	346.8	294.8	315.9	326.4	294.8	315.8	326.7	298.7	290.4	300.3	302.2	305.6	288.5	297.3	297.2	303.5	301.8
Total hydrocarbon recovery	117.0	126.4	161.8	177.6	173.6	182.5		168.9	177.4	183.1	179.8	178.7	178.8	177.9	174.2	155.2	147.7
Theoretical recovery	206.3	206.0	203.7	202.9	203.6	203.2	202.6	202.4	201.3	200.7	200.9	200.8	201.7	201.4	200.6	201.7	202.0
Overall weight balance...pct.	79.0	86.3	93.5	97.0	97.5	100.1		97.9	97.8	100.3	99.0	99.3	98.2	97.8	97.9	92.5	87.2
Hydrocarbon recovery, lb./1,000 ft.^{3} fresh gas	5.3	7.1	8.8	9.5	10.3	10.4		9.7	10.3	10.6	10.3	10.3	10.4	10.1	10.0	8.9	8.3
Weight-percent of hydrocarbons:																	
C_1-C_2	14.0	15.8	21.4	23.4	17.8	20.9		22.6	26.9	27.2	29.0	28.8	26.0	28.0	28.9	26.3	27.4
C_3	0.8	2.9	3.1	3.0	4.1	4.1		5.9	7.5	8.3	5.2	9.6	7.8	8.5	8.7	7.5	7.7
Gasoline (C_4 - 204° C.)	44.9	64.5	60.3	58.0	65.9	59.3		59.7	56.1	54.0	55.2	52.8	55.4	54.7	54.7	58.1	57.5
Diesel fuel (204°-316° C.)	12.6	11.3	10.4	11.3	9.8	12.0		9.9	8.8	9.6	9.6	7.9	9.7	8.0	7.0	7.1	6.2
Fuel oil (316°-450° C.)	11.6	1.3	1.0	1.2	0.4	0.9		0.4	0.1	0.2	0.2	0.2	0.3	0.1	0.1	0.2	0.2
Wax (> 450° C.)	16.1	4.2	3.8	3.1	2.0	2.8		1.5	0.6	0.7	0.8	0.7	0.8	0.7	0.6	0.8	1.0

[illegible] drocarbons.

TABLE 10. - Operating conditions and results, experiments 11, 12, and 13

	Experiment 11			Experiment 12	Experiment 13							
Period	A	B	C	A	A	B[1]	C	D	E	F[2]	G	H[2]
Hours of synthesis	168-216	236-284	318-366	186-258	392-464	528-557	619-691	747-819	873-945	969-1,017	1,041-1,089	1,113-1,145
Fresh gas space velocity....vol./vol.-hr.	265	244	241	300	799	802	801	601	600	601	601	600
Recycle:fresh gas ratios:												
Total	162	81	82	54	46	37	43	36	35	31	31	17
Cold	2.6	2.8	2.8	2.5	2.2	2.2	2.1	1.5	1.0	0.7	0.6	0.6
CO_2 scrubbed	2.6	2.8	2.8	2.5	2.2	2.2	2.1	1.5	1.0	0.7	0.6	0.6
Cold recycle, pct. of total	2	3	3	4	5	6	5	4	3	2	2	4
H_2:CO ratios:												
Fresh gas	1.41	1.23	1.32	1.33	1.40	1.46	1.36	1.40	1.47	1.35	1.33	1.53
Recycle gas	4.32	3.58	2.65	6.22	8.31	0.88	1.45	1.65	1.43	1.74	1.36	1.26
Usage	0.96	1.02	1.22	1.11	1.13	1.73	1.35	1.36	1.49	1.43	1.31	1.85
Reactor pressure..p.s.i.g.	400	400	401	400	399	401	400	400	402	401	402	402
Catalyst pressure drop in. H_2O/ft. catalyst	8	2	9	1	11	16	28	25	49	36	43	39
Catalyst temperature, ° C.:												
Top (inlet)	349	380	379	310	310	310	310	310	310	311	311	311
Bottom (outlet)	359	399	401	330	330	330	330	330	330	330	330	329
Differential temperature	10	19	22	20	20	20	20	20	20	19	19	18
Average	353	390	390	318	319	319	319	319	318	318	319	318
H_2 conversion..........pct.	60.2	75.9	86.8	80.2	77.6	81.2	84.6	81.7	70.9	63.7	57.4	54.9
CO conversion..........pct.	87.3	92.0	93.7	95.9	96.2	68.6	85.7	84.4	70.1	60.3	58.3	45.4
H_2+CO conversion......pct.	71.5	83.1	89.7	87.0	85.3	76.0	85.1	82.8	70.6	62.3	57.8	51.1
Water vapor in recycle gas..............vol.-pct.	3.2	2.5	2.4	6.8	5.1	7.3	6.3	7.6	9.2	9.5	9.0	10.0
Exit gas analysis (dry basis, vol.-pct.):												
H_2	63.1	49.7	32.8	57.2	58.2	38.4	39.9	43.1	46.8	46.2	48.9	50.1
CO	14.6	13.9	12.4	9.2	7.0	43.9	27.5	26.2	32.7	37.4	35.9	39.7
N_2	0.5	1.1	2.4	1.6	0.5	0.4	0.1	0.1	0.9	0.5	0.5	0.6
CO_2	4.8	5.0	4.9	5.0	5.3	1.6	3.7	5.0	4.8	4.9	4.9	2.3
C_1	10.0	19.0	40.2	14.7	15.9	6.6	14.8	13.6	7.3	5.6	4.7	3.1
C_2=	1.2	1.6	1.0	2.1	2.2	1.4	0.9	0.8	0.9	0.8	0.7	0.9
C_2	2.5	5.1	4.1	3.8	3.6	2.0	4.7	4.1	2.0	1.4	1.2	0.9
C_3=	1.7	2.2	0.7	3.4	3.8	3.2	3.5	3.0	2.2	1.7	1.6	1.3
C_3	0.8	1.5	1.2	1.0	1.1	0.6	2.1	1.8	0.7	0.5	0.5	0.3
C_4=	0.5	0.6	0.2	1.1	1.5	1.4	1.7	1.6	1.1	0.8	0.4	0.6
C_4	0.2	0.2	0.1	0.3	0.4	0.2	0.5	0.4	0.2	0.1	0.2	0.1
C_5=	0.1	0.1	0.0	0.3	0.4	0.2	0.5	0.4	0.3	0.1	0.3	0.1
C_5	0.0	0.0	0.0	0.2	0.1	0.0	0.1	0.1	0.1	0.0	0.1	0.0
C_6=	0.0	0.0	0.0	0.1	0.0	0.1	0.0	0.0	0.0	0.0	0.1	0.0
C_6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Product yield, g./m.^{3} H_2+CO converted:												
C_1	36.9	42.7	72.1	20.9	27.2	17.4	27.2	28.3	25.6	27.0	26.7	22.3
C_2=	8.7	7.7	3.6	7.3	7.7	7.2	3.1	3.1	6.2	7.6	7.8	15.4
C_2	18.8	26.2	17.3	13.0	13.4	10.7	17.1	17.1	14.5	14.0	14.3	13.0
C_3=	18.4	15.5	5.5	18.3	19.9	24.9	18.3	18.5	23.2	25.5	27.5	27.6
C_3	8.9	10.6	9.6	5.1	6.2	4.6	11.1	11.1	7.7	7.5	8.9	6.5
C_4=	10.4	10.7	3.5	9.8	10.3	14.5	12.5	13.0	16.0	16.8	10.3	17.5
C_4	7.8	5.4	3.2	5.0	4.2	3.1	4.5	3.8	3.3	2.6	5.1	3.2
C_5=	3.8	3.6	0.6	3.7	3.4	3.5	4.7	4.6	6.0	3.2	9.0	4.3
C_5	1.3	1.8	0.0	2.4	2.9	0.4	1.1	1.0	2.0	0.0	3.2	0.0
C_6=	0.0	0.0	0.0	0.6	0.0	1.4	0.0	0.0	0.0	0.0	3.8	0.0
C_6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Oil	85.4	30.2	6.3	83.9	86.3	97.0	75.8	71.2	74.9	71.0	69.7	81.0
Aqueous phase	109.3	73.2	67.7	121.3	128.4	208.2	157.0	148.7	168.1	167.1	168.1	196.4
C_1-C_3OH[3]	6.2	5.2	1.0	7.3	7.2	2.2	3.5	3.1	2.0	2.5	2.6	1.5
Other oxygenates[3]	2.6	2.3	0.7	2.4	1.8	0.4	0.3	0.4	0.7	0.6	1.0	2.1
H_2O	100.5	65.7	66.0	111.6	119.4	205.6	153.2	145.2	165.4	164.0	164.5	192.8
CO_2	376.8	377.5	353.3	308.8	318.8	100.6	215.8	225.9	194.0	185.2	190.2	101.2
Total hydrocarbon recovery	209.2	159.9	123.4	179.7	188.5	187.3	179.2	175.2	102.1	178.3	189.9	191.4
Theoretical hydrocarbon recovery	199.9	198.3	194.2	202.8	202.1	204.2	201.4	201.5	202.1	202.2	202.0	202.9
Overall weight balance ....pct.	101.2	92.1	90.3	94.5	98.9	97.2	94.9	94.9	98.9	96.3	95.8	99.6
Hydrocarbon recovered, lb./1,000 ft.^{3} fresh gas.	9.3	8.3	6.9	9.8	10.0	8.9	9.5	9.0	8.0	6.9	6.8	6.1
Weight-percent of hydrocarbons recovered:												
C_1-C_3	30.1	48.3	75.4	22.9	25.6	18.8	26.4	27.7	25.4	27.2	25.7	24.9
C_4	4.2	6.8	7.8	2.9	3.3	2.5	6.2	6.4	4.3	4.2	4.7	3.4
Gasoline (C_5= - 204° C.)	46.4	42.1	12.0	54.9	57.7	66.2	59.3	58.5	61.0	59.8	61.2	61.7
Diesel oil (204°-316°C.)	7.1	2.5	0.9	11.4	6.7	7.7	4.8	5.7	5.1	6.3	5.9	6.9
Fuel oil (316°-450° C.)	6.2	0.3	3.9	6.3	3.9	3.5	2.1	1.4	2.8	2.1	1.8	1.8
Wax (> 450° C.)	6.0	0.0	0.0	1.6	2.8	1.3	1.2	0.3	1.4	0.4	0.7	1.3

[1] Alumina reformer in service.

[2] Activated carbon traps in service.

[3] Calculated as hydrocarbons.

TABLE 11. - Operating conditions and results, experiments 14 and 15

	Experiment 14		Experiment 15				
Period	A	B	A	B	C[1]	D	E
Hours of synthesis	153-225	369-393	108-180	396-468	492-564	572-644	668-740
Fresh gas space velocity, vol./vol.-hr.	268	1,000	250	1,002	1,002	998	1,000
Recycle:fresh gas ratios:							
Total	96	43	77	46	45	43	42
Cold	1.9	0.9	2.1	2.7	2.6	2.7	3.5
CO_2 scrubbed	1.9	0.9	2.1	2.7	2.6	2.7	1.3
Cold recycle, pct. of total	2	2	3	6	6	6	8
H_2:CO ratios:							
Fresh gas	0.98	1.84	0.99	1.40	1.39	1.41	2.01
Recycle gas	0.84	4.88	1.02	9.82	6.88	12.23	18.32
Usage	1.03	1.52	0.98	1.29	1.29	1.29	1.78
Reactor pressure, p.s.i.g.	400	401	400	399	400	400	399
Catalyst pressure drop, in. H_2O/ft. catalyst	6	47	7	62	66+	66-	66+
Catalyst temperature, ° C.:							
Top (inlet)	270	310	265	310	310	310	310
Bottom (outlet)	280	330	275	330	329	330	330
Differential temperature	10	20	10	20	19	20	20
Average	274	317	270	318	319	321	321
H_2 conversion, pct.	79.7	74.3	73.9	90.3	92.1	90.4	87.6
CO conversion, pct.	76.2	90.1	74.5	98.2	98.9	98.6	98.4
H_2-CO conversion, pct.	77.9	79.9	74.0	93.6	94.9	93.8	91.2
Water vapor in recycle gas, vol.-pct.	4.2	14.6	3.5	3.9	3.0	4.0	4.6
Exit gas analysis (dry basis, vol.-pct.):							
H_2	35.7	58.6	42.3	28.5	24.1	26.9	34.8
CO	42.5	12.0	41.3	2.9	3.5	2.2	1.9
N_2	0.6	1.2	0.7	1.7	1.4	2.2	2.1
CO_2	5.2	4.1	5.1	4.9	4.7	4.8	4.8
C_1	8.1	11.3	5.5	40.6	42.6	41.3	38.3
C_2=	0.2	1.0	0.4	0.3	0.4	0.2	0.4
C_2	2.7	3.8	1.7	9.6	10.4	10.0	8.1
C_3=	1.4	3.5	1.0	1.1	1.4	1.3	0.7
C_3	1.6	1.4	0.6	6.1	6.2	6.8	5.5
C_4=	0.9	1.8	0.6	1.2	1.4	1.2	0.7
C_4	0.6	0.5	0.3	2.1	2.5	2.1	1.7
C_5=	0.3	0.6	0.3	0.4	0.6	0.4	0.4
C_5	0.1	0.1	0.1	0.5	0.5	0.5	0.5
C_6=	0.1	0.1	0.1	0.1	0.2	0.1	0.1
C_6	0.0	0.0	0.0	0.0	0.1	0.0	0.0
Product yield, g./m.^{3} H_2-CO converted:							
C_1	21.5	28.2	14.5	59.6	60.3	65.2	71.4
C_2=	1.1	4.7	2.5	0.8	1.0	0.5	1.3
C_2	14.3	18.6	10.2	27.9	28.7	31.1	29.2
C_3=	11.8	24.9	10.6	4.7	5.6	5.7	3.6
C_3	12.5	10.1	5.6	25.9	25.3	30.8	29.2
C_4=	10.0	17.1	8.1	6.8	7.5	7.1	4.8
C_4	7.2	5.1	4.3	12.3	13.8	12.9	12.0
C_5=	4.1	7.1	5.2	2.7	3.9	3.0	3.4
C_5	1.2	1.2	2.0	3.5	3.4	3.7	4.3
C_6=	1.4	1.5	1.6	0.8	1.5	0.9	1.0
C_6	0.0	0.0	0.0	0.0	0.8	0.0	0.0
Oil	60.1	70.0	97.9	34.2	26.1	23.9	18.3
Aqueous phase	93.9	195.6	95.5	92.9	102.5	98.1	151.9
C_1-C_5OH[2]	7.7	4.1	6.8	3.0	1.8	1.7	1.8
Other oxygenates[2]	1.7	0.7	1.7	0.2	0.2	0.2	0.3
H_2O	84.5	190.8	87.0	89.7	100.5	92.2	149.8
CO_2	305.6	118.2	337.8	310.1	298.2	307.7	172.0
Total hydrocarbon recovery	154.5	193.3	170.9	172.4	180.2	186.9	180.6
Theoretical hydrocarbon recovery	202.5	201.1	204.3	193.5	193.4	192.4	192.0
Overall weight balance, pct.	87.7	92.5	91.4	96.7	97.9	99.8	100.6
Hydrocarbons recovered, lb./1,000 ft.^{3} fresh gas	7.5	9.6	7.9	9.9	10.7	10.9	10.3
Weight percent of hydrocarbons recovered:							
C_1-C_2	23.9	26.7	15.9	51.2	50.0	51.9	56.4
C_3	8.1	5.2	3.3	15.0	14.0	16.5	16.2
Gasoline (C_3= - 204° C.)	60.3	59.4	53.7	32.6	34.8	30.5	26.8
Diesel fuel (204°-316° C.)	5.4	5.7	12.4	1.0	1.1	1.0	0.6
Fuel oil (316°-450° C.)	2.0	2.1	8.3	0.2	0.1	0.1	0.0
Wax (> 450° C.)	0.3	0.9	6.4	0.0	0.0	0.0	0.0

[1] Alumina reformer in service.
[2] Calculated as hydrocarbons.

TABLE 12. - Operating conditions and results, experiments 16 and 17

	Experiment 16							Experiment 17						
Period	A	B	C[1]	D	E	F	G	A	B[2]	C[3]	D	E[3]	F[3]	G
Hours of synthesis	156-228	344-416	464-536	632-704	712-760	804-852	872-920	219-267	291-339	363-411	423-471	507-579	591-639	663-711
Fresh gas space velocity ... vol./vol.-hr.	250	1,001	1,000	1,001	1,000	1,001	1,000	1,002	997	1,001	1,002	1,002	1,002	1,501
Recycle:fresh gas ratios:														
Total	93	44	44	17	44	16	16	43	40	21	41	45	46	21
Cold	2.0	2.6	2.6	2.6	2.6	4.5	4.4	2.4	2.4	2.4	2.5	2.5	2.5	2.5
CO_2 scrubbed	2.0	2.6	2.6	2.6	2.6	0.0	0.0	2.4	2.4	2.4	2.5	2.5	2.5	2.5
Cold recycle, pct. of total	2	6	6	15	6	28	28	6	6	11	6	6	6	12
H_2:CO ratios:														
Fresh gas	1.40	1.42	1.40	1.36	1.35	2.91	2.94	1.38	1.38	1.42	1.40	1.37	1.40	1.40
Recycle gas	2.55	8.78	8.28	10.03	9.74	22.35	16.64	7.74	8.78	10.93	11.82	16.55	11.67	9.36
Usage	1.18	1.28	1.31	1.26	1.23	2.61	2.67	1.17	1.25	1.28	1.34	1.20	1.23	1.17
Reactor pressure ... p.s.i.g.	401	403	400	400	400	401	400	401	400	400	400	400	400	400
Catalyst pressure drop, in. H_2O/ft. catalyst	7	27	46	17	59	12	14	26	15	8	33	34	39	20
Catalyst temperature, ° C.:														
Top (inlet)	265	310	310	285	285	290	310	310	310	295	300	300	300	300
Bottom (outlet)	275	331	330	325	305	330	350	330	330	330	320	320	320	340
Differential temperature	10	21	20	40	20	40	40	20	20	35	20	20	20	40
Average	270	319	319	305	295	313	334	321	323	316	309	309	309	317
H_2 conversion ... pct.	70.1	88.1	92.1	90.8	89.0	88.0	89.0	81.8	88.6	88.6	87.5	86.6	86.2	80.9
CO conversion ... pct.	83.4	97.9	98.2	98.5	98.1	98.4	98.1	96.5	98.0	98.4	98.1	98.7	98.0	96.8
H_2+CO conversion ... pct.	75.6	92.1	94.7	94.0	92.9	90.7	91.3	88.0	92.6	92.6	91.9	93.7	91.2	87.5
Water vapor in recycle gas ... vol.-pct.	4.0	4.0	3.2	4.3	4.2	4.9	5.0	5.5	5.8	6.1	5.7	5.7	5.7	5.3
Exit gas analysis (dry basis, vol.-pct.):														
H_2	56.9	36.0	26.5	29.1	34.1	38.0	36.6	51.1	42.6	43.7	44.9	48.0	49.0	58.0
CO	22.3	4.1	3.2	2.9	3.5	1.7	2.2	6.6	4.8	4.0	3.8	2.9	4.2	6.2
N_2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO_2	4.9	4.9	4.8	5.0	5.0	2.4	1.7	4.9	4.8	4.8	4.8	4.7	4.7	4.9
C_1	9.2	34.8	42.0	41.1	37.4	39.6	42.0	22.6	28.6	28.4	26.8	25.6	23.4	17.4
C_2=	0.2	0.0	0.1	0.1	0.0	0.0	0.0	0.5	0.7	0.4	0.3	0.5	0.7	1.4
C_2	2.6	8.9	10.0	9.9	9.2	9.1	8.7	5.6	7.1	7.6	7.6	7.1	6.5	3.9
C_3=	0.9	1.1	1.1	0.7	0.7	0.5	0.7	2.8	3.7	3.4	2.9	3.9	4.4	3.8
C_3	1.6	6.4	7.8	7.4	7.0	5.9	5.4	2.7	3.9	3.8	4.6	3.6	3.2	1.4
C_4=	0.5	1.2	1.3	0.9	0.8	0.5	0.4	1.5	2.0	2.0	1.9	1.8	2.1	1.7
C_4	0.5	1.8	2.2	2.1	1.7	1.7	1.6	0.9	1.0	1.0	1.3	1.0	1.0	0.6
C_5=	0.3	0.4	0.5	0.4	0.3	0.2	0.2	0.5	0.5	0.5	0.6	0.6	0.6	0.5
C_5	0.2	0.4	0.6	0.4	0.3	0.4	0.3	0.2	0.2	0.2	0.3	0.2	0.2	0.1
C_6=	0.1	0.0	0.1	0.0	0.0	0.0	0.1	0.1	0.1	0.2	0.2	0.1	0.1	0.1
C_6	0.0	0.0	0.0	0.0	0.0	0.0	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Product yield, g./m.^{3} H_2+CO converted:														
C_1	25.0	52.5	55.4	56.6	51.7	72.8	73.3	36.1	33.2	34.3	32.8	31.2	29.8	26.1
C_2=	1.5	0.0	0.2	0.4	0.0	0.0	0.0	1.6	1.6	1.0	0.6	1.3	1.8	4.0
C_2	15.2	25.8	25.4	26.7	25.2	32.0	29.0	18.2	16.3	17.4	18.6	17.3	16.0	11.7
C_3=	8.5	4.4	4.1	2.6	2.9	2.5	3.3	13.3	12.4	11.4	10.2	13.8	15.7	16.1
C_3	12.0	27.3	29.3	29.2	28.5	30.9	26.9	13.5	13.4	12.7	16.7	13.1	11.7	6.8
C_4=	6.5	6.5	6.5	4.5	4.5	3.3	2.6	9.3	8.7	9.1	9.0	8.5	9.9	9.5
C_4	6.2	11.0	11.3	12.2	9.9	11.9	10.7	7.2	5.6	5.6	6.9	5.9	5.9	4.7
C_5=	5.1	2.7	3.0	2.6	1.8	1.7	1.6	3.8	2.7	2.7	3.4	3.4	3.6	3.7
C_5	3.2	3.3	2.5	3.2	2.0	3.5	2.4	1.5	1.1	1.1	1.7	1.2	1.2	0.8
C_6=	1.5	0.0	0.8	0.6	0.0	0.0	1.0	1.0	0.6	1.3	1.4	0.6	0.6	0.9
C_6	0.0	0.0	0.0	0.6	0.0	0.0	1.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Oil	62.0	38.7	33.5	35.3	36.4	13.9	12.4	78.8	68.9	70.9	74.2	80.1	84.9	96.3
Aqueous phase	102.8	102.9	107.8	105.8	104.5	209.8	208.8	129.6	139.6	143.1	137.1	139.1	141.6	137.0
C_1-C_5OH[4]	5.3	2.7	1.3	2.1	3.2	3.3	1.8	7.6	9.0	7.2	8.6	8.8	9.5	7.0
Other oxygenates[4]	0.0	0.3	0.2	0.4	0.2	0.5	0.3	1.7	1.9	1.2	1.5	1.5	1.5	2.0
H_2O	97.5	99.9	106.3	103.3	101.1	206.0	206.7	120.3	128.7	134.7	127.2	128.8	130.6	128.0
CO_2	316.6	314.8	298.7	307.6	314.0	10.2	6.2	301.2	278.5	280.6	286.5	284.6	286.8	304.2
Total hydrocarbon recovery	152.0	175.2	173.5	177.0	166.3	176.3	166.3	193.6	175.4	175.9	185.4	186.7	192.1	189.6
Theoretical hydrocarbon recovery	201.8	194.9	194.4	194.2	195.3	191.2	191.4	199.3	200.2	199.8	199.9	200.5	200.8	202.0
Overall weight balance ... pct.	91.8	98.8	97.5	97.9	94.9	95.3	93.6	98.6	96.4	98.2	98.6	97.4	99.5	99.3
Hydrocarbons recovered, lb./1,000 ft.^{3} fresh gas	7.2	10.1	10.2	10.4	9.6	10.0	9.5	10.6	10.1	10.1	10.6	10.7	10.9	10.3
Weight-percent of hydrocarbons recovered:														
C_1-C_2	27.4	44.7	46.7	47.3	46.2	59.5	61.5	28.6	29.1	30.0	28.0	26.7	24.8	22.0
C_3	7.9	15.6	16.9	16.3	17.2	17.5	16.2	7.0	7.7	7.2	9.0	7.0	6.1	3.6
Gasoline (C_5= - 204° C.)	55.8	37.8	34.7	35.5	36.6	23.0	22.3	52.6	53.6	55.5	54.2	54.6	56.1	54.2
Diesel fuel (204°-316° C.)	8.9	1.7	1.6	0.7	0.0	0.0	0.0	7.0	4.8	5.7	6.0	8.0	7.9	9.6
Fuel oil (316°-450° C.)	0.0	0.2	0.1	0.0	0.0	0.0	0.0	3.1	1.5	1.4	2.6	3.3	4.0	5.4
Wax (> 450° C.)	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.5	3.3	0.2	0.1	0.6	1.1	5.2

[1] Alumina reformer in service.
[2] Split recycle stream.
[3] Added 6 g. of potassium.
[4] Calculated as hydrocarbons.

GLOSSARY OF TERMS

Fresh gas	Synthesis gas composed essentially of hydrogen and carbon monoxide.
Fresh gas ratio	Volumetric ratio of hydrogen-to-carbon monoxide, dimensionless.
Usage ratio	Volumetric ratio of hydrogen consumed-to-carbon monoxide consumed during synthesis, dimensionless.
Hourly space velocity of gas, S.V.H. (in tables)	Volumes (S.T.P.) of gas per hour per volume of catalyst. Cubic feet per hour per cubic foot of catalyst. (S.T.P.) refers to 0° C. and 760 mm. Hg pressure.
Conversion, percent	Extent of conversion of raw materials such as H_2+CO: $\frac{(H_2+CO) \text{ in} - (H_2+CO) \text{ out} \times 100,}{(H_2+CO) \text{ in}}$ dimensionless.
Yield	Weight of any component or components per unit volume of feed or converted gas, pound component per cubic foot of gas or gram component per cubic meter of gas.
Specific yield	Yield of given product per cubic meter of converted synthesis gas, gram per cubic meter converted (H_2+CO).
Space-weight yield	Volume of gas converted per hour per weight of catalyst, cubic foot (S.T.P.) per hour per pound of catalyst.
Catalyst activity	Quantitative relationship between extent of conversion and operating temperature at any time during synthesis, usually judged by temperature required for a certain conversion at a given space velocity.
Activity coefficient	Indication of catalyst activity--the larger the coefficient, the more active the catalyst.

$A_v = 1.241 \times 10^{-8} \times S \times 10^{4130/T} \times \log_{10} (1/1-c)$ where[27]

S = hourly space velocity

T = absolute temperature, ° K.

C = conversion

$A_{Fe} = A_v$ divided by catalyst density (gram per cubic centimeter)

Induction	Pretreatment of catalyst usually with a feed gas of $1H_2$+1CO at a low temperature of 260°-270° C.
Massive iron catalysts	Catalysts made from steel in the form of turnings, shavings, wool, and plates as opposed to fused iron or precipitated catalysts.
Classification of products:	Boiling range at atmospheric temperature, ° C.:
Gasoline	C_3 to 204
Diesel oil	204 - 316
Heavy distillate	316 - 450
Wax	> 450
C_1 plus C_2	Includes methane, ethane, ethylene
C_3	Propane.

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[27] Anderson, R. B., Seligman, B., Shultz, J. F., Kelly, R., Elliott, M. A., Fischer-Tropsch Synthesis--Some Important Variables of the Synthesis on Iron Catalysts: Ind. Eng. Chem., vol. 44, 1952, pp. 391-397.