MWK-MPR-22

RESEARCH AND DEVELOPMENT DEPARTMENT

DEVELOPMENT OF KELLOGG COAL GASIFICATION PROCESS

Contract No. 14-01-0001-380

May 31, 1966

Progress Report No. 22

APPROVED: [signature]
Project Manager

[signature]
DIRECTOR

[signature]
MANAGER
RESEARCH & DEVELOPMENT DEPARTMENT

CONTENTS

		Page No.
I.	INTRODUCTION	2
II.	SUMMARY	4
III.	PROCESS RESEARCH	7
IV.	CHEMICAL ENGINEERING STUDIES AND DEVELOPMENT	12
V.	MECHANICAL DEVELOPMENT	16
VI.	MANPOWER AND COST ESTIMATES (to be transmitted with invoices)	

1. INTRODUCTION

The objective of this contract with the Office of Coal Research is to develop the Kellogg Gasification Process to the point where it will be able, on a commercial scale, to convert coal into pipeline gas at a cost of 50¢/MSCF or hydrogen at 25¢/MSCF. Five raw materials are to be studied -- an anthracite, a high-volatile bituminous coal, a sub-bituminous coal, a lignite, and a char. Although Kellogg's experimental work will not extend beyond the production of raw synthesis gas, the overall project must make engineering evaluations for four ultimate end products -- pipeline gas, hydrogen, synthesis gas, and transport gas.

Basis for the Kellogg Gasification Process is the reaction between steam and fine coal in a molten salt bath to form synthesis gas, a mixture of hydrogen and carbon monoxide, according to the reaction:

$$\underset{\text{(coal)}}{C} + \underset{\text{(steam)}}{H_2O} + \text{heat} \longrightarrow \underset{\text{(synthesis gas)}}{H_2 + CO}$$

The necessary heat of reaction is supplied by circulating a heated molten salt stream. In addition, the molten salt mixture is chosen to catalyze the gasification reaction so that it may be carried out at a relatively low temperature.

The program is divided into three phases of study extending over a five year period. Phase I, which is now in progress, involves several concurrent efforts:

1. Bench-scale process research -- to investigate melt properties, reaction kinetics, and the effect of process variables.

2. Chemical engineering studies and development -- to determine the optimum process flowsheet and operating conditions and to coordinate experimental work with overall project objectives.

3. Mechanical development -- to find acceptable materials of construction and develop techniques for handling the molten salt and powdered coal.

Phase I will be concluded by the design of a pilot plant to gasify 24 tons of coal per day.

Phase II will be devoted largely to the construction and operation of a pilot plant to convert a variety of raw materials into raw synthesis gas. The effect of operating variables found to be significant in Phase I will be investigated to obtain data for design of a commercial plant.

Phase III will involve the detailed process design of a commercial plant to produce 250 million standard cubic feet a day of product gas, including cost estimates and projected economics for those areas of the country that appear to offer commercial possibilities.

II. SUMMARY

This progress report is the twenty-second since the awarding of the contract. It is concerned with the first phase of the contract and summarizes the progress that has been made in the three principal areas now being studied: process research, chemical engineering studies and mechanical development.

Gasification experiments have been continued in the 2-inch-diameter test reactor at three atmospheres pressure to study the effect of steam orifice size, coke particle size, ash content of the melt, initial carbon concentration in the melt, and melt depth.

The size of the steam inlet orifice was decreased by a factor of three in an attempt to produce smaller bubbles and thereby lessen the effect of bed depth (and the subsequent tendency for large bubbles to form) on gasification rate. No noticeable change in rate was observed indicating that either bubble size in the 2-inch unit is not critical to the operation or that the actual bubble size is not determined by orifice size but rather by gas impingement on the reactor bottom (due to downflow steam feed).

Increasing coke particle size from the standard 12/20 mesh to 4/8 mesh (a factor of 2.8) decreased gasification rate by 25 percent, and decreasing particle size to 80/100 mesh (a factor of 7.7) increased rate by 24 percent. It is felt that this change is due to the effect of particle size on the efficiency of carbon-melt contacting rather than that of increasing surface area.

Experiments studying the effect of ash level on gasification rate were continued. Maxima in the rates were attained at 7 percent and 4 percent ash at superficial gas velocities of 0.5 and 1.0 feet per second, respectively. A more significant percentage increase was noted at the low gas velocity indicating that the effect of ash appears to be one of improving the mixing wherein ash helps distribute the carbon throughout the melt.

The initial carbon concentration in the melt was varied between 1.6 and 7.9 percent and its effect on gasification rate noted. At concentrations at and above about 4 percent there appears to be little change in the rate constant. However, a rather substantial increase in rate is noted as the initial level is lowered to the range of 2 percent.

One new run was made to study the effect of bed depth on gasification rate. A 2-inch bed depth was used and the result indicates only a slight increase in the previously reported rate in a 3-inch bed.

A paper entitled "Bench-Scale Studies of the Kellogg Coal Gasification Process I. Gasification" was prepared for presentation at the American Chemical Society 1966 Fall Meeting.

Several studies were made to investigate the economic consequences of varying several of the design bases and assumptions used in formulating the conceptual process designs. These investigations included varying carbon and ash concentrations in the melt, gasification rate, superficial velocity, sodium carbonate loss, steam/carbon ratio, gasification temperature and gasifier-combustor configuration. The results obtained can be summarized as follows:

1. Decreasing carbon concentration in the melt to 2 percent (from 4 percent) increases pipeline gas cost by 3.2¢/MSCF if the increased melt volume is compensated for by increasing the number of gasifiers. If simply bed depth is increased, an increase of only 0.7¢/MSCF is incurred.

2. If the ash level in the melt is reduced from 8 percent to 4 percent, gas cost is increased by 2.7¢/MSCF.

3. Variations in gasification rate of a factor of two result in only small changes in manufacturing cost. If, however, rate were reduced by a factor of ten, the result is a catastrophic increase of 18.4¢/MSCF of pipeline gas.

4. Increasing or decreasing superficial velocity in the gasifier by a factor of two decreases gas cost by 1.4¢/MSCF and increases it by 3.2¢/MSCF, respectively.

5. Doubling sodium carbonate losses increases gas cost by 1.8¢/MSCF.

6. If the steam/carbon ratio is increased from 2/1 to 3/1, a cost increase of 2.3¢/MSCF is incurred.

7. Lowering the gasification temperature from 1830°F to 1740°F adds 1.4¢/MSCF to manufacturing cost.

8. There is no economic difference between designing the gasifier-combustor system as separate vessels with melt circulated by external gas lift or as a single vessel with melt circulated internally by density differences between the two sections.

Corrosion Test #3 under oxidizing conditions was resumed and continued uninterrupted to a total of 1,000 hours in a simulated combustion atmosphere. Monofrax A showed an average overall corrosion rate of 0.05 inches per year after the full 1,000 hours of the test. It is thus re-confirmed that this material will be an excellent choice for the intended use.

Another 1,000 hour test has been begun, this time in a reducing atmosphere. Two blocks of Monofrax A have been clamped together under a load of 40 psi and placed in the unit to ascertain the effect of corrosion on a joint in a Monofrax A wall.

The 4 1/4-inch reactor has been modified to provide a bottom outlet for melt withdrawal, and a test program was begun to provide information on melt quenching.

A material to simulate the melt is being sought in an effort to determine the flow properties of the melt to aid in the design of any future pilot plant. Presently, DuPont's "Hitec" is being investigated.

Work has continued on the 5-3/4-inch reactor test facility. The reactor has been received and furnace delivery is scheduled for late July.

III. PROCESS RESEARCH

A. Accomplishments

The effects of the following variables on rate of gasification of bituminous coke were investigated in the 2-inch ID Inconel reactor:

1. Orifice size of steam inlet.
2. Particle size of coke.
3. Ash content in the melt
4. Initial carbon concentration in the melt.

The run conditions and results are summarized in Table I. All these runs have been made at 1740°F, 3 atmospheres total pressure and, with one exception, a 4-inch quiescent molten salt bed height.

A paper entitled "Bench-Scale Studies of the Kellogg Coal Gasification Process I. Gasification" was prepared for presentation at the American Chemical Society 1966 Fall Meeting, Fuel Division Symposium on Gasification.

1. Effect of Orifice Size of Steam Inlet

In run 9997, the steam inlet orifice size was changed from the standard 0.152 inch to 0.047 inch. This was an attempt to vary the size of the inlet steam bubbles and determine whether steam bubble size is a factor in the change in rate constant with bed height described in previous summaries. As illustrated in the following tabulation, this change did not have a significant effect on the reaction rate constant.

Run J-	Orifice Size-in.	Reaction Rate Const.-k_o
9993	0.152	1.57
9997	0.047	1.48

Apparently, steam bubble size does not account for the bed height effect and, as previously stated, the effect must still reside with carbon distribution. However, it must be pointed out that the steam inlet points downward and is about 1/2 inch from the reactor bottom. Bubble impingement on the bottom may change the bubble size created by the orifice and cancel to some extent the effect of variations in orifice size.

2. Particle Size of Coke

Variation in particle size of bituminous coke from the standard 12/20 mesh to a larger size, 4/8 mesh, and a smaller size, 80/100 mesh, had an appreciable effect on the rate constant. Results of identical runs except for coke size are summarized in the following tabulation.

Run J-	Coke Size Mesh	Coke Size mm.(1)	React. Rate Const. k_o
9999	4/8	3.57	1.59
9986	12/20	1.26	2.14
9998	80/100	0.163	2.65

(1) Based on arithmetic average of sieve openings.

An increase in coke size by a factor of about 2.8 over the standard 12/20 mesh decreases the reaction rate constant by 25%. A decrease in coke size to 13% of the standard increased the reaction rate constant by 24%.

Previous attempts to determine the effect of coal or coke size on rate were inconclusive. The effect illustrated above suggests two concepts which help to explain some of the previous data. First of all, a decrease in rate with increasing coke particle size does imply poorer distribution of coke in the molten salt. This coke is porous and the difference in surface area due to size is probably much smaller than the rate change. Secondly, as the coke is gasified, the particles get smaller and, as expected, the rate of gasification should increase. Every first order plot for gasification runs verifies this--carbon remaining gasifies faster as the run proceeds.

3. Ash Level in the Melt

A series of runs were completed in which 0 to 12% bituminous coal ash was added to the melt. Other operating conditions were 1740 F, 3 atmospheres total pressure, 4 inch quiescent bed height, 4% initial carbon concentration, and two superficial gas velocities, 0.5 and 1.0 ft./sec. The results of these runs are illustrated in Figure 1.

The results directly indicate that a maximum rate constant is reached at 7% ash for 0.5 ft./sec. velocity and at 4% ash for 1.0 ft./sec. At and above 8% ash, superficial velocity appears to have no effect on rate constant. Bed expansion and temperature profile must be factors in determining the shape of the curves. It is difficult to believe that rate would be diminished below the non-ashed melt as it was for the 1 ft./sec. curve. In the region where temperature profiles and bed expansion are approximately constant, addition of ash significantly improves the rate of gasification as is illustrated in the following tabulation.

Gas Velocity ft./sec.	Reaction Rate Const. - k_o No Ash	Max. Value	at % Ash
0.5	1.13	1.80	7
1.0	1.80	2.22	4

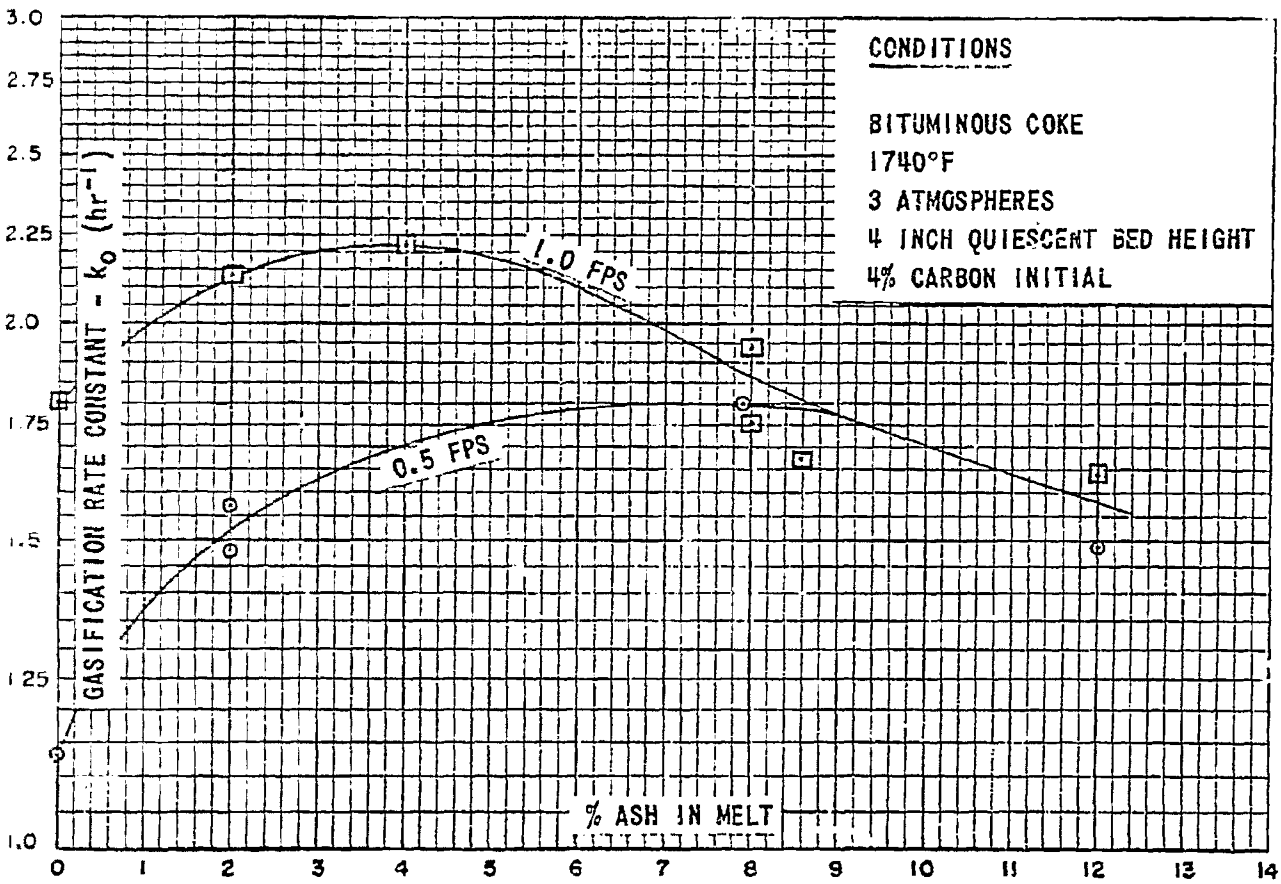
Since the improvement is much greater at the lower velocity, the effect appears to be more of an action upon mixing of coke and steam where ash helps to distribute the coke. At high velocity the mixing is greater and the effect of ash is lessened accordingly.

4. Effect of Initial Carbon Concentration

The previously described effects which implied that the distribution of coke appears to be a rate controlling phenomenon suggested that the effect of initial concentration of carbon in the melt be considered.

FIGURE I

EFFECT OF ASH CONTENT IN MELT ON GASIFICATION RATE CONSTANT



This was performed in runs H-11 and 12 and can be compared to an earlier run 9986 at the standard carbon level of 4%. Also inadvertedly in run H-6 only 1.6% carbon concentration got into the bed although 4% was desired--this was at the 12% ash level and can be compared to run H-7 at 4% carbon initially. All of these runs were at 1 ft./sec. superficial velocity and they are shown in the following tabulation:

Run No.	% Initial Carbon	% Ash	React. Rate Const. - k_o
H-11	2.1	2	2.47
9986	4	2	2.14
H-12	7.9	2	2.09
H-6	1.6	12	3.46
H-7	4	12	1.64

At the 2% ash level, a significantly higher rate constant was obtained at the lower carbon concentration of 2% but the rate was apparently constant above 4% carbon. The result with 12% ash at the low carbon level of 1.6% was an exceptionally high rate constant when compared to the 4% carbon run. This however also reflects the fact that less bed expansion occurred and that the temperature profile may have been more normal.

Again these results give further support to the theory that carbon distribution has been one of the major rate controlling factors in the experimental bench scale process research.

5. Completion of Melt Height Effect on Gasification Rate

One additional run (H-9) was made to complet. the study of the effect of height of melt in the 2-inch diameter reactor on the gasification rate of bituminous coke. This run was made at 2-inch bed height and required a correction in rate constant for an 11°F lower than normal reaction temperature. The result is plotted in Figure 2 and indicates very little change over that at 3-inch bed height.

B. Projections

Evaluation of the gasification rate of other coals and lignites will now be made. A char has been requested from OCR for similar evaluation. Upon completion of this work combustion work under pressure will be started.

FIGURE 2

EFFECT OF MELT HEIGHT ON GASIFICATION RATE OF BITUMINOUS COKE

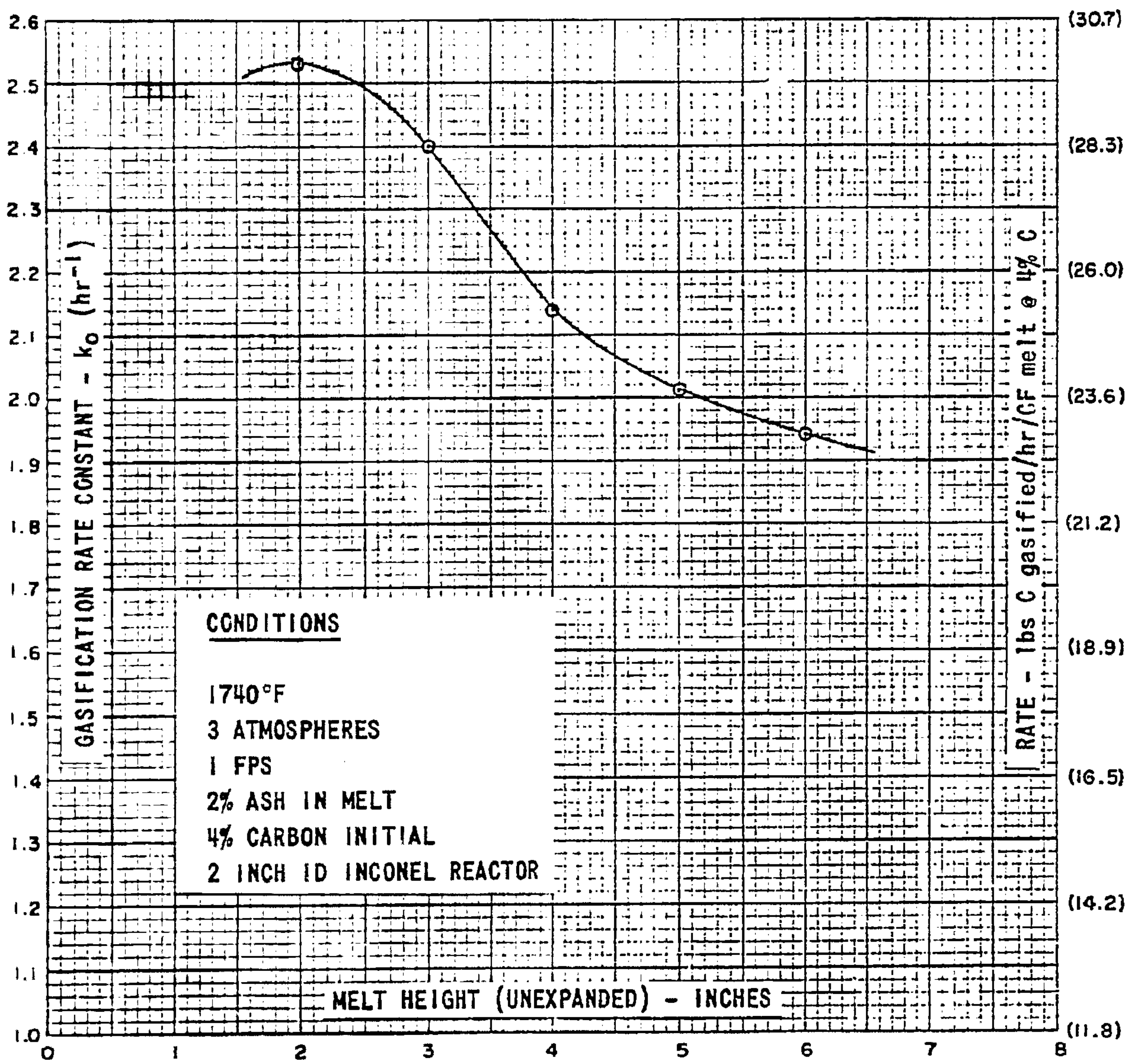


TABLE I

GASIFICATION OF BITUMINOUS COKE IN MOLTEN SODIUM CARBONATE[1]

Run No. J-	9996	9997	9998	9999	H-1	H-2	H-3	H-4	H-5	H-6	H-7	H-8	H-9	H-10	H-11	H-12
Date	4/29	5/3	5/5	5/5	5/9	5/9	5/11	5/11	5/13	5/16	5/17	5/17	5/20	5/24	5/24	5/25
Feed	Coke IV	→	←	←	←	←	←	←	Coke VI	→	→	→	→	→	→	→
% Total Carbon	93.2	→	←	←	←	←	←	←	93.2	→	→	→	→	→	→	→
% Vol. Matter	0.6	→	←	←	←	←	←	←	0.6	→	→	→	→	→	→	→
% Ash	6.2	→	←	←	←	←	←	←	6.2	→	→	→	→	→	→	→
gms. charged	19.0	→	←	←	19.0	→	→	→	→	6.8	19	19	9.5	19	9.5	38
mesh size	12/20	→	80/100	4/8	←	←	←	←	12/20	→	→	→	→	→	→	→
Na_2CO_3 - gms.	405.7	→	405.7	→	(3)	(3)	380.9	(3)	(3)	(3)	(3)	(3)	202.9	414	(3)	(3)
Ash - gms.	8.3	→	8.3	→			33.1						4.1	-		
% in melt	2.0	→	2.0	→	3	4	8	8	8.55	12	12	12	2	-	2	2
% C in melt-init.	4.0	→	4.0	→	4.0	4.0	4.0	4.0	4.0	[illegible]	4.0	4.0	4.0	4.0	2.1	7.9
Bed Heights-in.	4	4	4	4	4	4	4	4	4	4	4	4	2	4	4	4
Conditions	(2)	(2)														
Temp.-°F avg.	1740	1739	1735	1736	1738	1736	1740	1736	1736	1744	1739	1738	1729	1735	1732	1742
Pressure-psia	45	45.2	45.2	45.9	45.4	45.1	45	44.7	44.9	[illegible]	45.4	45.2	45.7	45.7	45.6	45.3
Steam Pressure-psia	--	41.1	40.0	41.6	40.1	40.6	40	40.4	40.6	41.4	41.1	40.4	41.7	41.4	41.4	41.2
Gas in Steam	←	←	←	←	←	←	←	N_2	→	→	→	→	→	→	→	→
Ft./Sec., Stm.+Gas	1.0	0.51	0.97	0.97	0.50	1.00	0.5	1.05	1.05	1.03	1.05	0.49	1.03	1.05	1.03	1.02
Minutes to 0% CO	--	40	25	35	55	35	45	35	30	15	35	50	25	30	20	35
Total Run - min.	--	45	30	45	55	35	55	40	40	[illegible]	40	50	25	35	25	45
cc H_2O in/hr.	--	580	1118	1112	559	1137	565	1108	1192	1131	1197	553	1186	1197	1192	1190
cc N_2 in/min.	--	1197	2200	2132	1514	2610	1329	2635	2582	2486	2570	1368	2360	2515	2546	2444
Results																
Total % C to Oxides	--	100	95	102	101	96	74	98	96	92	101	100	99	102	101	97
Gasif. Rate Const.- k																
Basis Input	--	1.48	2.37	1.64	1.48	2.07	--	1.64	1.54	3.00	1.65	1.48	2.25	1.82	2.47	1.96
Basis Output	--	1.48	2.65	1.59	1.48	2.22	--	1.70	1.67	3.46	1.64	1.48	2.31	1.80	2.47	2.09
Rate-lbs C/hr/CF																
at 4% C in bed	--	18	31	19	18	26	--	20	20	41	19	18	30	21	29	25
Salt Carryover-gms	--	4.4	6.6	5.9	7.9	11.3	8.5	12.7	9.5	7	3.5	0.2	5.5	9.5	4.0	7.9
Notes					(4)		(5)						(6)			

(1) Island Creek #27 bituminous coal coked at 950°C (Cokes IV and VI); coke charged in N_2 atmosphere at 0.05-0.1 ft./sec. superficial velocity to 2-inch ID Inconel reactor with standard 0.15 inch orifice for steam.

(2) Used 0.047 inch orifice for steam; outlet plugged from salt buildup at start of run J-9996.

(3) Used previous run's melt plus makeup.

(4) % Ash in melt was in doubt.

(5) Gas leak at exit union gave poor balance.

(6) Correction of output rate constant from 1729°F to 1740°F gives 2.53.

IV. CHEMICAL ENGINEERING STUDIES AND DEVELOPMENT

A. Accomplishments

A number of studies have been made to investigate the economic consequences of varying several of the design bases and assumptions from the base case. The studies undertaken are listed below:

Case	Economic Study
I	Carbon in melt equals 2% instead of 4%
I-A	Compensate by increasing number of gasifiers
I-B	Compensate by increasing melt height
II	Ash in melt equals 4% instead of 8%
III	Variation in gasification rate
III-A	Assume rate decreased by factor of 2 ($k = 1.15\ hr.^{-1}$)
III-B	Assume rate increased by factor of 2 ($k = 4.6\ hr.^{-1}$)
III-C	Assume rate decreased by factor of 10 ($k = 0.23\ hr.^{-1}$)
IV	Variation of superficial velocity
IV-A	Assume velocity halved (V = 0.62 fps)
IV-B	Assume velocity doubled (V = 2.5 fps)
V	Assume sodium carbonate losses doubled
VI	Steam/carbon ratio increased from 2/1 to 3/1
VII	Gasification temperature is 1740°F instead of 1830°F
VIII	Design gasifier and combustor as separate units

The results of these investigations are summarized in Table II, for the case of pipeline gas from subbituminous coal. As can be seen, halving the allowable carbon concentration in the melt to 2 percent results in a rather modest increase in gas cost. If this reduction in concentration (hence increased melt volume requirement for the same gasification rate) is compensated for by increasing the number of gasifiers a larger cost is incurred than if the melt height is correspondingly increased. The reason for this is the increased labor required to operate the larger number of operating trains in Case I-A.

Decreasing the ash level in the melt to 4 percent adds 2.7¢/MSCF to gas cost. The bulk of this increase is attributed to increases in the investment of the ash removal section and in sodium carbonate losses (due to the larger amount of melt handled). The increased investment of the gasification section was due to an expected decrease in rate according to a correlation previously reported [(1)]. However results reported in this Summary (Figure 1) indicate that at a velocity of 1.25fps there may be an increase in rate resulting from such a decrease in ash level. Thus, the increase in gasifier investment shown in Table II may not be a real one.

The effect of halving or doubling the gasification rate is a rather small one (Cases III-A and B). On the other hand, if the rate were actually one-tenth that used as the design basis, the result is a catastrophic increase of 18.4¢/MSCF. Such a drastic rate reduction does not appear probable, however, based upon the experimental results to date and our current techniques of extrapolation to commercial design conditions.[(2)]

Present experimental results have indicated an effect of gas velocity on gasification rate, most probably because of its influence on the degree of melt turbulence (and hence mass transfer). Halving or doubling the average superficial velocity in the gasifier would add 3.2¢/MSCF and save 1.4¢/MSCF, respectively. These cost differentials reflect changes in investment and operating labor due to increased (or decreased) gas volume requirements.

Doubling sodium carbonate loss would add 1.8¢/MSCF to the pipeline gas manufacturing cost.

If the steam/carbon ratio would have to be increased to 3/1 to achieve the design gasification rate, a cost increase of 2.3¢/MSCF would be incurred. The major contributors to this increase are a higher investment in the gasification section (because of the increased quantity of gas handled), and a decrease in the electric power credit. Since more steam is required for usein the process, less is available for use in generating electricity.

---

(1) Progress Report No. 16, Contract 14-01-0001-380, November 30, 1965, P. 19.

(2) Progress Report No. 21, Contract 14-01-0001-380, April 30, 1966, P. 7.

If gasification temperature were lowered to 1740°F and a corresponding decrease in rate assumed, gas cost would rise by 1.4¢/MSCF. The other major factor (besides the lower rate) in this increase is a decrease in electric credit. Since less energy is put into the gasification system, there is less to recover and be converted into electricity.

Finally, there is no economic difference whether the gasifier and combustor are designed as separate vessels with melt circulated by external gas lift or whether they are combined into a single vessel separated by a dividing wall with melt circulated by a difference in the degree of aeration between the two sections.

B. Projections

1. Pipeline Gas

The process package for pipeline gas from subbituminous coal will be completed. The process flowsheet, material balance, and utilities summary have all been completed and capital cost estimating of the plant, now essentially done, will be finished.

2. Hydrogen

Preparation of the process flowsheet for the production of 250 MM SCFD of hydrogen from bituminous coal will be continued.

TABLE II

EFFECT OF VARIATIONS IN DESIGN BASES ON COST OF PIPELINE GAS FROM SUBBITUMINOUS COAL

Basis: 250 MM SCFD of pipeline gas
Cost of Coal: $2 per ton
Value of by-product power: 4.5 mills/kwh
Maintenance and fixed charges at 11.2% per year
Gasification temperature: 1830°F
Carbon in melt: 4%
Ash in melt: 8%
Gasification rate: 21.2 # C/hr.,- CF melt
Superficial gas velocity in gasifier: 1.25 fps
Steam/carbon ratio: 2#/#
Na_2CO_3 loss: 2.4% per pass

CASE	I		II	III			IV		V	VI	VII	VIII
	A	B		A	B	C	A	B				
Investment $MM												
Coal Handling	--	--	--	--	--	--	--	--	--	--	-0.68	--
Gasification	8.7	5.02	0.74	5.02	-2.34	75.9	8.7	-3.41	--	2.79	3.74	6.97
Shifting	--	--	--	--	--	--	--	--	--	-0.01	0.28	--
Ash Removal	0.35	--	1.3	--	--	0.9	0.35	-0.24	--	0.15	-0.10	0.19
Offsites	--	--	--	--	--	--	--	--	--	-0.29	-2.46	--
Net Change	+9.05	+[illegible]2	+2.04	+5.02	-2.34	+76.8	+9.05	-3.65	--	+2.64	+0.78	+6.78
Manufacturing Cost ¢/MSCF												
Coal	--	--	0.4	--	--	--	--	--	--	0.2	-1.0	--
Na_2CO_3	--	--	1.9	--	--	--	--	--	1.8	--	-0.1	--
Labor & Ovhd	2.0	--	--	--	--	8.0	2.0	-0.9	--	0.4	--	-0.9
Fixed Chgs	1.2	[illegible].7	0.4	0.7	-0.3	10.4	1.2	-0.5	--	0.4	0.1	0.9
Elec Credit	--	--	--	--	--	--	--	--	--	1.3	2.4	--
Net Change	+3.2	+[illegible].7	+2.7	+0.7	-0.3	+18.4	+3.2	-1.4	+1.8	+2.3	+1.4	0.0

V. MECHANICAL DEVELOPMENT

A. Accomplishments

1. Corrosion Testing of High Temperature Materials

The replacement parts for the furnace used in Oxidation Test #3 were received, and the test continued without further interruption to 1,000 hours. The results of this 1,000-hour test were as follows:

a. Monofrax A (High-purity cast alumina) -

This sample was in the reactor for the full 1,000 hours, and on this basis was found to have corroded at an average corrosion rate of 0.05 in/year. (This rate, while less than that reported at 846 hours, i.e., 0.08 in/year, is consistent. The decrease in corrosion rate is due to almost no change in specimen size during the interval from 846 to 1,000 hours, hence a lower average rate). The specimen, while still intact and in good condition, shows some areas of localized corrosion where the grain size was largest.

b. Monofrax M (High-purity cast alumina with 1.1% SiO_2) -

This sample also had been in the reactor for the full 1,000 hours, with an average corrosion rate of 0.283 in/year on this basis. The specimen broke at the mounting hole during the test and was found at the bottom of the reactor. The corrosion was not local but rather general in nature and resulted in pronounced rounding of the specimen's edges and planes.

c. Inconel 600 - This sample was included at the 500-hour mark and remained during the latter 500 hours of testing. It corroded at an average rate of 0.23 in/year based on 500 hours. The specimen appears to be still in good condition, with some rounding of edges and roughening of plane surfaces.

d. Morganite Triangle RR (High-purity alumina) -

The Morganite sample had been included in the test at the 200-hour mark and remained for 800 hours. Based upon the 800 hours of testing, an average corrosion rate of 0.13 in/year was determined. The recovered specimen had broken into four separate segments which could be fitted back together with ease. The general appearance of each segment was good in that the edges were still sharp and the plane surfaces were still quite smooth. The fact that the specimen was broken in such a manner could be explained by thermal shock.

Monofrax A still appears to be the best material for the intended use. Inconel 600 might also be of some use in spite of its higher corrosion rate as the corrosion seems to be only a surface effect.

A second 1,000 hour test has been started, this time, however, in a reducing atmosphere. Two blocks of Monofrax A have been clamped together under a load of 40 psi and placed in the test environment to ascertain the effect of corrosion on a joint in a Monofrax A wall. This clamped pair of blocks will be removed from the test periodically and examined, but will not be taken apart until conclusion of the 1,000 hours.

2. Mechanical Characteristics Testing

The 4-1/4" reactor has been modified to provide a bottom outlet for melt quench testing. This test program was begun during the latter portion of the report period, and should provide information as to the effect of quenching the hot melt in water. In addition, information will be acquired as to possible grinder requirements.

A material to simulate the molten sodium carbonate-ash system is being sought. Currently, E. I. DuPont's "Hitec" is being investigated as to its ability to duplicate the melt's behavior. This work is being conducted at low temperature, 300 to 700°F, in a reactor similar to the 4-1/4" reactor previously described.

Work is continuing on the 5-3/4" reactor test facility. The reactor has been received, as have many of the other items required for this unit. Furnace delivery is scheduled for late July with facility erection and testing shortly thereafter.

B. Projections

1. Environmental Testing of High Temperature Materials

The 1,000 hour block joint corrosion test currently in progress will continue at least well into the next report period.

2. Mechanical Characteristics Testing

Investigation of quenching of the hot melt will continue as well as work to ascertain what the grinder requirements might be for the proposed pilot plant.

Work will also continue to find a suitable system for measuring the flow parameters for the molten salt. This information is needed to insure proper design of the proposed pilot plant.

In addition, the erection of the 5-3/4" test facility will begin in the near future.

VI. MANPOWER AND COST ESTIMATES

Figure 3 shows the projected breakdown for Phase I for 1966 as well as the actual effort that was made. It can be seen that a 10 man-effort was made during May.

Figure 4 shows the expenditures during May. For the month $18,644 was expanded, not including fee and G & A. The total expenditures through May were $427,841. Including fee and G & A the total expenditures were $489,656. This is 82% of the encumbered funds.

FIGURE 3
MANPOWER FOR PHASE I
1966

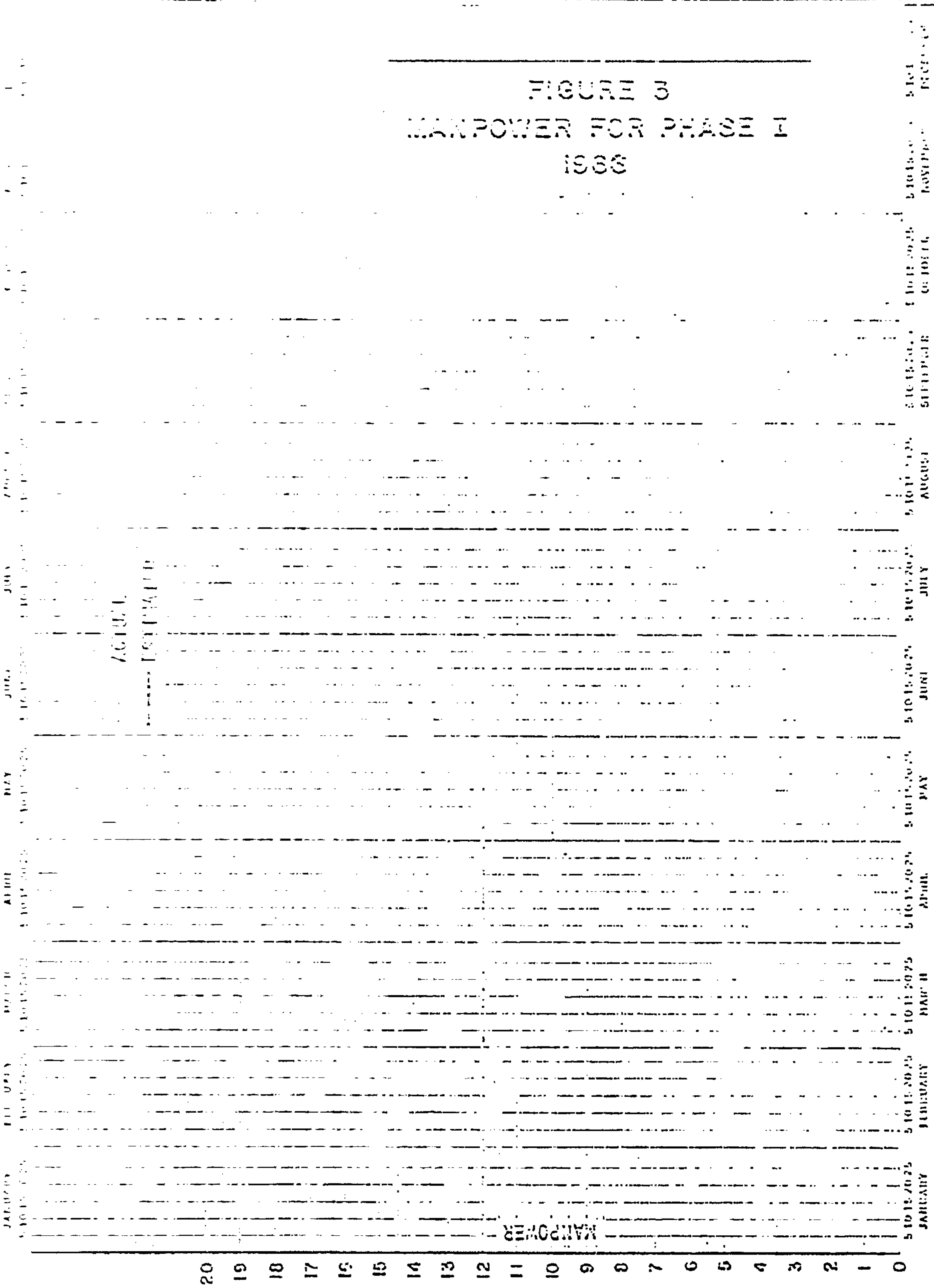


FIGURE 4
COST ESTIMATED FOR PHASE I
1966