

MWR-MPR-11

RESEARCH AND DEVELOPMENT DEPARTMENT



DEVELOPMENT OF KELLOGG COAL GASIFICATION PROCESS

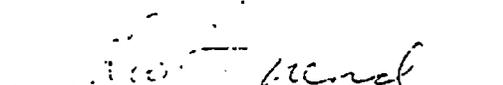
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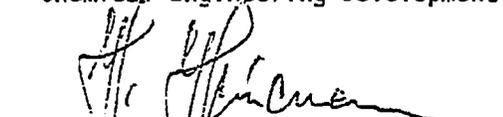

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I. INTRODUCTION

The evaluation of gas purification processes for pipeline gas and hydrogen has been completed. For both pipeline gas and hydrogen, the optimum purification scheme is the Fluor process using propylene carbonate. The major advantage over MEA, Vetrocoke and Sulfinol is substantial savings in reboiling steam.

Recent experimental data on ash removal and sodium recovery are being evaluated, and a flowsheet for this section of the plant is being developed using the data.

It has been found that the pressure of steam has a significant effect on gasification rate. A change from 0.3 atm to 0.9 produced about a two fold increase in gasification rate.

An increase in the temperature of the molten salt has had a substantial effect on gasification rate. Increasing temperature from 1740°F to 1830°F increased gasification rate by a factor of 2.3 and increasing temperature to 1940°F improved it by a factor of five.

The switch from a coke to bituminous coal feed has been successfully made.



II. PROCESS DEVELOPMENT

A. Accomplishments

1. Evaluation of Gas Purification Processes

The evaluation of alternate gas purification schemes for both pipeline gas and hydrogen has been completed. The regenerable solvent purification processes under consideration for these two applications were:

1. Hot potassium carbonate
2. Catacarb (organic-promoted K_2CO_3)
3. Rectisol (refrigerated methanol)
4. Monoethanolamine
5. Vetrocoke (arsenic-promoted K_2CO_3)
6. Acetone
7. Sulfinol (sulfolane)
8. Fluor (propylene carbonate)

Neither hot carbonate, Catacarb, nor Rectisol has been given detailed attention for these studies. The two carbonate-based schemes (hot carbonate and Catacarb) were screened out after qualitatively comparing them with another carbonate system, namely Vetrocoke. Based on previous experience with Vetrocoke and hot carbonate, Vetrocoke has been found to be the better of the two for several reasons. First, the presence of the arsenic promoter improves the stripping equilibrium relationships and also increases the rate of mass transfer within the system. This results in rather substantial decreases in the size of the towers and hence in the plant cost. Also, steam requirements for Vetrocoke (for reboiling the rich solvent stream) are considerably lower than for hot carbonate^(1, 2), due to the improved stripping equilibria.

- (1) Riesenfeld, F. C. and J. F. Mallowney, "Giammarco-Vetrocoke Processes," *Petroleum Refiner*, 38, No. 5, May, 1959, p. 161.
- (2) Hoogendoorn, J. C., "A Comparison of Some Carbon Dioxide Removal Processes," *Trans. Instn. Chem. Engrs.*, 41, 1963, p. 264.



Finally, the presence of the promoter tends to make the alkaline solution less corrosive than plain carbonate solutions, and operating difficulties due to such problems are largely eliminated (as well as reducing the cost by enabling the use of carbon steel in areas where stainless would be required in the hot carbonate system.) The Catacarb process was eliminated based on a detailed study made by Kellogg comparing it with Vetrocoke. It was concluded that for all practical purposes the two schemes are the same insofar as investment and operating costs are concerned, and any conclusions drawn concerning the use of Vetrocoke for the present applications could be applied equally to the Catacarb process.

Process information requested from Lurgi (developers of the Rectisol process), including plant investment figures, has not as yet been received. For this reason, Rectisol has not been included in the present evaluations. If and when such information is received, Rectisol will also be evaluated. However, based on the results of previous studies it is not expected that this process will prove to be the most economical one.

a. Pipeline Gas

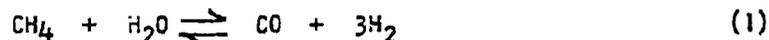
The purpose of the gas purification section in the pipeline gas plant is to remove bulk quantities of CO_2 and essentially all sulfur compounds from the shifted synthesis gas in preparation for the methanation step. In all of the cases considered, the CO_2 concentration was reduced to 1.0 mole percent and total sulfur was reduced to 0.004 grains/100 SCF. In addition to the regenerable solvent system for CO_2 and bulk sulfur removal, the processing sequences included sponge iron (iron oxide) and activated carbon for residual H_2S and organic sulfur removal, respectively. Since the sulfur concentration of the treated gas is the same for each case, the costs associated with the final "clean-up" steps (sponge iron and activated carbon) are essentially constant and have not been included in the comparison. Therefore, the selection of the optimum purification scheme was based on the CO_2 -bulk sulfur removal process which yielded the most attractive economics. It should also be noted that the cost of removing CO_2 and bulk sulfur is much higher than that for residual sulfur removal, so any minor differences in the cost of this latter step are unimportant with respect to the overall purification step.



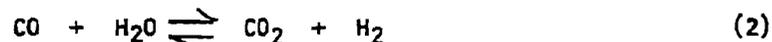
A table of economics, including investments and operating costs, for each of the five gas purification schemes given detailed consideration is presented in Table I. The Fluor process results in the lowest total operating cost, and is therefore the optimum purification scheme for use in this application. The major advantage which the Fluor process has over MEA, Vetrocoke, and Sulfinol is that of substantial savings in reboiling steam. The Fluor system has only a very small steam requirement since the CO_2 is removed from the rich solvent by flashing at low pressures instead of reboiling. Acetone, on the other hand, while it has the advantage of low steam costs, has the disadvantage of high costs due to solvent losses. These losses could, at best, probably be cut down to about 1¢/MSCF (operating cost of 4.9¢/MSCF) by increasing the efficiency of the solvent recovery step of the process. However, even with this improvement (neglecting the increase in investment which would necessarily result), the operating cost still would not be as low as that attainable with Fluor.

b. Hydrogen

The gas purification scheme used in the hydrogen plant is somewhat different from that used in making pipeline gas. Following gasification, the bulk of the sulfur is removed by a regenerable solvent process, and complete sulfur removal (to 0.004 grains/100 SCF) is effected using sponge iron and activated carbon. This gas is then fed to a steam reformer, where the methane concentration is reduced to about 5 percent via reaction 1:



The methane concentration of the reformer feed gas is determined by assuming that all the hydrogen in the gasified coal goes to methane. This, then, represents the maximum methane concentration attainable and is therefore somewhat conservative for the case where hydrogen is the product. After cooling, the gas enters a shift converter using low temperature shift catalyst where the CO level is reduced to about 0.6% via reaction 2:



The shifted synthesis gas, containing appreciable quantities of CO_2 , then flows to a regenerable solvent CO_2 removal step (the same type as

TABLE 1
ECONOMIC SUMMARY
ALTERNATE GAS PURIFICATION SCHEMES
250 MM SCFD of Pipeline Gas
90% Steam Efficiency

Process	HEA	VETROCOKE	ACETONE	SULFINOL	FLUOR
Total Fixed Investment	\$24,300,000	\$17,050,000	\$17,850,000	\$10,100,000	\$12,300,000
<u>Operating Costs, ¢/MSCF of Pipeline Gas</u>					
Steam at 30¢/H lbs.	5.60	3.90	0.19	3.17	0.07
Power at 0.7¢/KWH	0.50	0.67	0.59	0.56	0.74
Cooling Water at 1.5¢/H Gal.	0.97	0.68	0.03	0.42	0.03
Chemicals	0.08	0.06	5.00	0.67	0.21
Sulfur Credit at \$25/ton	-	(1.20)	-	-	-
Operating Labor at \$3.20/Man-Hour	0.43	0.37	0.15	0.06	0.15
Maintenance at 3% of Bare Cost	0.79	0.56	0.58	0.33	0.40
Supplies at 15% of Maintenance	0.12	0.08	0.09	0.05	0.06
Supervision at 10% of Operating Labor	0.04	0.04	0.02	0.01	0.02
Payroll Overhead at 10% of Operating Labor + Supervision	0.05	0.04	0.02	0.01	0.02
General Overhead at 50% of Operating Labor + Supervision + Maintenance + Supplies	<u>0.69</u>	<u>0.52</u>	<u>0.42</u>	<u>0.22</u>	<u>0.32</u>
Plant Operating Expenses	9.27	5.72	7.09	5.50	2.02
Depreciation at 5% of Fixed Investment	1.47	1.03	1.08	0.61	0.75
Taxes and Insurance at 3% of Fixed Investment	<u>0.88</u>	<u>0.62</u>	<u>0.65</u>	<u>0.37</u>	<u>0.45</u>
Sub-Total	11.62	7.37	8.82	6.48	3.22
Contingencies	<u>0.23</u>	<u>0.15</u>	<u>0.18</u>	<u>0.13</u>	<u>0.06</u>
TOTAL OPERATING COST	11.85	7.52	9.00	6.61	3.28





is used to remove bulk sulfur), where the CO_2 level is reduced to 1.0 mole percent. The residual CO and CO_2 are then methanated over a nickel catalyst, and the resulting gas stream, containing about 94.0 percent H_2 , leaves the plant as product. This hydrogen concentration was chosen for the present evaluation because it is readily attainable with the processing scheme selected (steam reforming of methane impurity.) The major impurities are methane and nitrogen, which are considered inerts for many hydrogen applications. If higher concentrations of hydrogen are desired (99+% H_2), the methane in the raw synthesis gas would have to be completely removed by partial combustion with oxygen or by low temperature absorption. Both of these alternatives are more costly and would yield higher hydrogen costs than the presently-proposed scheme.

On the other hand, if even lower concentrations of hydrogen can be tolerated (as low as 86%), the reforming step can be eliminated and a corresponding reduction in investment realized. However, because the product gas now contains more methane and since each mole of methane is capable of producing four moles of hydrogen (according to reaction 1), the case will have the disadvantage of operating at lower coal utilization efficiencies. The net result of these opposing factors is that there is but marginal benefit to be gained in producing the low (86%) purity hydrogen product.

In evaluating the alternate gas purification schemes for hydrogen, the costs associated with sponge iron, activated carbon, and reforming were taken into account in addition to the costs of the regenerable solvent schemes. The sulfur removal costs were found to be small, but the cost of reforming was found to be rather substantial and sensitive to relatively small variations in gas composition brought about by the different CO_2 -bulk sulfur removal systems. One other thing should be pointed out before discussing economics. As is obvious from the above process description, there is a considerable amount of gas cooling and heating taking place. Thus there is a rather large amount of waste heat available to the process. Credit for this waste heat has not been taken at this time since overall plant energy balances have not yet been made. This is not important to the present evaluation though, since the waste heat liberated in all of the purification schemes is essentially the same.

Estimated economics for the five purification schemes considered (including reforming and sulfur removal) are shown in Table 2. The economics are calculated on the basis of 250 MM SCFD of 100% hydrogen,

TABLE 2
ECONOMIC SUMMARY
ALTERNATE GAS PURIFICATION SCHEMES
250 MM SCFD of Hydrogen
90% Stream Efficiency

(All cases include reforming, iron oxide, activated carbon in addition to the indicated purification process.)

Process	HEA	VETROCOKE	ACETONE	SULFINOL	FLUOR
Total Fixed Investment	\$24,000,000	\$21,190,000	\$17,350,000	\$18,940,000	\$15,780,000
<u>Operating Costs, c/MSCF of H₂</u>					
Reforming Coal at \$4/ton	2.69	2.22	2.20	2.69	2.20
Steam at 30¢/M Lbs.	3.30	2.54	0.10	2.00	0.05
Power at 0.7¢/KWH	0.29	0.74	0.39	0.33	0.47
Cooling Water at 1.5¢/M Gal	0.58	0.51	0.02	0.30	0.02
Boiler Feed Water at 25¢/M Gal.	0.24	0.24	0.23	0.24	0.23
Chemicals	0.07	0.04	2.90	0.44	0.13
Sulfur Credit at \$25/Ton	-	(0.33)	-	-	-
Operating Labor at \$3.20/Han-Hour	0.34	0.49	0.31	0.28	0.34
Maintenance at 3% of Bare Cost	0.78	0.69	0.56	0.62	0.51
Supplies at 15% of Maintenance	0.12	0.10	0.08	0.09	0.08
Supervision at 10% of Operating Labor	0.03	0.05	0.03	0.03	0.03
Payroll Overhead at 10% of Operating Labor + Supervision	0.04	0.05	0.03	0.03	0.04
General Overhead at 50% of Operating Labor + Supervision + Maintenance + Supplies	0.63	0.66	0.49	0.51	0.48
Plant Operating Expenses	9.11	8.00	7.34	7.56	4.58
Depreciation at 10% of Fixed Investment	2.91	2.57	2.10	2.30	1.91
Taxes and Insurance at 3% of Fixed Investment	0.87	0.77	0.63	0.69	0.57
Sub-Total	12.89	11.34	10.07	10.55	7.06
Contingencies	0.26	0.23	0.20	0.21	0.14
TOTAL OPERATING COST	13.15	11.57	10.27	10.76	7.20





even though the actual product gas is about 94% hydrogen. The reason for doing this is that the different CO₂ removal schemes operate at several levels of residual CO₂. Therefore, the amount of hydrogen consumed by methanation is different in the various schemes. The production rate of elemental hydrogen has thus been kept constant at 250 MM SCFD, while the total product gas rate will vary slightly from case to case. One additional point should be noted here. Depreciation has been taken at 10% of investment, instead of 5% as was done for pipeline gas, because it is felt that hydrogen production parallels more closely a chemical or refining operation than it does a public utility.

Once again, as for pipeline gas, the Fluor process is the most attractive route, primarily because of its low investment and utilities costs. MEA, Vetrocoke, and Sulfinol continue to show high steam costs, while acetone is hampered by large solvent make-up costs.

2. Ash Removal and Sodium Recovery

Recent experimental data on ash removal and sodium recovery are being evaluated, and a flowsheet for this section of the plant is being developed using these data. The objective of this effort is to determine whether any further experiments are necessary at this time to extend the range of variables studied -- type of ash, ratio of quench liquor to melt, melt particle size, time of leaching, and ratio of wash water to ash.

B. Projections

Development of the ash removal-sodium recovery flowsheet will continue. Further data (if any) needed to firm up the process design will be requested.

Process design of the gasification section will be revised to incorporate recent data on gasification rate, observations on the effect of melt viscosity on quality of gas-melt contacting, and a different base-case method for supplying heat (direct combustion instead of fired-tube heater). Complete process flowsheets from bituminous coal feed to product pipeline gas or hydrogen will be prepared to permit more accurate estimates of plant investment and a study of waste heat utilization.



III. PROCESS RESEARCH

A. Accomplishments

1. Gasification Kinetics

This has been a very informative period in the study of coal gasification. The most notable results are:

1. Pressure of steam has a significant effect on gasification rate. A change from 0.3 atm to 0.9 atm produced approximately a two-fold increase in gasification rate. Thus justification for evaluation of gasification under pressure has a kinetic as well as economic basis.
2. An increase in the temperature of the molten salt has had a profound effect on gasification rate. A jump from 1740°F to 1830°F improved the rate by a factor of 2.3 and a further increase to 1940°F improved it by a factor of five. Thus gasification rates of 30 pounds of coal per hour per cubic foot of molten salt appear attainable.
3. A successful switch in feed from coke to bituminous coal has been made and the data appear reliable. The two gasification units now in use give excellent agreement in gasification rates.

The data for these conclusions and others which appear in the following discussion are presented in Table 3.

After three runs with Coke II, made from bituminous coal at 600°C, seven excellent runs have been made feeding bituminous coal directly. In order to accomplish this, the charge size was lowered from 35 grams to 20 grams of 12/20 mesh material. Naturally, when this is dropped into the hot molten salt, an initial burst of gas is produced which fortunately is not excessive for the gas meter used.

TABLE 3

GASIFICATION OF BITUMINOUS COAL AND COKE IN HOLTEN Na_2CO_3 (1)

Run No. -J-	9790	9791	9792	9793	9794	9810	9811	9812	9813	9814
Date of Run - 1965	6/8	6/10	6/11	6/15	6/16	6/21	6/22	6/24	6/25	6/28
Feed	←-----Coke 11-----→			←-----Coal-----→						
% Total Carbon		86.6					81.6			
% Volatiles		7.6					37.3			
% Ash		5.8					3.8			
gms. charged	35.0	35.0	35.0	20.0	20.0	20.0	20.0	20.0	20.0	20.0
mesh size	12/20	-200	12/20	12/20	12/20	12/20	12/20	12/20	12/20	12/20
Additive	-	-	-	-	-	-	K_2CO_3 10%	-	Fe_2O_3 1%	-
Conditions										
Temperature - °F	1730	1750	1730	1750	1740	1740	1745	1940	1740	1830
% Steam in N_2	90	90	90	90	90	30	90	90	90	90
min. to 0% CO	>55	95	105	90	90	110	90	<45	75	50
Unit No.	2	1	2	1	2	1	2	1	2	1
Total Time Run - min.	55	125	120	120	120	155	110	65	105	75
Results										
% C to $\text{CO} + \text{CO}_2$	70	64	90	92.1	91.1	84.8	95.6	99.4	99.0	91.2
% C to CH_4	-	-	-	6.3	6.1	4.6	5.0	7.5	2.6	5.0
% C to tar, etc.	-	-	-	6.0	6.0	6.0	6.0	6.0	6.0	6.0
Total % C	70	64	90	104.4	103.2	95.4	106.6	112.9	107.6	102.2
% C to 1/2 ($\text{H}_2 + \text{CO}$)	67	49	90	81	84	72	79	79	83	70
Ratio H/O Prod. Gas	1.9	1.5	2.0	1.7	1.8	1.7	1.6	1.4	1.6	1.5
Specific Gasification Rate(2)										
Basis $\text{CO}, \text{CO}_2, \text{CH}_4$ - Input	0.58	(0.4)	0.55	0.55	0.55	0.32	0.69	3.0	0.69	1.29
Basis $\text{CO}, \text{CO}_2, \text{CH}_4$ - Output	-	0.75	0.67	0.55	0.53	0.33	0.62	2.6	0.62	1.29
Basis 1/2 $\text{H}_2, 1/2 \text{CO}, \text{CH}_4$ - Input	0.53	(0.3)	0.55	0.47	0.55	0.31	0.47	2.6	0.69	0.86
Basis 1/2 $\text{H}_2, 1/2 \text{CO}, \text{CH}_4$ - Output	-	0.78	0.67	0.55	0.55	0.42	0.58	3.0	0.82	1.29
Notes	(3)	(4)					(5)	(6)		

- (1) Atm. Pres., 0.5 ft./sec. superficial gas velocity, 4" quiescent bed height, 414 gms. Na_2CO_3 , Island Creek # 27 coal.
(2) k in hr^{-1} from $k = -\frac{1}{t} \log \frac{C_t}{C_0}$, assuming first order carbon reaction.
(3) Outlet plugged with 7% CO still in exit gas.
(4) Some feed was held up in the reactor.
(5) Estimated % C to CH_4 .
(6) Calc. based on 25 minutes of run.



However, a tar collects on the cooling surfaces and in a glass wool scrubber. At times a fog of finely divided material has been noticed passing all the way through the wet test meter and into the stack. Collection of the tarry and sooty material yielded one gram, which is assumed to be carbon and represents 6% of the initial carbon charged. Consequently, rates on an output basis are now calculated to 94% gasification. Bottom feeding and a longer molten bed would decrease or eliminate this material. A composite first five minute gas sample is taken which allows calculation of the carbon to methane and gives a reliable carbon monoxide figure for this first condition.

a) Effect of Steam Pressure

In the last report (No. 10, pg. 7) two runs were presented which showed no effect of changing steam from 30 to 90% in nitrogen at one atmosphere and combined rate of 0.5 ft./sec. superficial gas velocity in the reactor. Due to a number of tests, it has now become clear that this is not so. Both runs were made in Unit Number 1, but J-9785 was made with the old reactor which after this run was retired because the 1/4 inch Inconel tubes for the inlet and thermowell had corroded through. It now appears that part of the steam was not bubbling through the molten salt in this and some of the previous runs which may account for the poorer results for 90% steam. This is substantiated by a temperature profile under gasification conditions on the old reactor versus the new reactors in Unit 1 and Unit 2.

<u>Inches from Bottom</u>	<u>Temperature °F</u>		
	<u>Unit 1</u>		<u>Unit 2</u>
	<u>Old Reactor</u>	<u>New Reactor</u>	<u>New Reactor</u>
1	1620	1740	1730
3	1780	1740	1730
5	1860	1740	1740
7	1900	1790	1760
9	1930	1830	1750
11	1910	1810	1700
13	1810	1680	1550

In the old reactor, it appears that steam was not agitating the salt, hence the non-uniform temperature profile in the reaction zone whereas with the new reactors the first 6 inches are at constant temperature.



Run J-9790 was an initial attempt to commission Unit 2 but the outlet plugged causing termination of the run in 55 minutes. Only an input gasification rate was determined and the run was repeated in J-9792. The results of this run are compared in the following table to the 30% steam run in Unit 1 using a new reactor. Also shown are similar runs using coal as feed.

<u>Run J-</u>	<u>Feed</u>	<u>% Steam in N₂</u>	<u>Gasification Rate -k Output Oxide Basis</u>
9792	Coke II	90	0.67
9787	Coke II	30	0.29
9793	Coal	90	0.55
9810	Coal	30	0.33

These results show the approximately two-fold advantage in rate caused by the higher steam pressure, 0.9 atm versus 0.3 atm. This gives additional justification for the construction of a gasification unit which will allow a study of the effect of steam pressure to 400 psi.

b. Comparison of Units 1 and 2

The duplicability of Unit 1 versus Unit 2 was established with coal in Runs J-9793 and 9794. Identical results at a 0.54 gasification rate constant were obtained.

c. Effect of Particle Size

A further attempt to establish the effect of particle size was made in Run J-9791 but as previously the difficulty of testing has clouded the results. At this time it appears that the effect of particle size is minor in the region of 12/20 to 40/60 mesh and that -200 mesh material has not been adequately tested.

d. Effect of Temperature

The effect of increasing temperature has led to the most significant improvement in rate observed to date. The results are summarized in the following table.



<u>Run J-</u>	<u>Temperature</u>	<u>Gasification Rate k</u> <u>Output Basis</u>
9794	1750	0.55
9814	1830	1.29
9812	1940	2.6 - 3.0

The 80°F jump in temperature from 1750°F improved the rate by a factor of 2.3, while the 190°F increase to 1940°F upped the gasification rate by a factor of five. It should be noted that the temperature is that of the molten salt and that the steam-gas product may not be at this temperature but may be at a lower temperature. Activation energy for these temperature changes calculates to 50 kcal, which indicates that chemical reaction is controlling. It should also be mentioned that carryover of salt did not appear to be any different than usual at these higher temperatures.

e. Effect of Additives

A homogeneous additive, potassium carbonate, and a heterogenous additive, ferric oxide, were added to the sodium carbonate melt and evaluated in separate runs. Potassium carbonate in 10% concentration was added because literature indicates potassium carbonate reacts faster with carbon than does sodium carbonate⁽³⁾. Ferric oxide has been quoted in the literature as showing catalytic power for the carbon-steam reactions. The ferric oxide, as supplied by Columbian Carbon Company, was Mapico 297, a finely divided solid which is 90% less than one-half micron in size. Approximately 1% ferric oxide was present in the molten salt.

<u>Run J-</u>	<u>Additive</u>	<u>Gasification Rate k</u> <u>Output Basis</u>
9794	None	0.53
9811	10% K ₂ CO ₃	0.62
9813	1% Fe ₂ O ₃	0.62

The 17% increase in rate is not considered to be a significant amount and it would not justify the use of relatively expensive potassium carbonate and difficultly useable iron oxide.

(3) G. L. Putnam, Ind. and Eng. Chem. 30, 1136 (1938).



2. Pressure Test Unit

A basic design of a reactor to operate at 1600-2000°F and up to 400 psig has been completed. The following lists the pertinent physical features of the reactor vessel:

1. Cylindrical reaction space 3 inches in diameter by 5 feet high,
2. A six-inch layer of zirconia refractory to contain the molten salt as well as to insulate the stainless steel shell from the high reaction temperature,
3. Heat supplied by passing an electric current between one electrode, which forms the bottom of the reaction chamber, and another located in the reactor wall at the one-foot level,
4. A centrally located, vertical inlet pipe to introduce solid and gaseous feed components to the melt at a level one inch above the bottom, and
5. Means for measuring temperature at any vertical level in the system.

The design has been checked against requirements when studying combustion and has been found satisfactory.

A process flowsheet has been drawn to incorporate the necessary feed and product handling facilities, variable measurement, and safety features. Ranges of extrinsic variables have been specified for meter and control system sizing.

Both designs have been turned over to the pilot plant design group to prepare detailed fabrication and erection drawings and to determine cost and time of completion.

B. Projections

Gasification work will establish the effect of bituminous coal ash content on rate. Preliminary evaluation of combustion will also be made in the atmospheric unit. Effect of carbon in the melt on the viscosity of a melt containing coal ash will be evaluated.



IV. MECHANICAL DEVELOPMENT

A. Accomplishments

1. Environmental Testing of High Temperature Materials

All the materials required for Test #3 have been procured and the test equipment has been set up. The test specimens having been prepared and loaded into the test vessel consist of three samples each of:

1. RA 446
2. Thermalloy 38
3. Zircofrax "0" (carborundum)
4. Thermalloy "28"
5. Zirconia Flame Sprayed RA 330
6. Ritex CB (Grefco)
7. Chromex B (Harbison-Walker)
8. Monofrax K-3 (Harbison-Carborundum)
9. Monofrax A (Harbison-Carborundum)
10. Zirconia Y-1027 (Zirconium Corporation of America)

2. Coal Feeding Studies

The experimental set up outlined previously has been modified to permit operation with coal as the solid material and nitrogen as the transporting medium. Some twenty runs have been conducted with the set up thus modified, at the following conditions:

Pressure in transport tube - 1 & 20 psig
Gas Velocity in transport tube - 25, 30, 40, & 50 fps
Solids loading (coal) approximately - 0.04 to 0.43 lb/ft³ of N₂



The coal used was a bituminous coal with a sieve analysis of:

<u>Sieve</u>	<u>Microns</u>	<u>Wt. %</u>
20	840	69.60
60	250	17.50
100	149	4.53
200	74	3.71
325	44	1.65
Pan	-	<u>3.02</u>
		100.0

The results of these tests appear as Figures 1 and 2

In addition, a star feeder has been designed for use with the bench-scale reactor. A drawing of this feeder appears as Figure 3. All component parts and materials of construction for this feeder have been ordered and are now on hand.

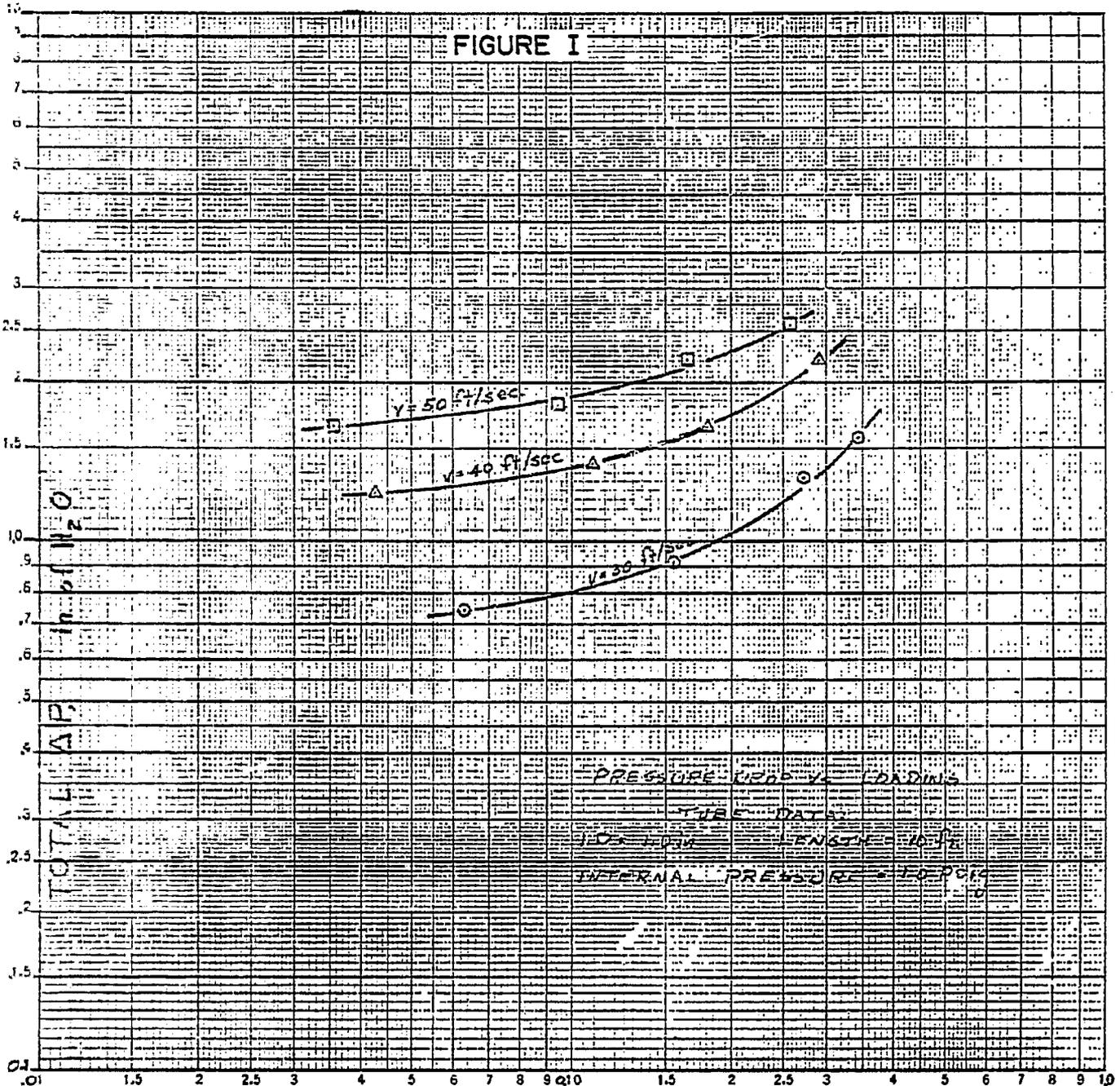
B. Projections

1. Environmental Testing of High Temperature Materials

Extended testing of the materials at temperature in the reactor will begin immediately.

2. Coal Feeding Studies

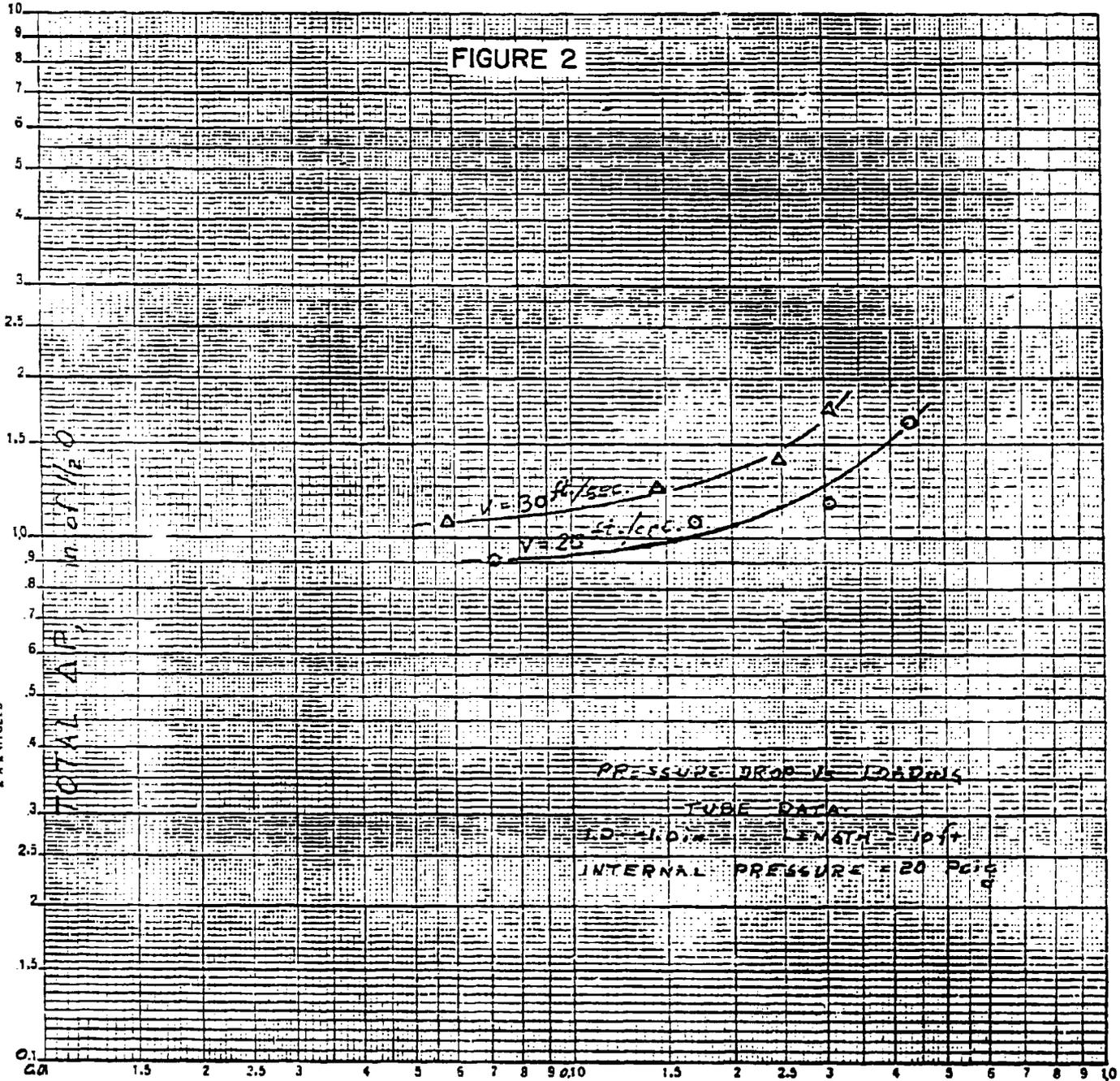
Work will commence on the fabrication of the feeder for the bench scale reactor. This feeder will then be tested for operation prior to installation in the coal feeding system.



COAL LOADING, lb/ft³ of N₂

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