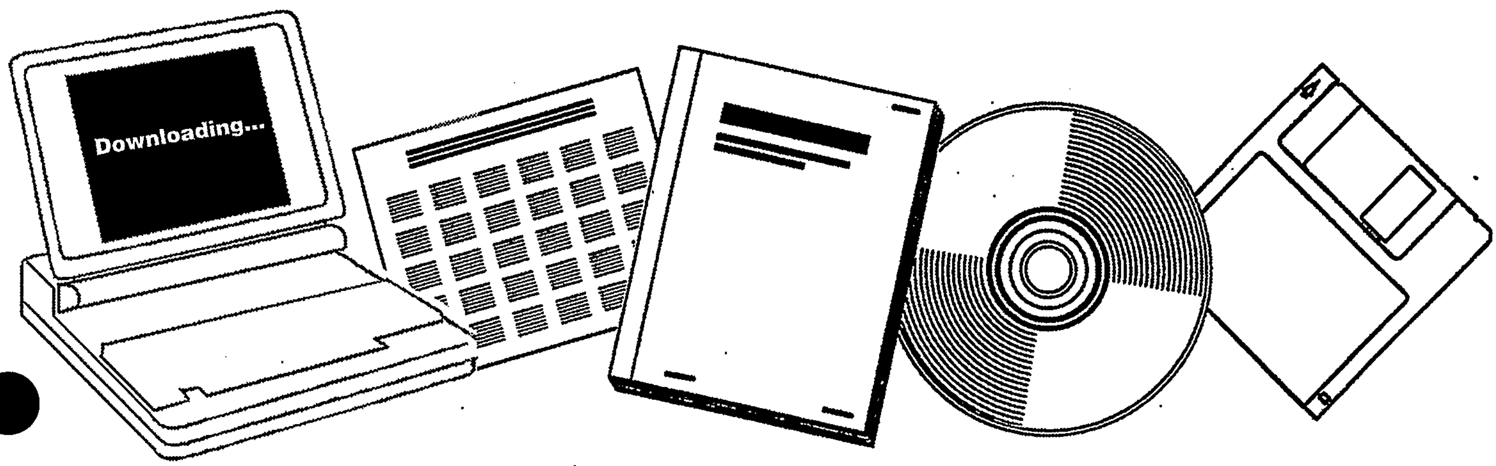
PIPELINE GAS FROM COAL: HYDROGENATION (IGT HYDROGASIFICATION PROCESS). PROJECT STATUS REPORTS, JANUARY--DECEMBER 1970

INSTITUTE OF GAS TECHNOLOGY, CHICAGO, ILL

1970



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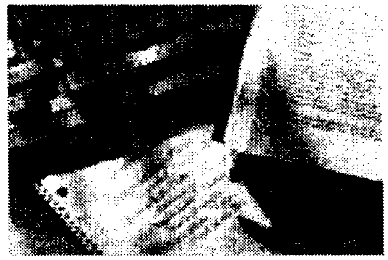

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MASTER

PIPELINE GAS FROM COAL - HYDROGENATION (IGT HYDROGASIFICATION PROCESS)

Project Status Reports for the
Period January - December 1970

NOTICE

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Institute of Gas Technology
IIT Center
Chicago, Illinois 60616

Prepared for

Office of Coal Research
U. S. Department of the Interior
and
American Gas Association

OCR Contract No. 14-01-0001-381*
AGA Project No. IU-4-1

*This contract evolved into OCR Contract No. 14-32-0001-1221

FE--381-T-7

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IGT-MPR--1/70

INSTITUTE OF GAS TECHNOLOGY · IIT CENTER - CHICAGO 60616

Project Status Report
For
OFFICE OF COAL RESEARCH
and
AMERICAN GAS ASSOCIATION

Report For January 1970
OCR Report No. 64

Project Title Pipeline Gas From Coal — Hydrogenation (IGT Hydrogasification Process)

OCR Contract No. 14-01-0001-381 (1) A.G.A. Project No. IU-4-1

I. Project Objective

The overall objective of this project is a process for production of pipeline gas from coal that is economically attractive for supplementing natural gas supplies. The present objective is the design, construction, and operation of a large integrated pilot plant to obtain scale-up data and operating experience. Developmental research, engineering studies, and economic evaluations are in progress to help attain this objective.

II. Achievements

HIGH-PRESSURE METHANATION

Results from a series of runs at low flow rates and low pressures checked well with results at high pressures. These tests are aimed at extending the range of methanation rate correlation.

ENGINEERING ECONOMICS STUDIES

A computer program was developed to estimate the cost of vessels as a function of vessel dimensions and configurations. We began a study of the effect of using air cooling to reduce the requirements for cooling water in pipeline-gas plants.

DEVELOPMENT UNIT STUDIES

Results from a free-fall thermal treatment at 1300°F of lignite showed 14% carbon gasification using nitrogen as a sweep gas. The degree of gasification at 280 psi is comparable to that from another run at 1000 psi, indicating devolatilization is the only reaction occurring.

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Hydrogasification of lignite at 500 psi with hydrogen and steam showed 41% carbon gasification, indicating no significant loss of reactivity from the 1000-psi operation.

A series of pretreatment runs this month produced char for hydrogasification and also data to define details of pretreater operation in the HYGAS pilot plant.

Four tests were made in the electrothermal gasifier at 1000 psi with a concentric electrode arrangement. All four tests were stopped due to difficulties in equipment such as feeders and the pressure-relief valve. These difficulties, although minor, all led to a loss of fluidization, which caused arcing.

Work was begun on evaluating the performance of spray devices that would be used in the HYGAS pilot plant for distributing the coal feed slurry to the slurry dryer. Equipment is being set up to study the dry letdown of char from high pressure.

NEW PROCESS STUDIES

Design and cost estimates are being prepared for gas transfer ducts and current collection equipment. Electrical overload and protection systems are being designed.

PILOT PLANT CONSTRUCTION

Engineering is 90% complete, purchasing is 83% complete, and material receipt is 35% complete. The hydrogasifier reactor vessel was shipped from Struthers-Wells Corp. on January 17 and should arrive on the site the first week of February. The package hydrogen plant is also complete and en route from C&I/Girdler. Construction is 20% complete. Plant completion date, barring labor shortage and unexpected cold weather, is early July 1970.

III. Problems

No major problems were encountered this month.

IV. Recommendations

We recommend that the project proceed into the areas defined in the contract amendment.

V. Status of Funding

1. A.G.A. Funding

A.	1970 Funds Allocated	$300,000
B.	Funds Expended This Month (estimated)	$ 36,600
C.	Funds Expended to Date (estimated)	$ 36,600

2. OCR Funding

A.	Funds Expended This Month (estimated)	$ 262,000
B.	Funds Expended Since Contract Amendment No. 1 (estimated)	$ 3,664,000

As a result of personally reviewing the pertinent data and information reasonably available, it is our opinion that the project's objective will be attained within the contract term and the funds allocated.

Approved ______________
Frank C. Schora, Jr.
Director

Signed ______________
Bernard S. Lee
Manager

____
Jack Huebler
Vice-President

Appendix. Achievements in January

COAL CHARACTERIZATION

A revised plot of the volatile matter versus vitrinite reflectance of coals used in hydrogasification tests is shown in Figure 1. These points may be

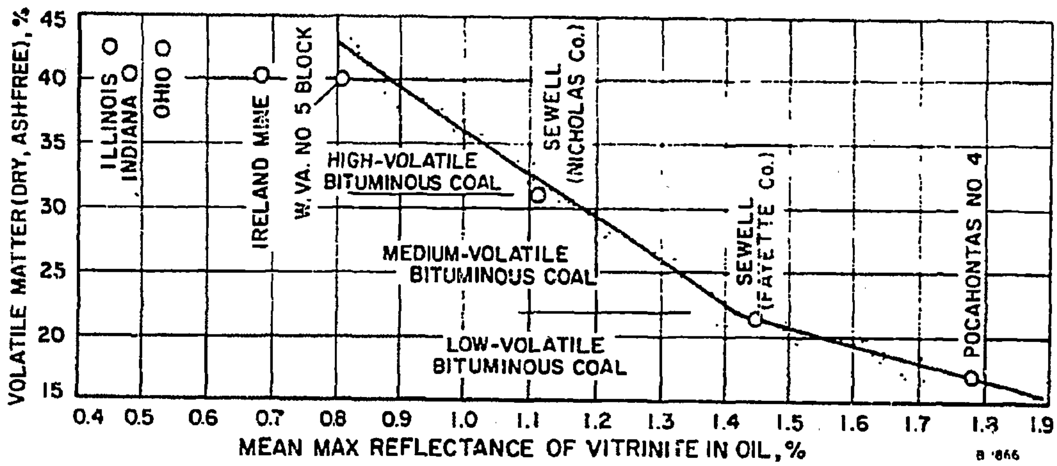

Figure 1. VOLATILE MATTER AND VITRINITE REFLECTANCE OF TEST COALS

compared to results, also shown, compiled from a number of petrography laboratories.* The Sewell seam coal from Nicholas County, W.Va., has been added since the plot was shown in the Fourth Quarter, 1967, Project Status Report. This exhibits the coverage of the range of rank of bituminous coals we have achieved.

HIGH-PRESSURE METHANATION

To extend the range of the rate of methane formation, data were obtained at low flow rates and low pressures. The results, Runs 292 to 295, are in good agreement with high-pressure data obtained previously. (See Table 1.) Run 296 was made to check the catalyst activity after the catalyst was in use

* Bayer, J. L., "Report on Comparative Coal Petrographic Analyses by U.S. Laboratories." Paper presented at meeting of Coal Petrographers at the Illinois State Geological Survey, Urbana, Illinois, October 8, 1964.

Table 1. LOW PRESSURE-LOW FLOW RATE METHANATION EXPERIMENTAL DATA

Run No.	Temp, °F	Pressure, psig	Flow Rate, SCF/hr	r_{CH_4}, (lb mole/hr-g cat.) X 10^{-4}
292	575	103	3.35	1.13
293	575	103	1.62	0.85
294	575	102	0.84	0.56
295	575	102	0.83	0.58
296	575	594	1.25	1.02

since Run 291. We found that the rate was lower by 15% than the average of previous runs at the same conditions. Analysis of the catalyst after Run 296 showed that the carbon content increased from 4.9 to 6.1% by weight, the significance of which is being determined. Changes in sulfur or nickel content were negligible.

ENGINEERING ECONOMICS

A computer program for estimating the cost of vessels used in plant estimates has been developed. The program calculates the wall thickness and weight of shell plus ellipsoidal heads, adds the cost of attachments and fittings, and determines the cost of a shop-fabricated vessel from a weight vs. cents/lb relationship corrected for cost-index changes.

As a corollary program, the size and cost of reactor vessels such as methanation or shift reactors can be calculated from design gas linear and space velocities, including allowance for internal insulation.

We are starting to work on the effect of reducing cooling water requirements by the use of air cooling where feasible in the lignite pipeline gas plant design.

DEVELOPMENT UNIT STUDIES

Hydrogasification Tests

We are presenting the results of two gasification tests conducted in previous months in the high-temperature balanced-pressure development unit (Runs HT-234 and HT-239) with untreated North Dakota lignite from the Glenharold mine. Run HT-234 (Fourth Quarter, 1969, Project Status Report) was designed to obtain a limited conversion of the lignite (20% or less, if possible) by thermal treatment of the lignite while in free fall through the 15-1/2-ft heated section of the hydrogasification reactor tube, with no feed gas. The objective of Run HT-239 (Fourth Quarter, 1969, Project Status Report) was to study the hydrogasification reactivity of a North Dakota lignite with hydrogen and steam in a fluidized bed at a system pressure of 500 psig. Feed lignite for these tests was prepared by crushing and screening to a –10+80 mesh size and air-drying in the fluidized-bed coal pretreatment unit at 220°-240°F.

Operating conditions and results of these two runs are presented in Tables 2 and 3. Compositions and screen analyses of the feeds and residues are given in Table 4. Liquid products and compositions are shown in Table 5.

The thermal treatment of North Dakota lignite in free fall in Run HT-234 (Table 2) resulted in the gasification of 24.5% of the lignite on a moisture-, ash-free basis. Gasification of carbon in the lignite was limited to 13.7%. This includes the formation of hydrocarbons (primarly methane), carbon monoxide, and carbon dioxide. Additional lignite was converted to liquid products with 7.11% going into the formation of water and 1.91% producing oils. The carbon appearing in the oil represents 2.41% of the carbon in the lignite (Table 5). Based on an ash balance, 0.728 lb of partially gasified lignite was recovered for each pound of lignite fed. The volatile matter content of the lignite was reduced from 40.0 to 11.9%, while the ash content increased from 7.66 to 10.54% as a result of the carbonization (Table 4). The results of this test are similar to those obtained in Run HT-229 (August 1969 Project Status Report), the first limited conversion test performed with a Montana lignite at similar conditions, but at a system pressure of 1004 psig.

The hydrogasification reactivity of a North Dakota lignite with hydrogen and steam remains relatively high at a system pressure of 500 psig, as shown by the results in Table 3. On a moisture-, ash-free basis, over 54% of the lignite was gasified, with 41.4% of the carbon in the lignite converted to gaseous products. For comparison purposes, key results of this test are presented in Table 6 with those of Run HT-145 (February 1967 Project Status Report), conducted at generally similar conditions, but at a system pressure of 1000 psig. Carbon gasification at 500 psig was only somewhat less than the 42.1% obtained at 1000 psig. However, there was a shift in the proportion of the carbon in the lignite converted to hydrocarbons and to carbon oxides as shown by the SCF/lb yield data for these gaseous products. The hydrocarbon yield at 500 psig was only 86% of that at 1000 psig, while the carbon

Table 2. OPERATING CONDITIONS AND RESULTS OF THE CARBONIZATION OF A NORTH DAKOTA LIGNITE IN HIGH-TEMPERATURE ADIABATIC REACTOR FOR RUN HT-234

Coal	North Dakota Lignite
Source	Glenharold Mine
Sieve Size, USS	−10+80
Run No.	HT-234
Duration of Test, hr	3-1/4
Steady-State Operating Period, min[a]	134-212
OPERATING CONDITIONS	
Bed Height, ft	Free-fall
Reactor Pressure, psig	282
Reactor Temperature, °F[b]	
Inches From Bottom	
62-1/2	955
67-3/4	1200
73	1280
78-1/4	1310
83-1/2	1380
89	1265
94-1/2	1190
100	1400
104	1300
114	1400
124-1/2	1265
135	1430
145	1280
155-1/2	1420
166	1425
176	1310
187-1/2	1360
197	1355
207	1255
217-1/2	1060
Average	1290
Lignite Rate, lb/hr[c]	19.68
Sweep Nitrogen Rate, SCF/hr	185.1
Superficial Sweep Nitrogen Velocity, ft/s[d]	0.0975
Lignite Space Velocity, lb/cu ft-hr[e]	14.37
OPERATING RESULTS	
Product-Gas Rate (nitrogen-free), SCF/hr	59.59
Net Btu Recovery, 1000 Btu/lb	1.404
Product-Gas Yield, SCF/lb	3.027
Hydrocarbon Yield, SCF/lb	0.902
Carbon Oxides Yield, SCF/lb	1.659
Residue, lb/lb lignite[f]	0.728
Liquid Products, lb/lb lignite	
Water	0.0711
Oil	0.0191

Table 2, Cont. OPERATING CONDITIONS AND RESULTS OF THE CARBONIZATION OF A NORTH DAKOTA LIGNITE IN HIGH-TEMPERATURE ADIABATIC REACTOR FOR RUN HT-234

	HT-234
Net MAF Lignite Gasified, wt %[g]	24.48
Carbon Gasified, wt %	13.68
Overall Material Balance, %	103.1
Carbon Balance, %	104.3
Hydrogen Balance, %	89.3
Oxygen Balance, %	102.7
PRODUCT GAS PROPERTIES	
Gas Composition (nitrogen-free), mole %	
Carbon Monoxide	14.3
Carbon Dioxide	40.5
Hydrogen	15.4
Methane	25.0
Ethane	1.2
Propane	3.6
Benzene	Trace
Total	100.0
Heating Value, Btu/SCF[h]	456
Specific Gravity (Air = 1.00)	0.975
Nitrogen Purge Rate, SCF/hr	183.5

a. From start of lignite feed.

b. Tube wall temperatures. Heated reactor length of 15.5 ft. Sweep nitrogen inlet at 62-in. level.

c. Operating conditions and results based on weight of dry feed.

d. (CF/s sweep nitrogen at reactor pressure and temperature)/cross-sectional area of reactor.

e. Based on 1.37-cu-ft heated reactor volume.

f. By weight of residue recovered.

g. 100(wt product gas/wt of moisture-, ash-free lignite).

h. Gross, gas saturated at 60°F, 30-in. Hg pressure. SCF: dry gas volume in SCF at 60°F, 30-in. Hg pressure.

Table 3. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF A NORTH DAKOTA LIGNITE IN HIGH-TEMPERATURE ADIABATIC REACTOR FOR RUN HT-239

Coal	North Dakota Lignite
Source	Glenharold Mine
Sieve Size, USS	−10 +80
Run No.	HT-239
Duration of Test, hr	5-1/4
Steady-State Operating Period, min[a]	165-307
OPERATING CONDITIONS	
Bed Height, ft	3.5
Reactor Pressure, psig	478
Reactor Temperature, °F[b]	
Inches From Bottom	
62-1/2	1310
67-3/4	1480
73	1500
78-1/4	1515
83-1/2	1620
89	1550
94-1/4	1520
100	1680
104	1630
Average	1535
Lignite Rate, lb/hr[c]	32.07
Hydrogen Gas Rate, SCF/hr	368.8
Steam Rate, lb/hr	8.69
Steam, mole % of hydrogen-steam mixture	33.1
Hydrogen/Lignite Ratio, % of stoichiometric[d]	34.2
Hydrogen/Steam Ratio, mole/mole	2.02
Bed-Pressure Differential, in. wc	--
Lignite Space Velocity, lb/cu ft-hr	103.7
Feed-Gas Residence Time, min[e]	0.294
Superficial Feed-Gas Velocity, ft/s[f]	0.199
OPERATING RESULTS	
Product-Gas Rate, SCF/hr	694.8
Net Btu Recovery, 10^3 Btu/lb	4.579
Product-Gas Yield, SCF/lb	21.67
Hydrocarbon Yield, SCF/lb	4.33
Carbon Oxides Yield, SCF/lb	3.68
Net Reacted Hydrogen, SCF/lb	3.33
Residue, lb/lb coal[g]	0.439
Liquid Products, lb/lb coal[h]	0.261
Net MAF Lignite Hydrogasified, wt %[i]	54.2
Carbon Gasified, wt %	41.4
Steam Decomposed, lb/hr[j]	1.49
Steam Decomposed, % of steam fed	17.16
Steam Decomposed, % of total equivalent fed[k]	56.9

Table 3, Part 2. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF A NORTH DAKOTA LIGNITE IN HIGH-TEMPERATURE ADIABATIC REACTOR FOR RUN HT-239

	HT-239
Overall Material Balance, %	97.3
Carbon Balance, %	97.5
Hydrogen Balance, %	92.0
Oxygen Balance, %	94.5
PRODUCT GAS PROPERTIES	
Gas Composition, mole %	
Nitrogen	24.8
Carbon Monoxide	9.4
Carbon Dioxide	7.6
Hydrogen	37.7
Methane	18.7
Ethane	0.9
Propane	0.4
Butane	--
Benzene	0.4
Hydrogen Sulfide	0.1
Total	100.0
Heating Value, Btu/SCF[m]	377
Specific Gravity (Air = 1.00)	0.606
Nitrogen Purge Rate, SCF/hr	172

Table 3, Part 3. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF A NORTH DAKOTA LIGNITE IN HIGH-TEMPERATURE ADIABATIC REACTOR FOR RUN HT-239

a. From start of coal feed.

b. Tube wall temperatures. Bottom of coal bed at 62 in.

c. Operating conditions and results based on weight of dry feed.

d. Percent of the stoichiometric hydrogen/char ratio – the net feed hydrogen/char ratio required to convert all the carbon to methane.

e. Coal bed volume/(CF/min feed gas at reactor pressure and temperature).

f. (CF/s feed gas at reactor pressure and temperature)/cross-sectional area of reactor.

g. By ash balance.

h. Includes condensed, undecomposed steam.

i. 100 (wt of product gas-wt hydrogen in-wt decomposed steam-wt nitrogen in/wt of moisture-, ash-free coal).

j. Computed as difference between steam feed rate and the measured liquid water rate leaving the reactor.

k. Computed as difference between the total equivalent steam feed rate (includes moisture content of feed char and bound water corresponding to oxygen content of feed char) and the measured liquid water rate leaving the reactor.

m. Gross, gas saturated at 60°F, 30-in. Hg pressure. SCF: dry gas volume in SCF at 60°F, 30-in. Hg pressure.

Table 4. CHEMICAL AND SCREEN ANALYSES OF LIGNITE FEED AND RESIDUE

Run No.	HT-234		HT-239	
Sample	Feed	Residue	Feed	Residue
Proximate Analysis, wt %				
Moisture	1.1	0.2	2.1	1.0
Volatile Matter	40.0	11.9	38.3	5.6
Fixed Carbon	51.3	77.4	51.8	75.1
Ash	7.6	10.5	7.8	18.3
Total	100.0	100.0	100.0	100.0
Ultimate Analysis (dry), wt %				
Carbon	65.3	79.3	65.7	76.9
Hydrogen	4.35	2.35	4.12	0.86
Nitrogen	0.93	1.42	1.10	0.39
Oxygen	21.07	5.69	20.26	2.74
Sulfur	0.69	0.70	0.81	0.65
Ash	7.66	10.54	8.01	18.46
Total	100.00	100.00	100.00	100.00
Screen Analysis, USS, wt %				
+20	25.7	24.5	22.0	13.4
+30	19.0	11.6	14.7	22.8
+40	23.2	20.8	21.9	26.5
+60	20.4	22.3	23.7	22.5
+80	7.6	10.3	10.3	6.9
+100	2.1	3.5	3.4	2.2
+200	1.6	4.6	3.0	4.1
+325	0.2	1.1	0.5	1.1
−325	0.2	1.3	0.5	0.5
Total	100.0	100.0	100.0	100.0

Table 5. COMPOSITION OF LIGNITE CARBONIZATION LIQUID PRODUCTS

Run No.	HT-234	HT-239
Sample	Condenser	Condenser
Liquid Products,		
lb/lb lignite	0.0902	0.2609
Composition of Liquid		
Products, wt %		
Water	78.79	86.03
Oil	21.21	13.97
Total	100.00	100.00
Composition of Oil		
Fraction, wt %		
Carbon	82.2	85.7
Hydrogen	7.63	7.49
Total	89.83	93.19
Carbon in Oil Fraction,		
lb/lb lignite	0.0157	0.0312
wt % of carbon in lignite	2.41	4.75

Table 6. COMPARISON OF NORTH DAKOTA LIGNITE HYDROGASIFICATION RESULTS AT 1000 AND 500 psig

Run No.	HT-145	HT-239
Reactor Pressure, psig	1048	478
Lignite Bed Temp Average, °F	1675	1535
Lignite Bed Height, ft	3.5	3.5
Lignite Feed Rate, lb/hr	85.76	32.07
Hydrogen Feed Rate, SCF/hr	710.3	368.8
Steam Feed Rate, lb/hr	15.35	8.69
Steam/Lignite Ratio, lb/lb	0.179	0.271
Hydrogen/Lignite Ratio, % of stoichiometric	24.7	34.2
Steam Concentration in Feed Gas, mole %	31.2	33.1
Steam Decomposed, % to total equivalent steam fed	75.9	56.9
Carbon Gasified, %	42.1	41.4
MAF Lignite Gasified, %	58.1	54.2
Hydrocarbon Yield, SCF/lb	5.06	4.33
$CO + CO_2$ Yield, SCF/lb	3.22	3.68
Carbon in Oil Fraction, % of carbon in lignite	7.39	4.75
Product-Gas Composition (nitrogen-free), mole %		
Carbon Monoxide	11.2	12.5
Carbon Dioxide	14.2	10.1
Hydrogen	34.9	50.2
Methane	37.1	24.9
Ethane	1.6	1.2
Propane	0.8	0.5
Benzene	0.2	0.5
Hydrogen Sulfide	--	0.1
Total	100.0	100.0
Product-Gas Heating Value (nitrogen-free), Btu/SCF	574	501

oxides yield at the lower pressure was 115% of that at the higher pressure. The higher hydrogen-to-lignite ratio of Run HT-239 (34.2% of stoichiometric) compared to that of Run HT-145 (24.7% of stoichiometric) is partially responsible for the lignite gasification at 500 psig being similar to that at 1000 psig. Carbon conversion to oil at 500 psig (4.75% of the carbon in the lignite) was 64% less than that converted to oil at 1000 psig (7.39% of the carbon in the lignite). A lower methane concentration in the product gas of the test at 500 psig is as expected because of lower equilibrium methane concentration at this pressure compared with that at 1000 psig.

During the month we conducted a series of coal pretreatment operations (Run FP-141) in which we lightly pretreated 1250 lb of Pittsburgh No. 8 seam, Ireland mine bituminous coal to provide a feed for the hydrogasification development unit. The coal was pretreated with air and nitrogen in a fluidized bed at 750°-800°F. Laboratory agglomeration tests of the pretreated coal showed that all of the caking characteristics were not destroyed. The coal will, therefore, have to be treated a second time before it can be used for hydrogasification. During the pretreatment operation we collected and separated about 85 lb of oils and tars from the condensed liquids for detailed laboratory chemical and physical analysis. We also measured the scrubbing water rate to the pretreatment unit venturi scrubber. This averaged 9950 lb/hr. Data on the oils and tars and the scrubbing water will be used as guides for operation of the coal pretreater section of the hydrogasification pilot plant currently under construction.

ELECTROTHERMAL GASIFICATION

Four tests were conducted in the electrothermal gasifier during the month. The tests were conducted with the concentric electrode configuration at 1000 psig using a hydrogasified high-volatile bituminous char. The four tests were terminated before steady-state operating conditions were attained due to various operational difficulties.

The first run, EG-39, was terminated when the reactor discharge screw jammed, preventing the continual flow of solids through the unit, and the steam-transfer line heater to the orifice meter failed, causing erratic steam flow rate measurement.

A pressure upset occurred during the heat-up period of Run EG-40, causing unloading of char from the feed hopper to the reactor, settling of the lower section of the char bed, and subsequent arcing near the electrode tip. The test was terminated when the char bed bridged in the reactor and prevented the discharge of char from the unit.

Inspection of the unit following the run showed the electrode to be slightly melted along a 3-inch length from the tip. A 2-3 inch hole was burned through the 6-in.-ID stainless steel shell. We think the burnout was caused by the loss of fluidization in the bed near the electrode tip.

Run EG-41, conducted at the same conditions as the previous tests, was terminated during the reactor heat-up period when a sudden increase in the bed temperatures (1300° - 1800°F) near the electrode tip occurred soon after establishing the steam flow rate into the reactor. Removal of the electrode showed a burned spot 1 inch above the tip opposite the hole in the reactor wall left over from the previous run.

The center electrode was shortened 6 in. prior to Run EG-42 to avoid any adverse effect of the hole in the reactor wall on the operation of the unit. Also, the reactor heat-up rate was maintained at a slower rate to prevent overheating of the electrode tip. Following a smooth heat-up to 1300°F, the steam flow rate was established and the power input was increased to 15 kW to reach 1900°F. As the bed temperature approached run conditions several thermocouples indicated a hot spot near the reactor tip. The run was terminated when the overall resistance of the system decreased from 0.7 to less than 0.5 ohm and the bed temperature near the electrode tip rose to 2200°F. Removal of the electrode following the run showed it to be severely melted along one side for 15 inches from the tip. A large hole was burned through the 6-in. reactor shell opposite the melted area of the electrode. The damage to the reactor liner will require its replacement. The extent of the burned section is shown in Figure 2.

A suspected cause of the high current density which caused the reactor damage was the venting of steam through a spring-loaded high-pressure relief valve at the exit of the steam superheater. This occurred downstream from the steam flow orifice meter; the actual flow of steam to the reactor was insufficient to fluidize the char bed. The steam was observed to be venting during the shutdown of the unit, although the superheater was operating well below the relief setting of the valve. A successful leak test was performed prior to the run at room temperatures, but the valve may have unseated when

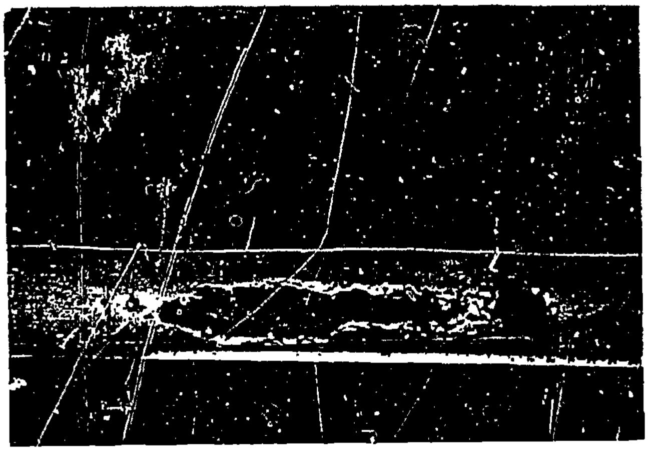

Figure 2. 6-in. REACTOR LINER REMOVED FOLLOWING RUN EG-42

subjected to operating conditions. The valve will be closely observed during future tests to ensure its correct operation. A new 6-inch Type-316 stainless steel reactor liner is being installed in the pilot unit.

Coal Slurry Pumping

Work has resumed on the coal slurry feed system. After consultation with the Spraying Systems Company, three different-sized nozzles have been ordered for evaluation.

We plan photographic determination of the median slurry particle size at various flow rates through each nozzle.

Dry-Char Letdown

Work has begun on assembling the equipment needed to gather data on the letdown of dry char from high pressure. A modified Conoflow valve has been installed at the exit of our high-pressure vessel. A cyclone separator has been installed to retain the exit char. The study will attempt to determine the ratio of solid to gas flow and the attrition of char particles.

FUEL CELL ENGINEERING STUDY

Errata

Total PEC for estimate B shown in Table 8 of the Fourth Quarter, 1969, Project Status Report should be read as 65.5 in place of 56.5.

Gas Duct System

The design of a package of 8 batteries each consisting of 100 cells was completed last month and an estimation was made for Purchased Equipment Cost (PEC) of the package. The packages are arranged in rows of 8, with each row designated as a power unit. A pair of power units is supplied with gases through a single main duct. The duct walls are composed of mild steel casing lined with a refractory and insulating material. The cost of the duct work for supply of gases to all the power units is estimated to be about $12/kW. A cost estimation of duct work for burners, heat exchangers, and other equipment is in progress.

Current Collection and Battery Protection

The packages in the power unit are connected in series to build up 4800 V. All the power units are then connected in parallel. Since the packages are operating at 1400°F, the stainless steel electrical terminals have to be cooled to 700°F or lower before copper conductors can be used. In case a package fails during operation, the package has to be isolated from the circuit for maintenance. Circuit breakers and isolating switches should therefore be incorporated in the circuit. Design and cost estimations are being prepared.

Engineering design of the heat transfer and recovery system is also in progress. Design of other components such as blowers, burners, and gas stream splitters will be considered next.

PILOT PLANT CONSTRUCTION

Engineering

A major portion of the engineering effort during this report period was instrumentation details. Electrical detail drawings are 95% complete and instrument drawings are 40% complete. Eighty-five percent of the piping plans and details have been issued. The total project detailed design and drafting is 92% complete.

Procurement

The hydrogasifier reactor was shipped on January 17 and is scheduled to arrive during the first week in February. Figure 3 shows the completed vessel

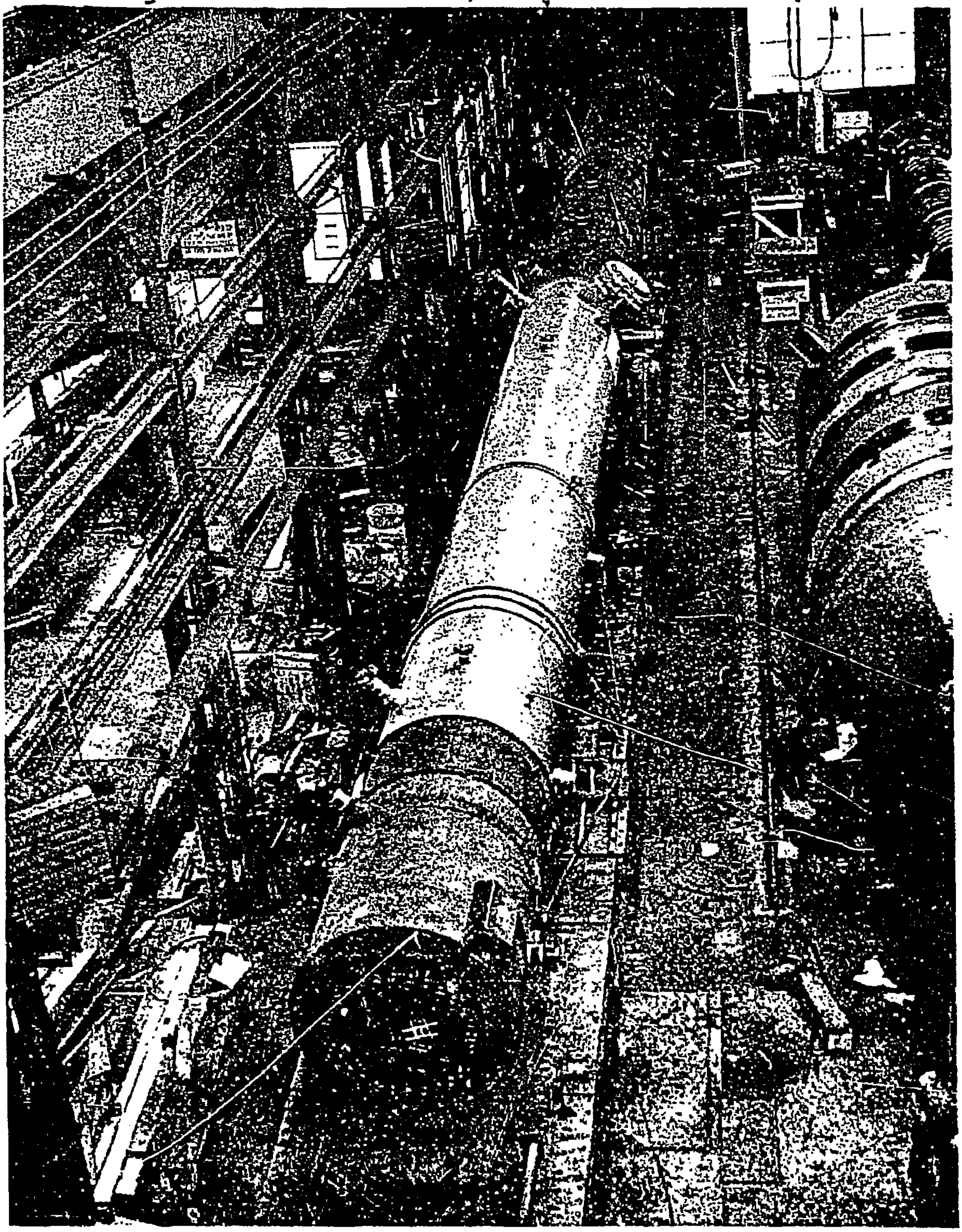

Figure 3. HYDROGASIFIER REACTOR READY FOR SHIPPING

ready for shipping. Other major equipment is arriving as promised; no items have been delayed. Several large package plants (i.e., hydrogen and sulfur plants) are to be delivered during this report period.

The alloy pipe fabrication purchase order was issued and delivery of all alloy pipe has been promised by April 1. The subcontract for the electrical work will be let during the week of January 19.

The percent completion reported above for purchasing includes subcontracts. Purchasing, excluding subcontracts, is 94% complete.

Construction

Major equipment is being erected as received. Field pipe fabrication is 5% complete. Concrete pipe rack bents are 60% fabricated and 40% erected. The gin poles are being readied for the reactor erection. We anticipate the necessity of beginning overtime construction during the next report period to maintain the manpower level required to complete the project on schedule.

Schedule

The project schedule was updated and reviewed with Procon. The computer run was made and reflects a negative float of approximately 15 days, based on using July 1, 1970, as the target completion date. Procon feels that there are some construction activity durations that can be reduced as field work progresses to reduce the total negative float for the project.

It should be noted that the project schedule does not include any cold-weather time. This time could cause an extention of the completion date. Five such cold-weather days have occurred to date.

Figure 4 shows the overall plant in its present state of completion, looking west.

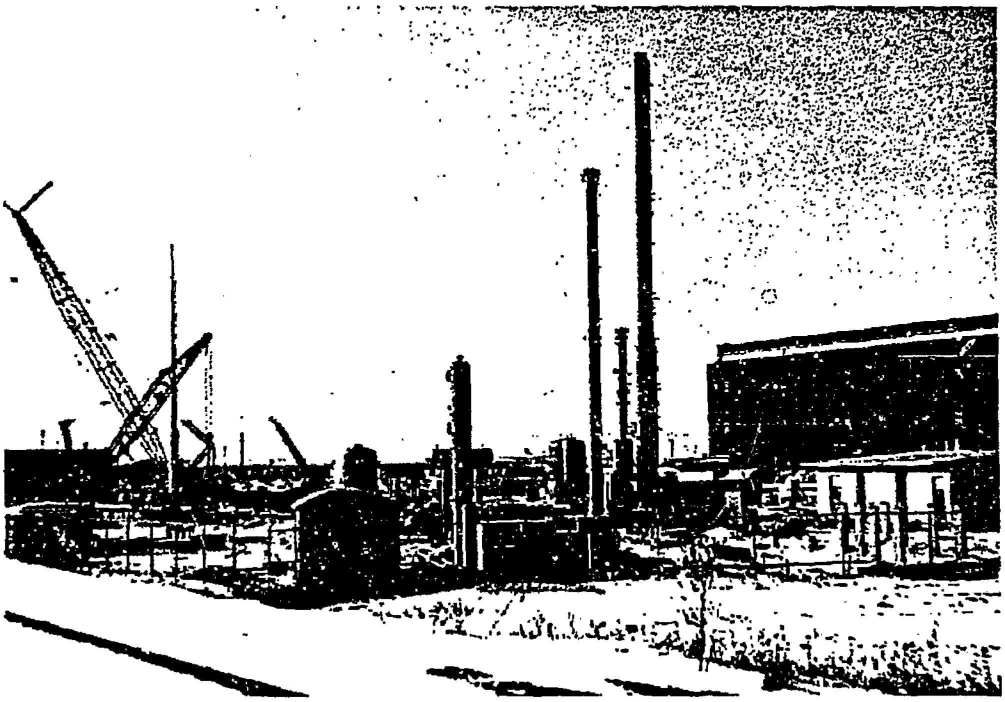

Figure 4. OVERALL PLANT, LOOKING WEST

IGT-MPR--2/70

INSTITUTE OF GAS TECHNOLOGY - IIT CENTER - CHICAGO 60616

Project Status Report
For
OFFICE OF COAL RESEARCH
and
AMERICAN GAS ASSOCIATION

Report For February 1970
OCR Report No. 65

Project Title Pipeline Gas From Coal – Hydrogenation (IGT Hydrogasification Process)

OCR Contract No. 14-01-0001-381 (1) A.G.A. Project No. IU-4-1

I. Project Objective

The overall objective of this project is a process for production of pipeline gas from coal that is economically attractive for supplementing natural gas supplies. The present objective is the design, construction, and operation of a large integrated pilot plant to obtain scale-up data and operating experience. Developmental research, engineering studies, and economic evaluations are in progress to help attain this objective.

II. Achievements

COAL CHARACTERIZATION

A test for rating the attrition resistance of coal chars is being developed. We are studying the behavior of compounds formed from minor coal constituents in the HYGAS Process steps, for example, ammonia formed from nitrogen in the coal.

HIGH-PRESSURE METHANATION

The methanation rate expression has been extended to cover regions with large excesses of hydrogen and methane. An improved correlation was obtained.

$$r = \frac{k_1 p_{CO} p_{H_2}^{0.5}}{1 + k_2 p_{H_2} + k_3 p_{CH_4}}$$

Data at low conversions and near equilibrium are being collected to test this correlation.

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A small laboratory is being set up to measure the sulfur tolerance of the type of nickel catalyst that will be used in the pilot plant. Sulfur concentrations in the low-ppm range will be used under conditions simulating plant operation.

ENGINEERING ECONOMICS STUDIES

A study using air in place of water cooling in a HYGAS plant shows that the cooling water requirement can be reduced by as much as 63%. This assumes air cooling to 140°F with final cooling by water.

The effects of financial factors on the return on equity for gas utility financing are discussed and the results presented in graphical form.

DEVELOPMENT UNIT STUDIES

Results of lignite gasification at 500 psi with synthesis gas-steam and hydrogen-steam mixtures show that about 5% more carbon (36 vs. 41%) was gasified with the hydrogen-steam mixture. However, either gas mixture is adequate for the HYGAS Process in terms of obtaining the required gasification.

Minor mechanical difficulties first in the exit-gas quench system and then in the coal feed system caused shutdown of two runs in the electrothermal gasifier. Modifications were made that eliminated the problems in the quench system.

NEW PROCESS STUDIES

Design work for accessories in the 400-MW fuel cell power plant is complete. Capital investment ranges from $94 to $168 per kW for two power densities and two sets of equipment cost factors. A power cost of 4-5 mills/kWhr is estimated for a high-power density of 300 watts/sq ft.

PILOT PLANT CONSTRUCTION

Engineering is 95% complete, purchasing 86% complete, and material receipt 72% complete. The hydrogasifier reactor and the hydrogen plant were received onsite. All major equipment, except for less than 10 small items, have been received. No delays are expected due to equipment arrival.

Construction is 24% complete. Overtime construction, 50 hr/wk, began on February 10 in an effort to attract the pipefitter/welder manpower required to maintain the construction schedule.

III. Problems

No major problems were encountered this month.

IV. Recommendations

We recommend that the project proceed into the areas defined in the contract amendment.

V. Status of Funding

1. A.G.A. Funding

A. 1970 Funds Allocated	$ 300,000
B. Funds Expended This Month (estimated)	$ 36,600
C. Funds Expended to Date (estimated)	$ 73,000

2. OCR Funding

A. Funds Expended This Month (estimated)	$ 993,000
B. Funds Expended Since Contract Amendment No. 1 (estimated)	$4,693,000

As a result of personnaly reviewing the pertinent data and information reasonably available, it is our opinion that the project's objective will be attained within the contract term and the funds allocated.

Approved

Frank C. Schora, Jr.
Director

Signed

Bernard S. Lee
Manager

Jack Huebler
Vice-President

Appendix. Achievements in February

COAL CHARACTERIZATION

A test for rating the attrition resistance of hydrogasification and other chars is being developed. A microtumbler test with results on coal correlated with Hardgrove grindability was investigated. In this test 2 grams of a 20-30-mesh sieve fraction are placed in a 12-in. long, 1-in. ID cylinder with 12 steel balls; the cylinder is turned end-over-end for 800 turns at a specified rate. In the test used on coal, the amount passing a 100-mesh sieve is then determined. Results found on Ireland mine coal and on several chars are shown in Table 1. The value found on Ireland mine coal agrees with those reported in the literature on Pittsburgh seam coal. Reproducibility is satisfactory, but so much of the char is ground in the test that its use for characterization of behavior in a fluidized bed is very questionable. Tumbling with glass balls is being investigated.

Table 1. ATTRITION TESTS

Sample	Ash, wt %: Original Sample	Ash, wt %: −30+40 Sieve Fraction	Sieve Analysis of Tumbled Sample, %: USS 30	−30+100	−100+200	−200
Ireland Mine	--	--	0.8	37.9	24.7	36.6
			0.4	38.2	24.9	36.5
			0.6	38.2	23.7	37.5
Residue HT-210, Ireland Mine Coal	19.4	23.3	0.0	5.0	19.5	75.5
			0.1	3.4	19.2	77.3
			0.0	3.6	17.8	78.7
Residue HT-192, N. D. Lignite	14.6	12.5	0.0	16.9	17.7	65.5
			0.0	16.0	18.3	65.7
			0.0	8.3	19.9	71.7
			0.0	10.4	22.5	67.1
Residue EG-37	27.3	28.4	0.0	4.4	16.2	79.4
			0.2	3.6	14.9	81.3

The effects of nitrogen and minor elements in coal on the hydrogasification process are being investigated. Ammonia is formed in the process, as shown by the presence of a solid ammonium salt (presumably the bicarbonate) in some of the condensates from runs in the experimental unit. Thermodynamic calculations indicate that no solid product of ammonia, carbon dioxide, and water will form in any section of the plant ahead of the quench tower. The effect of the ammonia on a solution of carbon dioxide and hydrogen sulfide in the quench water and the effect of this on the process is being investigated.

Many elements are present in coal in amounts varying from a few parts per million to as much as several percent. Some of these are combined with the organic part of the coal, e.g., germanium, which is present in some coals to as much as 0.2% of the ash. We are investigating which of the elements may have stable volatile forms under the conditions in the hydrogasifier and thus may become trapped therein.

HIGH-PRESSURE METHANATION

Kinetic Study

The methanation reactor was overhauled. New seal rings, bearings, and thermocouples were installed. Analytical instruments, the gas partitioner and infrared analyzer, were calibrated. A total of 138 data points were obtained at 575°F for kinetic studies to date. About 30% of these data were collected to find the order of CO and 35% to find the order of H_2; the remainder were to find the effect of excess H_2 and CH_4 on the rate of methanation and of the presence of an inert (He) and of benzene in the feed gas.

The final set of data used to find a rate model for methanation is presented in Table 2. The order of CO was found to be approximately 1.0 and that of H_2 about 0.45. When the rate model, $r = kp_{CO}p_{H_2}^{0.5}$, was used to fit the data the representation was not good, especially at higher partial pressures of H_2. The data were analyzed for the effect of excess H_2 and CH_4 by plotting

$\frac{p_{CO}p_{H_2}^{0.5}}{r}$ vs. p_{H_2} and $\frac{p_{CO}p_{H_2}^{0.5}}{r}$ vs. p_{CH_4}.

From this study, we found the equation —

$$r = \frac{k_1 p_{CO} p_{H_2}^{0.5}}{1 + k_2 p_{H_2} + k_3 p_{CH_4}}$$

where

k_1 = rate constant = 2×10^{-5} at 575°F

k_2 = 0.1

k_3 = 0.05

p = partial pressure, psig

r = rate of methane formation, $\frac{\text{lb-mole}}{\text{hr-g catalyst}} \times 10^{-4}$

Table 2. METHANATION DATA AT 575°F USED FOR ANALYSIS OF RATE EQUATION

$$r = \frac{2 \times 10^{-5} p_{CO} p_{H_2}^{0.5}}{1 + 0.1 p_{H_2} + 0.05 p_{CH_4}}$$

Run No.	Feed Composition CO, mole %	Feed H$_2$, mole %	Feed CH$_4$, mole %	Feed He, mole %	Press., psig	Flow Rate, SCF/hr	Product Composition CO, mole %	Product H$_2$, mole %	Product CH$_4$, mole %	Rate$_{exp}$, lb mole/hr-g cat X 10^{-4}	Rate$_{calc}$	% Deviation, $\frac{r_e - r_c}{r_e}$ 100
159	--	--	--		573	5,945	0.6	97.9	0.7	0.56	0.74	-33
160	--	--	--	--	575	7,502	7.2	86.5	3.2	3.12	--	--
166	--	--	--	--	575	5,600	7.1	71.6	10.6	7.66	6.39	+15
167	--	--		--	571	6,835	3.0	81.4	5.8	4.81	4.26	+12
168	--	--	--	--	578	5,033	2.3	85.5	5.8	3.75	3.06	+18
169	--	--	--	--	573	6,935	1.3	85.3	4.9	4.32	4.71	-9
170	--	--	--	--	576	5,333	5.4	80.1	7.3	4.94	5.44	-10
171	--	--	--	--	573	6,382	1.7	92.1	3.1	2.49	1.96	+21
172	--	--	--	--	573	5,576	4.0	84.0	6.0	4.23	4.23	0
173	8.4	73	0	18.6	576	5,486	3.9	81.2	6.0	4.17	4.12	+1
174	6.0	55	0	38.4	573	4,311	3.1	83.6	6.1	3.83	4.31	-13
175	--	--	--		576	3,907	4.3	79.1	8.2	4.12	4.20	-2
176	--	--	--	--	576	5,303	1.1	92.7	3.0	2.01	1.63	+19
177	--	--	--	--	576	5,660	0.7	[illegible]	1.7	1.24	0.85	+32
178	--	--	--	--	575	6,235	3.1	84.7	5.1	4.32	4.74	-10
179	8.1	37	0	54.9	575	7,034	4.2	85.1	5.2	4.71	4.56	+3
180	--	--	--	--	283	3,155	4.1	53.6	6.0	3.21	2.33	+27
181	7.3	23	0	69.7	280	3,736	3.9	87.1	4.3	2.10	2.29	-9
182	8.2	31	0	60.8	278	3,617	3.4	84.9	[illegible]	2.55	2.12	+17
183	--	--	--	--	578	5,541	4.0	82.6	4.7	3.02	4.16	-38
184	8.6	69	0	22.3	570	3,534	3.1	53.6	6.1	2.77	4.36	-57
185	--	--	--	--	280	3,677	4.2	81.5	7.1	3.34	2.11	+37
186	4.9	12	0	83.1	281	3,473	3.9	13.7	1.2	0.52	--	--
187	--	--	--	--	281	3,700	1.1	83.6	6.2	2.10	2.19	+25
188	--	--	--	--	277	--	1.1	80.3	--	3.34	--	--
211	11.5	39.7	0	48.8	533	2,711	1.2	35.0	10.8	3.61	2.82	+22
212	7.3	31.8	0	60.2	573	3,464	3.5	31.0	6.2	3.88	2.61	+33
214	10.2	53.7	0	0	576	3,013	6.2	85.9	8.7	3.40	3.06	+10
215	6.1	40.7	0	53.2	602	4,941	3.4	31.7	3.0	--	--	--
218	9.3	61.6	0	26.7	600	2,840	3.8	52.2	11.3	3.25	2.26	+30
221	9.0	76.7	0	12.8	600	2,444	4.6	68.1	11.7	2.88	2.39	+15
227	10.2	28.0	61.5	0	600	3,006	5.7	19.8	51.0	2.18	1.70	+22
228	9.8	17.2	71.0	0	600	3,824	6.7	11.2	82.0	2.12	1.92	[illegible]
231	3.6	23.0	61.2	0	288	3,862	7.2	20.1	72.0	1.54	1.56	15
233	4.9	10.2	75.8	0	301	3,850	3.2	9.2	88.1	0.72	0.52	+21
234	3.9	11.8	80.1	0	300	3,751	3.1	[illegible]	85.5	0.71	0.67	[illegible]
236	9.3	33.8	57.4	0	298	3,152	7.3	[illegible]	70.0	[illegible]	1.07	28
237	9.3	25.0	62.5	0	601	3,177	6.1	13.8	78.1	[illegible]	[illegible]	33
244	4.2	25.8	61.0	0	577	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	+11
245	4.3	31.5	58.5	0	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	+12
251	4.3	41.1	55.8	0	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	22
252	4.0	49.1	17.0	0	[illegible]	[illegible]	--	[illegible]	[illegible]	[illegible]	[illegible]	+11
253	[illegible]	41.0	15.6	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
256	10.7	27.5	[illegible]	[illegible]	301	[illegible]	7.1	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
258	10.0	[illegible]	[illegible]	[illegible]	600	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
259	10.2	40.5	[illegible]	[illegible]	602	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
260	[illegible]	20.1	71.1	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	--
261a	[illegible]	30.0	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
261b	[illegible]	40.0	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	

Table 2, Cont. METHANATION DATA AT 575°F USED FOR ANALYSIS OF RATE EQUATION

$$r = \frac{2 \times 10^{-5} p_{CO} p_{H_2}^{0.5}}{1 + 0.1 p_{H_2} + 0.05 p_{CH_4}}$$

Run No.	Feed Composition CO (mole %)	Feed Composition H_2 (mole %)	Feed Composition CH_4 (mole %)	Feed Composition H_2O (mole %)	Press., psig	Flow Rate, SCF/hr	Product Composition CO (mole %)	Product Composition H_2 (mole %)	Product Composition CH_4 (mole %)	Rate, exp.	Rate, calc.	% Deviation
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]	[illegible]

[illegible]

This equation fits all the data in Table 2 with a standard deviation of 11% and an average deviation of 20%. (The results are presented in Table 2 and in Figure 1.) Data at low conversions and near-equilibrium conditions are being obtained to test this rate equation.

Sulfur Resistance Studies

Laboratory equipment is being set up to measure the resistance of the methanation catalyst to various sulfur compounds to provide information for operation of the pilot plant. In addition, the two process gas chromatographs which will be used in the pilot plant are being installed in the same laboratory. A schematic diagram of this equipment is shown in Figure 2.

A gas mixture having the same composition as that expected in the pilot plant methanator feed will be doped with various sulfur compounds in the low-ppm concentration range. This gas will be fed to a small fixed-bed reactor containing 1/8-inch pellets of nickel-on-kieselguhr catalyst.

The operating conditions and gas loading will duplicate those expected in the pilot plant. The two chromatographs will be tested for general performance as well as used for analysis of feed and product gases. In addition, the interfacing of these analyzers with the data-logging computer will be worked out under conditions approximating actual operation. This should help improve data handling during initial plant operation and point out potential problem areas.

ENGINEERING ECONOMICS STUDIES

Substitution of Air for Cooling Water in Pipeline Gas Plants

We are studying the use of air cooling in the pipeline gas-from-lignite plant design. In the original design water cooling is used requiring 229,220 gpm within a 85°-115°F cooling range, of which 88,400 gpm was for turbine steam condensation. The plant cooling water makeup amounts to 8022 gpm or 11,552,000 gal/day. An additional 3640 gpm is required for the quench tower, bringing the total cooling water makeup to 16,793,000 gal/day.

A series of specifications for the application of air coolers to the above design was prepared and submitted to Hudson Engineering Corporation, suppliers of air cooling equipment, for its estimate of capital and operating costs. Heat transfer curves and condensate formation versus temperature

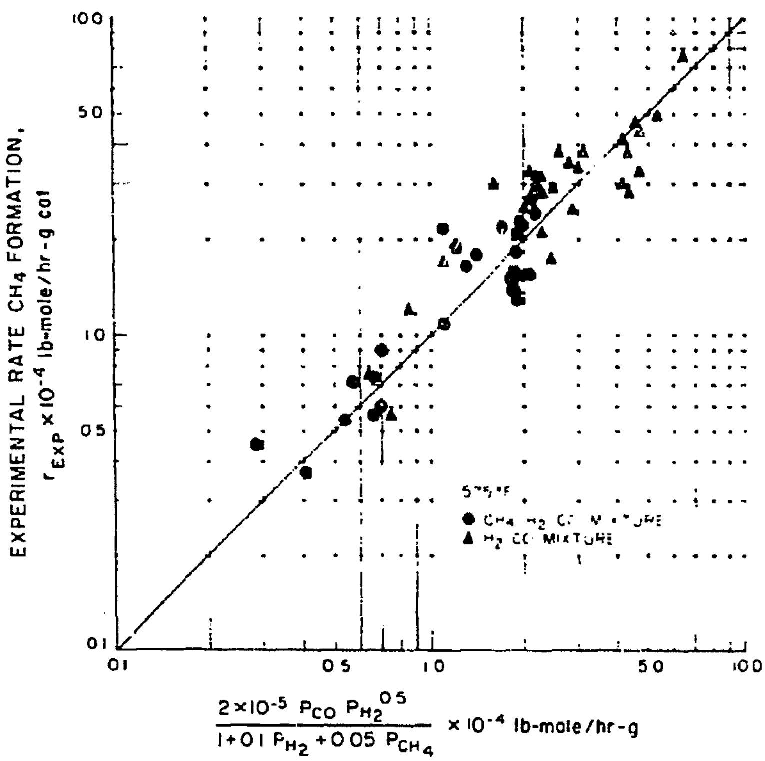

Figure 1. ANALYSIS OF METHANATION DATA

plots were prepared. Other pertinent design data, such as composition and density of vapor phases, were also submitted.

Most of the requirements are for cooling down to 100°F. Usually air cooling down to this temperature is not economical; final cooling is done with water. The breakpoint between air and water cooling depends on costs for the two types. However, we expect air cooling at least down to 140°F to be feasible.

On the basis of cooling to 140°F with air, the amount of cooling water and resultant makeup is greatly reduced. Process cooling water is reduced from 140,820 to 16,960 gpm, resulting in a saving of 4335 gpm makeup water. If, instead of cooling hydrogasifier effluent from 425° to 100°F by quench water, air cooling is used to first cool to 140°F and then the final cooling is done by direct contact, an additional 3640 gpm makeup water is saved. These figures assume that the 88,400 gpm cooling water is still required for turbine steam condensation.

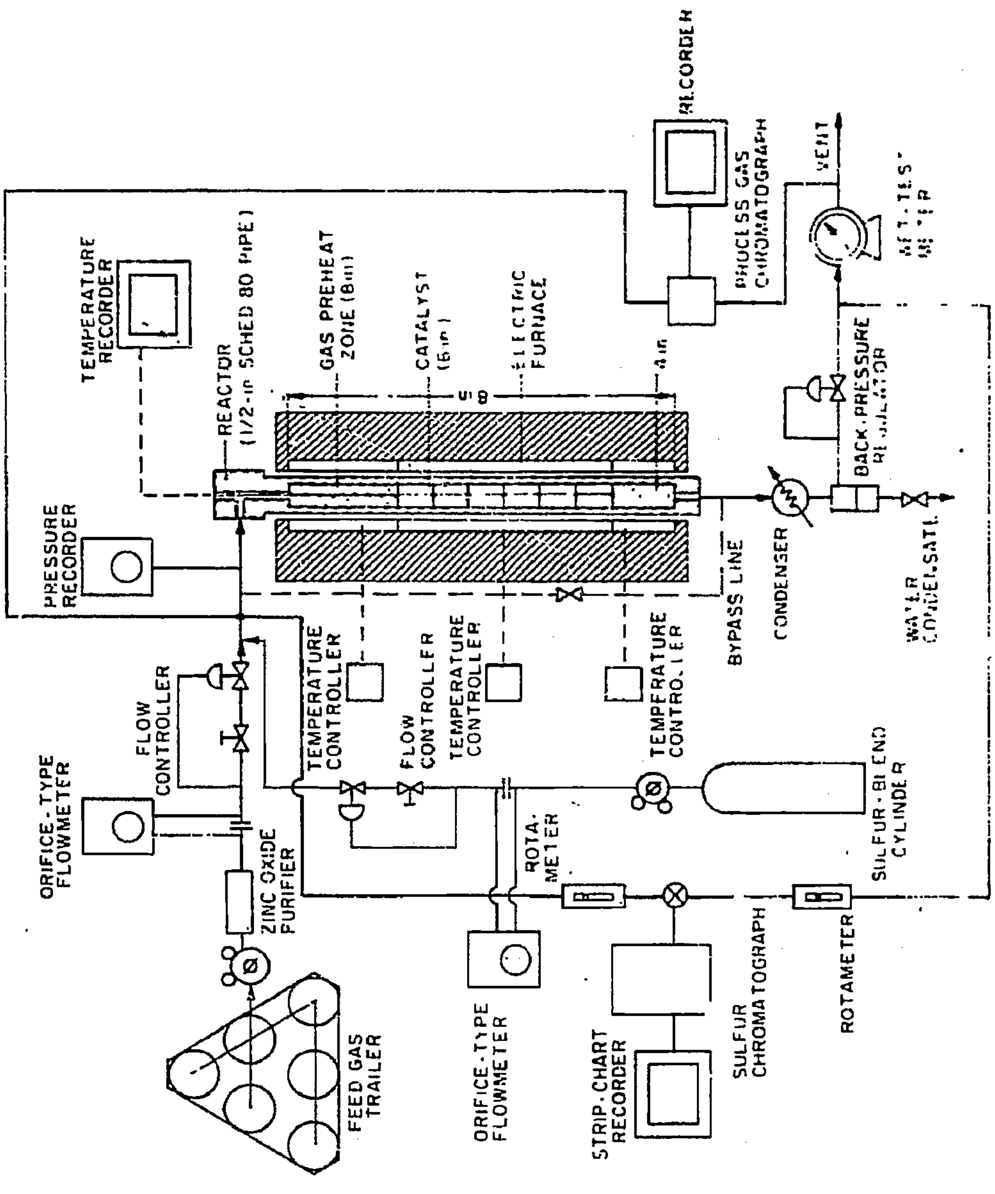

Figure 2. UNIT TO STUDY METHANATION CATALYST SULFUR RESISTANCE

In addition to the above savings in makeup water, with air cooling it may be feasible to recover water from the lignite drying operation. This is not practical with water cooling because water recovered from the lignite dryer would be counterbalanced by equivalent evaporation losses in the plant cooling tower. The preliminary figures are summarized in Table 3.

Table I. WATER MAKEUP REQUIREMENTS IN 500 BILLION Btu PIPELINE GAS-FROM-LIGNITE PLANT BY WATER AND AIR COOLING

	Water Cooling	Air Cooling to 140°F; Water Cooling to 100°F (including steam turbine condensers)
Heat Exchangers, gpm	8023	3688
Quench Tower, gpm	3640	--
Process Water, gpm	4253	4253
Total	15,916	7941
Recovered From Lignite Dryer, gpm	--	2075
Net Makeup, gpm	15,916	5866
Plant Makeup, gal/day	22,919,000	8,447,000

A reduction of 63% in makeup-water requirements appears possible by air cooling. We expect to obtain costs for the air coolers and engineering confirmation of this potential from contacts with the air-cooler manufacturers.

Effect of Financial Factors on the Return on Equity

The November 1969 Project Status Report presented results of the effects of varying financial factors on the price of gas by the accounting procedure we have traditionally used. In the calculation procedure, the return on equity is not an input, but is dependent on the factors selected for input. Certainly the rate of return on the rate base should be selected to give reasonable rates of return to both debt and equity capital. But in the mechanics of computing the gas price based on assumed values of debt-to-equity ratio, return on rate base, interest, and income, equity return does not appear in the revenue requirement calculation.

In the memorandum by J. F. Kavanagh outlining the procedure, he presented a computation of return on equity. This computation shows that the cash flow, depreciation plus net income, exceeds the sum of debt retirement plus net income (dividends) by a fixed amount. Surplus cash flow is used to liquidate the outstanding equity, just as the debt is retired in equal annual installments over a 20-year period.

Table 4 summarizes the pertinent financial data for the pipeline gas plant using electrothermally generated synthesis gas. The annual surplus equals depreciation minus 5% of the initial debt.

Table 4. CALCULATION OF PERCENT RETURN ON EQUITY

Installed Equipment or Bare Cost, $			76,920,000
Total Fixed Investment, $			87,009,000
Total Capital Investment, $			93,208,000
Debt/Equity Ratio			0.65
Initial Debt, $			60,585,000
Initial Equity, $			32,623,000
Return on Rate Base, %			7
Interest on Outstanding Debt, %			5
Annual Depreciation, $			4,350,000
Annual Debt Retirement, $			3,029,000
Annual Surplus, $			1,321,000

	End of 1st yr	End of 20th yr	20-yr Avg
Equity, Liquidation, $	31,302,000	6,203,000	18,752,000
Net Income, $	3,191,000	282,000	1,737,000

20-yr Avg Return on Outstanding Equity $\frac{\$1,737,000}{\$18,752,000} \times 100 = 9.3\%$

20-yr Avg Return, Income Plus Surplus, on Original Equity

$$\frac{\$1,737,000 + \$1,321,000}{\$31,623,000} \times 100 = 9.4\%$$

In these calculations, equity for a given year is the end-of-year amount just as the end-of-year undepreciated investment is used in the rate base. Equity is not completely liquidated; average return is 9.3%. Another return can be calculated, the net income plus the annual surplus as a percentage of original equity, amounting to 9.4% for this case.

Using the computerized A.G.A. accounting procedure we calculated the effect of varying financial factors on the return on equity. We used the investment and operating costs for the plant in which synthesis gas is generated by electrothermal gasification of spent char. To show sensitivity to investment, the installed equipment cost was doubled from $76,900,000 to $153,800,000. For these equipment costs the ranges of financial factors are —

Debt Fraction	0.6 to 0.8
Interest Rate	5, 6, 7, 8, and 9%
Return on Rate Base	7, 8, and 9%

Percentage returns on both average outstanding equity and on original equity were calculated. Results are shown in Figures 3 to 5, which present plots of percent return on equity versus percent interest on debt at a constant debt fraction, with families of curves for a fixed return on the rate base.

As shown in Figures 3 and 4, percentage return on average equity is very insensitive to the level of investment. The greatest sensitivity occurs at the higher debt ratio. For return on initial equity the effect is so slight that we have not included a second plot.

All the families of curves (constant return on rate base) show a similar pattern. As the interest rate increases, return on equity drops. As the debt ratio increases from 0.6 to 0.8, the sensitivity of equity return to interest charges more than doubles. The sensitivities of equity return to changes in interest rate and return on rate base are summarized as follows:

Change in Equity Return for 1% Change in Interest Rate

Debt Ratio	% Return on 20-yr Avg Equity ($76.9 to $153.8 X 10^6 Installed Cost)	% Return on Original Equity
0.6	1.4 to 1.5	0.8
0.7	2.1 to 2.25	1.25
0.8	3.2 to 3.6	2.10

Change in Equity Return for 1% Change in Return on Rate Base

Debt Ratio	% Return on 20-yr Avg Equity ($76.9 to $153.8 X 10^6 Installed Cost)	% Return on Original Equity
0.6	2.25 to 2.35	1.25
0.7	2.85 to 3.0	1.75
0.8	3.9 to 4.2	2.55

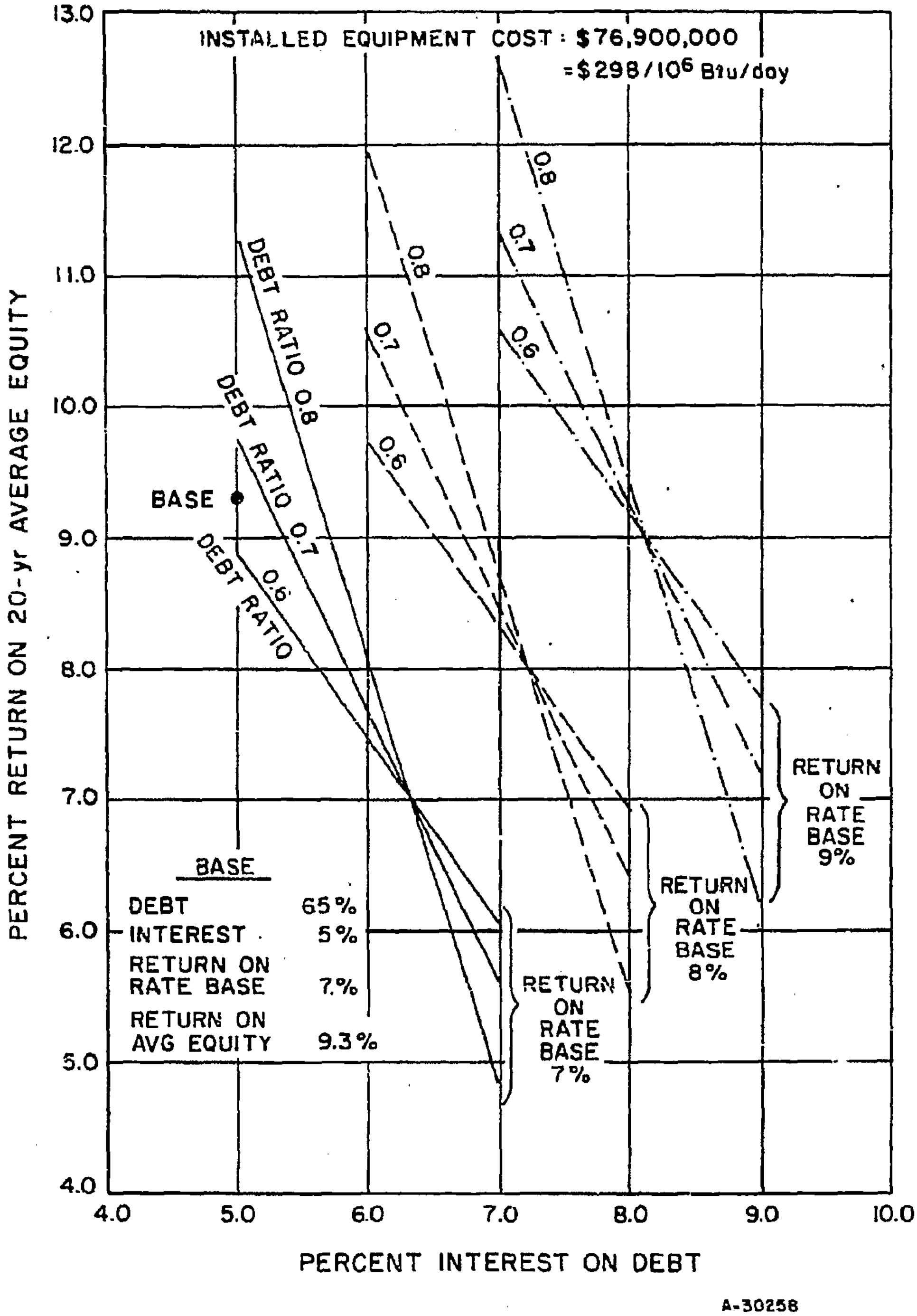

Figure 3. EFFECT OF VARYING FINANCIAL FACTORS ON THE RETURN ON AVERAGE OUTSTANDING EQUITY (9.3%)

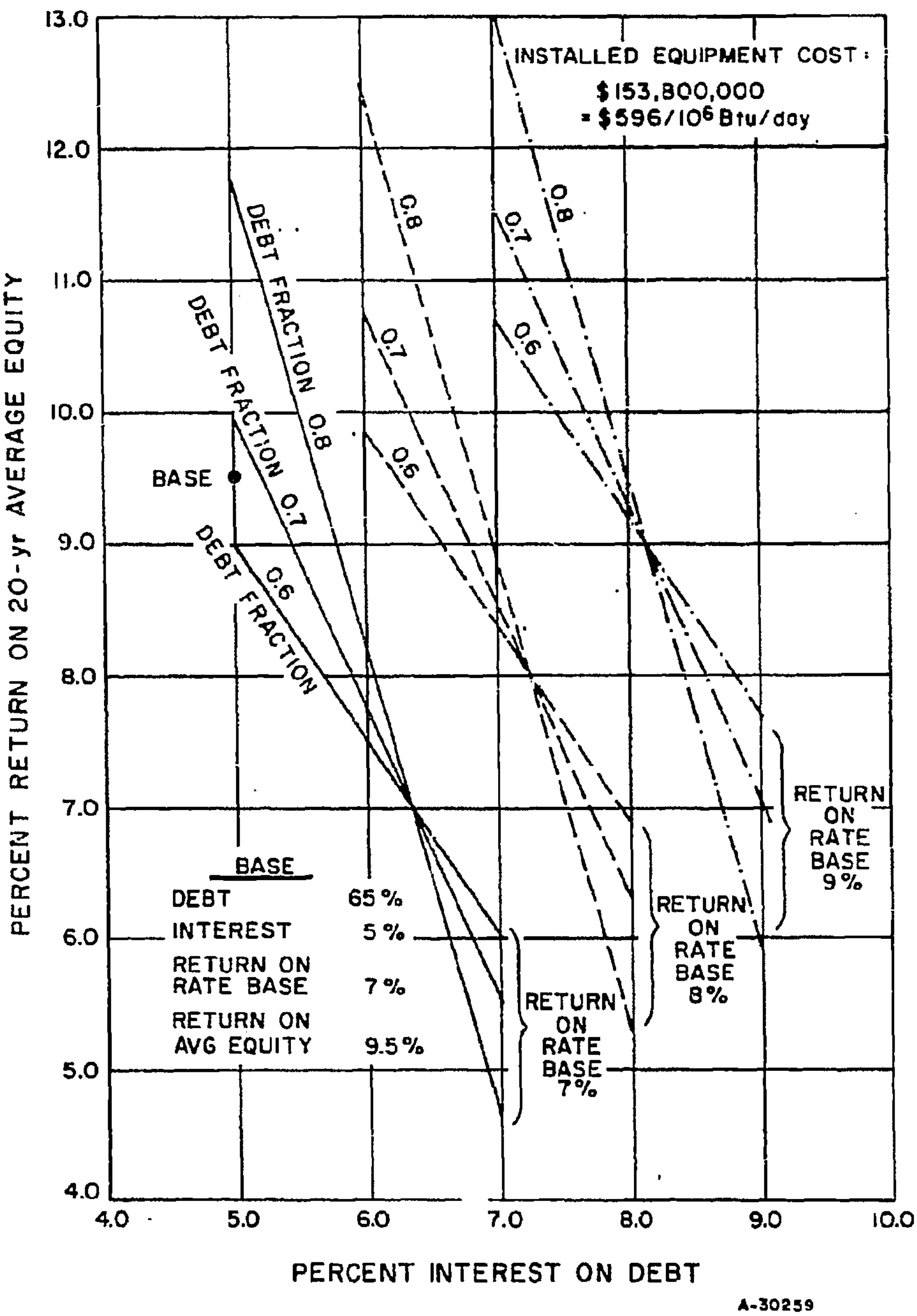

Figure.4. EFFECT OF VARYING FINANCIAL FACTORS ON THE RETURN ON AVERAGE OUTSTANDING EQUITY (9.5%)

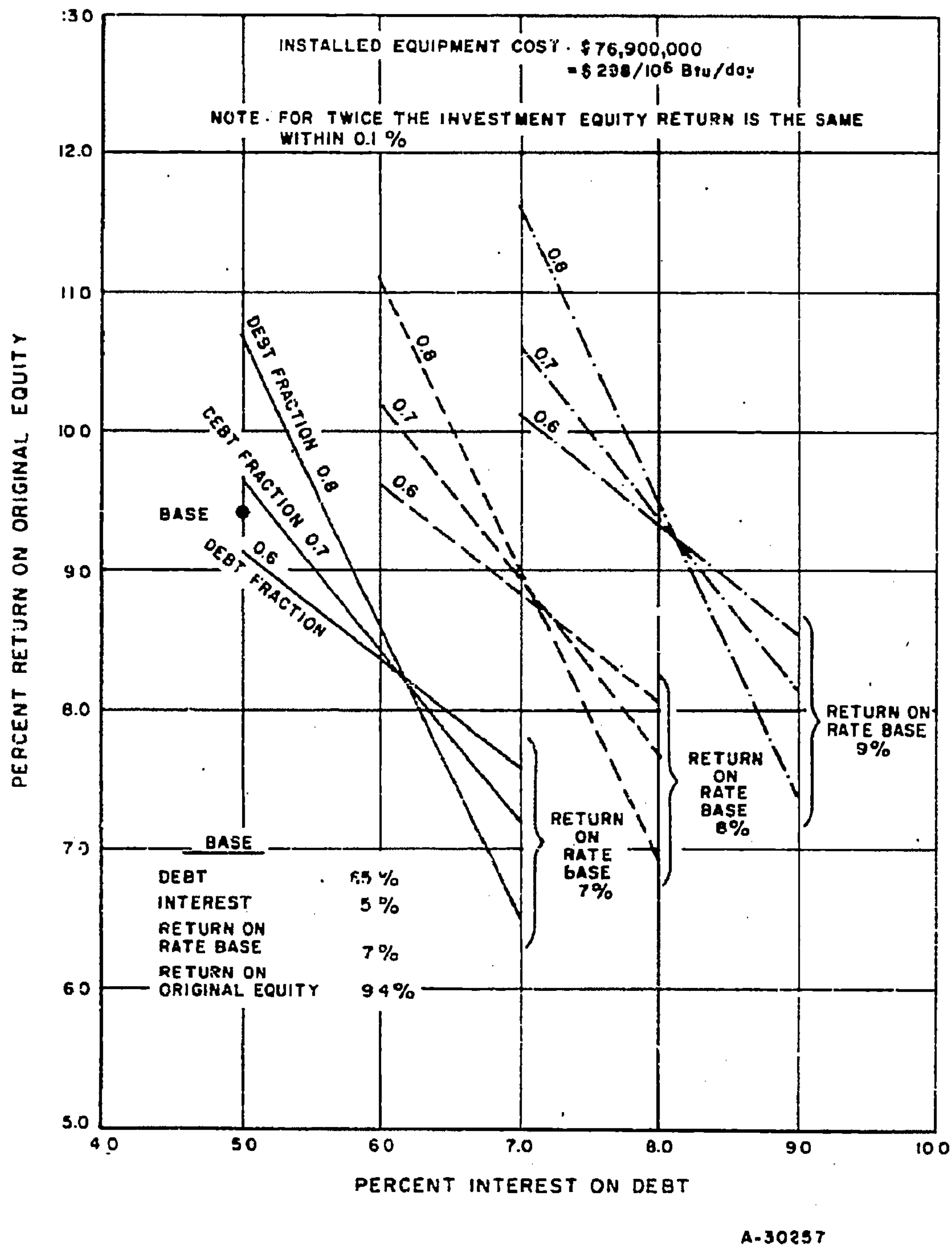

Figure 5. EFFECT OF VARYING FINANCIAL FACTORS ON THE RETURN ON ORIGINAL EQUITY

Changes in return on the rate base have a larger effect on equity return than changes in interest, but the effects are of the same general level. Changes are greater for the percent return on 20-year average equity. The greater effects at higher debt fractions illustrate the "leverage" effect of high debt utility financing. For any given debt ratio the lines are all parallel.

Note that for any given return on rate base there is a particular interest rate where the lines intersect where there is only one equity return regardless of debt fraction. This point is independent of the investment level.

DEVELOPMENT UNIT STUDIES

Hydrogasification Tests

We are presenting the hydrogasification results of Run HT-241 conducted in December 1969 in the high-temperature balanced-pressure development unit with untreated North Dakota lignite from the Glenharold mine. The objective of this test was to study the hydrogasification reactivity of the lignite with a synthesis gas and steam at a system pressure of 500 psig. For this test the lignite was crushed and screened to a −10+80 mesh size and air-dried in the fluidized-bed coal pretreatment unit at 220°-240°F.

Operating conditions and results of this run are presented in Table 5. Compositions and screen analyses of the feed and residue are given in Table 6. Liquid products and composition are shown in Table 7.

Hydrogasification of the lignite in a 3.5-ft fluid bed with synthesis gas and steam at an average lignite bed temperature of 1590°F resulted in the gasification of 51.6% of the moisture-, ash-free lignite and 36.0% of the carbon in the lignite. These conversions were obtained with a synthesis-gas rate giving a hydrogen-to-lignite ratio of 16.9% of the stoichiometric ratio, and a steam concentration of 49.8% in the feed gas (Table 5). Additional lignite was converted to liquid products containing oil that included 2.72% of the carbon in the lignite. Based on an ash balance, 0.531 lb of lignite residue was recovered for every pound of lignite fed.

For comparative purposes, key results of Run HT-241 are presented in Table 8 with those of Run HT-239 (January 1970 Project Status Report), conducted at 500-psig system pressure with the same North Dakota lignite, but with a hydrogen-steam feed gas. With synthesis gas about 5% less carbon was gasified than with hydrogen, although the moisture-, ash-free

Table 5, Part 1. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF NORTH DAKOTA LIGNITE IN HIGH-TEMPERATURE ADIABATIC REACTOR

Coal	North Dakota Lignite
Source	Glenharold Mine
Sieve Size, USS	−10+80
Run No.	HT-241
Duration of Test, hr	5
Steady-State Operating Period, min[a]	139-304
OPERATING CONDITIONS	
Bed Height, ft	3.5
Reactor Pressure, psig	498
Reactor Temperature, °F[b] Inches From Bottom	
62-1/2	1465
67-3/4	1510
73	1640
78-1/4	1590
83-1/2	1655
89	1660
94-1/2	1510
100	1665
104	1595
Average	1590
Lignite Rate, lb/hr[c]	23.92
Feed Gas Rate, SCF/hr	276.7
Steam Rate, lb/hr	13.08
Steam, mole % of hydrogen-steam mixture	49.8
Hydrogen/Lignite Ratio, % of stoichiometric[d]	16.9
Hydrogen/Steam Ratio, mole/mole	0.529
Bed-Pressure Differential, in. wc	--
Lignite Space Velocity, lb/cu ft-hr	77.3
Feed-Gas Residence Time, min[e]	0.297
Superficial Feed-Gas Velocity, ft/s[f]	0.196

Table 5, Part 2. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF NORTH DAKOTA LIGNITE IN HIGH-TEMPERATURE ADIABATIC REACTOR

	HT-241
OPERATING RESULTS	
Product Gas Rate, SCF/hr	786.5
Net Btu Recovery, 10^3 Btu/lb	4.775
Product-Gas Yield, SCF/lb	32.88
Hydrocarbon Yield, SCF/lb	3.62
Carbon Oxides Yield, SCF/lb	3.59
Net Reacted Hydrogen, SCF/lb	nil
Residue, lb/lb coal[g]	0.531
Liquid Products, lb/lb coal[h]	0.393
Net MAF Lignite Hydrogasified, wt %[i]	51.6
Carbon Gasified, wt %	36.0
Steam Decomposed, lb/hr[j]	4.22
Steam Decomposed, % of steam fed	32.3
Steam Decomposed, % to total equivalent fed[k]	86.0
Overall Material Balance, %	99.5
Carbon Balance, %	100.1
Hydrogen Balance, %	97.0
Oxygen Balance, %	98.6

PRODUCT GAS PROPERTIES

Gas Composition, mole %	Feed	Product
Nitrogen	--	36.2
Carbon Monoxide	43.0	14.2
Carbon Dioxide	4.5	13.6
Hydrogen	52.5	24.8
Methane	--	10.2
Ethane	--	0.5
Propane	--	0.3
Butane	--	--
Benzene	--	0.2
Hydrogen Sulfide	--	--
Total	100.0	100.0
Heating Value, Btu/SCF[m]	304	249
Specific Gravity (Air = 1.00)	0.521	0.786
Nitrogen Purge Rate, SCF/hr	285	

Table 5, Part 3. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF NORTH DAKOTA LIGNITE IN HIGH-TEMPERATURE ADIABATIC REACTOR

a. From start of lignite feed.

b. Tube wall temperatures. Bottom of coal bed at 62 in.

c. Operating conditions and results based on weight of dry feed.

d. Percent of the stoichiometric hydrogen/char ratio — the net feed hydrogen/char ratio required to convert all the carbon to methane.

e. Lignite bed volume/(CF/min feed gas at reactor pressure and temperature).

f. (CF/s feed gas at reactor pressure and temperature)/cross-sectional area of reactor.

g. By ash balance.

h. Includes condensed, undecomposed steam.

i. 100 (wt of product gas-wt feed gas in-wt decomposed steam-wt nitrogen in/wt of moisture-, ash-free lignite).

j. Computed as difference between steam feed rate and the measured liquid water rate leaving the reactor.

k. Computed as difference between the total equivalent steam feed rate (includes moisture content of feed char and bound water corresponding to oxygen content of feed char) and the measured liquid water rate leaving the reactor.

m. Gross, gas saturated at 60°F, 30-in. Hg pressure. SCF: dry gas volume in SCF at 60°F, 30-in. Hg pressure.

Table 6. CHEMICAL AND SCREEN ANALYSES OF LIGNITE FEED AND RESIDUE

Run No.	HT-241	
Sample	Feed	Residue
Proximate Analysis, wt %		
Moisture	2.6	0.9
Volatile Matter	34.1	5.[illegible]
Fixed Carbon	55.0	77.5
Ash	8.3	15.9
Total	100.0	100.0
Ultimate Analysis (dry), wt %		
Carbon	68.9	79.6
Hydrogen	3.94	0.95
Nitrogen	1.19	0.57
Oxygen	16.71	2.48
Sulfur	0.72	0.24
Ash	8.54	16.08
Total	100.00	100.00
Screen Analysis, USS, wt %		
+20	36.0	24.4
+30	20.6	22.6
+40	16.1	21.5
+60	16.0	20.0
+80	7.3	6.7
+100	2.3	1.9
+200	1.5	2.5
+325	0.1	0.2
–325	0.1	0.2
Total	100.0	100.0

Table 7. COMPOSITION OF HYDROGASIFICATION LIQUID PRODUCTS

Run No.	HT-241
Sample	Condenser
Liquid Products,*	
lb/lb lignite	0.393
Composition of Liquid Products, wt %	
Water	94.29
Oil	5.71
Total	100.00
Composition of Oil Fraction, wt %	
Carbon	83.70
Hydrogen	7.19
Total	90.89
Carbon in Oil Fraction,	
lb/lb lignite	0.01876
wt % of carbon in lignite	2.72

* Includes condensed, undecomposed steam.

gasification was less by only 2.6%. Partially responsible for the lower level of lignite gasification in Run HT-241 is the lower hydrogen plus carbon monoxide-to-lignite ratio, 30.7% compared to 34.2% in Run HT-239. The lower carbon gasification of Run HT-241 is also reflected in the significantly smaller hydrocarbon yield, compared to that of Run HT-239. The carbon oxides yields in both tests were nearly the same at 3.59-3.68 SCF/lb of lignite. Product-gas compositions of the two runs show large differences. With synthesis gas as the feed gas, the methane concentration was 16.0%, compared to 24.9% with hydrogen as the feed gas, while the carbon oxides concentration at 43.6% was nearly twice that of the test with hydrogen.

Run FP-141 coal pretreatment operations, conducted on a one shift per day basis, were concluded in the pilot plant fluid-bed coal pretreatment unit. Over 2000 lb of lightly pretreated Pittsburgh No. 8 seam high-volatile-content bituminous coal was produced. The coal was pretreated with air and nitrogen

Table 8. COMPARISON OF NORTH DAKOTA LIGNITE HYDROGASIFICATION RESULTS AT 500 psig WITH SYNTHESIS GAS-STEAM AND WITH HYDROGEN-STEAM FEED GASES

Run No.	HT-239	HT-241
Feed Gas	Hydrogen + Steam	Synthesis Gas + Steam
Reactor Pressure, psig	478	498
Lignite Bed Temp Average, °F	1535	1590
Lignite Bed Height, ft	3.5	3.5
Lignite Feed Rate, lb/hr	32.07	23.92
Feed-Gas Rate, SCF/hr	368.8	276.7
Steam Feed Rate, lb/hr	8.69	15.08
Steam/Lignite Ratio, lb/lb	0.271	0.547
Hydrogen, or Hydrogen + Carbon Monoxide/ Lignite Ratio, % of stoichiometric	34.2	30.7
Steam Concentration in Feed Gas, mole/hr	33.1	49.8
Steam Decomposed, % of total equivalent fed	56.9	86.0
Carbon Gasified, %	41.4	36.0
MAF Lignite Gasified, %	54.2	51.6
Hydrocarbon Yield, SCF/lb	4.33	3.62
$CO + CO_2$ Yield, SCF/lb	3.68	3.59
Carbon in Oil Fraction, % of carbon in lignite	4.75	2.72
Product-Gas Composition (nitrogen-free), mole %		
Carbon Monoxide	12.5	22.3
Carbon Dioxide	10.1	21.3
Hydrogen	50.2	38.8
Methane	24.9	16.0
Ethane	1.2	0.8
Propane	0.5	0.5
Benzene	0.5	0.3
Hydrogen Sulfide	0.1	--
Total	100.0	100.0
Product-Gas Heating Value (nitrogen-free), Btu/SCF	501	390

in a fluidized bed at 750°-800°F. Laboratory tests showed that the coal was still lightly agglomerating; therefore, it was treated a second time in another series of operations (Run FP-142) at the same conditions as in Run FP-141. On a one shift per day basis, Run FP-142 extended over a period of 6 days.

ELECTROTHERMAL GASIFICATION

Two tests were conducted in the electrothermal gasifier and some minor modifications were made on the unit during the month. The tests were conducted with the concentric electrode configuration at 1000 psig using a hydrogasified high-volatile bituminous char. Both tests had to be terminated due to operational difficulties before steady-state operating conditions were reached.

Run EG-43 had a normal heat-up except for some small pressure upsets which made it difficult to control bed height. Once steam was introduced into the reactor the pressure upsets become worse, and the source of the difficulty was traced to the erratic flow of the quench-water exit stream. The test was terminated, and a pressure equalizing line was installed between the scrubber and the quench-water collecting pot. A check of the system under pressure revealed little improvement. To ensure proper control of the quench-water exit stream, the water drainline from the scrubber to the quench-water collecting pot was increased in inside diameter from 1.0 to 1.5 inches. To prevent further accumulation of fines in the liquid sight gage, the gage was tapped directly into the quench-water collecting pot. (See Figure 6.)

The heat-up period of Run EG-44 was one of the smoothest to date, although the feed screw jammed several times in the beginning due to moisture in the coal feed. The modifications on the quench-water exit eliminated the pressure upsets that occurred during draining. After the introduction of steam the feed screw jammed once again, and during a manual attempt to free it, the shear pin in the driveshaft broke, causing the test to be terminated. To rectify this recurring problem, strip heaters are being placed on the feed hopper and covered with sufficient insulation to prevent steam condensation during small pressure upsets. In addition, a continuous nitrogen purge will prevent steam from seeping into the feed hopper during normal operation of the unit. Inspection of the reactor after both tests revealed no damage to the walls or the electrode.

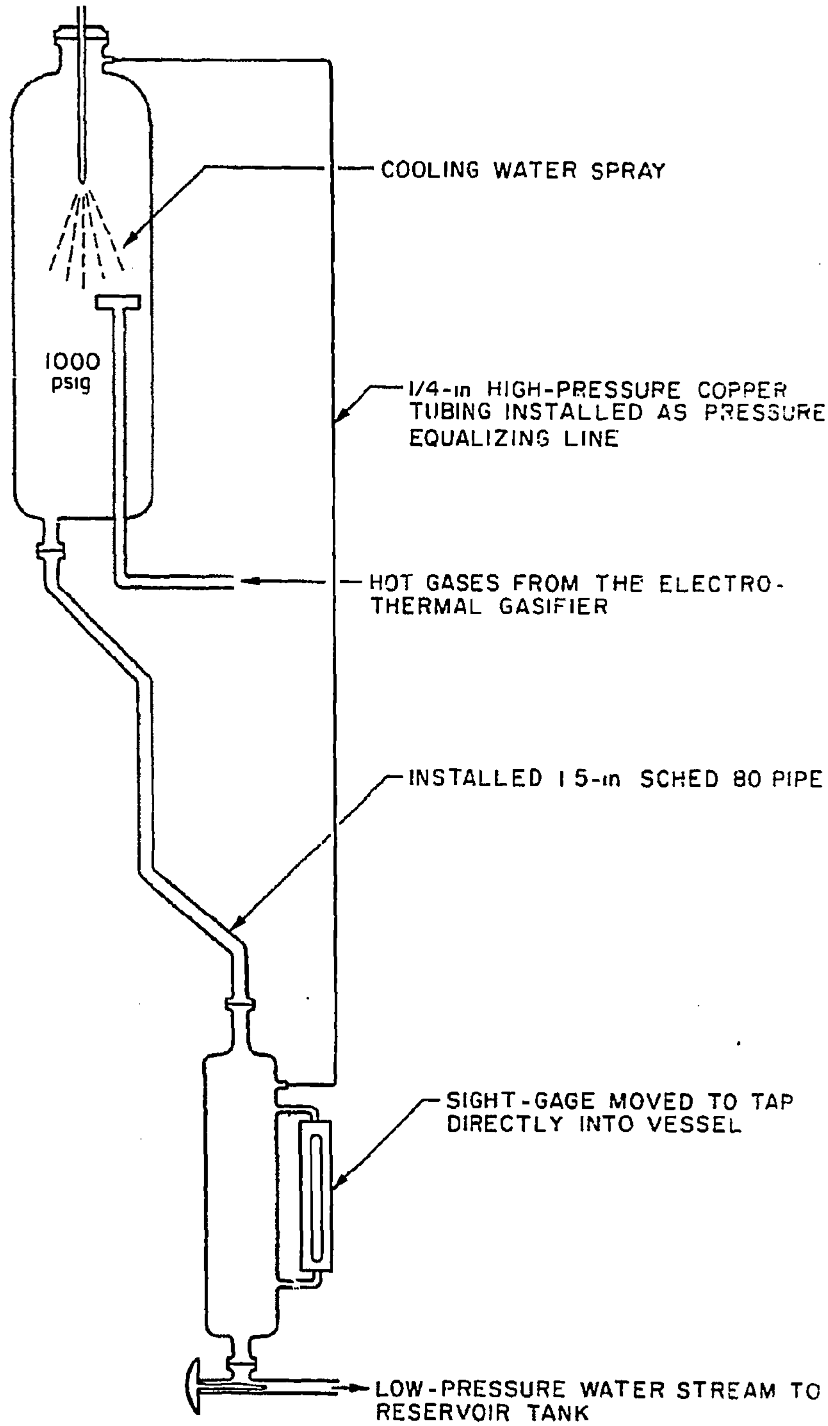

Figure 6. MODIFICATIONS IN THE QUENCH-WATER EXIT SYSTEM

An electrode of silicon carbide was bought and modified to our power source in preparation for future tests. A silicon carbide tube, to be used as an outer electrode, has been ordered. Tests will be made to determine the suitability of silicon carbide for electrothermal gasification.

FUEL CELL ENGINEERING STUDY

Design and cost estimation for ductwork and electrical connections is completed. Total capital investment is estimated for cell ratings of 150 and 300 W/sq ft. Table 9 shows the details of the estimate. Cost of fuel cell components constitute about 40-50% of total capital investment. Operating cost and revenue requirement per kilowatt-hour of energy are shown in Table 10. The price of anode gas was assumed to be 25¢/million Btu. Depreciation was calculated on the basis of the following useful life of plant components:

Component	Useful Life, yr
Electrolyte	3
Electrodes	3
Distributor Plate	10
Spacer	10
Casing Components	20
All Other Items	20

Waste gases from the fuel cell, available at 1385°F, can be used to generate steam for the HYGAS plant. Waste-heat credit for energy above 300°F is subtracted from operating cost to arrive at net operating cost. Total revenue requirements amount to 4-5 mills/kWhr for a cell performance of 300 W/sq ft.

PILOT PLANT CONSTRUCTION

Engineering

A major portion of the engineering effort during this report period has been instrumentation and related details. Electrical detail drawings are complete; instrument drawing is 65% completed. All piping plans and details have been issued. Total project detailed design and drafting is 95% complete.

Procurement

The reactor arrived on February 6th. All major equipment and materials scheduled to be received have arrived, with the following exceptions: A portion of the switchgear and a compressor motor were delayed due to the General Electric Company strike; the pretreater reactor, several small vessels, and

Table 9. COST ESTIMATE FOR 400-MW COAL-FUELED MOLTEN CARBONATE FUEL CELL POWER PLANT

Item	Estimate A (150 WSF-F_1)*		Estimate B (150 WSF-F_2)†		Estimate C (300 WSF-F_1)		Estimate D (300 WSF-F_2)	
	$/kw	%	$/kw	%	$/kw	%	$/kw	%
Purchased Equipment Cost of Package (PEC)	81.98	100.0	56.50	100.0	51.29	100.0	34.80	100.0
Installation (15% PEC)	12.30	15.0	8.47	15.0	7.69	15.0	5.22	15.0
Subtotal	94.28	115.0	64.97	115.0	58.98	115.0	40.02	115.0
Ductwork	11.30	13.8	11.30	20.0	7.90	15.4	7.90	15.4
Electrical	4.48	5.5	4.48	7.9	4.48	8.7	4.48	12.9
Blowers	8.00	9.7	8.00	14.1	8.00	15.6	8.00	23.0
Burners	1.10	1.3	1.10	1.9	1.10	2.1	1.10	3.2
Instrumentation and Control	1.50	1.8	1.50	2.7	1.50	2.9	1.50	4.3
Land, Structures and Yard Improvements	5.00	6.1	5.00	8.9	5.00	9.7	5.00	14.4
Total Physical Cost	125.66	153.2	96.35	170.5	86.96	169.4	68.00	195.5
Engineering and Construction (7%)	8.77	10.7	6.67	11.9	6.09	11 9	4.76	13.7
Direct Plant Cost (DPC)	134.43	163.9	103.02	182.4	93.05	181.3	72.76	209.2
Contigency (10% DPC)	13.44	16.4	10.30	18.2	9.31	18.1	7.28	20.9
Interest during construction 7%/yr	14.12	17.2	10.82	19.2	9.77	19.0	7.64	22.0
Fixed Capital Investment (FCI)	161.99	197.5	124.14	219.8	112.13	218.4	87.68	252.1
Working Capital	6.50	6.9	6.50	10.0	6.50	11.0	6.50	16.2
Total Capital Investment (TCI)	168.49	204.4	130.64	229.8	118.63	229.4	94.18	268.3

* F_1 = Upper limit cost factor

† F_2 = Lower limit cost factor

Table 10. ANNUAL OPERATING COST AND REVENUE REQUIREMENTS

	Estimate C (300 w/sq ft; F_1) mills/kwh	Estimate C (300 w/sq ft; F_1) %TRR	Estimate D (300 w/sq ft; F_2) mills/kwh	Estimate D (300 w/sq ft; F_2) %TRR
Raw material (anode gas) at 25¢/MBTU	2.820	55.5	2.820	67.6
Labor and supervision cost	0.129	2.4	0.129	3.1
Maintenance (3% Direct Plant Cost)	0.279	5.5	0.218	5.2
Material Supplies (15% Maintenance)	0.042	0.8	0.033	0.8
Depreciation*	1.411	27.8	0.893	21.4
Insurance, Local Taxes (1% FCI)	0.128	2.5	0.100	2.4
Subtotal	4.809	94.5	4.193	100.5
Contigency (2% Subtotal)	.096	1.9	0.084	2.1
Operating Cost	4.905	96.4	4.277	102.6
Waste Heat credit at 20¢/MBTU	1.235	24.3	1.235	29.6
Net Operating Cost	3.670	72.1	3.042	73.0
Gross Return (7% TCI)	0.946	18.6	0.752	18.0
Federal Income Tax (50% Gross Return	0.473	9.3	0.376	9.0
Total Revenue Requirements (TRR)	5.089	100	4.170	100

*Calculated by straight line method and useful life of individual components.

a portion of the material handling equipment will arrive late due to recent design changes and additions. The above delayed deliveries can be worked into the construction schedule without extending the completion date.

We have anticipated possible delayed delivery of both carbon steel and alloy shop-fabricated pipe which would directly affect the construction schedule. Therefore, a maximum expediting effort is being made to maintain promised deliveries.

Construction

Overtime construction began on February 10th in an effort to attract the pipefitter/welder manpower required to maintain the construction scheduled. No immediate increase occurred; however, the manpower situation in this craft will be watched closely to determine if further measures are required to increase the manpower level.

Preparations for the reactor erection are almost complete. Structural steel and major equipment have been and are being erected. A total of 10 cold-weather days was accumulated to date; 5 occurred in this report period. On these days no significant outside progress was made.

Schedule

The project schedule was updated and reviewed with IGT. The computer run was made and reflects a negative float of approximately 8 days based on using July 1, 1970, as the target completion date.

The precast pipe-rack bents have been erected, except in those areas where erection of equipment requires an open pipe-rack area. First-level pipe-rack piping has begun, but is proceeding slowly due to lack of manpower. A delay in the start of process area piping is foreseen due to lack of manpower.

The electrical subcontractor has moved to the field and is presently working in the motor control center.

PROGRESS REPORT NO. 19

COAL HYDROGASIFICATION PILOT PLANT

FOR

INSTITUTE OF GAS TECHNOLOGY

CHICAGO, ILLINOIS

W-1784

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DATE:	February 15, 1970
PROCON PROJECT MANAGER:	T. A. Taylor

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W-1784

Coal Hydrogasification Pilot Plant

I. SUMMARY

Engineering	95%
Purchasing	86%
Material Receipt	72%
Construction	24%

A. ENGINEERING

A major portion of the engineering effort during this report period has been instrumentation related details. Electrical detail drawings are complete and instrument drawing completion is 65 percent. All piping plans and details have been issued. Total project detailed design and drafting is 95 percent complete.

B. PROCUREMENT

The reactor arrived on February 6th. All major equipment and materials that are scheduled to be received have arrived, with the following exceptions: a portion of the switchgear and a compressor motor which were delayed due to the General Electric Company strike; the pretreater reactor, several small vessels, and a portion of the material handling equipment which will arrive late due to recent design changes and additions. The above delayed deliveries can be worked into the construction schedule without extending the completion date.

We have anticipated possible delayed delivery of both carbon steel and alloy shop fabricated pipe which would directly affect the construction schedule. Therefore, a maximum expediting

B. PROCUREMENT (continued)

effort is being made to maintain as promised deliveries.

C. CONSTRUCTION

Overtime construction began on February 10th as an effort to attract the pipefitter/welder manpower required to maintain the construction scheduled. No immediate increase occurred, however, the manpower situation in this craft will be watched closely to determine if further measures are required to increase the manpower level.

Preparations for the reactor erection are almost complete. It is now scheduled to be erected on the 18th. Structural steel and major equipment have been and are being erected.

We have experienced a total of ten cold weather days, five of which occurred in this report period. On these days no significant outside progress was made.

II. SCHEDULE AND S-CURVE REPORT

The project schedule was updated and reviewed with IGT. The computer run was made and reflects a negative float of approximately eight days based on using July 1, 1970 as target completion date.

Bar Chart - The bar chart in this report shows the critical path construction activities with scheduled early start and finish dates. It should be noted that this bar chart does not include any cold weather time. This time could cause an extention of the completion date.

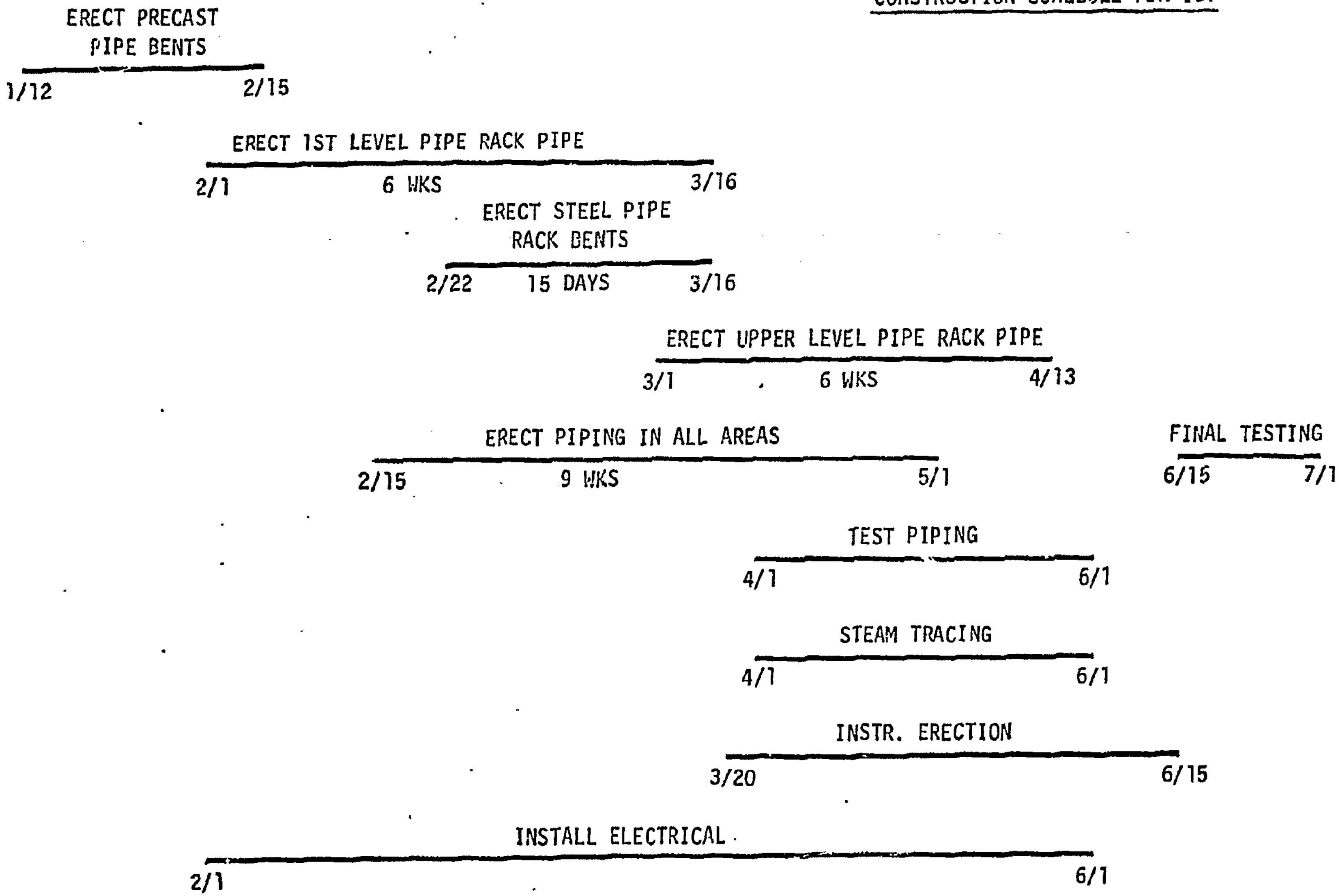
The precast pipe rack bents have been erected except in those areas where erection of equipment requires open pipe rack area.

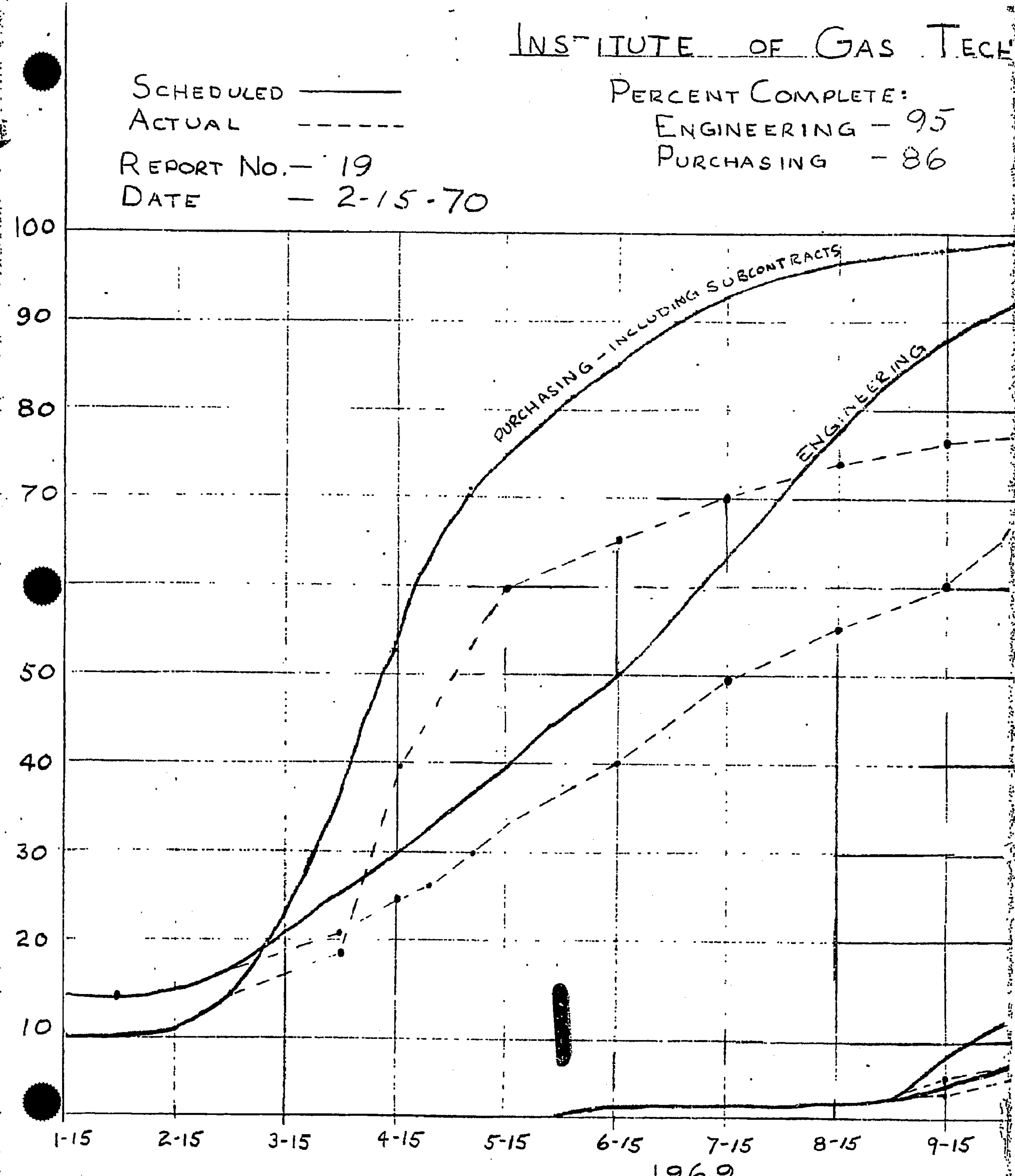
First level pipe rack piping has begun, but is proceeding slowly due to lack of manpower.

A delay in the start of process area piping is foreseen due to lack of manpower.

The electrical subcontractor has moved to the field and is presently working in the motor control center.

CONSTRUCTION SCHEDULE FOR IGT

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ACTUAL
REPORT NO. – 19
DATE – 2-15-70
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ENGINEERING – 95
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PURCHASING - INCLUDING SUBCONTRACTS
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MATERIAL RECEIVED - 72
CONSTRUCTION - 24

REV 0 - 3-15-69
REV 1 - 4-15-69
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REV. 3 - 8-15-69
REV. 4 - 1-15-70

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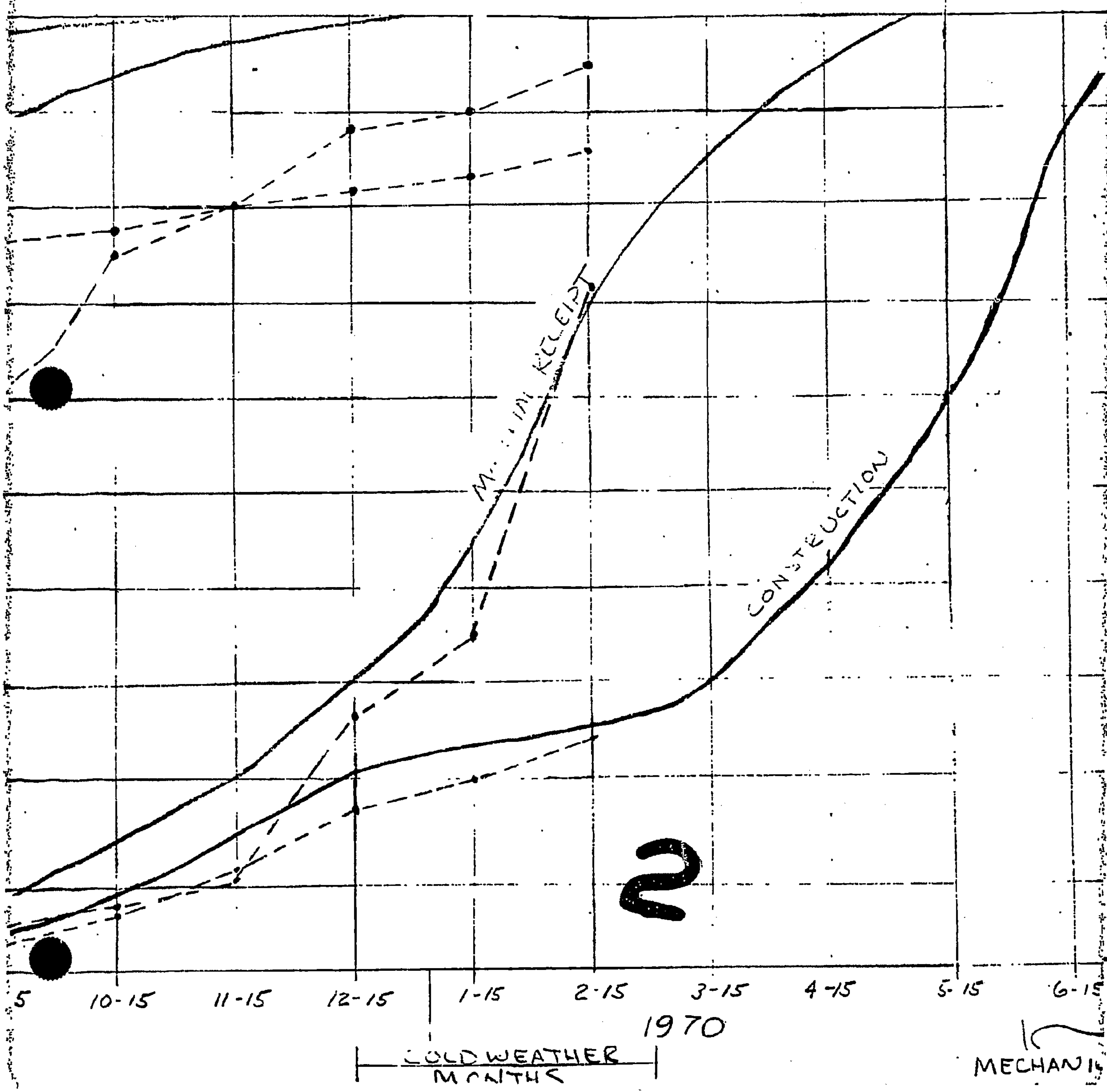

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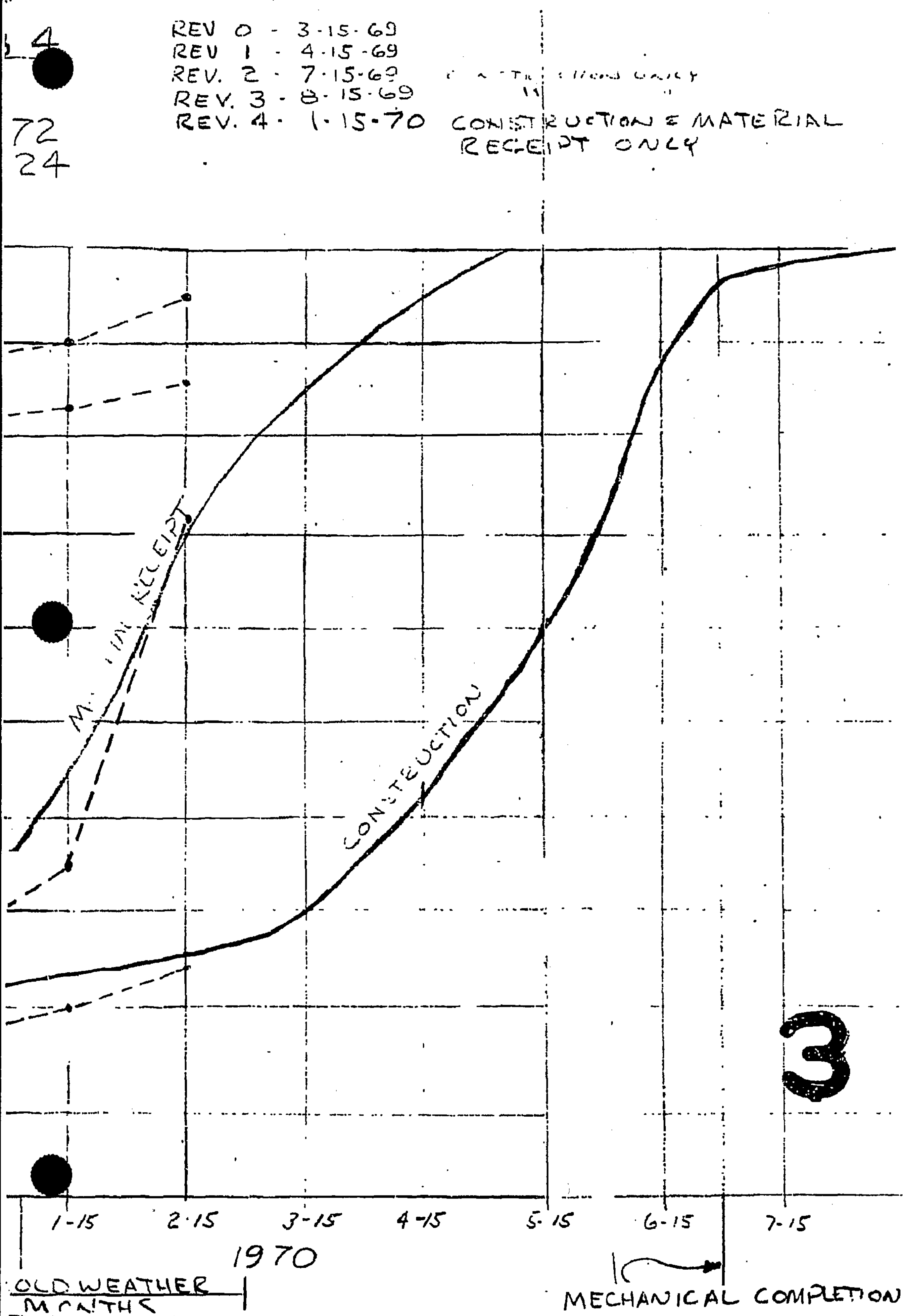


REV 0 - 3-15-69
REV 1 - 4-15-69
REV. 2 - 7-15-69
REV. 3 - 8-15-69
REV. 4 - 1-15-70 CONSTRUCTION & MATERIAL RECEIPT ONLY
72
24
RECEIPT
CONSTRUCTION
1-15
2-15
3-15
4-15
5-15
6-15
7-15
1970
OLD WEATHER MONTHS
MECHANICAL COMPLETION
3

III. CONTRACT FINANCIAL REPORT

Procon's portion of Form No. 80R0178 has been completed and reflects actual cost incurred through the last calendar month; estimated costs during this month; and the estimated total cumulative cost through this month. All costs have been rounded off to the nearest thousand dollars.

CONTRACT FINANCIAL REPORT
(Dollars in thousands)
(See instructions before preparation)

1 For Month Ended: February 15, 1970

2 No. of Work Days

3 Contract No.

Form Approved
Budget Bureau
No. 80R0178
Sheet ____ of ____

4 To:

5 From:

6 Contract Value
$

7 Contract Type

8 Funded Contract Amount
$

9 Amounts Billed
$

10 Program/Scope of Work

11 Signature and Title of Authorized Representative

12 Preparation Date

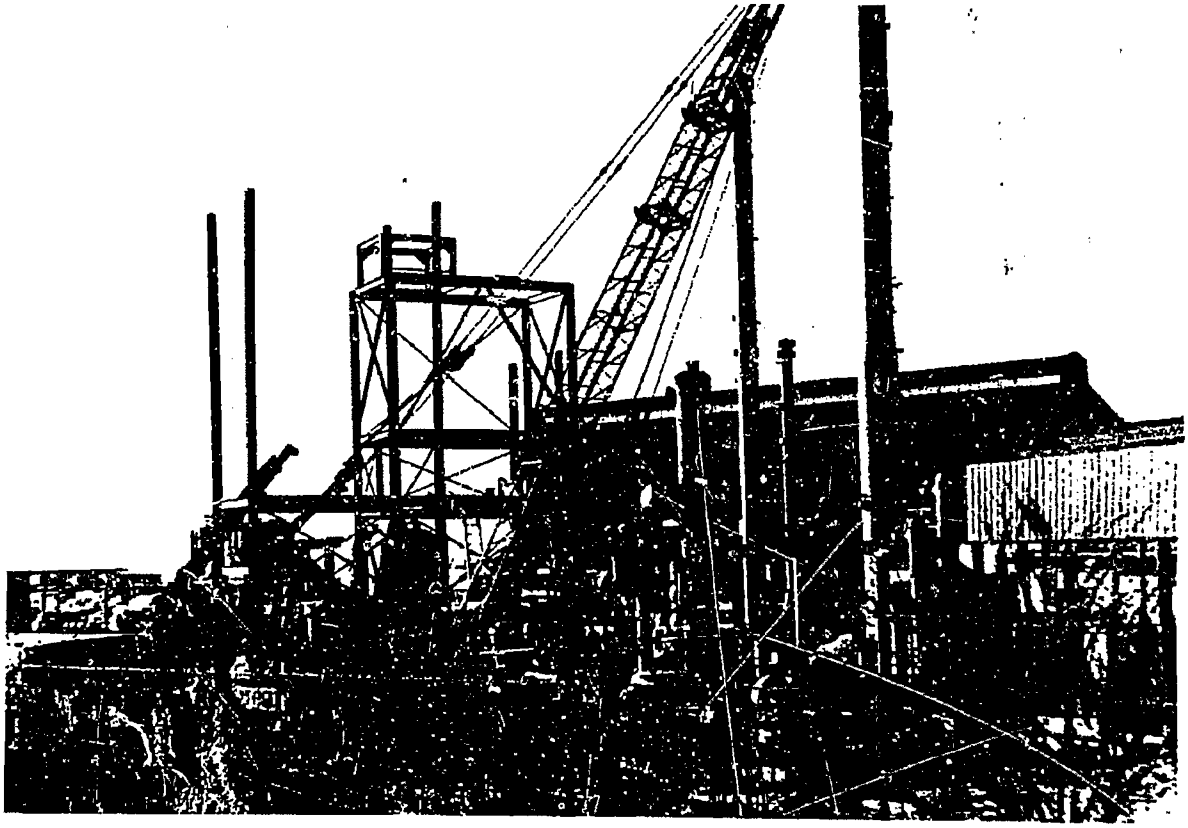
13 Payments Received
$

14 Appropriation (or Fund Citation) and/or Reporting Category	15 Cost Incurred/Contract Earnings			16 Planning Data (For Agency use only)			
	Cum. Actual End of Prior Mo. a	Actual/ Estimated Current Month b	Cumulative Actual/ Estimated To Date c	a	b	c	d
PROCON INCORPORATED W-1784	$ 3,081	$ 600	$ 3,681				
LESS: 10% RETENTION AS APPLICABLE	305	60	365				
	$2,776	$540	$3,316				
ABOVE INCLUDES 66 THOUSAND DOLLARS PREVIOUSLY REPORTED AS W-1784 -X-1							
17 Total							

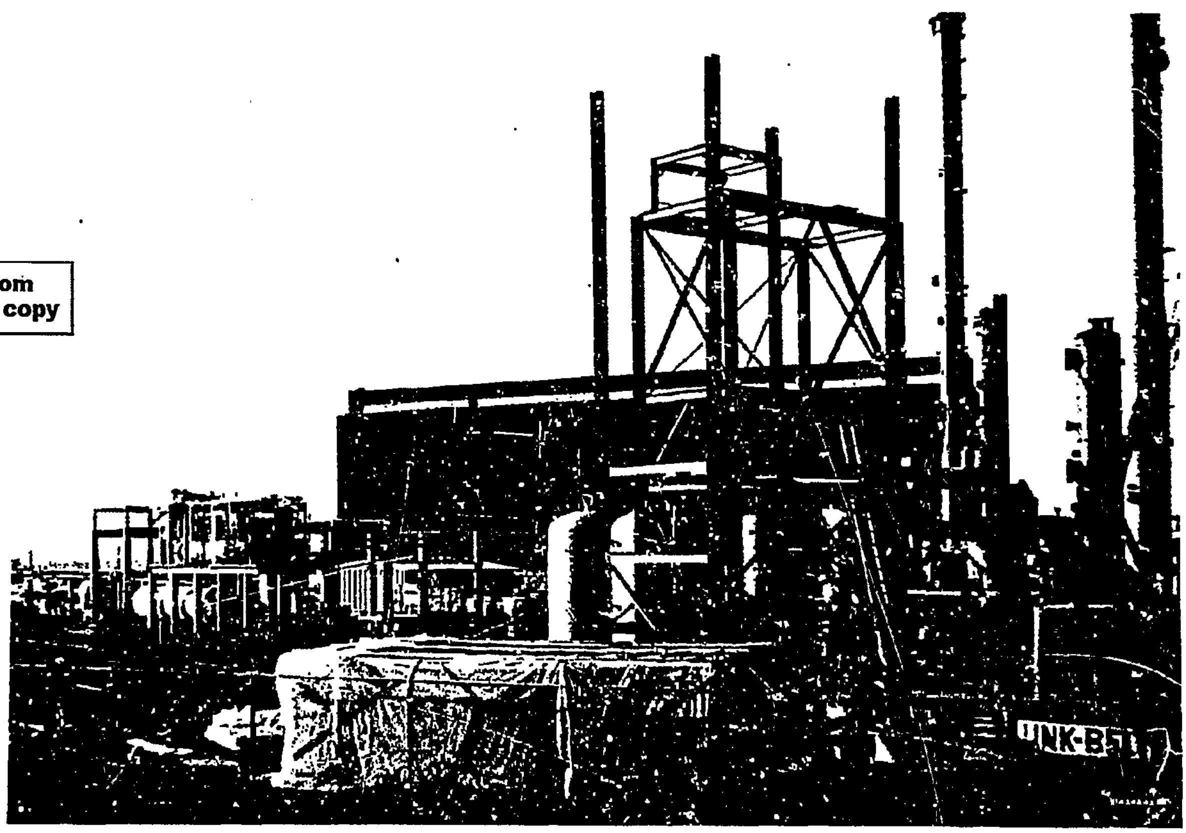
"The undersigned certifies that the amount is due and payable to Procon, in accordance with the terms of the contract up to the date of this Certificate and that the contractor has fully complied with the terms and conditions of the contract."

Progress Report No. 19, February 15, 1970

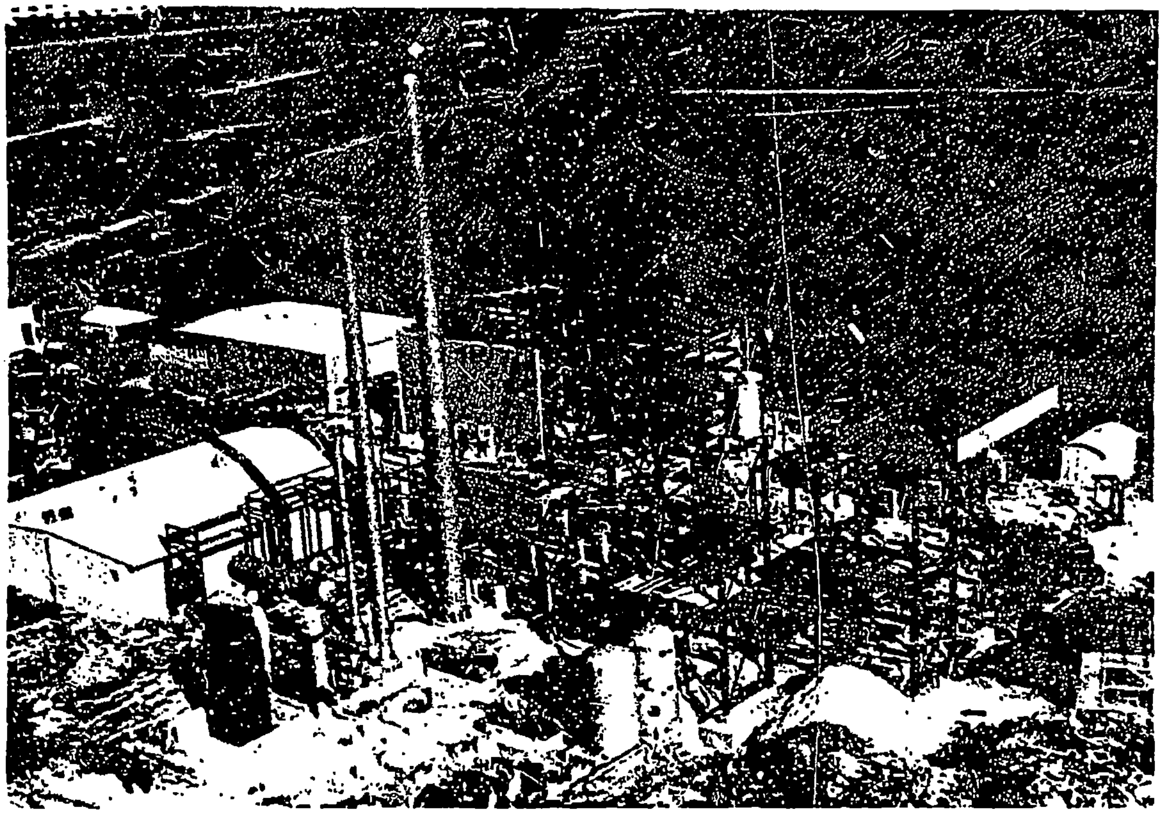
T. A. Taylor



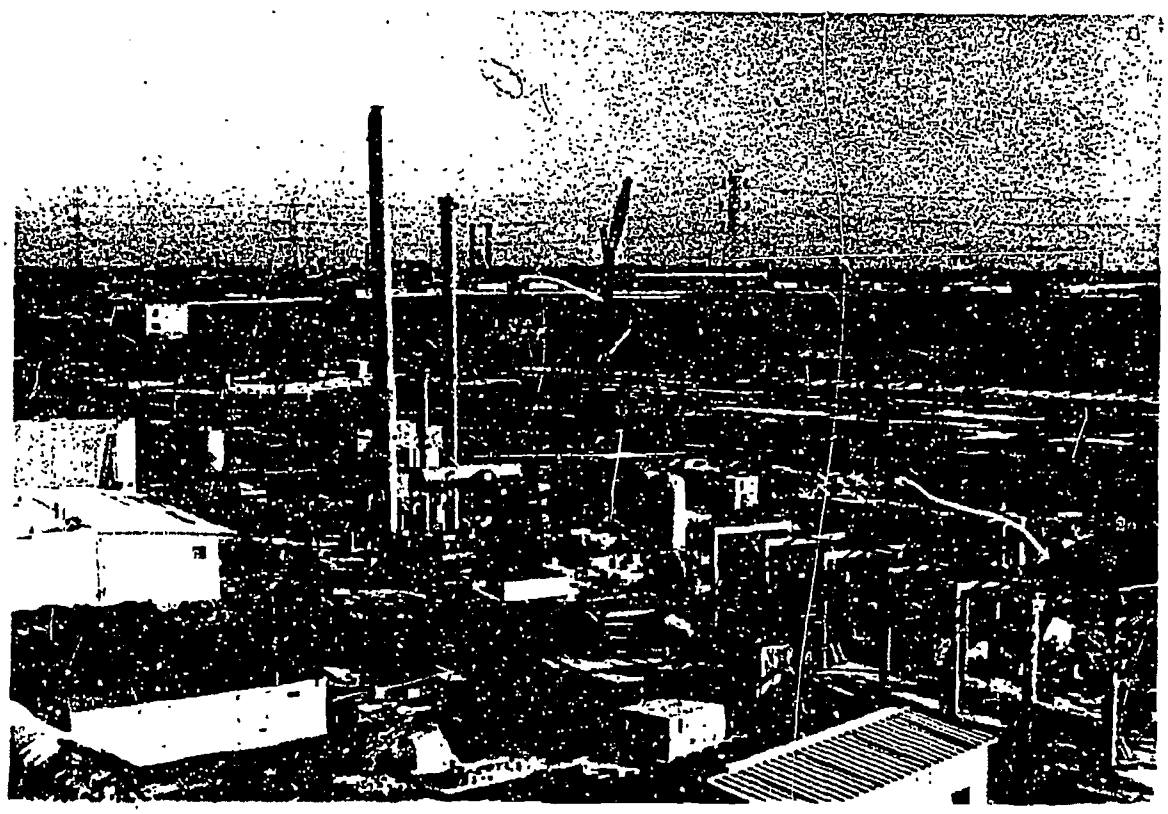
PROCESS AREA - LOOKING NORTHWEST



MATERIAL HANDLING STRUCTURE - LOOKING NORTHWEST



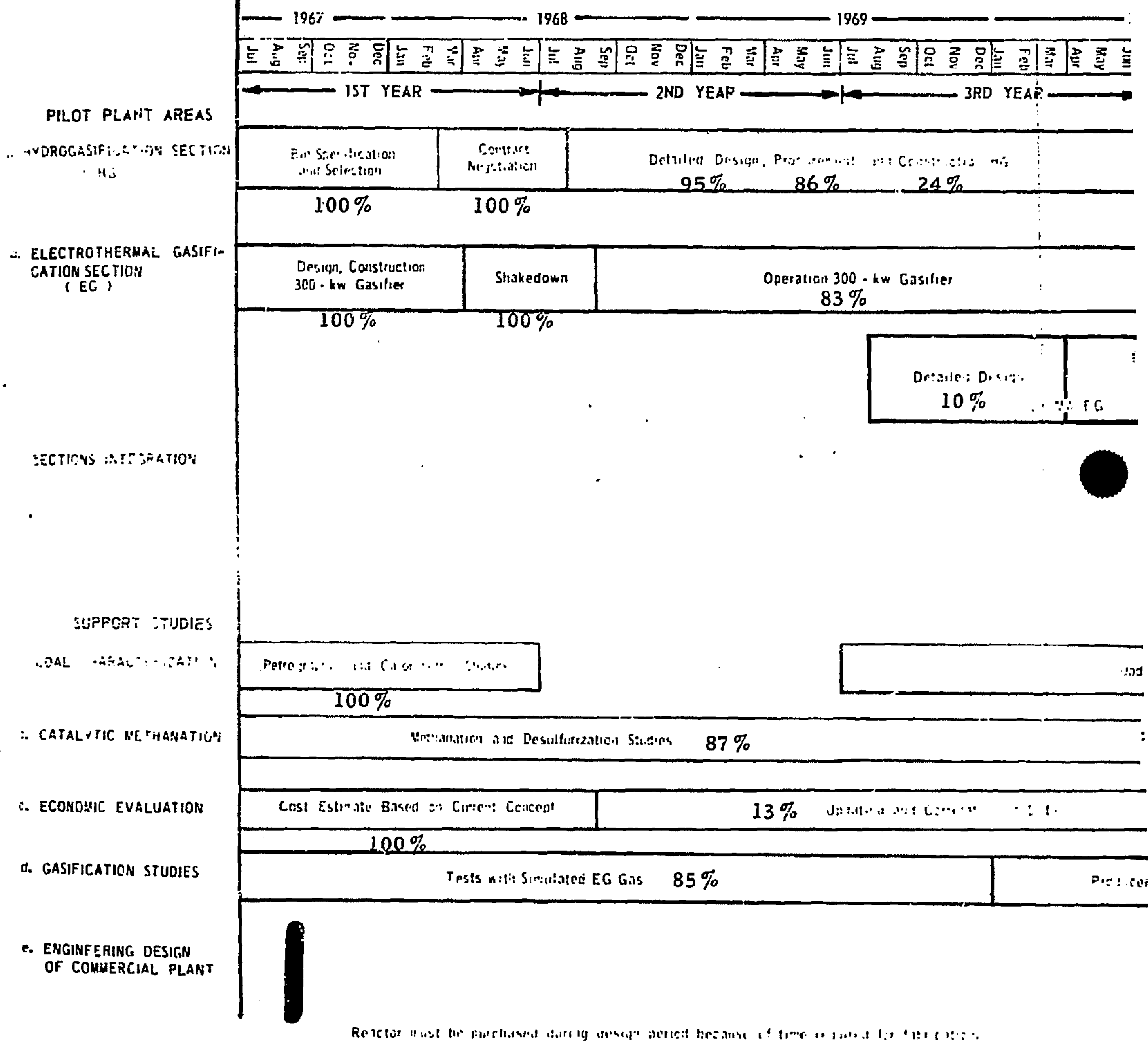
OVERALL PLANT



OVERALL PLANT - LOOKING SOUTHEAST

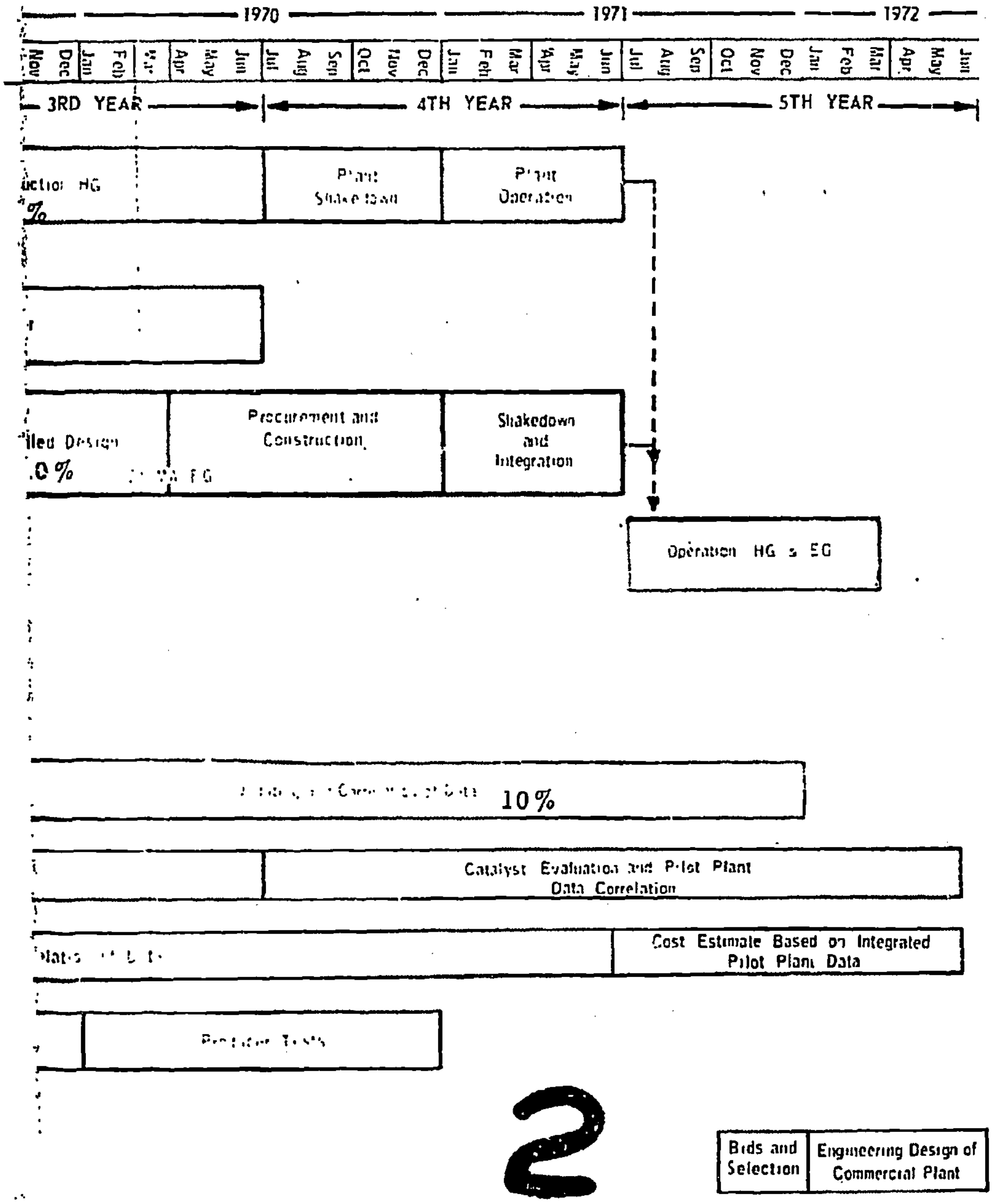
PILOT PLANT PROGRAM OF IGT HYDROGA

OCR CONTRACT No. 14-01-0001-381 (1)



IGT HYDROGASIFICATION PROCESS

381 (1)   AGA : IU-4-1



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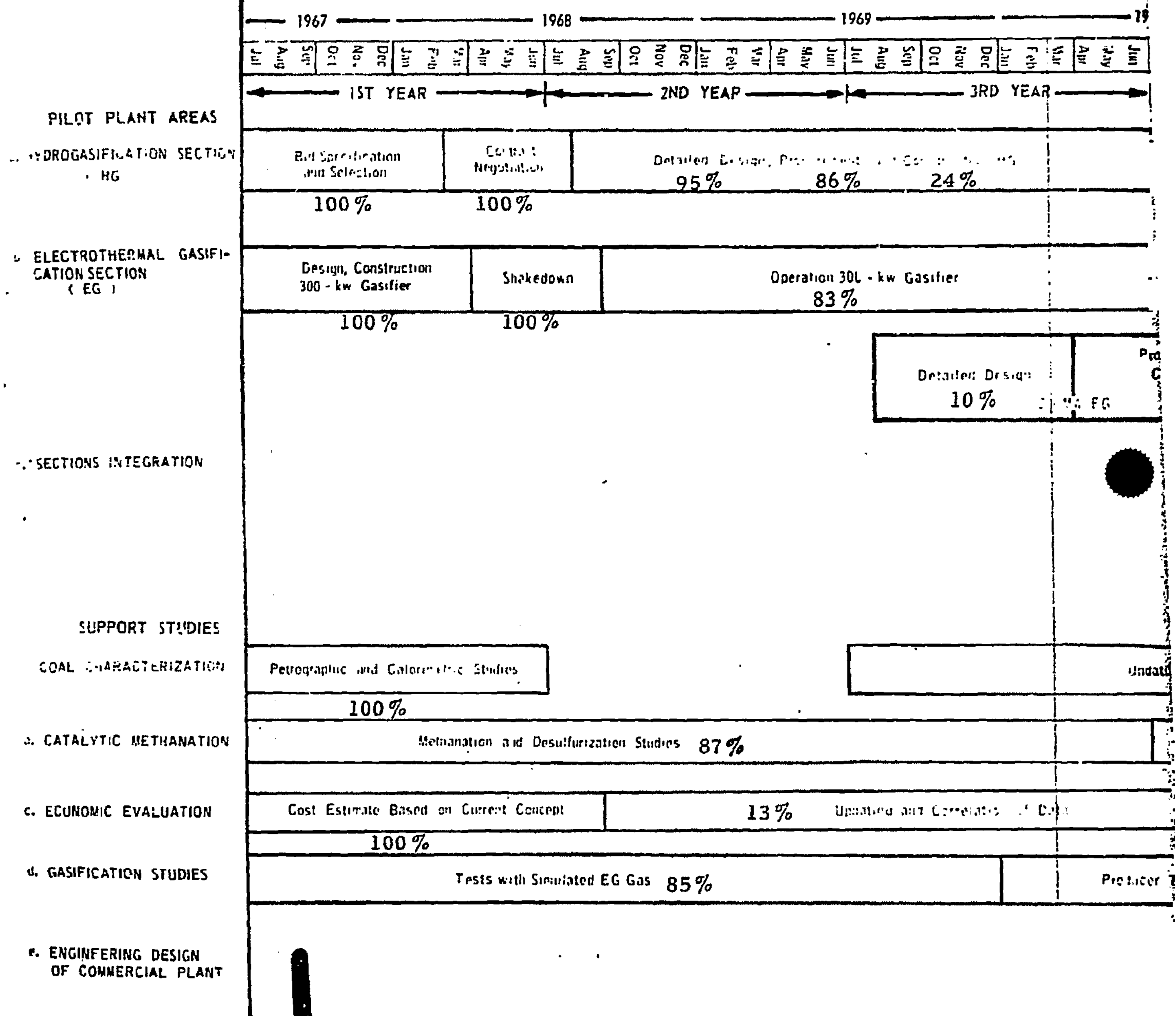
PILOT PLANT PROGRAM OF IGT HYDROGA

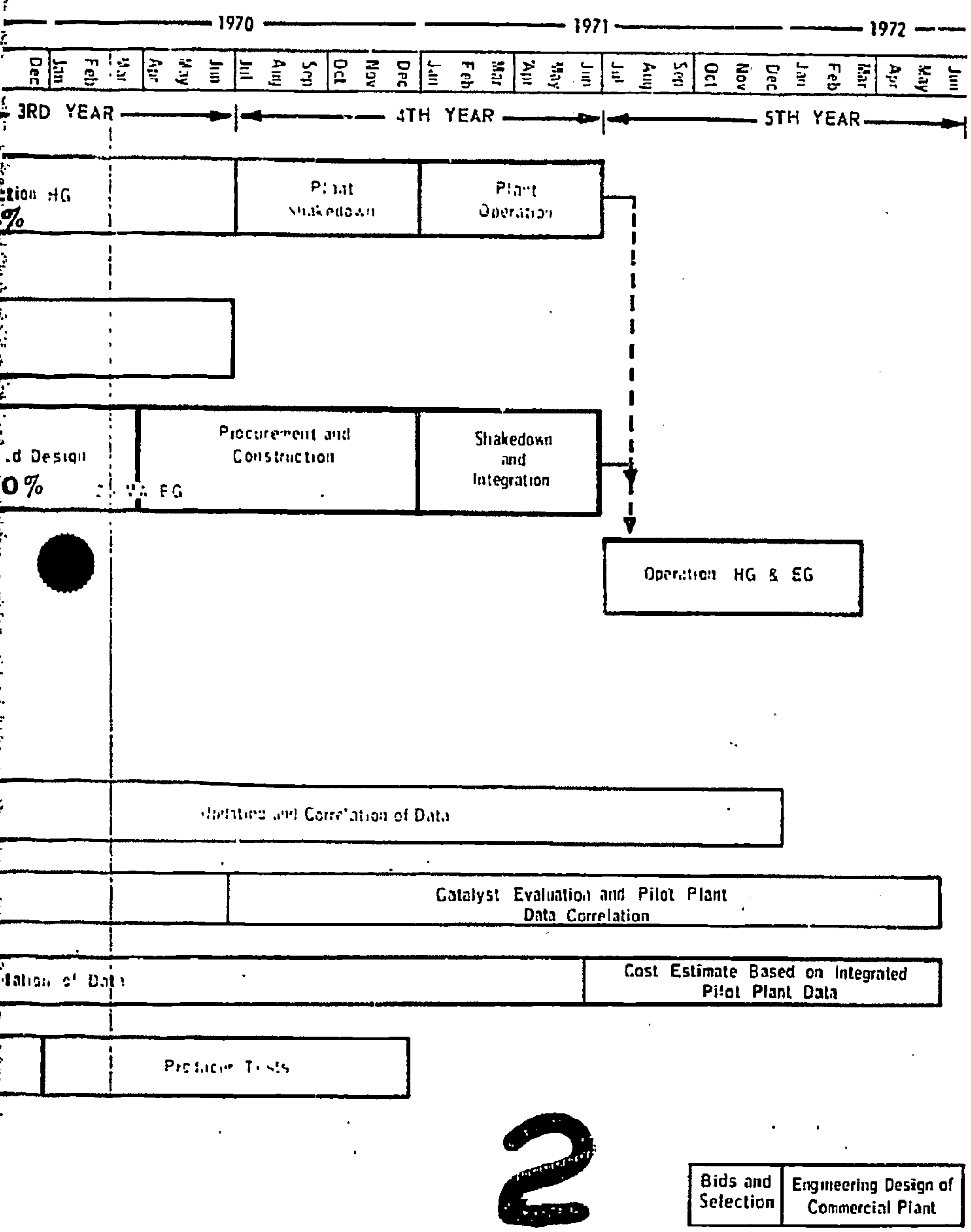
OCR CONTRACT No. 14-01-0001-381 (1)



Reactor must be purchased during design period because of time required for fabrication.

IGT HYDROGASIFICATION PROCESS

381 (1) AGA: IU-4-1



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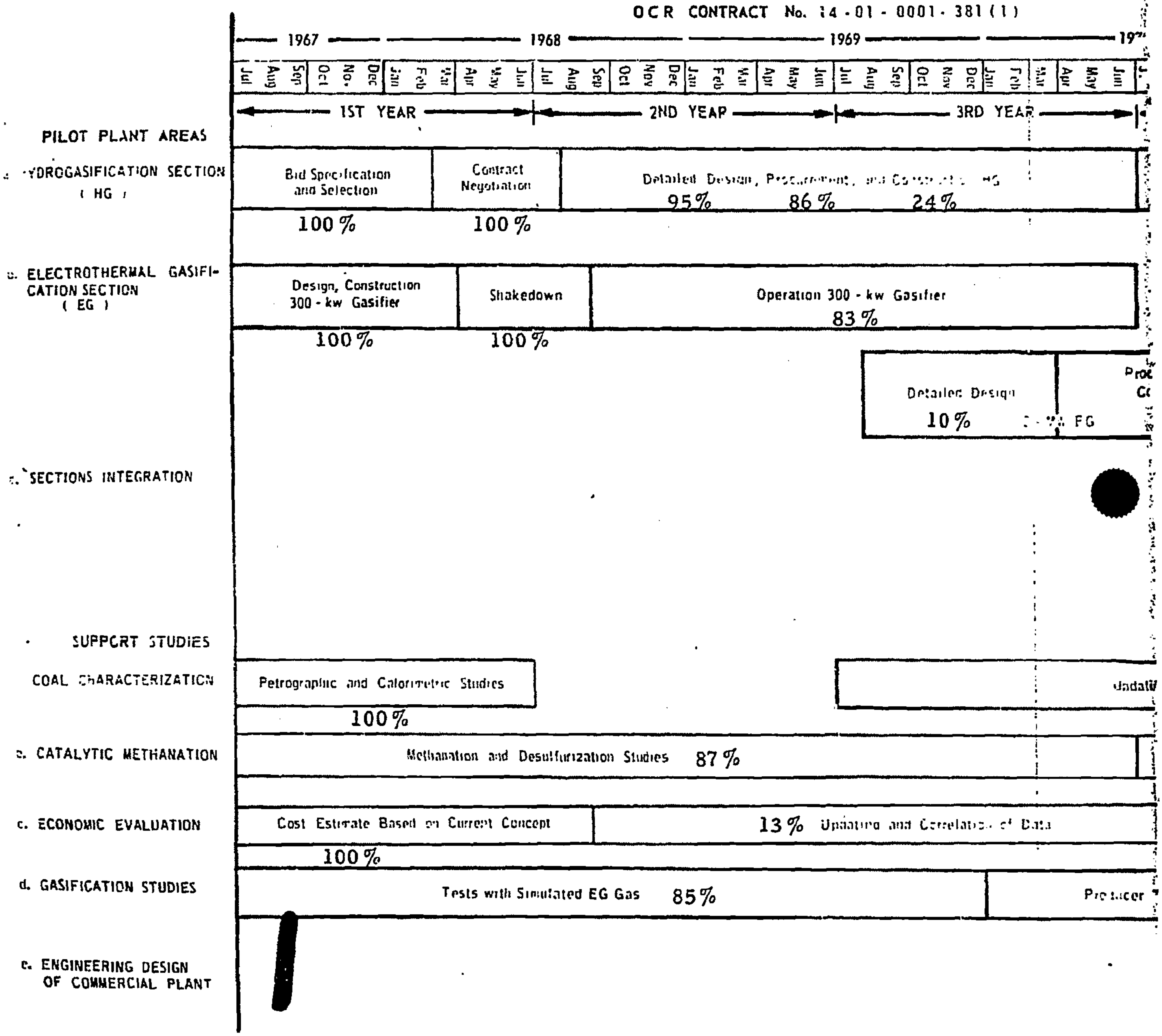
PILOT PLANT PROGRAM OF IGT HYDROGAS

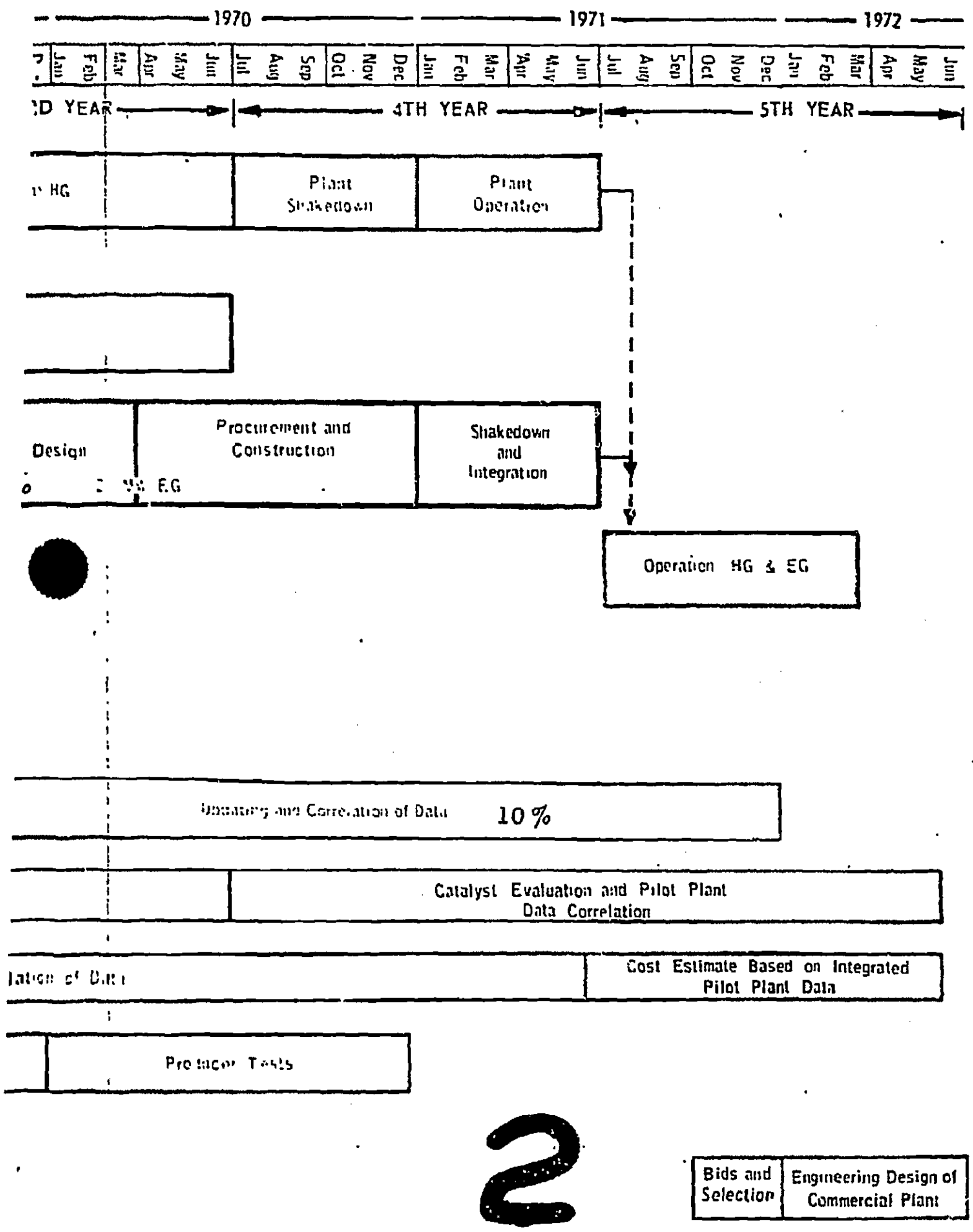
OCR CONTRACT No. 14-01-0001-381(1)



Reactor must be purchased during design period because of time required for fabrication.

GT HYDROGASIFICATION PROCESS

1(1) AGA: IU-4-1



ion.

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IGT-GTPR--1-3/70

INSTITUTE OF GAS TECHNOLOGY · IIT CENTER · CHICAGO 60616

Project Status Report
For
OFFICE OF COAL RESEARCH
and
AMERICAN GAS ASSOCIATION

Report For First Quarter, 1970
OCR Report No. 66

Project Title Pipeline Gas From Coal — Hydrogenation (IGT Hydrogasification Process)

OCR Contract No. 14-01-0001-381 (1)   A.G.A. Project No. IU-4-1

I. Project Objective

The overall objective of this project is a process for producing pipeline gas from coal that is economically attractive for supplementing natural gas supplies. The present objective is the design, construction, and operation of a large integrated pilot plant to obtain scale-up data and operating experience. Developmental research, engineering studies, and economic evaluations are in progress to help attain this objective.

II. Achievements

COAL CHARACTERIZATION

A test for rating the attrition resistance of coal chars is being developed. Initial results from studying the distribution of minor coal constituents in the HYGAS Process show that all the nitrogen removed from the coal appears as ammonia; over 75% of it is in the recycle quench water. If a cooling tower is used, then most of the ammonia would be released to the atmosphere. We are studying means of recovering the ammonia and avoiding atmospheric pollution.

HIGH-PRESSURE METHANATION

Results from a series of runs at low flow rates and low pressure agreed well with results at high pressures. The methanation rate expression has been extended to cover regions with large excesses of hydrogen and methane. An improved correlation was obtained.

$$r = \frac{k_1 p_{CO} p_{H_2}^{0.5}}{1 + k_2 p_{H_2} + k_3 p_{CH_4}}$$

Data at low conversions and near equilibrium are being collected to test this correlation. Initial data indicate that the correlation requires modification for conditions near equilibrium.

Preliminary tests will be made to determine the feasibility of using the heat release during ethylene hydrogenation to start up the HYGAS hydrogasifier.

The small laboratory for measuring the sulfur tolerance of the catalyst is essentially completed. Initial poisoning tests will be done with a typical methanation feed gas doped with 100 ppm of H_2S.

ENGINEERING ECONOMICS STUDIES

A computer program was developed to estimate the cost of vessels as a function of their dimensions and configurations. The effects of financial factors on the return on equity for gas utility financing were calculated and the results presented in graphical form.

A study using air in place of water cooling in a HYGAS plant shows that the cooling water requirement can be reduced by as much as 63%. This assumes air cooling to 140°F with final cooling by water.

The economics of lock hopper and slurry systems for feeding pretreated char to the hydrogasifier were compared. Recent data on lock hoppers were used in the study. The results indicate that the lock hopper system could show a gas price advantage of 3¢/million Btu if a reasonable life of the control valves can be expected. These valves must seal against 500 psi differential pressure and must handle solids flowing through them. Further probing is planned.

DEVELOPMENT UNIT STUDIES

Results from a free-fall thermal treatment of lignite at 1300°F showed 14% carbon gasification using nitrogen as a sweep gas. The degree of gasification at 280 psi is comparable to that from another run at 1000 psi, indicating that devolatilization is the only reaction occurring.

Hydrogasification of lignite at 500 psi with hydrogen and steam showed 41% carbon gasification, indicating no significant loss of reactivity from the 1000-psi operation. Results of lignite gasification at 500 psi with synthesis gas-steam and hydrogen-steam mixtures show that about 5% more carbon (36 vs. 41%) was gasified with the hydrogen-steam mixture. However, either gas mixture is adequate for the HYGAS Process in terms of obtaining the required gasification.

Four runs were made this month. The one using pretreated Ireland mine coal and a synthesis gas-steam mixture at 1500 psi was successful. The other three tests with Montana subbituminous coal at 500 psi were only partially successful due to mechanical difficulties.

Minor mechanical difficulties first in the exit-gas quench system and then in the coal feed system caused the shutdown of a number of runs in the electrothermal gasifier. After modifications, three successful runs were made this month at 1900°F and 1000 psi using our own hydrogasified char. A silicon carbide electrode is being installed for the next run.

Work is in progress to evaluate the performance of spray devices that would be used in the HYGAS pilot plant for distributing the coal feed slurry to the slurry dryer. Equipment is being set up to study the dry letdown of char from high pressure.

NEW PROCESS STUDIES

The fuel cell engineering study was completed this month. Details of power plant configuration and cost calculations are given. A bus bar power cost of 4.5 and 5.4 mills/kWhr is estimated for fuel cell power densities of 300 and 150 watts/sq ft. Capital investment is estimated at $123 and $99/kW for cell power densities of 150 and 300 watts/sq ft.

PILOT PLANT CONSTRUCTION

Engineering is 97% complete, purchasing 89% complete, and material receipt 86% complete. The hydrogasifier vessel was erected on February 21. No delays are expected due to equipment delivery.

Construction is 32% complete. Overtime construction, 50 hr/wk, began on February 10 to attract the pipefitter/welder manpower needed to maintain the construction schedule.

III. Problems

No major problems were encountered this month.

IV. Recommendations

We recommend that the project proceed in the areas defined in the contract amendment.

V. Status of Funding

1. A.G.A. Funding

A.	1970 Funds Allocated	$ 300,000
B.	Funds Expended This Month (estimated)	$ 36,600
C.	Funds Expended to Date (estimated)	$ 110,000

2. OCR Funding

A.	Funds Expended This Month (estimated)	$ 590,000
B.	Funds Expended Since Contract Amendment No. 1 (estimated)	$7,320,000

As a result of personally reviewing the pertinent data and information reasonably available, it is our opinion that the project's objective will be attained within the contract term and the funds allocated.

Approved
Frank C. Schora, Jr.
Director

Signed
Bernard S. Lee
Manager

Jack Huebler
Vice-President

Appendix. Achievements in March

COAL CHARACTERIZATION

Quench Water and Condensate Contamination

The product gas from the hydrogasification step contains ammonia, carbon dioxide, hydrogen sulfide, light oil, and other contaminants that dissolve to some extent in the steam condensate (or quench water if it is used) when the gas is cooled. The amount and disposition of these materials is of concern from the standpoint of water pollution and by-product recovery, and of air pollution if the quench water is recycled through a cooling tower.

The yield of ammonia in hydrogasification has been determined on only one run in the past. A yield of 0.00501 lb NH_3/lb char was found for Run No. 42 made in March 1959 with Consolidation Coal Co. char as feed. This corresponds to a yield of 23% of the nitrogen in the feed.

We are determining the amount of ammonia in the water condensate of current runs. A yield of 0.0096 lb NH_3/lb dry coal was found for Run HT-243 with Montana subbituminous coal as feed. This corresponds to 65% of the nitrogen in the coal.

The conversion of nitrogen in the coal was calculated for all runs on the Montana subbituminous coal and on Ireland mine pretreated coal (Runs HT-94 to HT-169). Results are plotted in Figure 1 against carbon conversion. According to this correlation the expected nitrogen conversion for Run HT-243 is 60%, compared to the determined ammonia yield of 65%. Thus, it appears that substantially all gasified nitrogen is in the form of ammonia.

The 500 billion Btu/day lignite plant in a study by Tsaros et al.* was used as a case study of the flow of ammonia, hydrogen sulfide, light oil, and the accompanying carbon dioxide. Two levels of ammonia, 500 and 2000 moles per hour, were assumed. About 1600 moles per hour were expected for this plant if the correlation for subbituminous coal in Figure 1 holds for lignite, and if all converted nitrogen appears as ammonia.

The plant's quench water circuit with some minor changes in tank conditions is shown in Figure 2. The quench water was assumed to be saturated with carbon dioxide and hydrogen sulfide (corresponding to the composition of the gas entering the quench tower) at the temperature the water leaves the

* Reference 4, p. 20.

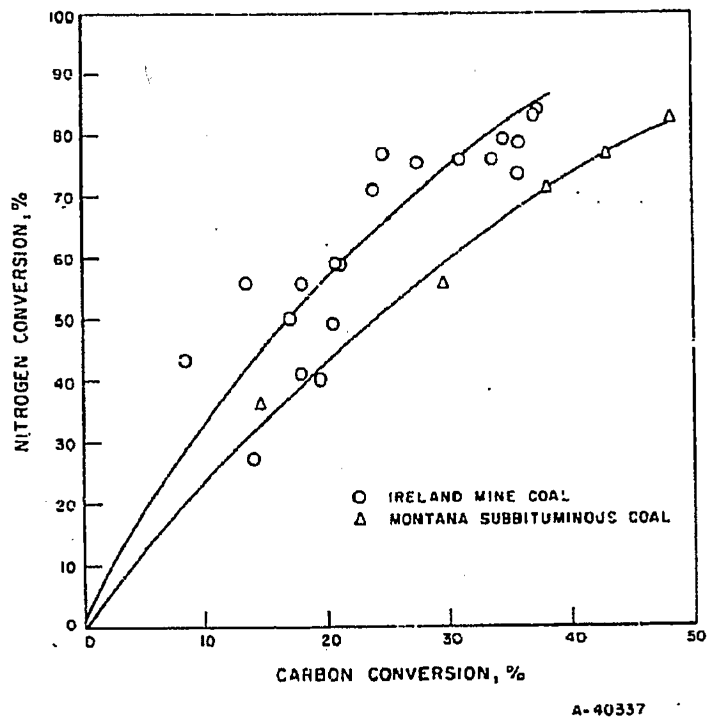

Figure 1. CONVERSION OF COAL NITROGEN IN HYDROGASIFICATION

quench tower. Complete absorption of ammonia was found to be reasonable. The water in the two tanks was flashed to equilibrium. In the recycle-water settling tank a partial pressure of 0.2 atmosphere was assumed for the carbon dioxide in order to remove a substantial part of the hydrogen sulfide and carbon dioxide.

The equilibriums were calculated in terms of the following ionization and Henry's law constants:

Component	Ionization Constant 100°F	Ionization Constant 250°F	Henry's Law Constant 100°F	Henry's Law Constant 250°F
NH_3	1.83×10^{-5}	1.09×10^{-5}	1.9	25
H_2S	9.5×10^{-8}	4.0×10^{-8}	830	1700
CO_2	4.5×10^{-7}	1.9×10^{-7}	2300	5400
H_2O	2.4×10^{-14}	0.98×10^{-12}		

Only the first constant for hydrogen sulfide and carbon dioxide was used. The ionization constants of carbon dioxide and hydrogen sulfude at 250°F were estimated. The solubility and vapor pressure of light oil were taken

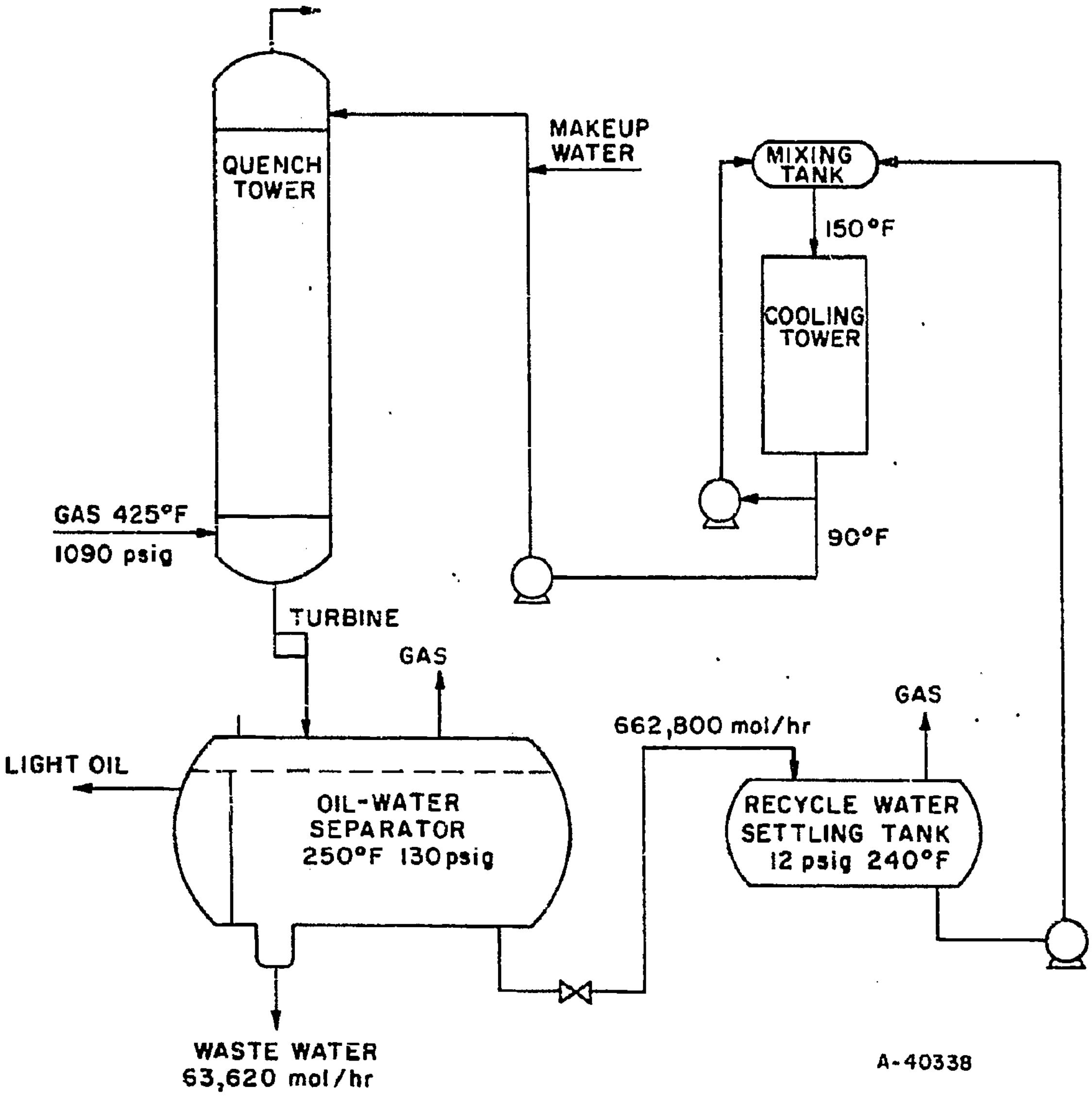

Figure 2. LIGNITE PLANT QUENCH WATER CIRCUIT

to be those of benzene; although, both solubility and vapor pressure will, in fact, be slightly lower.

Calculated amounts and concentrations of the contaminants in the streams of the quench water circuit are presented in Table 1. From these it appears that over three-fourths of the ammonia is retained in the recycle quench water. Most of this ammonia would be lost to the atmosphere in an open-cooling system along with a small amount of hydrogen sulfide and about 4 tons per hour of light oil.

Table. 1. CONTAMINANTS IN QUENCH WATER STREAMS

	CO_2		NH_3		H_2S		Dissolved Light Oil		Water Vapor	
	mole/hr									
NH_3 Rates	500	2000	500	2000	500	2000	500	2000	500	2000
In Quench Water	1870	2810	500	2000	83	94	--	--	--	--
In Gas From Oil-Water Separator	836	1082	1	7	23	24	501	650	334	433
In Gas From Recycle Water Settling Tank	893	1416	34	279	53	63	296	343	7590	11,190
In Recycle Water	132	286	424	1564	6	7	161	118	--	--
In Waste Water*	13	27	41	150	1	1	15	11	--	--
	Concentration, mole fraction X 10^3									
In Quench Water	2.58	3.87	0.69	2.75	0.11	0.13	--	--	--	--
In Water From Oil-Water Separator	1.43	2.38	0.69	2.75	0.08	0.10	0.65	0.65	--	--
In Recycle Water and Waste Water*	0.20	0.42	0.64	2.36	0.01	0.01	0.24	0.18	--	--

* Calculations were made as if the waste water were taken from the recycle water settling tank rather than from the oil-water separator.

HIGH-PRESSURE METHANATION

Kinetic Study

Results of experimental runs made at 575°, 630°, and 850°F and close to equilibrium conditions are being analyzed. The initial indication is that the rate equation proposed in the February 1970 Project Status Report requires modifications at conditions close to equilibrium. None of the reported methanation rate expressions in the literature apply at conditions close to equilibrium conversion.

Ethylene Hydrogenation

An apparatus was built for a preliminary study of ethylene hydrogenation at elevated pressures (up to 500 psig) and temperatures (700°-1100°F). The purpose of this study is to check the feasibility of using the heat release from ethylene hydrogenation to start up the HYGAS hydrogasifier.

The fluidized-bed reactor used for this study is shown in Figure 3. The mixture of gases enters the reactor and passes through a bed of sand or sand and Ni-Mo catalyst. The temperatures of the bed are recorded and the product gas is analyzed. The reactor is 1/2 inch in diameter and is 54 inches long. Thermowells are installed from both ends so that the temperatures of three positions in the reactor can be recorded.

Sulfur Resistance Studies

The laboratory unit is essentially completed, and gas mixtures that will be doped with sulfur compounds are being prepared. A meeting was held with Harshaw Chemical Co. recently, which provided information on the poisoning behavior of nickel-methanation catalysts. Based on this discussion we can probably limit our studies to H_2S poisoning, since Harshaw feels that all sulfur compounds are essentially equally toxic and will all be removed quantitatively by the catalyst.

ENGINEERING ECONOMICS STUDIES

Economic Comparison of Lock Hopper and Slurry Systems for Feeding Pretreated Char to the Hydrogasifier

Two systems to feed char from atmospheric pressure to the 1100-psig hydrogasifier have been used in previous designs: lock hoppers and slurry feed. The HYGAS pilot plant will use a slurry feed system because the

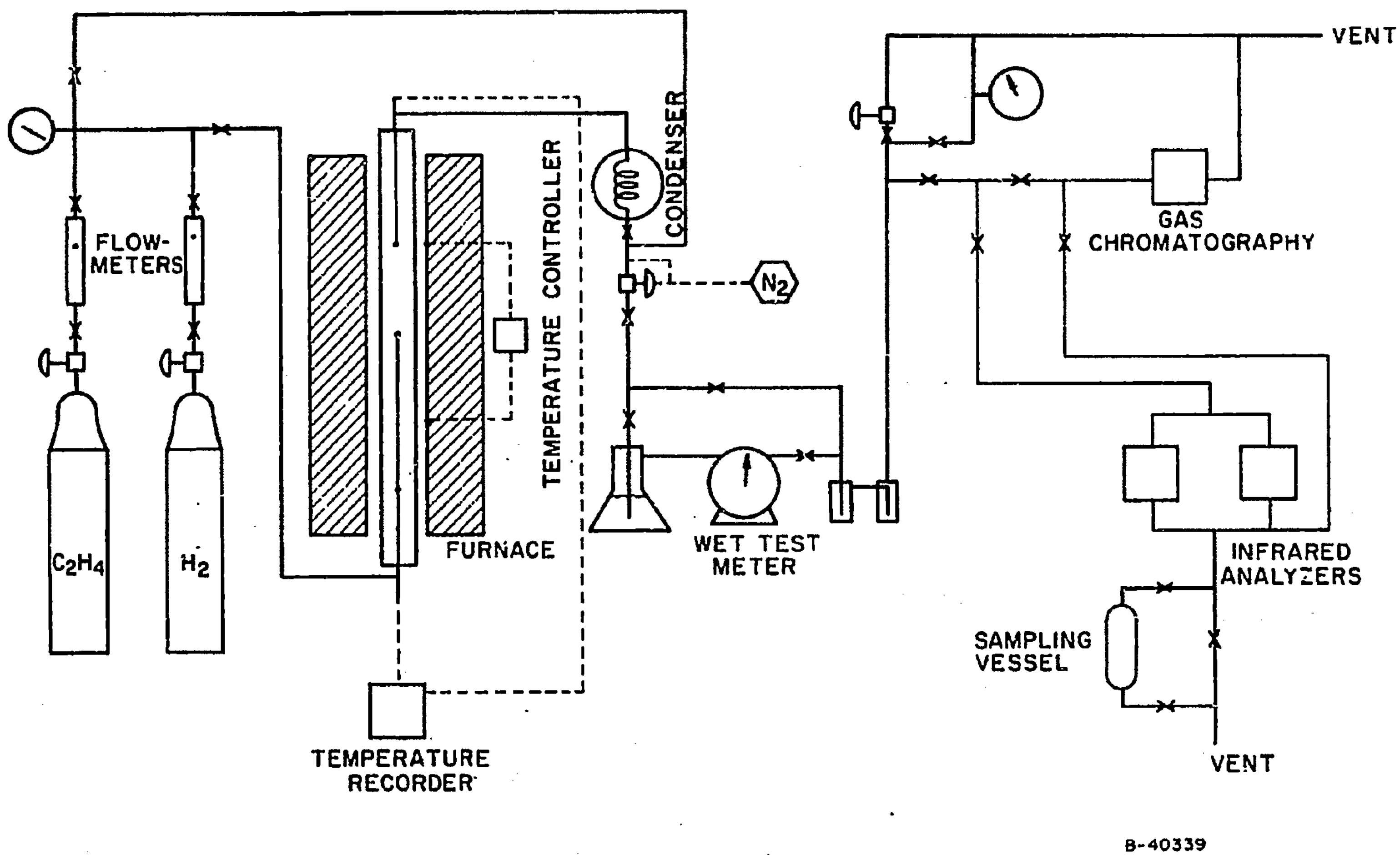

Figure 3. DIAGRAM OF ETHYLENE HYDROGENATION APPARATUS

operation is expected to be easier and, following suit, the cost estimate for the 500 billion Btu/day pipeline gas-from-lignite plant also used such a system. However, we did not have an economic comparison of lock hoppers versus slurry feed systems for a coal-based gas plant.

The original lock hopper system used in the earlier cost estimates was derived from one used by Kellogg in its hydrogasification system design for Project PB-23a.[2] The basic ideas in that design were also used in the three pipeline gas-from-coal plant estimates that preceded the design for lignite.[1,5,6] A lock hopper operating on a half-hour cycle discharged into a continuous feed hopper with a 30-minute holding capacity for the above designs; 8 pairs of such hoppers are required to handle the flow of char. Lock hoppers are pressurized either with hydrogen or methane, which after depressurization is vented to a large surge drum and then recompressed, requiring significant power consumption.

With some up-to-date information on lock hoppers and coal preparation costs, we decided to size and cost both systems to handle the process coal required for the gas plant using electrothermal gasification.[1] In the original electrothermal design, only part of the spent char flows from the hydrogasifier to the electrogasifier; the excess must be cooled and let down to atmospheric pressure as a by-product fuel. The same scheme was used for both cases compared. By-product char is quenched in a spray chamber from 1800° to 600°F by water sprays and the steam generated (145 million Btu/hr) used to pressure the letdown lock hoppers. Since this same scheme and credit for heat recovery is used for both char feeding systems, there is no effect on the relative economics of either.

Lock Hopper Feed System

Figure 4 shows the system for feeding char via lock hoppers. Ground coal from the preparation system is pneumatically transferred by air to the pretreatment feed hopper, which also acts as a ground-coal surge tank. Pretreated, hot char is pneumatically conveyed by low-pressure, lock hopper discharge steam. Since the char is hot there should be no condensation problem. This system avoids cooling the char, which would be necessary if conveyors were used.

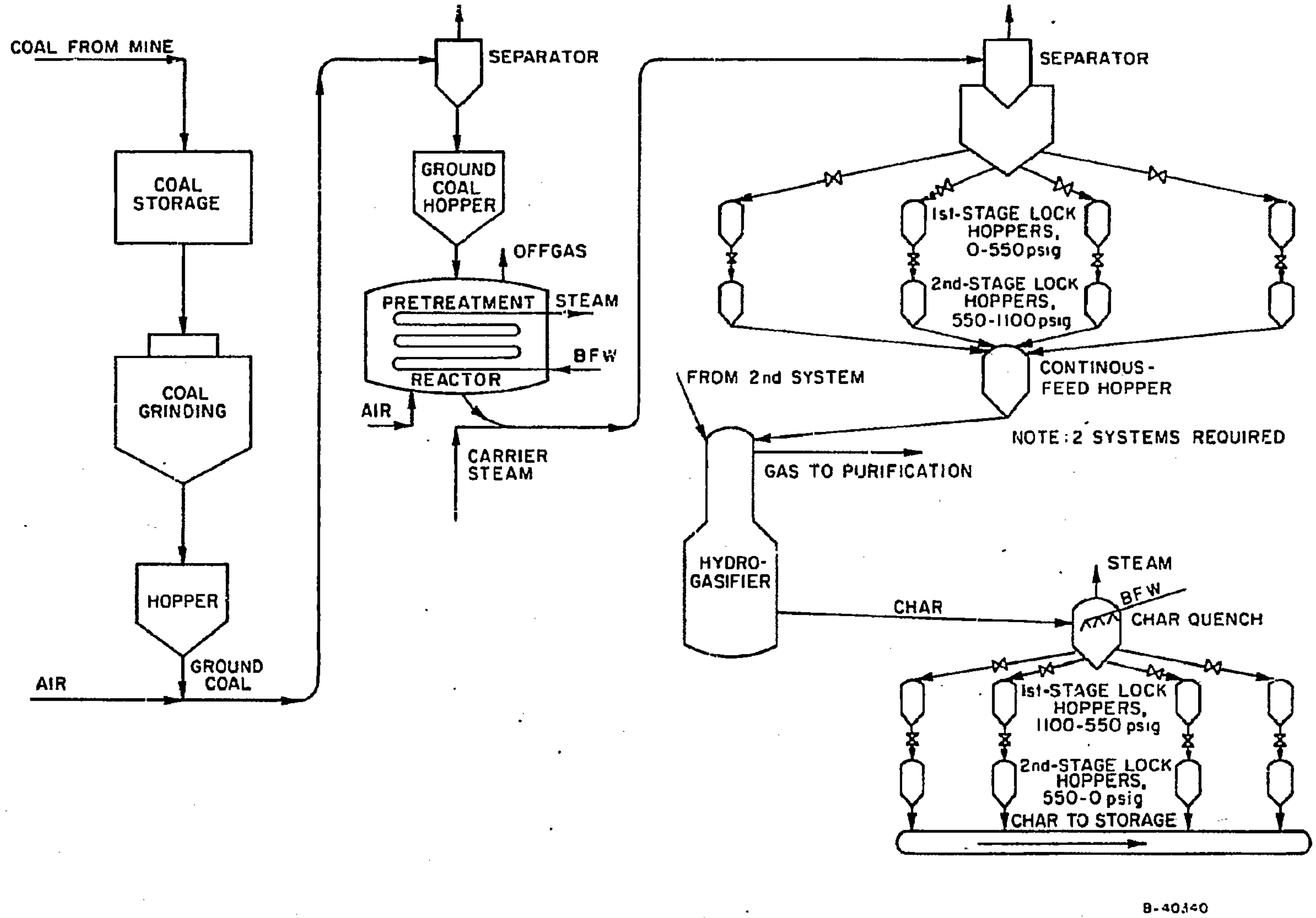

Figure 4. COAL PREPARATION AND FEEDING TO HYDROGASIFIER BY LOCK HOPPER SYSTEM

Because of valve problems, it may not be possible to use a single lock hopper stage between 0 and 1100 psig. Assuming a 550-psi differential to be feasible, we used two lock hopper stages. Steam is used to pressure the hoppers, each operating on a 15-minute cycle. The low- and high-pressure cycles are set so that steam from the high-pressure hopper (550-1100 psig) discharges into the low-pressure hopper (0-550 psig) during depressurization. As the low-pressure hopper depressurizes down to 50 psig, steam is fed to the main and used for conveying char and regenerating hot carbonate solutions.

With 15-minute staggered cycles, one continuous-feed hopper is used for 4 pairs of high- and low-pressure lock hoppers. These hoppers have a 6-ft ID, based on information obtained from Lurgi. Feed hoppers are 18.5-ft long, and by-product char hoppers are 14.5-ft long. Costs for the individual hoppers are based on the costs of metal plus an allowance of $10,000 and $6000 each for the internals for low- and high-pressure hoppers. These sizes necessitate a large number of hoppers. For each hydrogasifier train there are two continuous-feed hoppers each fed by 4 pairs of lock hoppers, and one continuous by-product char hopper feeding 4 pairs of both hoppers. Since there are four hydrogasifier trains, the total number of hoppers is large. By doubling the capacity of the individual hopper the number of feed units would be halved, with a reduction in cost.

In comparing costs for the revised and original feed systems (Table 2), the saving in the installed cost of the former is $6 million over the original system. Major savings for the revised system are in continuous hoppers (one for each 4 lock hoppers instead of one-for-one as in the original system), in the use of 15-minute cycles, and in the elimination of conveyors and the lock hopper gas recycle system. The net power saved is 2190 kW, reducing the annual purchased power cost to $8,358,200. There are some savings in feed hopper unit costs because of lower vessel unit costs used in the revised system. (The same savings in vessel unit costs are also applied in the slurry feed system.)

Table 3 summarizes the investment for the whole plant. (Compare with Table 5 in Reference 1 for the original lock hopper case.) Total installed equipment is reduced by $6,650,000 from the original. In addition to the above savings in the hydrogasification system there is a $2,725,000 increase

Table 2. COST COMPARISON OF REVISED AND ORIGINAL FEED SYSTEMS

	Original Lock Hopper System, $	Revised System, $
Feed Hoppers	9,600	--
Feed Cyclone	--	32,000
Dust Lock Hoppers	--	40,000
Feed Lock Hoppers	1,746,400	2,032,000
Continuous-Feed Hoppers	1,390,800	176,000
Hydrogasifiers (4)	3,198,000	3,198,000
Cyclones	160,000	160,000
By-Product Char Hoppers	1,650,000	1,083,000
Vent-Gas Compressor	525,000	--
Vent-Gas Surge Drum	49,600	--
Conveyors	135,000	--
Total Equipment	8,864,400	6,721,000
Installed Cost	25,100,000	19,000,000

in the sum of coal storage and grinding and a $3,280,000 decrease in offsite facilities. The former is due to updated costs. Savings in offsites are due to a big reduction in the cost of a fired heater for superheating reaction steam. (The cost is reduced from the boiler cost used in Reference 1, which was derived from an earlier figure now believed to be too high.)

Annual operating costs, calculated by the standard accounting procedure, are given in Table 4. In addition to regular maintenance at 3%, additional maintenance for the lock hoppers amounts to $426,000 per year. This is the 20-year average cost of replacing internals every 2 years. Coal feed lock hoppers have given few problems at the Lurgi plant at Sasol. Ash lock hoppers have given more problems.[3] The gas price with this new system is 50¢/million Btu, 1.1¢ less than the original estimate. In the latter no extra maintenance for lock hoppers was included, so on the same basis the difference would be 1.5¢ to 2¢/million Btu.

Slurry Feed System

For slurry feed (Figure 5) the system in the pipeline-gas-from-lignite study was used as a model.[4] In order to prepare slurry feed so that light oil does not vaporize, the temperature of pretreated char is lowered from 700° to 285°F in the char cooler. The char cooler is similar in design to the pretreater except that pretreatment off-gas instead of air is used for

Table 3. SUMMARY OF REVISED INVESTMENT FOR 258 BILLION Btu/DAY PIPELINE GAS – HYDROGASIFICATION WITH SYNTHESIS GAS GENERATED BY ELECTROTHERMAL GASIFICATION

	Revised Lock Hopper, $	Slurry Feed, $
Coal Storage and Handling	2,660,000	2,660,000
Coal Grinding and Preparation	4,005,000	4,005,000
Coal Pretreatment, Char Handling	4,650,000	5,796,000
Slurry Makeup and Feed System	--	4,958,000
Hydrogasification	19,000,000*	17,685,000 †
Quench Tower and Light Oil Recovery	--	5,897,000
Prepurification	15,210,000‡	11,760,000
Methanation and Drying	2,800,000	2,800,000
Gasification	7,510,000	7,510,000
Electrical Equipment	2,000,000	2,000,000
Offsite Facilities	12,430,000	14,800,000
Total Bare Cost	70,265,000	79,871,000
Eng Overhead and Profit, 7.73%	5,431,000	6,174,000
	75,696,000	86,045,000
Interest During Construction	3,785,000	4,302,000
Fixed Investment, 5%	79,481,000	90,347,000
Working Capital	6,110,000	6,326,000
Total Capital Investment	85,591,000	96,673,000

* Reactor plus feed and spent-char lock hoppers.

† Includes $5,570,000 for slurry vaporizer.

‡ Includes $3,450,000 for waste-heat recovery.

Table 4. ANNUAL OPERATING COSTS FOR 258 BILLION Btu/DAY PIPELINE GAS — HYDROGASIFICATION WITH SYNTHESIS GAS GENERATED BY ELECTROTHERMAL GASIFICATION

	Revised Lock Hopper, $	Slurry Feed, $
Raw Material	23,377,000	23,377,000
Other Direct Materials	250,000	250,000
Direct Operating Labor	883,000	883,000
Maintenance at 3% of Bare Cost	2,108,000	2,396,000
Added Lock Hopper Maintenance	426,000	--
Supplies, 15%	316,200	359,400
Supervision	88,300	88,300
Payroll Overhead	97,000	97,000
General Plant Overhead	1,697,800	1,862,900
Depreciation	3,974,000	4,517,400
Local Taxes and Insurance	2,384,000	2,710,400
Electric Power at 3 mills/kWhr	8,358,200	8,617,400
Operating Expense	43,959,500	45,158,800
Contingency at 2%	879,200	903,200
Total Operating Cost	44,838,700	46,062,000
By-Product Credit	7,030,000	6,469,900
Net Operating Cost	37,808,700	39,592,100
Capital Charges	4,548,600	5,097,500
Annual Revenue Requirement	42,357,300	44,689,600
20-Year Average Price of Gas, ¢/10^6 Btu	50.0	52.7

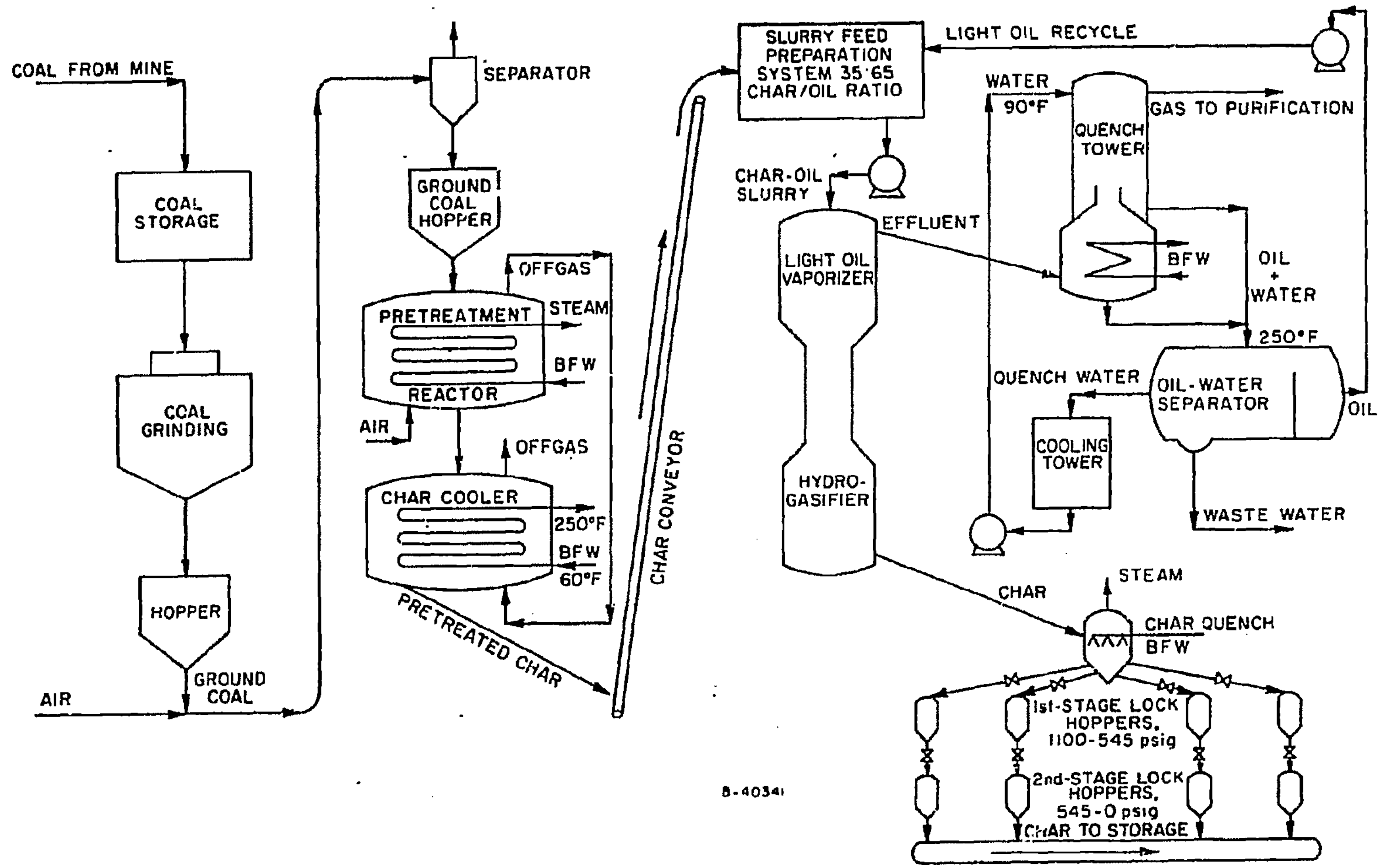

Figure 5, COAL PREPARATION AND FEEDING TO HYDROGASIFIER BY SLURRY FEED SYSTEM fluidizing char. Use of the char cooler recovers about 379 million Btu/hr from char and pretreatment off-gas by preheating boiler feedwater to 250°F through heat transfer coils in the vessel.

Char at 285°F is mixed with a recirculating stream of light oil at 115°F. With char from coal the benzene/char weight ratio is 65:35, compared to 35:65 for benzene/lignite. The high porosity of pretreated char makes it necessary to have a large proportion of light oil in the slurry in order to ensure pumpability and handling of the slurry. Benzene required for 1,212,500 lb/hr of char is 2,236,000 lb/hr compared to 1,740,000 lb/hr of benzene for 3,230,000 lb/hr lignite.[4] On the basis of slurry densities the volumetric rates for the two cases were about the same, so investment for slurry preparation was taken as the same. Costs for the slurry vaporizer, quench tower, and benzene recovery were designed and estimated individually. Centrifugal pumps are used to transfer slurry from makeup tanks to slurry holding tanks sized for 20-minute slurry-holding capacity.

The light oil vaporizer operates as in the lignite study. The light oil in the feed slurry is vaporized by the hot effluent gases from the low-temperature reactor section and is carried to the quench system. From the vaporizer the char flows to the low-temperature zone of the hydrogasifier. Because of the high oil-to-solids ratio in the light-oil vaporizer, a large amount of heat is used up in vaporizing the light oil, resulting in a lower temperature of light-oil vaporizer effluent than in the lignite case (500° vs. 625°F). This results in a lower amount of waste-heat recovery.

In order to recover some of the waste heat, a different approach from that used in the lignite study was used in designing the quench system. The new approach consists in having two sections for the quench tower: a) a heat recovery section and b) a packed tower section.

a. In the heat recovery section light-oil vaporizer effluent is cooled from 500° to 360°F accompanied by condensation of light oil and water vapor. This condensation results in a good heat transfer coefficient. Waste heat is transferred through heat transfer coils and used to generate 100-psig steam for the hot carbonate regeneration system.

b. The packed section of the quench tower is similar to the one designed for the lignite study. Effluent from the heat exchange section is brought in direct contact with quench water at 90°F in a packed-bed tower where effluent gases are cooled from 360° to 100°F; the quench water is heated to 250°F accompanied by condensation of light oil and water.

This modified quench system resulted in a waste-heat saving of 630 million Btu/hr plus a reduction in the water otherwise needed for the removal of this heat. The lower water requirement results in a smaller quench tower, cooling tower, pumps, storage tanks, and other related equipment. The fuel value of the heat saved by the new quench system is about $1 million per year. The mixture of oil and water at 250°F is reduced in pressure to 50 psig in hydraulic power recovery turbines and sent to an oil-water separator. Most of the oil at 250°F is recycled to slurry preparation.

Water from the oil-water separator is reduced in pressure to 15 psig to release acid gases, and stored in settling tanks for 10-15 minutes. Water from the settling tank at 250°F is sent to a direct-contact cooling tower by the cooling tower feed pump.

Economics of Lock Hopper and Slurry Feed Systems

Tables 3 and 4 compare investment and operating costs for the entire plant with the two different char feeding systems. The bare equipment cost with slurry feed is $9.6 million higher than with the revised lock hopper. The extra cost is due to the vessel for cooling the pretreated char and recovering the heat, the slurry makeup system, quench tower and light-oil recovery equipment, and to the larger fired heater (offsites). The fluidized-bed slurry vaporizer costs $1.3 million less than the lock hoppers; however, the quench tower and light-oil recovery systems add $5.9 million. With the lock hopper system there is $3.45 million for waste-heat recovery (in the original design), which occurs between the hydrogasifiers and the hot carbonate scrubbers. For slurry feeding, there are some waste-heat recovery coils in the char cooler and quench tower, the cost of which is included in these items. There is less waste heat to recover in this system.

Annual charges are higher for slurry feed because of both higher investment and operating costs. Total operating labor for both plants was assumed to be the same, although there might be a couple more operators for the slurry system. This difference would not be significant. For the lock hopper system there was actually a reduction in power required over the base case because of the elimination of the recycle compressor for lock hopper gas. For the slurry feed there is an increase in net power (8780 kW), which raises annual power cost to $8,617,400. Because of the loss of waste heat in the slurry feed system, more char must be burned, so the by-product credit is reduced from $7,030,000 to $6,469,900.

Both schemes can probably be improved: The size of the lock hoppers could be increased, with a reduction in the number of parallel systems. Lock hoppers have a disadvantage in the need to control a large pressure differential with valves subject to erosion and/or interference by solid particles. The slurry system would probably have less operating problems, but has the economic disadvantage of the slurry liquid recycle. A lot of equipment is required and energy wasted in making slurry, vaporizing light oil with valuable high-temperature waste heat, and then condensing it with cooling water. This results in increased investment and operating costs.

The 2.7¢ differential is a preliminary one; both systems merit further study to firm up costs.

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DEVELOPMENT UNIT STUDIES

Hydrogasification Tests

Four hydrogasification tests were performed this month in the high-temperature balanced-pressure development unit to study the effect of system pressure on the conversion of bituminous and subbituminous coal. In Run HT-242 the reactivity of a lightly pretreated Pittsburgh No. 8 seam bituminous coal from the Ireland mine to hydrogasification with a synthesis gas-steam feed gas at 1500 psig was determined. In Runs HT-243, HT-244, and HT-245 the hydrogasification performance of a dried Montana subbituminous coal from the Colstrip mine was investigated at 500 psig. The feed gas in Run HT-243 was a mixture of synthesis gas and steam; in the two other runs, it was hydrogen and steam. The test with the Pittsburgh seam bituminous coal was successful. The test with Montana subbituminous coal and a synthesis gas-steam feed gas was partially successful, being of shorter than normal duration. The two other tests with the subbituminous coal were terminated before steady-state operation was reached because of coal feeding difficulties.

Both of the coals were prepared for hydrogasification by first crushing and screening to a −10+80 mesh size. The bituminous coal was then lightly pretreated with air plus nitrogen at 750°-800°F in the fluidized-bed coal pretreatment unit. The subbituminous coal, after the crushing and screening, was dried with air plus nitrogen in the same unit at 220°-240°F from an as-received moisture level of 23% to one of 3%.

Significant features of the four tests are given in Table 5.

Run HT-242 is a repeat of Run HT-225 reported in the July 1969 Project Status Report. In Run HT-225 two of the reactor heating zones failed during the test and the coal-bed temperatures were much below the desired 1700°F level. Although we had all of the reactor heaters operating during Run HT-242, the coal-bed temperatures were not significantly higher than those of Run HT-225. In the upper 1-1/2 ft of the 3.5-ft fluidized coal bed the temperatures were 1555°-1600°F; in the lower part of the bed they were 1250°-1350°F. Bituminous coal was fed at a rate of 42.6 lb/hr, synthesis gas at a rate of 674 SCF/hr, and steam at a rate of 32.4 lb/hr. At these flow rates, the equivalent hydrogen (hydrogen plus carbon monoxide)-to-coal ratio was 40.9% of the stoichiometric ratio. The steam concentration in the

Table 5. FEATURES OF HYDROGASIFICATION TEST RESULTS FOR RUNS HT-242 TO HT-245

Run No.	Temperature, °F	Purpose of Run	Results
		Feed Solids: Pretreated Pittsburgh No. 8 Seam Bituminous Coal, Ireland Mine	
HT-242	1300-1700	To study the hydrogasification reactivity of a high-volatile-content bituminous coal with a synthesis gas and steam at 1500 psig	Successful
		Feed Solids: Dried Montana Subbituminous Coal, Colstrip Mine	
HT-243	1300-1700	To study the hydrogasification reactivity of a subbituminous coal with a synthesis gas and steam at 500 psig	Partially successful, short duration
HT-244	1300-1700	To study the hydrogasification reactivity of a subbituminous coal with hydrogen and steam at 500 psig	Leak at base of gas feed tube
HT-245	1300-1700	Same as HT-244	Coal feed screw jammed

feed gas was 50.3%. The test lasted 4-1/4 hr with 2-3/4 hr of this time at steady state. Preliminary results show that 24.5% of the carbon was gasified, compared to 17.4% attained in Run HT-225.

Dried, but otherwise untreated, Montana subbituminous coal fed at a rate of 21 lb/hr was gasified in a 3.5-ft fluidized bed in Run HT-243 with synthesis gas (265 SCF/hr) and steam (12.6 lb/hr). The equivalent hydrogen (hydrogen plus carbon monoxide)-to-coal ratio was 35% of the stoichiometric ratio; the steam concentration in the feed gas was 50 mole percent. The test lasted 3-1/4 hr, but only 20 minutes of this time was at steady state. The test was interrupted several times by a coal feeding stoppage as the coal packed at the outlet of the feed screw. Too low a nitrogen purge rate to the top of the reactor was partially responsible for the coal packing. Tentative results show that 30% of the carbon in the coal was gasified.

In Run HT-244 we were gasifying the Montana subbituminous coal in a 3.5-ft fluidized bed with hydrogen and steam. Nominal feed conditions were

51.6 lb/hr coal, 345 SCF/hr hydrogen, and 8.8 lb/hr steam. The hydrogen-to-coal ratio was 20% of the stoichiometric ratio; the steam concentration in the feed gas was 35 mole percent. This test had to be terminated just after steady-state operation was reached because of a failure in the discharge of the coal bed. A leak at the base of the gas feed tube allowed steam to leak and condense in the discharge screw housing. This wet the coal and kept it from flowing. Product-gas samples were taken before shutdown so that a conditional evaluation of the test could be made.

Run HT-245 was conducted at conditions similar to those of Run HT-244. After feeding coal for 12 minutes, the coal feed screw jammed. The test was terminated when efforts to clear the jamming by partial depressurization of the reactor failed. Analysis of the coal feed showed a moisture content of 8%. At this relatively high moisture level the coal particles tend to pack under the forces exerted by the feed screw and to jam at the mouth of the coal feed tube. To remedy this situation the coal will be redried to a 3% moisture level for the next hydrogasification test.

Complete hydrogasification results of Runs HT-242 and HT-243 will be presented when analyses of these tests are completed.

ELECTROTHERMAL GASIFICATION

During the month three successful tests were conducted in the electrothermal gasifier. The tests were made at 1900°F reactor temperature and 1000 psig pressure using an IGT hydrogasified high-volatile bituminous char. Minor modifications were made in the feed system to control fines entrainment.

Nominal test conditions were —

Run No.	Char Feed Rate, lb/hr	Steam Feed Rate, lb/hr	Reactor Temp, °F	Reactor Press., psig	Power Input, kW	Overall Resist., ohm	Steam Conver, wt %	Char Conver, wt %
EG-46	70	105	1902	1036	74.5	0.72	65	48
EG-47	74	138	1916	1003	64.1	[illegible]	62	44
EG-48	80	120	1906	1010	67.2	0.82	66	41

Run EG-46 was the first test at 1000 psig in which reactor temperatures in excess of 1900°F were attained. The reactor heat-up was very smooth and progressed at a rate of 1000°F/hr. The test was terminated after 1-1/2 hr of steady-state operation when the supply of feed char was exhausted. No complications were encountered with the feed screw following the installation of a continuous nitrogen purge. Considerable fines entrainment was encountered, making it unnecessary to discharge solids from the bottom of the reactor. Preliminary data from the test indicate nominal char gasification of 48% by weight, steam decomposition of 65% by weight, and a steady-state power input of 73.5 kW, accompanied by an overall resistance of 0.62 ohm.

Operating conditions for Run EG-47 were identical to the last run except the steam feed rate was increased in order to achieve a higher char conversion. A 1-1/4-hr steady-state operating period was obtained during which fines entrainment was again noticeable. Initial data indicate only a slightly higher char conversion of 49% by weight. Steam conversion was 52% by weight; an overall resistance of 1.29 ohm requiring an average power input of 63.1 kW was observed. A Vycor feed tube was extended into the reactor in an attempt to eliminate the fines entrainment (Figure 6).

The longest steady-state operating period was reached in Run EG-48. It was terminated after 2-1/2 hr of operation during which time the bed height was controlled by bottom discharge. Inspection of the quench water system after the test indicated some particle carry-over took place. Steam conversion was estimated at 66% and char conversion at 41% by weight. The average power input during steady state was 67.2 kW and the overall resistance was 0.82 ohm.

After Run EG-48 the Type-316 stainless steel electrode was removed from the reactor. The only physical wear on the electrode after six tests was a thin scale on its surface (Figure 7).

A silicon carbide tube is being installed in the reactor for the next test (Figures 6 and 8). Electrical properties of the material will be determined using a concentric electrode configuration.

The completed data and results of Runs EG-46, EG-47, and EG-48 will be reported as soon as they become available.

FUEL CELL ENGINEERING STUDY

Objective

The fuel cell engineering study was completed this month. This final report summarizes the results of a multimegawatt molten carbonate power

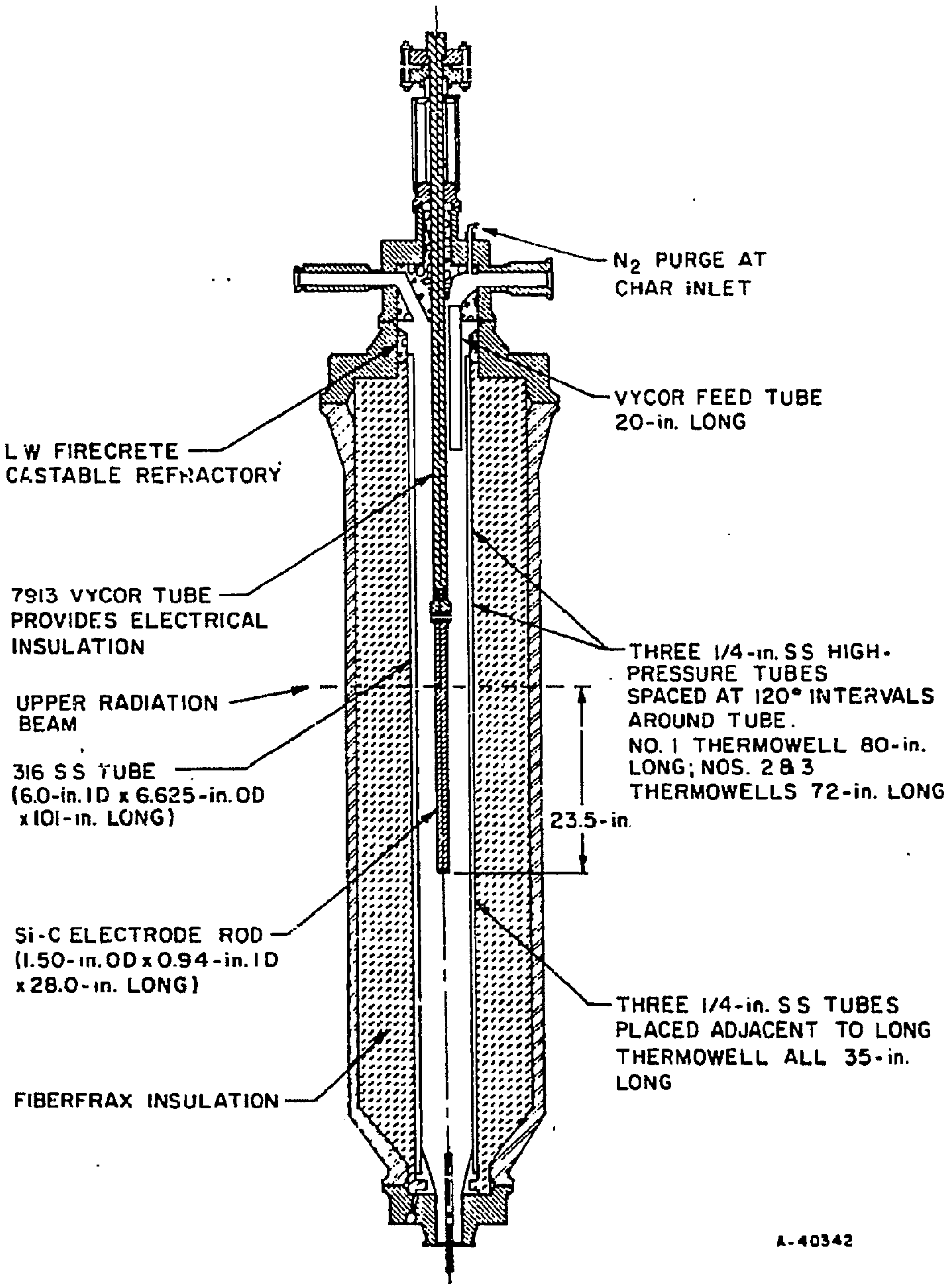


Figure 6. ELECTRODE CONFIGURATION FOR RUN EG-49

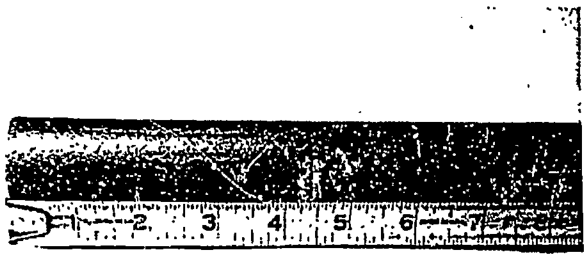


Figure 7. 316 SS ELECTRODE AFTER 6 RUNS, EG-43 TO EG-48

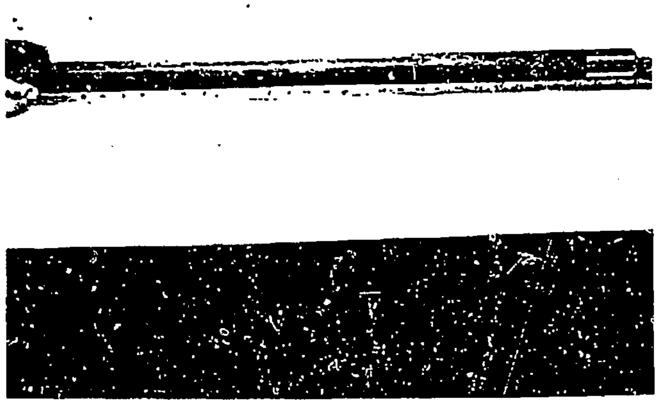


Figure 8. SILICON CARBIDE TUBE
(1.50-in. OD x 0.94-in. ID x 28.0-in. long)

plant design. This is a first round design and no attempt was made to optimize it.

During October 1969, work started on the engineering study of a molten carbonate fuel cell power plant which will supply 400 MW of d-c electrical energy. The objective of the study is to answer the question: Can we scale up a molten carbonate fuel cell power plant to multimegawatt capacity from the viewpoints of engineering, hardware, and costs?

Fuel Cell-HYGAS Plant Concept

The power plant is part of the coal-to-pipeline gas HYGAS plant. An integrated flow diagram is shown in Figure 9. The coal char exhaust is used to supply a producer-gas fuel for the fuel cell. In turn the electrical energy from the fuel cell powers the electrothermal gasifier unit of the HYGAS plant. The waste heat of the fuel cell can be recovered in two ways: steam generation and added electrical energy generated by prime movers. Both of these methods were studied. The results of the power plant study reported here are based on the latter method (Option B in Figure 9). We feel that this mixed fuel cell-prime mover power plant more clearly exposes the full potential and problems of a multimegawatt, molten carbonate fuel cell power plant.

Power Requirement and Design Base Points

The electrical power requirement of the electrothermal gasifier of the HYGAS plant and the fuel cell design base points are given in Table 6.

Table 6. POWER REQUIREMENT AND FUEL CELL DESIGN POINTS

Electrothermal Gasifier	
Power Requirement, MW	400
Voltage Requirement, V	5000
Single Cell	
Voltage Design Point, V	0.75
Current-Density Design Point I, A/sq ft	200
Current-Density Design Point II, A/sq ft	400
Fuel Cell	
Power-Density Design Point I, W/sq ft	150
Power-Density Design Point II, W/sq ft	300
Active Area of Single Cell, sq ft	10

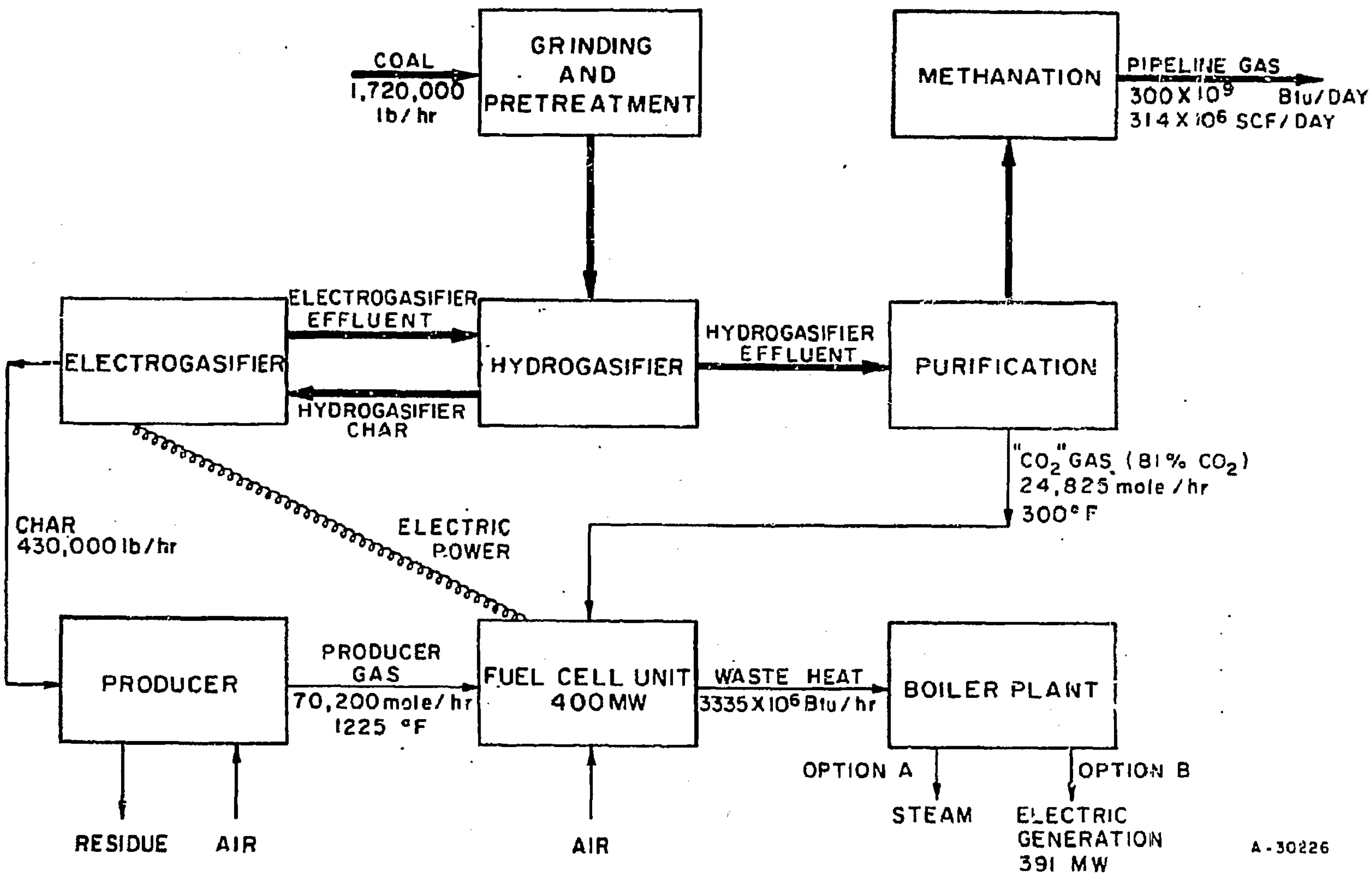


Figure 9. OVERALL BLOCK FLOW DIAGRAM FOR HYGAS PROCESS INTEGRATED WITH FUEL CELL POWER PLANT

The design calculations were based on two performance design points of the single fuel cell. The first design point of 0.75 volt at 200 A/sq ft (power density of 150 W/sq ft) can be achieved with present molten carbonate fuel cell technology. Based on the fuel and oxidant gas compositions and the plant operating conditions we feel that the fuel cell performance can reasonably be extrapolated to Design Point II, i.e., 0.75 volt at 400 A/sq ft (power density of 300 W/sq ft).

The single-cell design and costs are significant factors in the power plant design. The total costs of all the single cells could amount to 25-40% of the total capital investment cost of the power plant. Several detailed designs of the single cell were considered. A 10-sq-ft cell was chosen for this design study. This fuel-cell-size design point is of crucial importance; it must be confirmed in a future technical testing program.

Power Plant Specifications and Performance

The specifications and performance of the molten carbonate-prime-mover mixed generating plant are given in Table 7.

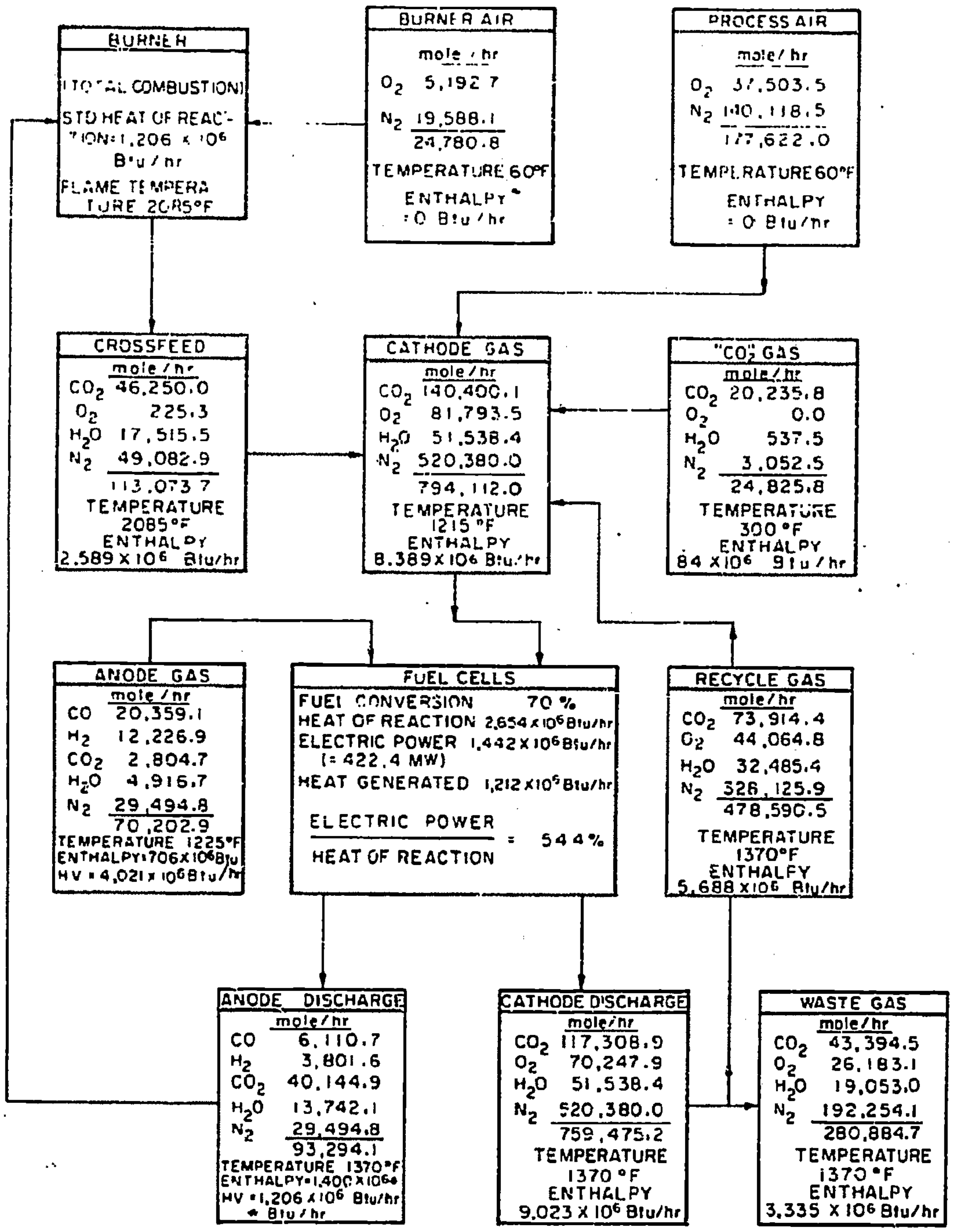
Table 7. OVERALL POWER PLANT SPECIFICATIONS AND PERFORMANCE

Fuel Cell Plant	
Gross Power, MW	422
Parasitic Power Requirement, MW	20
Net Power, MW	402
Power of Prime-Mover Waste-Heat Recovery Plant, MW	391
Total Power of Mixed Fuel Cell-Prime Mover Plant, MW	793
Total No. of Single Fuel Cells	281,600
No. of Cells in Basic Fuel Cell Power Package	800
No. of Basic Power Packages in Fuel Cell Plant	352
Specific Volume of Basic Fuel Cell Power Package, cu ft/kW	0.75
Specific Weight of Basic Fuel Cell Power Package, lb/kW	39

Overall Thermal Efficiency of Mixed Fuel Cell-Prime Mover Power Plant =

$$\frac{\text{Electrical Energy Output}}{\text{HHV of Fuel + Sensible Heat}} = 56\%$$

The overall material and energy balances of the fuel cell power plant are shown in Figure 10. The gas flow rates and heat fluxes at various points of the plant were used for sizing and heat transfer calculations of the components of the fuel cell stack.



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NOTE ENERGY BALANCES BASED ON STD STATE AT 60°F, WATER AS LIQUID

Figure 10. MATERIAL AND ENERGY BALANCES FOR 400-MW COAL-FUELED MOLTEN CARBONATE FUEL CELL POWER PLANT

The total of 280,000 single cells is subdivided into smaller groupings to render the cell-stacking problem more manageable. The basic power package is made up of 800 cells in 8 batteries of 100 each. The power plant layout of the 352 power packages is shown in Figure 11. The specific volume and weight of the power package (based on 150 W/sq ft performance) are 0.75 cu ft/kW and 39 lb/kW. The power plant layout in Figure 11 does not include the waste heat recovery-generator subsystem. The fuel cell power plant output is 422 MW, of which 20 MW is required as parasitic power, mainly for the gas blowers. The combined output of the fuel cell and the prime mover generator is 793 MW.

The overall thermal efficiency of the mixed fuel cell-prime mover plant is 56%. The fuel cell power plant alone has an efficiency of 28%, based on the higher heating value of the fuel gas and its sensible heat. If we neglect the sensible heat, the fuel cell's efficiency becomes 38%.

Basic Fuel Cell Power Package Cost Estimation

A cost factor method was used to estimate the purchased equipment cost (PEC) of the basic fuel cell power package. The detailed design of each component of the fuel cell stack was the basis for calculating the raw material weight of that component. Then the raw material cost of each component was calculated. The total raw material cost of the component was multiplied by a cost factor to yield the purchased equipment cost. Cost factors used for the various components of the fuel cell stack are shown in Table 8. Initially, the first-round set of cost factors, F_1, was used. Subsequently, several of the cost factors were reevaluated and modified. The second set of cost factors, F_2, was used in the results of the capital costs reported here.

Table 9 is the estimated PEC of the basic power package at design points of 150 and 300 W/sq ft. The PEC in dollars per kilowatt are $74.71 and $45.35. Note that the unit cell components amount to about three-fourths of the total package cost in both cases.

Total Capital Investment of the Power Plant

Tables 10 and 11 show the total capital investment costs (TCI) of the fuel cell-prime mover generating plant at fuel cell ratings of 150 and 300 W/sq ft. The costs in dollars per kilowatt in the fuel cell plant column are based on the net power produced by the fuel cell only (402 MW). The turbine plant

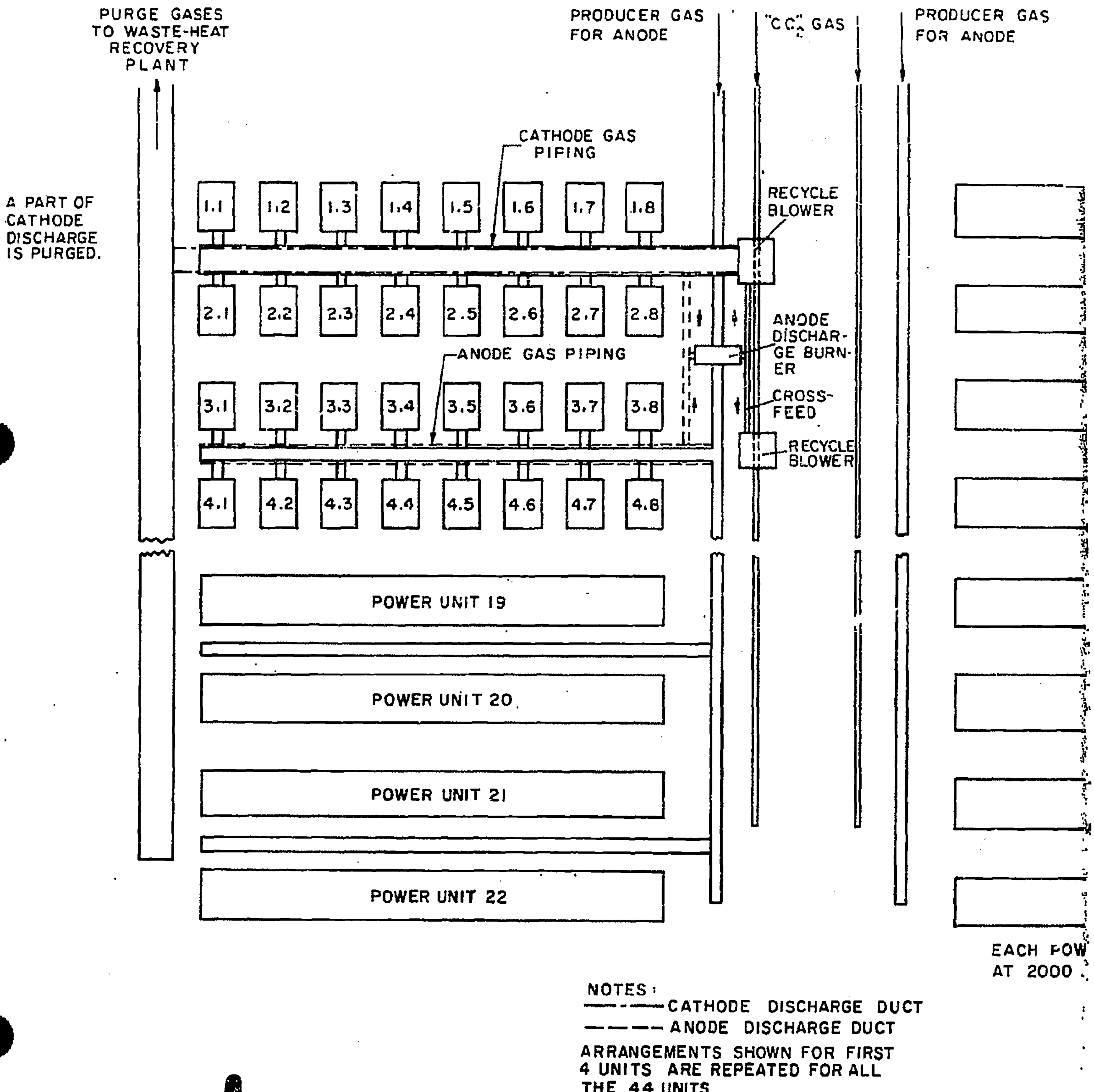

Figure 11. SIMPLIFIED LAYOUT OF POWER UNITS

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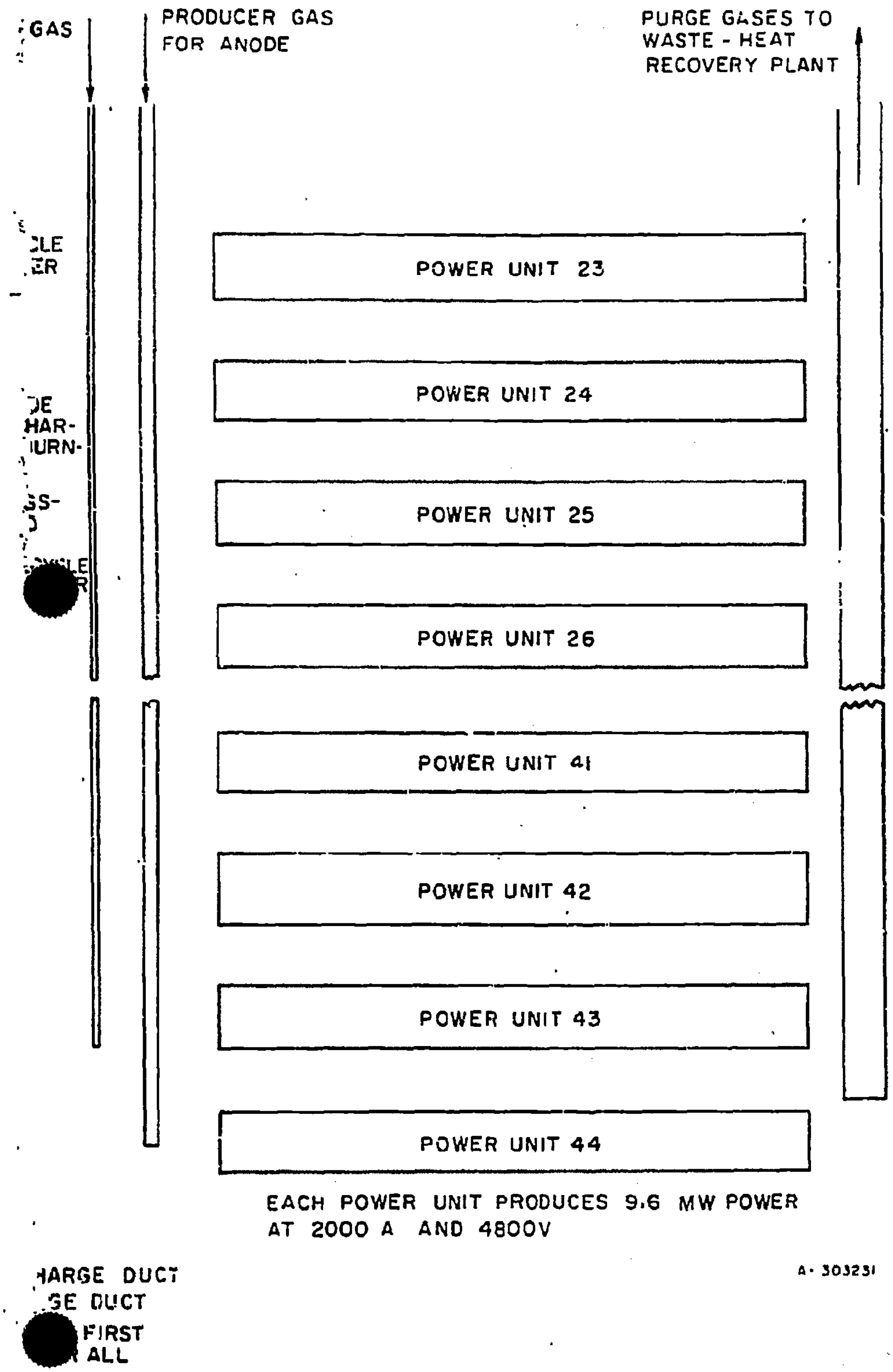


OF POWER UNITS

Table 8. COST FACTORS (F) FOR OBTAINING PURCHASED EQUIPMENT COST (PEC) OF INDIVIDUAL COMPONENTS OF POWER PACKAGE

PEC = (Component wt, lb) X (Cost, $/lb) X (Cost Factor, F)

Description	Material	First-Round Factors, F_1	Second-Round Factors, F_2
Cell Components			
Anode	Ni felt	4.0	8.0
Cathode	Ni felt	4.0	8.0
Electrolyte	Li-Na-K:CO_3 + Al_2O_3	2.0	5.0
Distributor Plate	SS 316 L	1.5	2.0
Module Components			
Spacer (Heat Transfer Gap)	SS 316 L	1.5	2.0
Battery Components			
End Plate	316 SS	1.5	1.5
Tie Rod	316 SS	1.5	1.5
Casing Components			
Gas Stream Separators	Alumina	2.0	5.0
Current Collectors	316 SS	1.5	1.5
Casing Inner Lining	Fireclay	2.0	5.0
Casing Shell	Mild Steel	2.0	5.0
Thermal Insulation	Kaolin Wool	2.0	5.0

Note: Estimates of subsequent tables are all based on second-round cost factors, F_2. Final results for F_1 are reported in parentheses at the end of subsequent tables.

Table 9. PURCHASE EQUIPMENT COST (PEC) FOR BASIC FUEL CELL POWER PACKAGE

	Power Density, W/sq ft			
	150		300	
Description	Weight, lb/pkg	PEC, \$/pkg	Weight, lb/pkg	PEC, \$/pkg
Unit Cell	31,496	64,540	32,625	76,400
Module Components	2,570	6,480	3,740	9,680
Battery Components	1,307	2,960	1,307	2,960
Gas Stream Separators	739	225	1,110	340
Current Collectors	792	1,400	1,190	2,100
Casing Shell, etc.	7,820	4,790	11,730	7,200
Miscellaneous	--	5,000	--	5,000
Total per Package	44,724	85,395	51,702	103,680
Net Power Produced, kW/pkg	1,143		2,286	
Weight, lb/kW	39.2		22.6	
PEC, \$/kW		74.71		45.35
Result for Cost Factors F_1, PEC and \$/kW		(47.21)		(28.0)

Table 10. TOTAL CAPITAL INVESTMENT FOR THE POWER PLANT

Fuel Cell Rating = 150 W/sq ft

Item	Basis of Investment per kW: Fuel Cell Plant, $/kW	Turbine Plant, $/kW	Total Generation, $/kW
Purchased Equipment Cost of Fuel Cell Package	74.71		
Installation (15% PEC)	11.21		
Subtotal	85.92		
Duct Work	11.30		
Electrical Connections and Protection	4.48		
Blowers	8.00		
Burners	1.10		
Instrumentation and Control	1.50		
Land Structures and Yard Improvements	5.00		
Subtotal, Fuel Cell Plant	117.30		
Boiler Plant		26.6	
Turbogenerator Plant		27.1	
Electrical Connections and Protection		5.5	
Miscellaneous Equipment		1.3	
Additional Land and Structures		4.5	
Subtotal, Turbine Plant		65.0	
Total Physical Cost (TPC)			91.44
Engineering and Construction (7% TPC)			6.40
Direct Plant Cost (DPC)			97.84
Contingency (10% DPC)			9.78
Interest During Construction, 7%/yr*			10.27
Fixed Capital Investment (FCI)			117.89
Working Capital			5.00
Total Capital Investment (TCI)			122.89
Total Capital Investment Based on F_1			(102.28)

* Charged for 50% of 3-year construction period.

Table 11. TOTAL CAPITAL INVESTMENT FOR THE POWER PLANT

Fuel Cell Rating = 300 W/sq ft

Item	Basis of Investment per kW: Fuel Cell Plant, $/kW	Turbine Plant, $/kW	Total Generation, $/kW
Purchased Equipment Cost of Fuel Cell Package	45.35		
Installation (15% PEC)	6.80		
Subtotal	52.15		
Duct Work	7.90		
Electrical Connections and Protection	4.48		
Blowers	8.00		
Burners	1.10		
Instrumentation and Control	1.50		
Land Structures and Yard Improvements	5.00		
Subtotal, Fuel Cell Plant	80.13		
Boiler Plant		26.6	
Turbogenerator Plant		27.1	
Electrical Connections and Protection		5.5	
Miscellaneous Equipment		1.3	
Additional Land and Structures		4.5	
Subtotal, Turbine Plant		65.0	
Total Physical Cost (TPC)			72.64
Engineering and Construction (7% TPC)			5.08
Direct Plant Cost (DPC)			77.72
Contingency (10% DPC)			7.77
Interest During Construction, 7%/yr*			8.16
Fixed Capital Investment (FCI)			93.65
Working Capital			5.00
Total Capital Investment (TCI)			98.65
Total Capital Investment Based on F_1			(85.67)

* Charged for 50% of 3-year construction period.

column costs are similarly based on the 391 MW produced by the turbine. The total generation column and the final total capital investment results are based on the total power output of 793 MW of the mixed power plant. For the presently achievable fuel cell rating of 150 W/sq ft we calculated a total capital investment cost of $123/kW. If the improved rating of 300 W/sq ft is realized, the TCI will drop to $99/kW. Although these results are contingent upon our initial assumptions, some of which must be confirmed by testing, we are very much encouraged. The capital investment cost of $123/kW coupled with an overall efficiency of 56% makes the prospect of a multimegawatt molten carbonate fuel cell power plant very bright. At 300 W/sq ft, which we expect to achieve in the not too distant future, the TCI of $99/kW is 19% below the average TCI of $125/kW of a conventional steam generating plant at 40% efficiency.

Bus Bar Electrical Energy Costs

The bus bar electrical energy costs for fuel cell ratings of 150 and 300 W/sq ft (Table 12) are 5.386 and 4.516 mills/kWhr. The contribution of the different cost components for the 300 W/sq ft case are shown diagramatically in Figure 12. The producer-gas fuel cost amounts of 33.0% and the depreciation of the generating equipment is 22.0% of the bus bar energy cost.

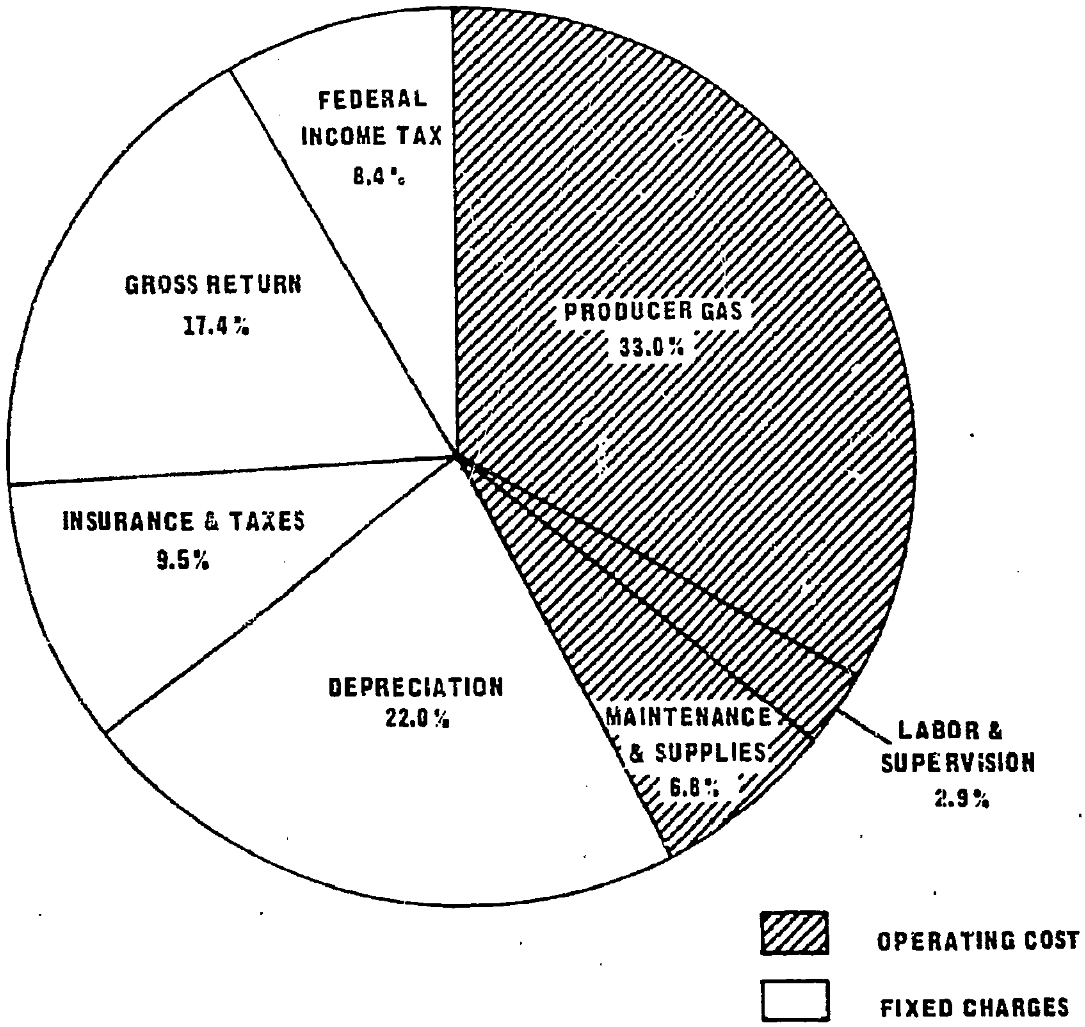
The data in Table 12 and Figure 12 set the producer-gas fuel cost at 25¢/ million Btu. The depreciation of the fuel cell generating plant is based on a straight-line method and a weighted average power plant life of 10 years. The assumed service life of the single-fuel-cell components is 3 years, while that for the other conventional components is 10-20 years.

Three parameters of the fuel cell power plant design which may be largely uncertain are the producer-gas cost, the installed cost of the fuel cell stack, and the service life of the plant components. In Figures 13, 14, and 15 these three parameters are plotted against the bus bar energy cost. Figure 13 shows that for the 150 W/sq ft case a fuel cost change of 10¢/million Btu will result in an 11% change in the bus bar electrical energy cost. A similar change in fuel cost for the 300 W/sq ft case will produce a bus bar energy cost change of 13%. The bus bar energy cost change is linear with respect to the fuel cost change.

Table 12. ENERGY COST FOR KILOWATT-HOUR PRODUCED* IN COMBINED FUEL CELL TURBINE PLANT

	Power Density, W/sq ft 150	300
	mills/kWhr	
1. Raw Material (Anode Gas) at 25¢/10^6 Btu	1.494	1.494
2. Labor and Supervision Cost	0.129	0.129
3. Maintenance (3% DPC)	0.335	0.266
4. Material Supplies (15% Maintenance)	0.050	0.040
5. Operating Cost	2.008	1.929
6. Depreciation	1.387	0.993
7. Insurance and Local Taxes (4% FCI)	0.538	0.428
8. Gross Return (7% TCI)	0.982	0.788
9. Federal Income Tax (48% Gross Return)	0.471	0.378
10. Fixed Charges (Nos. 6 to 9)	3.378	2.587
11. Energy Cost (Operating Cost Plus Fixed Charges)	5.386	4.516
12. Energy Cost Based on F_1	(4.506)	(3.967)

* 100% load factor is assumed for these estimates.



POWER DENSITY, W/sq ft	300
PRODUCER GAS, ¢/10^6 Btu	25
BUS BAR ENERGY COST mills/kWhr	4.516

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Figure 12. COMPONENTS OF BUS BAR ENERGY COST

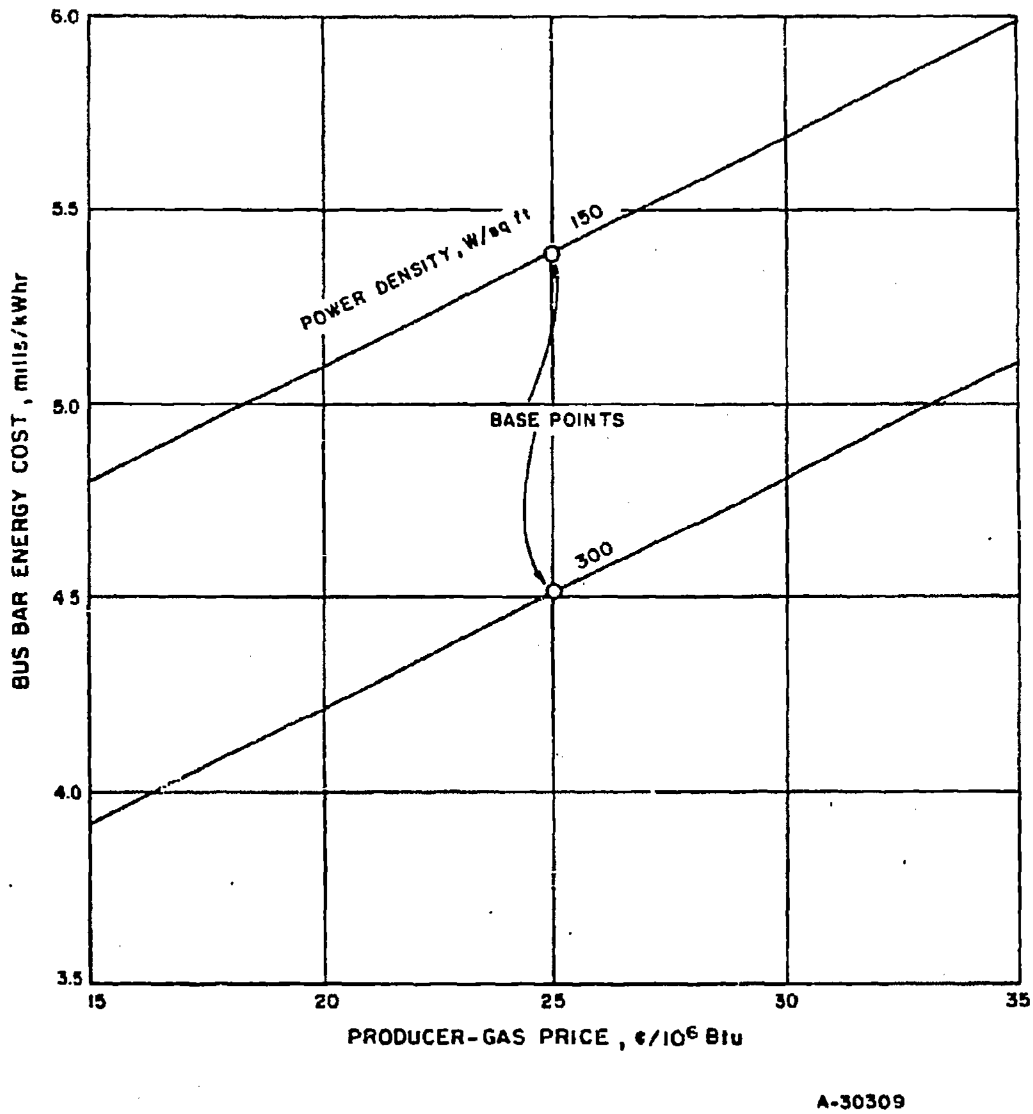

Figure 13. VARIATION OF BUS BAR ENERGY COST AS A FUNCTION OF GAS PRICE AND POWER DENSITY

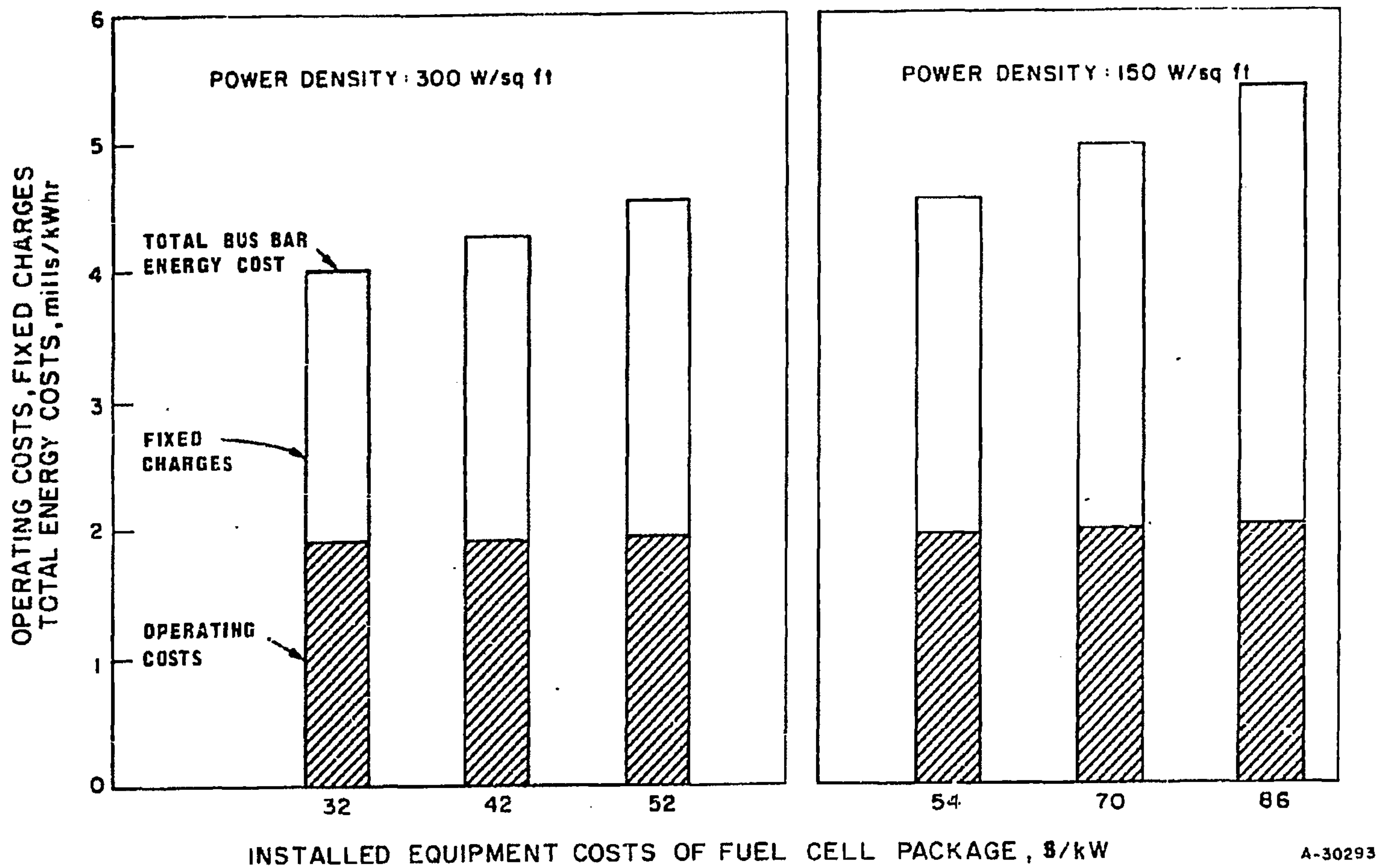

Figure 14. VARIATION OF BUS BAR ENERGY COST AS A FUNCTION OF PACKAGE COST

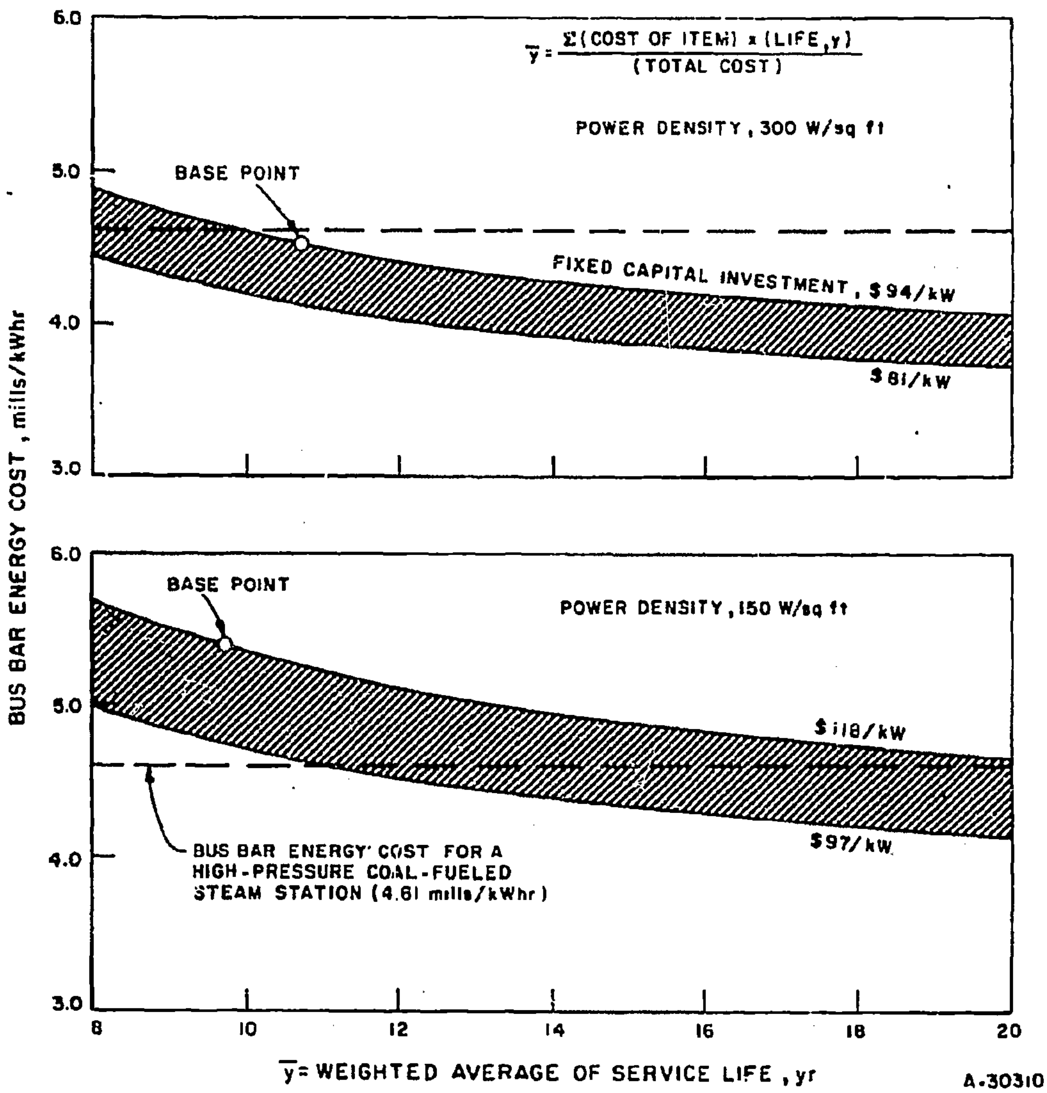

Figure 15. VARIATION OF BUS BAR ENERGY COST AS A FUNCTION OF SERVICE LIFE AND CAPITAL INVESTMENT

In the bar chart (Figure 14) the bus bar energy cost is plotted against the installed equipment cost of the basic fuel cell power package. For both power ratings of the fuel cell the bus bar electrical energy cost appears to be fairly insensitive to the installed equipment cost of the fuel cell power package. For an equipment cost change of $10/kW the bus bar electrical energy cost changes about 0.3 to 0.4 mill/kWhr for the 150 and 300 W/sq ft ratings.

Figure 15 shows the bus bar electrical energy cost as a function of the weighted average service life of the fuel cell power plant. In both the 150 and 300 W/sq ft cases, the horizontal dotted line is the bus bar energy cost of 4.61 mills/kWhr for a high-pressure, coal-fueled, steam central station. The broad-curve area indicates the possible range of the fixed capital cost of the plant equipment. For a 150 W/sq ft rated fuel cell, the weighted average service life must be more than 11 years before the bus bar energy achieved parity with the reference cost of 4.61 mills/kWhr. At the high, fixed capital cost range of $118/kW, parity is not attained until we have a weighted average life of 20 years. However, for a 300 W/sq ft rated fuel cell, equivalence with the 4.61 mills/kWhr reference is reached at 10 years service life, even for the high-range fixed cost of $94/kW. At all points beyond the 10-year weighted average service life, the bus bar energy cost is significantly below the reference cost. At 20 years, the fuel cell bus bar energy cost ranges from 3.7 to 4.0 mills/kWhr.

Problem Areas and Future Work

In the course of this engineering study we compiled a list of the problem areas in scaling up a molten carbonate fuel cell power plant to multimegawatt capacity. Some of the most significant problem areas are —

1. Mechanical limit of size and configuration of the electrolyte brick and study of techniques for its manufacture
2. Material technology testing of cell components
3. Technology testing of cell performance of fuel cell using anode and cathode gas of the specified composition
4. Optimization of recycle ratios in relation to fixed capital and operating cost
5. Determination of response of cell performance to fuel variation and investigation of allowable flow variations from cell to cell

6. Experimental determination of heat transfer coefficients, pressure drops, and conductivities under operating conditions

7. Prediction of performance under nonuniform temperature. Estimation of maximum tolerable temperature gradients and their effect on the life of the cell.

These and other problem areas can form the bases for the work tasks of a technology testing program.

PILOT PLANT CONSTRUCTION

Engineering

A major portion of the engineering effort during this report period has been instrumentation and related details. Electrical and instrument detail drawings are complete except for the process analyzer drawings. The emergency power system is being reexamined by IGT and Procon. Total project detailed design and drafting is 97% complete. Final work on the model is being done at the site.

Procurement

The instrument electrical subcontract price and the insulation subcontract bids will be received shortly. All major equipment and materials scheduled to be received have arrived, with the following exceptions: a portion of the switchgear and a compressor motor, which were delayed by the General Electric Company strike, and the pretreator reactor, several small vessels, and a portion of the material-handling equipment, which will arrive late due to recent design changes and additions. The above delayed deliveries can be worked into the construction schedule without extending the completion date.

We have anticipated the possible delayed delivery of carbon steel and alloy shop-fabricated pipe, which would directly affect the construction schedule. Promised deliveries of April 1 appear to be holding firm with the exception of some of the heavy wall pipe.

Construction

The reactor was erected on February 21. Major activities since that date have been the setting of equipment, erection of structural steel, platforms, and ladders, field pipe shop fabrication, and some area piping. All buildings have been erected except for the slurry filter building.

There are now 40 pipefitter/welders on the project, representing an increase of 10 during this report period. It is not known if it will be possible to man the job at the level required to meet the scheduled completion date.

We have experienced a total of 12 inclement weather days, 2 of which occurred in this report period. On these days no significant outside progress was made.

Figures 16-19 show the status of plant construction.

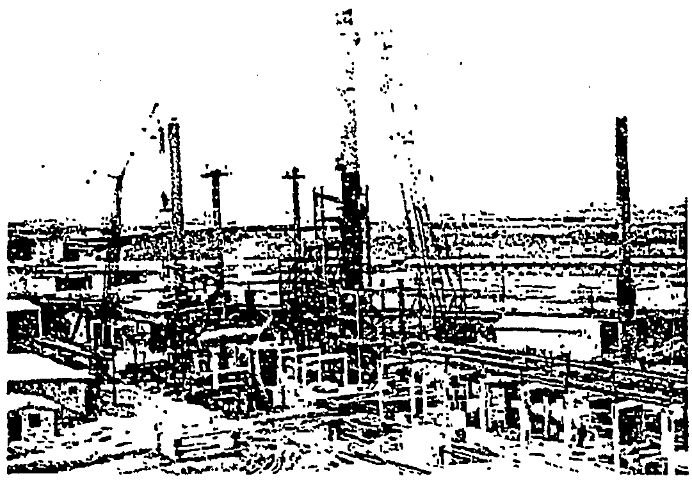

Figure 16. OVERALL PLANT — LOOKING SOUTHEAST

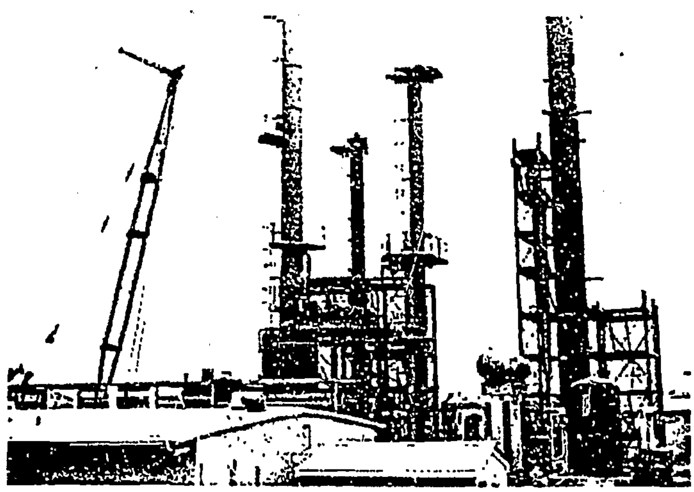

Figure 17. PROCESS AREA — LOOKING SOUTH-SOUTHEAST

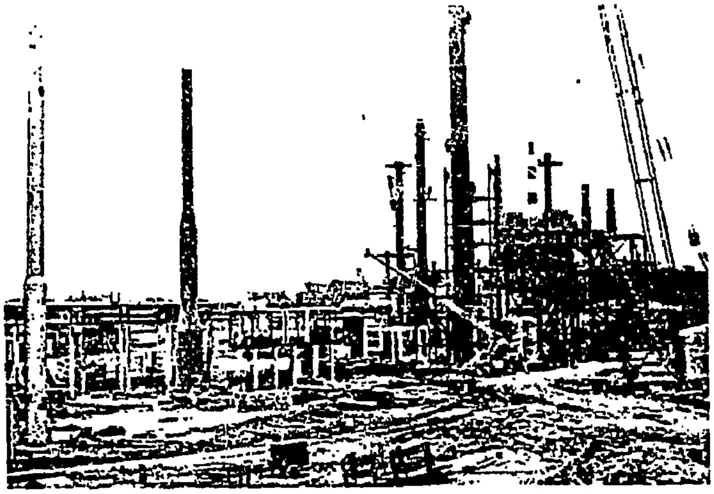

Figure 18. GAS CLEANUP AND PROCESS AREA – LOOKING NORTHEAST

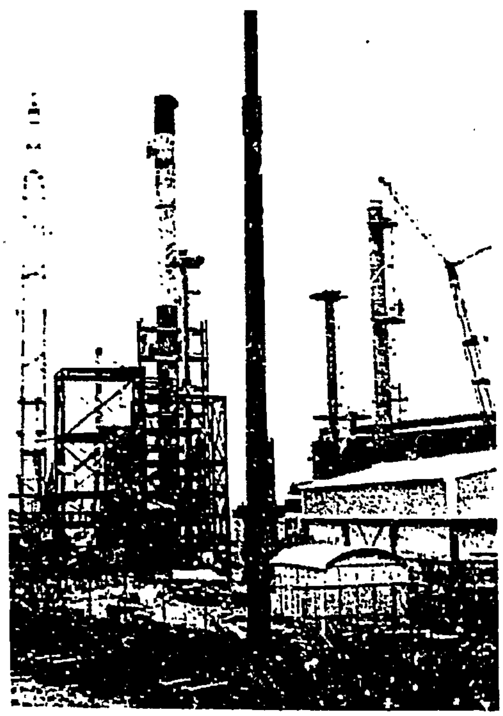

Figure 19. PROCESS AREA AND COMPRESSOR BUILDING – LOOKING WEST-NORTHWEST