PROGRESS REPORT NO. 20

COAL HYDROGASIFICATION PILOT PLANT

FOR

INSTITUTE OF GAS TECHNOLOGY

CHICAGO, ILLINOIS

W-1784

TABLE OF CONTENTS

PROGRESS REPORT

PROJECT:	Institute of Gas Technology Chicago, Illinois Coal Hydrogasification Pilot Plant Procon Job No. W-1784
REPORT NO:	20
DATE:	March 15, 1970
PROCON PROJECT MANAGER:	T. A. Taylor

DISTRIBUTION

INSTITUTE OF GAS TECHNOLOGY

Mr. F. C. Schora - 10

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Mr. C. J. Towle	1
Mr. T. A. Taylor	2
Field	1

W-1784

Coal Hydrogasification Pilot Plant

I. SUMMARY

Engineering	97%
Purchasing	89%
Material Receipt	86%
Construction	32%

A. ENGINEERING

A major portion of the engineering effort during this report period has been instrumentation related details. Electrical & instrument detail drawings are complete except for the process analyser drawings. The emergency power system is being re-examined by IGT and Procon.

Total project detailed design and drafting is 97 percent complete.

Final work on the model is being done at the site.

B. PROCUREMENT

The instrument electrical subcontract price will be received shortly, as will the insulation subcontract bids.

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.

B. PROCUREMENT (continued)

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

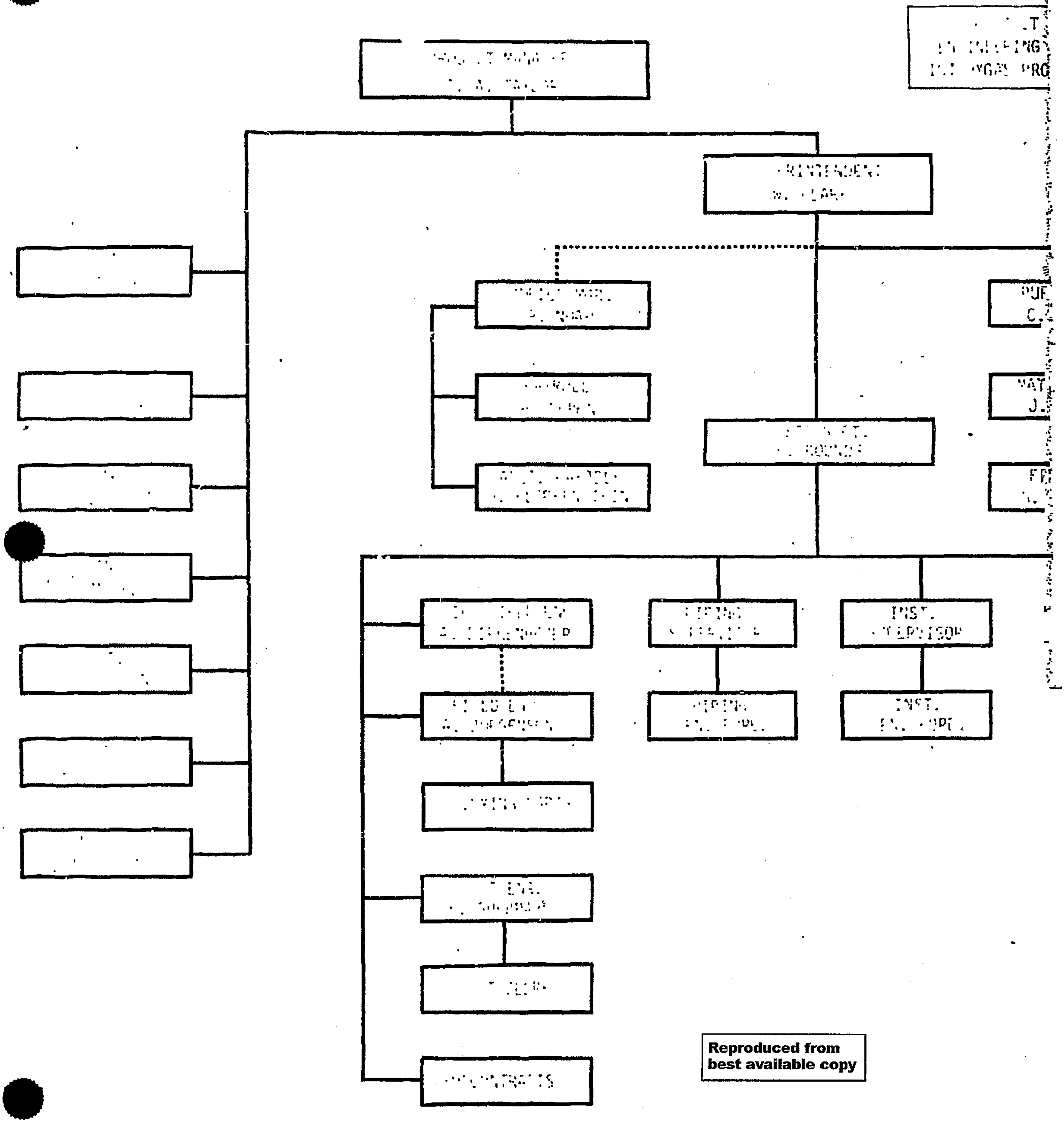
C. CONSTRUCTION

The reactor was erected on February 21st. Major activities since that date have been the setting of equipment, erection of structural steel and 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 forty pipefitter/welders on the project; represents an increase of ten during this report period. It is not yet 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 twelve inclimate weather days, 2 of which occurred in this report period. On these days no significant outside progress was made.

As requested by IGT, attached is an Engineering and Construction organizational chart for this project.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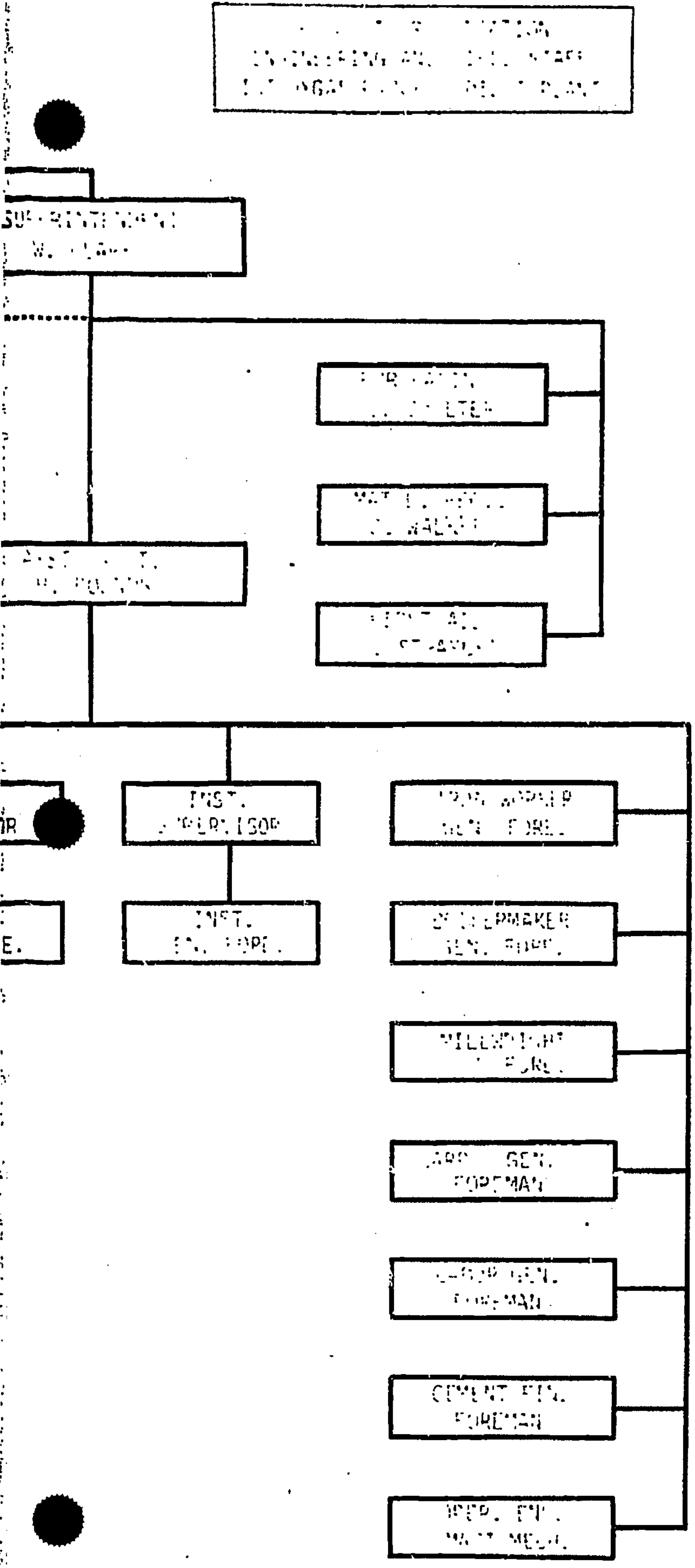
ENGINEERING AND ... STAFF
SUPERINTENDENT
W.
INST.
SUPERVISOR
IRON WORKER
GEN. FORE.
OR
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ENG. FORE.
E.
BOILERMAKER
GEN. FORE.
MILLWRIGHT
FORE.
CARP. GEN.
FOREMAN
LABOR GEN.
FOREMAN
CEMENT FIN.
FOREMAN
OPER. ENG.
MAST MECH.

2

II. SCHEDULE AND S-CURVE REPORT

The project schedule is presently under review and should be ready for discussion with IGT by April 1st.

The S-Curve Report has been updated to show present progress.

INSTITUTE OF GAS TEC

SCHEDULED ———

ACTUAL - - - - -

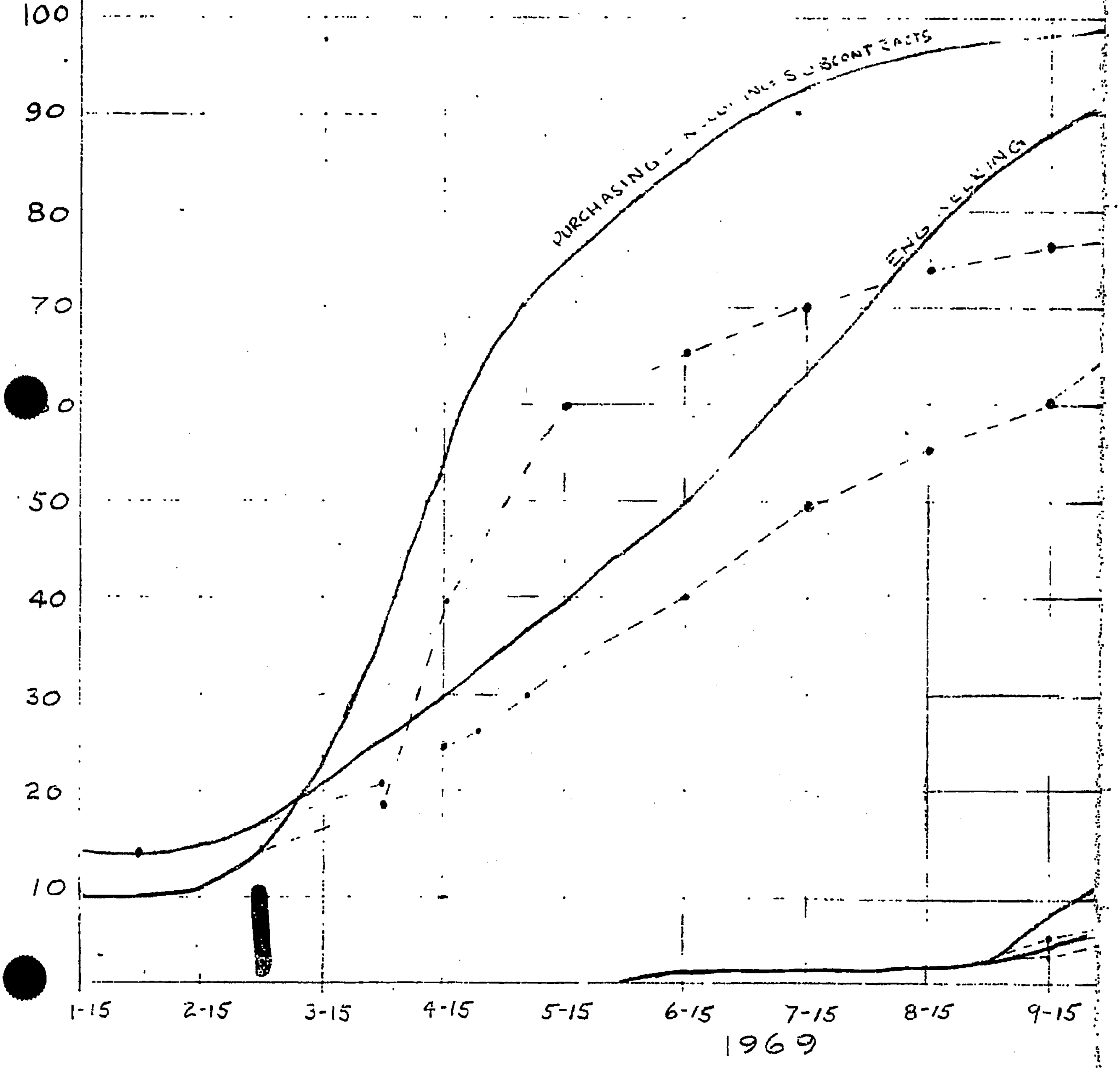
REPORT NO. – 20

DATE – 3-5-70

PERCENT COMPLETE:

ENGINEERING – 77

PURCHASING – 89



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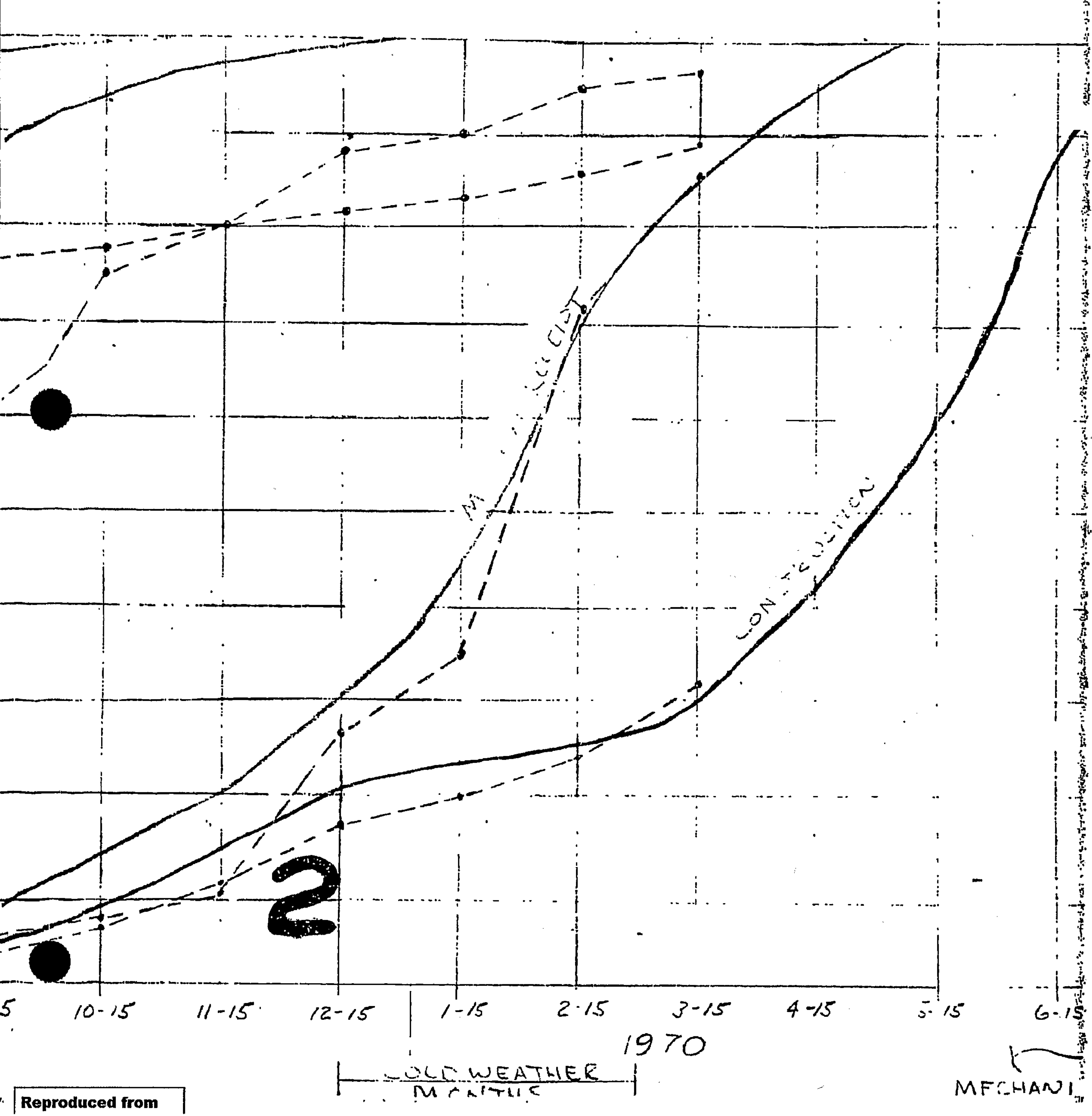
ECHNOLOGY - 1784

MATERIAL RECEIVED - 96
CONSTRUCTION - 32

REV 0 - 3-15-69
REV. 1 4-15-69
REV. 2 · 7-15-6
REV. 3 - 8-15-69
REV. 4 · 1-15-70 CONSTRUCTION
RECEIPT



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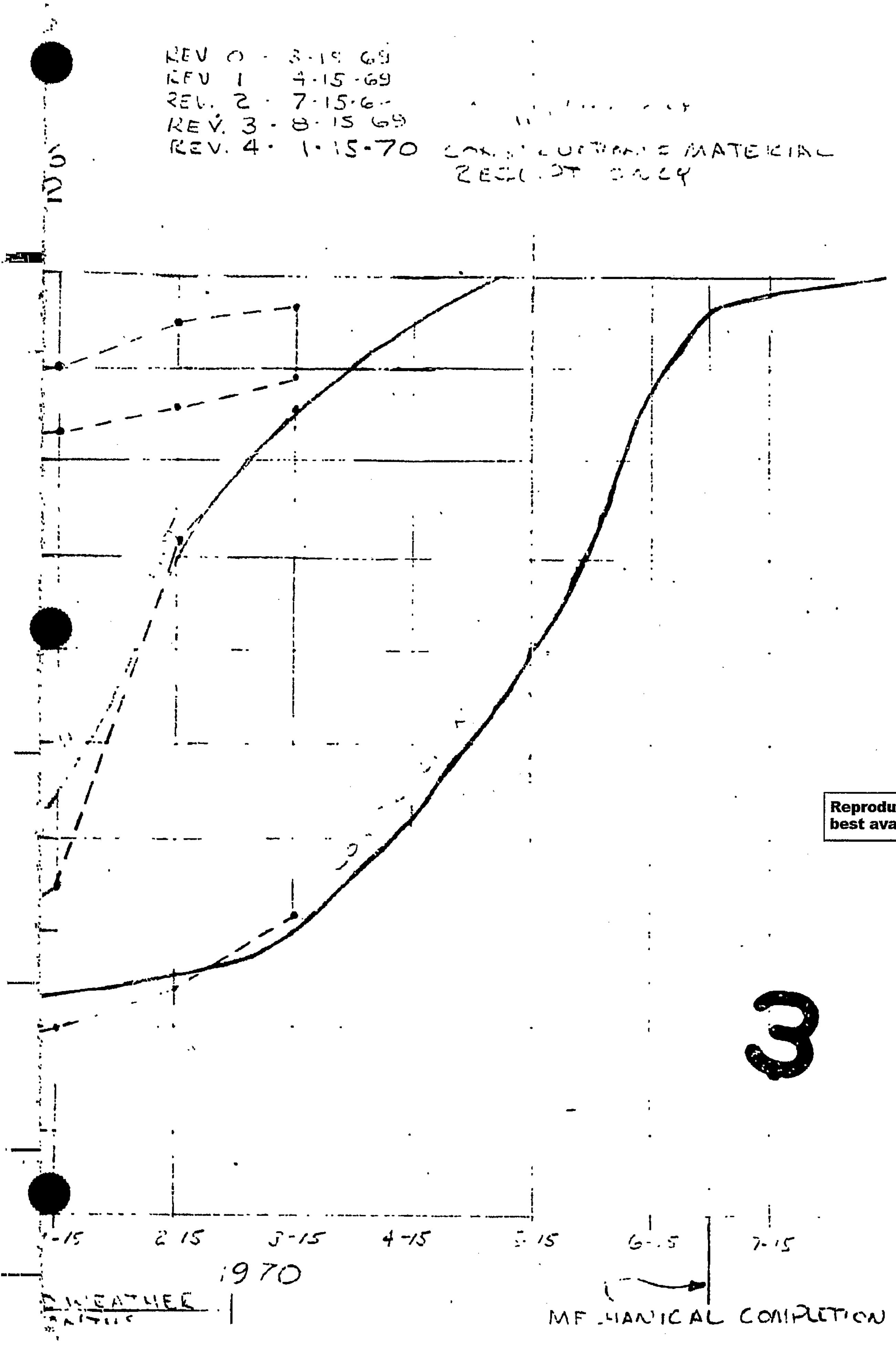

REV 0 · 3·15 69
REV 1 4·15·69
REV 2 · 7·15·6-
REV. 3 · 8·15 69
REV. 4 · 1·15·70 CONSTRUCTION & MATERIAL RECEIPT ONLY

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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 March 15, 1970

2 No. of Work Days

3 Contract No.

Form Approved
Budget Bureau
No. 60R0178
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

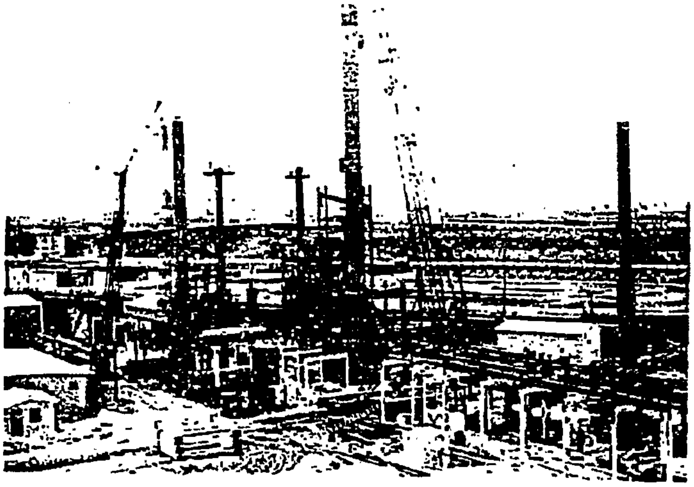
13 Payments Received $

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	$ 3,970	$ 800	$ 4,770				
Less: 10% Retention as applicable	392	80	472				
	$ 3,578	$ 720	$ 4,298				
(Above includes 66 Thousand Dollar 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."

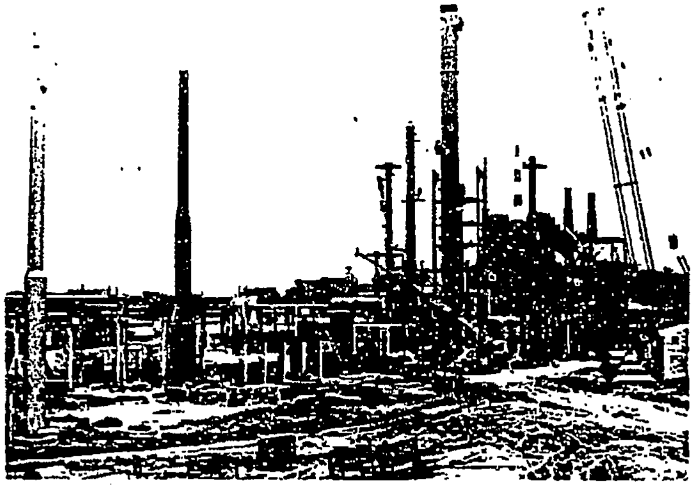
Progress Report No. 20, March 15, 1970

T. A. Taylor

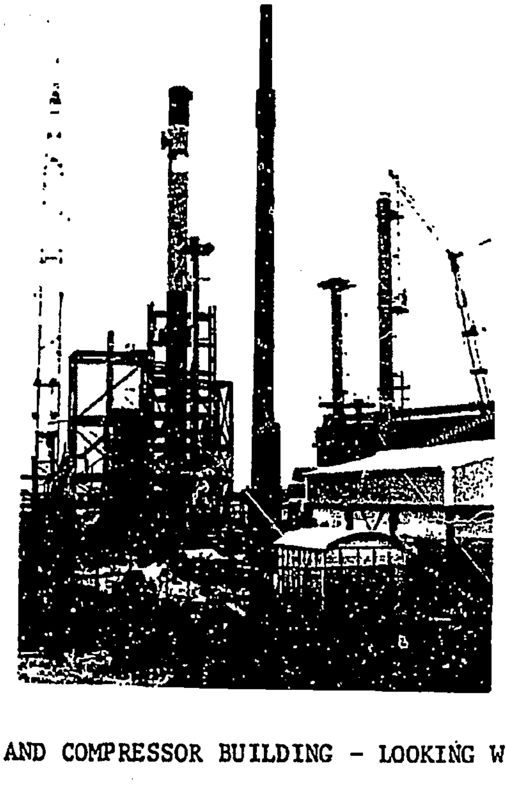


OVERALL PLANT - LOOKING SOUTHEAST

PROCESS AREA - LOOKING SOUTH-SOUTHEAST



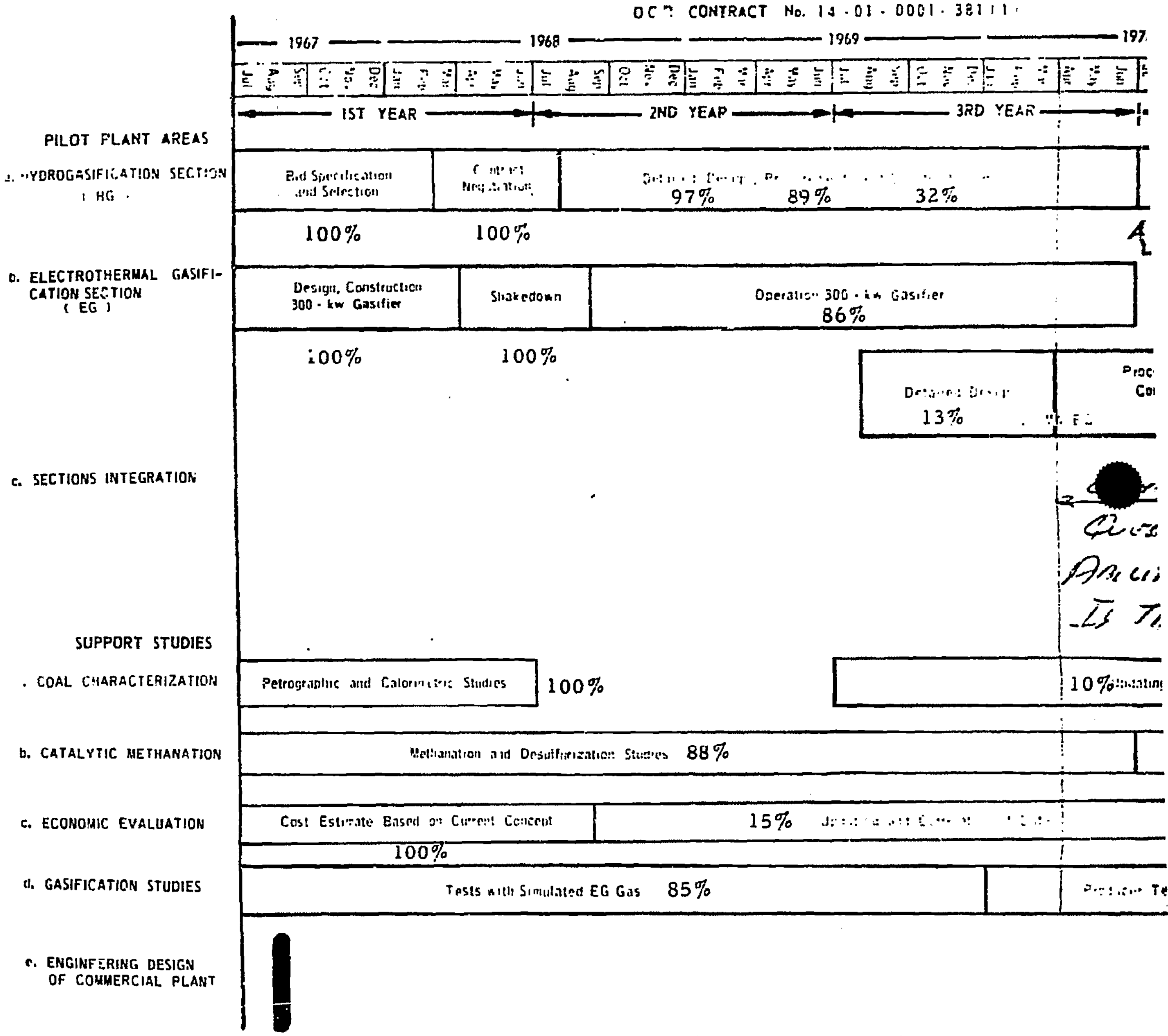
GAS CLEANUP AND PROCESS AREA - LOOKING NORTHEAST



PROCESS AREA AND COMPRESSOR BUILDING - LOOKING WEST-NORTHWEST

PILOT PLANT PROGRAM OF IGT HYDROGAS

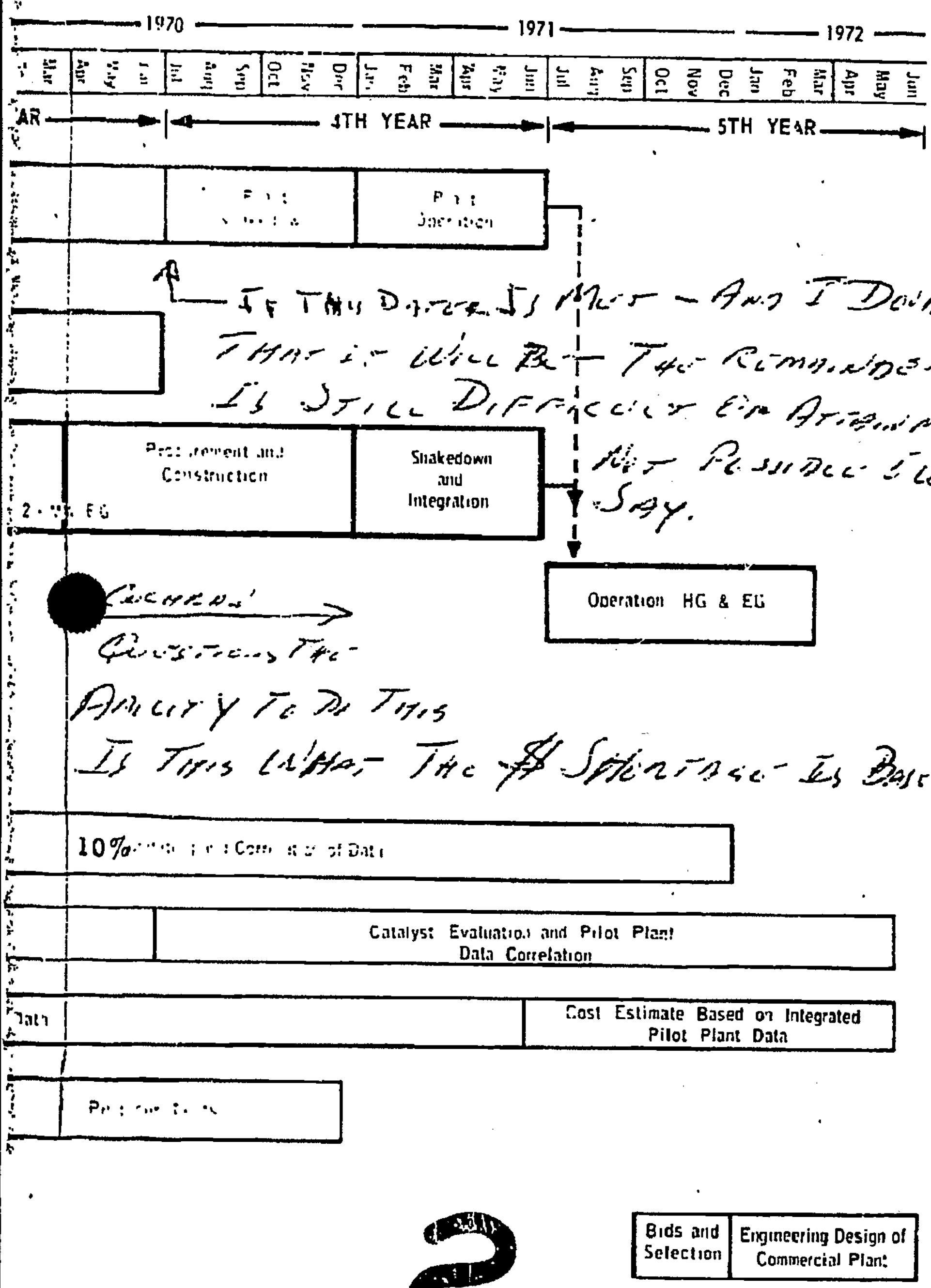
OCR CONTRACT No. 14-01-0001-38111



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

HYDROGASIFICATION PROCESS

AGA IU-4-1



If This Date Is Met – And I Doubt That It Will Be – The Remainder Is Still Difficult For Attainment. Not Possible I Would Say.

Cochran's → Questions The Ability To Do This

Is This What The $ Shortage Is Based On?

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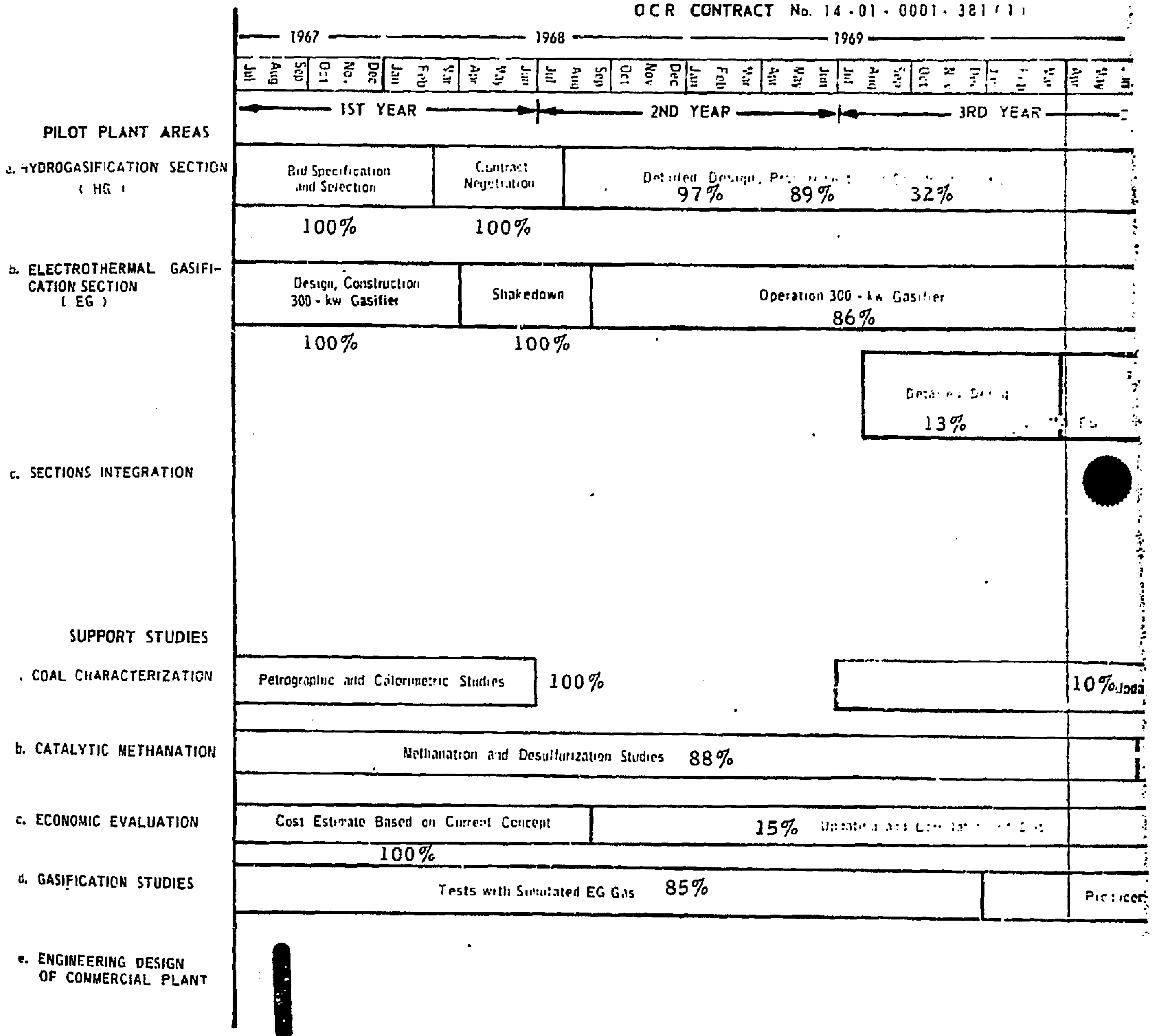
PILOT PLANT PROGRAM OF IGT HYDROG

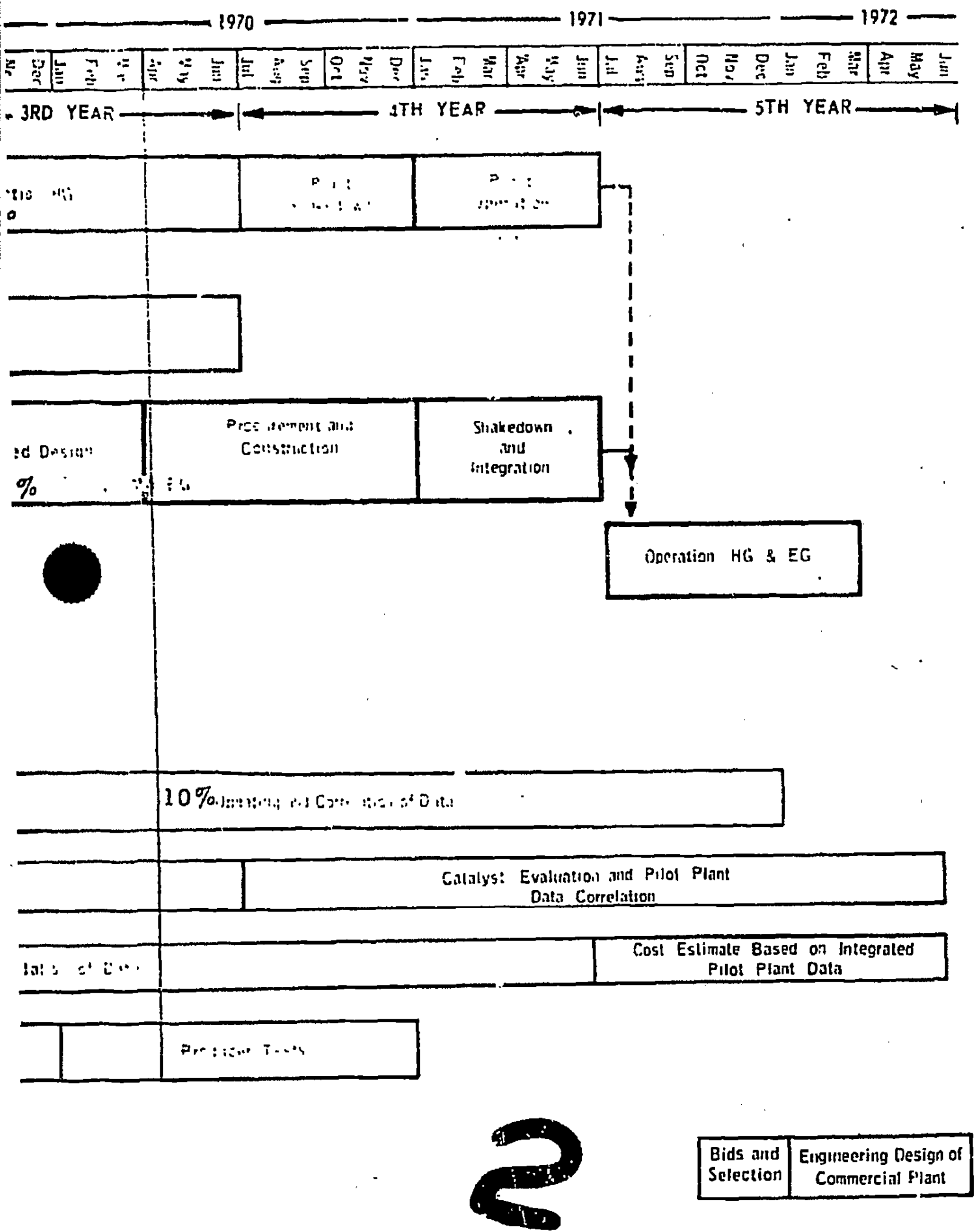
OCR CONTRACT No. 14-01-0001-381 (1)



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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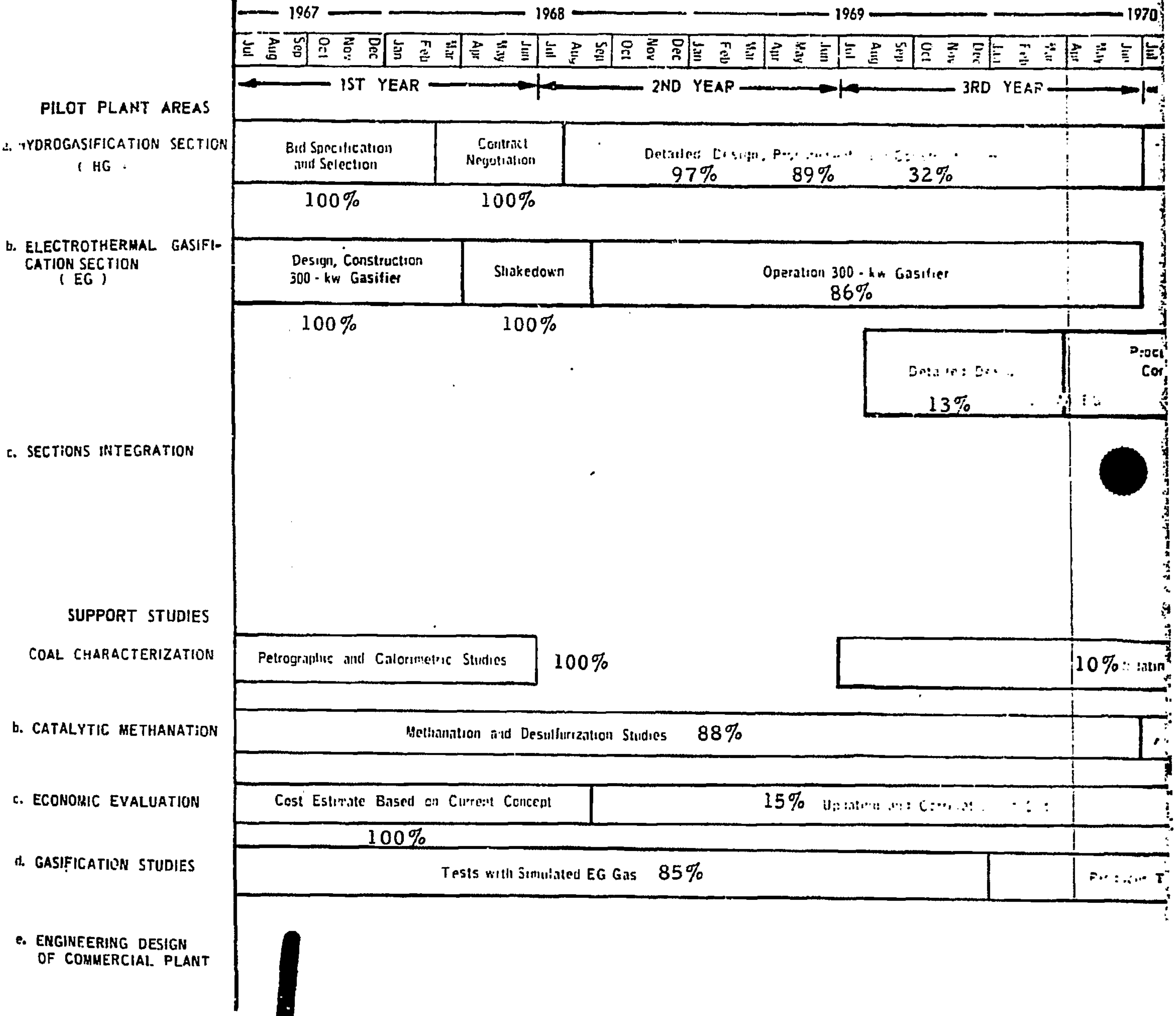
E - 1084

PILOT PLANT PROGRAM OF IGT HYDROGAS

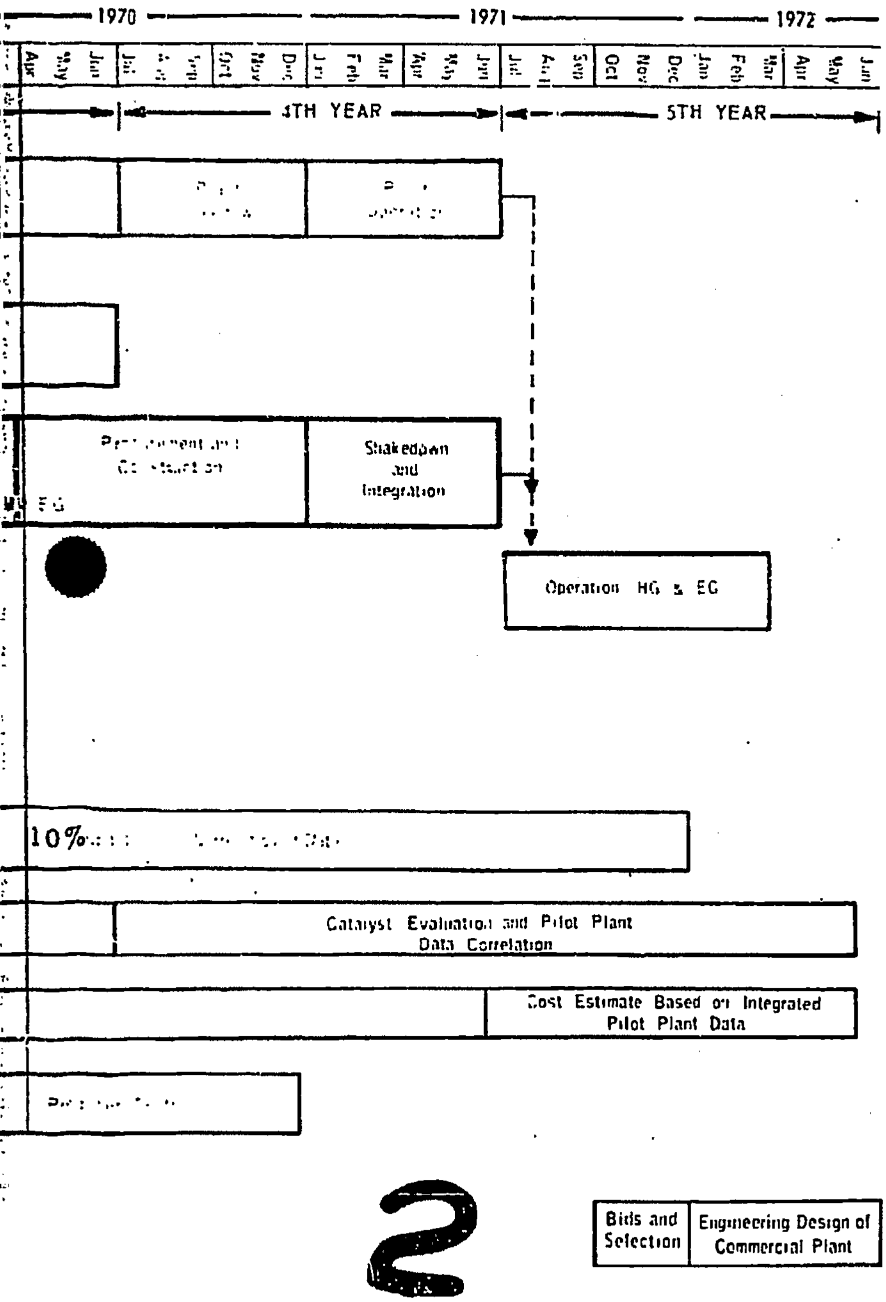
OCR CONTRACT No. 14-01-0001-381 (1)



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

ROGASIFICATION PROCESS

AGA IU-4-1



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IGT-MPR-- 4/70

IGT
EDUCATION · RESEARCH
INSTITUTE OF GAS TECHNOLOGY · IIT CENTER · CHICAGO 60616

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

Report For April 1970
OCR Report No. 67

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

Microtumbler tests on various hydrogasified chars were run to see if this method is suitable for quantifying attrition resistance.

HIGH-PRESSURE METHANATION

A series of tests with and without a Ni-Mo catalyst showed that with the catalyst ethylene hydrogenation begins at 600°F and is complete at 700°F or above. Without the catalyst, however, reaction did not occur below 800°F; hydrogenation occurred at 1100°F, accompanied by the formation of carbon and tar.

ENGINEERING ECONOMICS STUDIES

Further contact with a major supplier of air cooling equipment indicates that air cooling has real potential for large applications. Most new refineries do about 80% of their cooling by air.

DEVELOPMENT UNIT STUDIES

This month three tests were made in the hydrogasifier, two for producing char for the electrothermal gasifier. Tests with Montana subbituminous coal show that at 500 psig significantly lower methane and higher carbon oxides yields were obtained than at 1100 psig. This indicates that system pressures should be in the 1000-psi range to maximize methane formation in the hydrogasifier.

A silicon carbide tube tested as the central electrode proved to be physically brittle. A solid rod will be tried next. A run aimed at defining the electrical characteristics of the bed yielded much data. High- and low-frequency current and voltage fluctuations, along with transient response of the bed to step-changes in power input, should provide the necessary data for designing the power package of the 2-MW EG unit.

The nozzles from Spraying Systems Corp. appear to show no wear when dispersing coal-water slurries. Photographs were taken that should permit measurement of spray distribution.

PILOT PLANT CONSTRUCTION

Engineering is 98% complete and purchasing is 94% complete. The insulation subcontract was let. The instrument/electrical subcontract is being reviewed. Material receipt is 90% complete, with the delivery of shop-fabricated piping being delayed by 2 weeks. Field construction is 47% complete. Major field activities involve field piping and pipe fabrication, erection of platforms and ladders, area paving, and electrical work.

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)	$ 147,000

2. OCR Funding

A. Funds Expended This Month (estimated)	$ 630,000
B. Funds Expended Since Contract Amendment No. 1 (estimated)	$5,910,000

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
Jack Huebler
Vice-President

Appendix. Achievements in April

COAL CHARACTERIZATION

Several thermobalance tests for hydrogasification reactivity were run on samples of maceral concentrates obtained from Prof. W. Spackman of Pennsylvania State University. Results will be reported later.

The microtumbler test for attrition resistance was modified by substituting polyethylene balls for the steel balls used in a coal grinding test. Results are shown in Table 1. FMC char and residues from the hydrogasification of lignite are more resistant to attrition than residues from the hydrogasification of bituminous coal. Attrition resistance of the latter appears to increase with conversion. The results on the +30 sieve fraction do not necessarily show the true effect of increasing conversion; more highly reacted particles may break up in the process, leaving the more resistant particles including those with a high ash content. We plan to run microtumbler tests on other sieve fractions to investigate the subject more fully.

Work continued on ammonia and other contaminants in the quench-water circuit. Partial pressure-composition data for the ammonia-carbon dioxide-water system at 20°-60°C were compared with values calculated from ionization and Henry's law constants for solubilities of the individual gases. Poor agreement and some anomalies in the literature require further consideration.

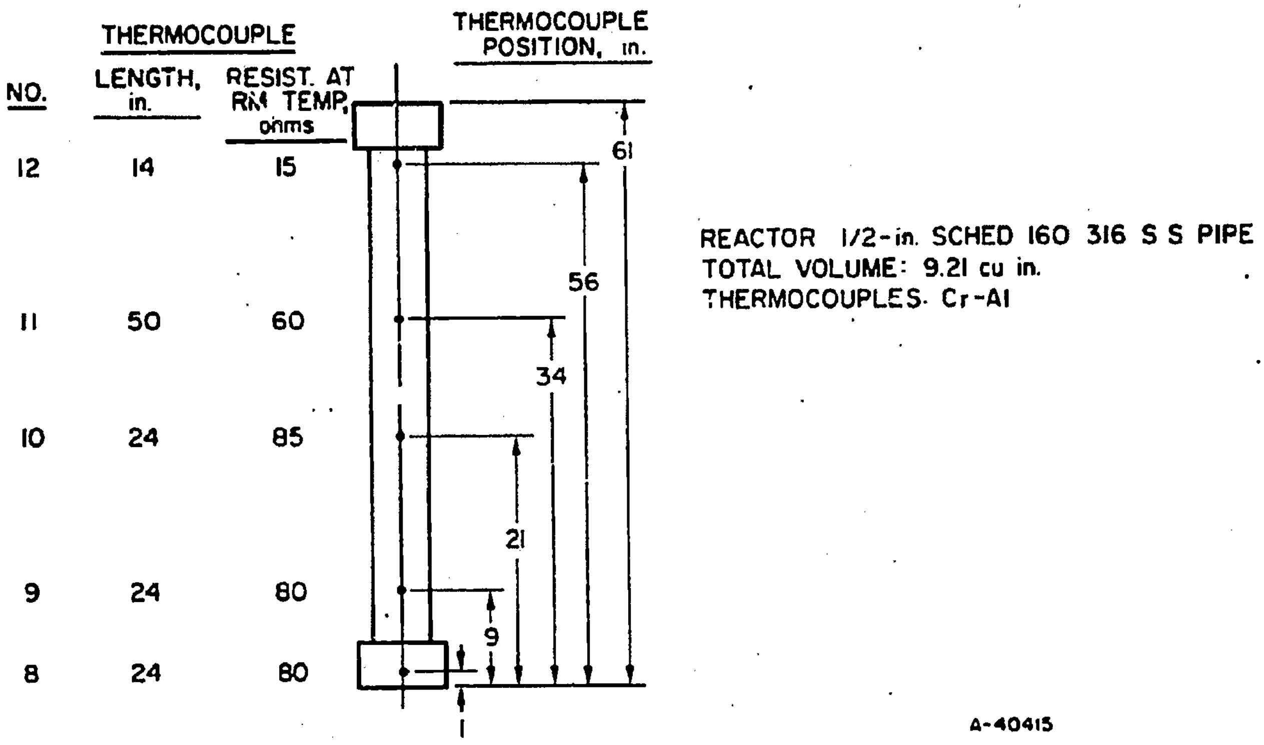
HIGH-PRESSURE METHANATION

Ethylene Hydrogenation

The reactor was assembled and pressure-tested to 1000 psig. The positions of the thermocouples in the reactor are shown in Figure 1. The temperature recorder was calibrated with a K-3 potentiometer; the results are presented in Table 2. The reactor is packed with sand for Runs EH-1 to EH-4 and with sand and catalyst (50 wt % each) for Runs EH-5 to EH-7. Some of the physical properties of the catalyst and of the sand are presented in Tables 3 and 4. Experimental results are presented in Table 5. Without the catalyst, ethylene was not hydrogenated below 800°F. At 1100°F hydrogenation took place in addition to the formation of heavy oil, tar, and carbon. When the catalyst was used, ethylene was hydrogenated to CH_4 and C_2H_6 at 600°F and almost totally converted to CH_4 at 700°F or higher (Runs EH-6 and EH-7).

Table 1. ATTRITION TESTS ON HYDROGASIFICATION RESIDUES AND RELATED SAMPLES

Sample	Carbon Conversion, wt %	USS Sieve Analysis of Tumbled Sample, %			
		+30	−30+100	−100+200	−200
Ireland Mine Coal	--	82.9	14.9	0.6	1.6
Ireland Mine Coal Residue, Run HT-154	24.6	19.8	52.6	12.6	15.0
Ireland Mine Coal Residue, Run HT-210	41.7	31.6	43.9	13.5	11.0
Indiana No. 6 Residue, Run HT-161	31.9	28.3	48.3	8.1	15.3
Lignite Residue, Run HT-192	33.9	45.5	44.4	4.0	6.1
Lignite Residue, Run HT-192	33.9	46.2	42.9	3.6	7.3
Electrogasification Residue, Run EG-34	37.2	42.0	39.8	6.8	11.4
Pretreated Pocahontas No. 4	--	27.6	53.5	9.7	9.2
FMC Char, Feed to EG-26	--	69.5	20.8	4.9	4.8



THERMOCOUPLE NO.	LENGTH, in.	RESIST. AT RM TEMP, ohms
12	14	15
11	50	60
10	24	85
9	24	80
8	24	80

Figure 1. POSITION OF THERMOCOUPLES IN REACTOR

Typical experimental run temperature data are presented in Figure 2. Temperature in the reactor was stabilized (−40 to 0 minutes). Hydrogen was introduced at 5 minutes, then ethylene was introduced at 19 minutes. Gas samples were taken after the temperatures in the reactor stabilized and were analyzed by mass spectrometry.

We will study the results to see if it would be feasible to use ethylene hydrogenation to start up the large hydrogasifier unit.

ENGINEERING ECONOMICS STUDIES

A visit was made to Hudson Products Corp. which is working on the application of air coolers to the lignite plant design. We were impressed with its development in this field. Its engineers will provide us with data on the cost and power consumption of air coolers to cool process streams to 140°F and also to 100°F where the latter is required. This includes the very large cooling requirements for the quench tower and the condenser for the turbine steam. The latter will probably be more expensive than a cooling tower.

Table 2. COMPARISON OF TEMPERATURES BY RECORDER AND POTENTIOMETER FOR THERMOCOUPLES IN ETHYLENE HYDROGENATION REACTOR

Thermocouple No.	Potentiometer		Chart	Potentiometer		Chart
	mV	°F		mV	°F	
2	7.37	358	358	15.35	707	705
3	8.05	388	420	15.51	714	705
8	1.42	95	90	2.23	131	120
9	5.52	275	270	11.76	552	550
10	9.84	468	470	16.17	742	740
11	8.49	408	405	15.82	727	720
12	4.70	238	230	9.51	453	450

Table 3. PHYSICAL PROPERTIES OF HARSHAW HT-100 CATALYST

Catalyst	wt %
NiO	3.8
MoO_3	16.8
SiO_2	1.5
Na_2O	0.02
Fe	0.03
Loss on Ignition at 480°C	1.4 wt %
Bulk Density	46 lbs/cu ft
Surface Area	175 sq m/g

Table 4. PHYSICAL PROPERTIES OF SAND

Size	−100 to +200 mesh
Bulk Density	1.57 g/ml

Table 5. EXPERIMENTAL DATA FROM ETHYLENE HYDROGENATION

	Run No.						
Feed	EH-1	EH-2	EH-3	EH-4	EH-5	EH-6	EH-7
Press., psig	505	505	510	505	500	515	510
Temp, °F							
Furnace	735	650	1065	1255	625	740	850
9	570	500	155	180	140	145	150
10	780	720	1070	1310	690	835	940
11	815	730	1110	1295	675	775	900
Flow Rate, SCF/hr	19.3	20.4	15.6	18.9	16.0	16.6	16.7
Composition, mole %							
H_2	81	76	90	84	88	83	80
C_2H_4	19	24	10	16	12	17	20
Product							
Temp, °F							
Furnace	780	650	1070	1250	625	755	870
9	560	500	150	200	255	490	480
10	800	715	1125	1295	730	840	1010
11	850	735	1150	1290	625	800	1000
Flow Rate, SCF/hr	18.3	17.6	13.8	14.5	14.2	12.5	12.9
Composition, mole %							
H_2	83.8	79.9	77.7	78.7	62.4	55.3	41.7
CH_4	--	--	1.7	3.4	11.5	42.3	55.3
C_2H_6	0.4	0.4	18.4	17.5	25.8	2.4	3.0
C_3H_8	--	--	1.0	--	--	--	--
C_2H_4	15.5	19.6	0.5	0.3	0.3	--	--
C_3H_6	0.1	--	0.2	0.1	--	--	--
C_2H_2	0.1	0.1	--	--	--	--	--
n-C_4H_{10}	--	--	0.5	--	--	--	--

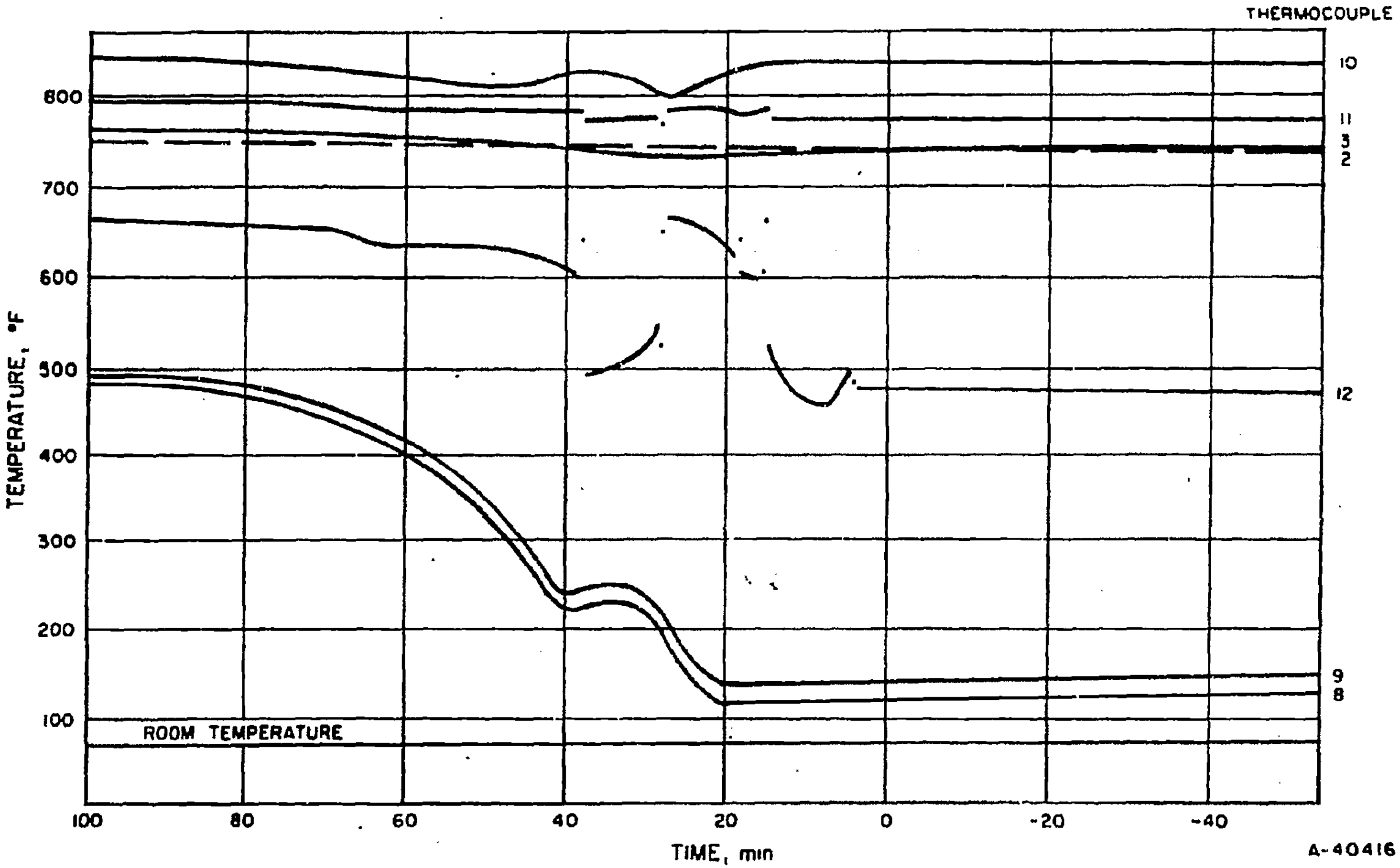


Figure 2. TEMPERATURE DISTRIBUTION INSIDE REACTOR FOR ETHYLENE HYDROGENATION RUN EH-6, PACKED BED: 50 wt % SAND + 50 wt % HT-100 CATALYST

The quench tower system is to be revised with air cooling to at least 140° F. The water that is used to cool the rest of the way to 100° F by direct contact will be cooled indirectly by either water or air, rather than in a separate cooling tower. Most new refineries now do about 80% of their cooling by air. Hudson believes there is a big potential for air cooling in power generation.

Previous reports discussed the effects of varying financial factors on the price of gas and on the return on equity. These were all based on the "electrothermal" design. While the effect of financial factors is almost entirely a function of the investment level, it is desirable to establish a more general relationship. We derived general equations for gas price and equity return and are attempting to develop nomographs. Results of the entire study will be presented in a report to OCR.

DEVELOPMENT UNIT STUDIES

Hydrogasification Tests

This month we ran three hydrogasification tests in the high-temperature balanced-pressure development unit. In one of these tests, Run HT-246, the reactivity of a Montana subbituminous coal to hydrogasification with hydrogen and steam at a system pressure of 500 psig was studied. The purpose of the two other tests, Runs HT-EG-6 and HT-EG-7, was to produce partially hydrogasified char from a Pittsburgh No. 8 seam bituminous coal for use as a feed in the electrogasifier development unit studies. Run HT-246 was terminated after feeding coal for only 1/2 hour when the coal began to hang up in the upper part of the reactor causing the coal feed screw to jam. The duration of Run HT-EG-6 was limited to less than 1 hour because of a light agglomeration of the coal in the lower portion of the reactor tube. Run HT-EG-7 was completely successful.

Significant features of the three tests are given in Table 6.

The subbituminous coal for Run HT-246 was redried to a moisture level of about 1%. Feeding difficulties with this coal in a previous test, Run HT-245 (March 1970 Project Status Report), were attributed to a relatively high-moisture content of 8% causing the coal to compact in the feed screw by forces of the screw flights acting on it. A drier coal would have less tendency to compact under similar forces. The subbituminous coal was gasified in a 3.5-ft fluidized bed with a mixture of hydrogen and steam. Nominal feed

Table 6. FEATURES OF HYDROGASIFICATION TEST RESULTS FOR RUNS HT-246, HT-EG-6, AND HT-EG-7

Run No.	Temperature, °F	Purpose of Run	Results
Feed Solids: Dried Montana Subbituminous Coal, Colstrip Mine Feed Gas: Hydrogen-Steam			
HT-246	1300-1700	To study the hydrogasification reactivity of a subbituminous coal with hydrogen and steam at 500 psig	Coal hang up in reactor, feed screw jammed
Feed Solids: Pretreated Pittsburgh No. 8 Seam Bituminous Coal, Ireland Mine Feed Gas: Hydrogen-Steam			
HT-EG-6	1300-1700	To produce a hydrogasified bituminous coal residue for use in the pilot plant electrothermal gasifier	Light agglomeration of coal. Short duration
HT-EG-7	1300-1700	Same as HT-EG-6	Successful

conditions were 51.6 lb/hr coal, 345 SCF/hr hydrogen, and 8.8 lb/hr of steam, the same as for Run HT-245. Within 20 minutes after coal feeding was started the coal feed screw jammed. Internal reactor temperatures indicated a holdup of the coal about 30 inches below the outlet of the coal feed tube. Feeding was resumed 14 minutes later when the holdup broke loose, but for only about 11 minutes before the feed screw jammed again. The test was terminated when the coal blockage could not be cleared. Light agglomeration of the coal was responsible for the coal holdup. This is apparently related to hydrogasification at a system pressure of 500 psig. This same coal in an earlier hydrogasification series at 1000 psig showed no tendency to agglomerate.

Operating conditions for Runs HT-EG-6 and HT-EG-7 were similar to those of previous bituminous coal char production runs. The lightly pretreated coal from the Ireland mine was fed at a nominal rate of 55 lb/hr and gasified with a mixture of hydrogen and steam at a system pressure of 1000 psig in a 3.5-ft fluidized bed. Hydrogen was fed at a nominal rate of 530 SCF/hr (25% stoichiometric hydrogen-to-coal ratio); steam was fed at 25.3 lb/hr (50 mole percent concentration in the feed gas). Run HT-EG-6 was ended after less than 1 hour of feeding coal, when a very small amount of the coal lightly agglomerated in the lower section of the reactor tube preventing the partially gasified coal from flowing to the discharge screw. Behavior of this small amount of coal is abnormal since the same batch of this pretreated coal was hydrogasified in Run HT-242 (March 1970 Project Status Report) with no agglomeration. After the coal was rescreened to remove the large fraction of fines (less than 80-mesh size) produced from previous handling of the coal, about 400 lb of the coal was successfully hydrogasified in Run HT-EG-7 over a period of 6-1/2 hours. Complete hydrogasification results of Run HT-EG-7 will be presented when analyses of this test are completed.

Complete hydrogasification results of two tests, Runs HT-242 and HT-243, conducted in February (March 1970 Project Status Report) are presented in this report. Run HT-242 studied the hydrogasification reactivity of a lightly pretreated Pittsburgh No. 8 seam bituminous coal from the Ireland mine with a synthesis gas and steam mixture at a system pressure of 1500 psig. The aim of Run HT-243 was to study the effect of a 500-psig system pressure on the conversion of a dried Montana subbituminous coal from the Colstrip mine when hydrogasified with a synthesis gas and steam mixture.

Operating conditions and results of these runs are presented in Table 7. Compositions and screen analyses of the feeds and residues are given in Table 8. Liquid products and compositions are shown in Table 9.

In Run HT-242, 29.2% of the moisture-, ash-free coal was gasified, and 24.5% of the carbon was converted to gaseous products (Table 7). Hydrocarbon (methane) yield was 3.46 SCF/lb coal, while the carbon oxides yield was 0.929 SCF/lb coal. Carbon converted to liquid products represented 2.29% of the carbon in the coal (Table 9).

For comparative purposes the key results of Run HT-242 are presented in Table 10 with those of a synthesis gas test conducted at 1000 psig (Run HT-169), and another test with synthesis gas conducted at 1500 psig (Run HT-225). Results of Run HT-169 were previously reported in the May 1967 Project Status Report and those of Run HT-225 in the July 1969 Project Status Report. A higher percentage of carbon gasification was attained in Run HT-242 (24.5%) than in Run HT-225 (17.4%) at 1500 psig, mostly due to the larger hydrogen-to-coal ratio, 24.8% of stoichiometric compared to 13.0% of stoichiometric. Product-gas composition of the two tests are not significantly different. At a system pressure of 1000 psig in Run HT-169, a carbon gasification of 23.8%, nearly equal to that of Run HT-242, was obtained with

Table 7, Part 1. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF PRETREATED BITUMINOUS COAL AND DRIED SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR

Coal	Ireland Mine Bituminous Coal	Montana Subbituminous Coal
Source	IGT Pretreater, FP-142	Colstrip Mine
Sieve Size, USS	—10+80	
Run No.	HT-242	HT-243
Duration of Test, hr	4-3/4	3-3/4
Steady-State Operating Period, min[a]	101-266	210-227
OPERATING CONDITIONS		
Bed Height, ft	3.5	3.5
Reactor Pressure, psig	1556	489
Reactor Temperature, °F[b] Inches From Bottom		
62-1/2	925	1250
67-3/4	1125	1435
73	1215	1535
78-1/4	1250	1575
83-1/2	1415	1700
89	1370	1570
94-1/4	1340	1535
100	1595	1700
104	1555	1635
Average	1310	1550
Coal Rate, lb/hr[c]	42.58	16.66
Feed Gas Rate, SCF/hr	674.0	249.2
Steam Rate, lb/hr	32.41	12.18
Steam, mole % of hydrogen-steam mixture	50.3	50.7
Hydrogen/Coal Ratio, % of stoichiometric[d]	24.8	24.9
Hydrogen/Steam Ratio, mole/mole	0.567	0.559
Bed Pressure Differential, in. wc	--	--
Coal Space Velocity, lb/cu ft-hr	137.6	53.85
Feed Gas Residence Time, min[e]	0.422	0.325
Superficial Feed Gas Velocity, ft/s[f]	0.138	0.180

Table 7, Part 2. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF PRETREATED BITUMINOUS COAL AND DRIED SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR

Run No.	HT-242	HT-243
OPERATING RESULTS		
Product Gas Rate, SCF/hr	1132.8	596.7
Net Btu Recovery, 10^3 Btu/lb	2.636	3.145
Product Gas Yield, SCF/lb	26.60	35.82
Hydrocarbon Yield, SCF/lb	3.46	3.33
Carbon Oxides Yield, SCF/lb	0.929	2.47
Net Reacted Hydrogen, SCF/lb	1.78	nil
Residue, lb/lb coal[g]	0.669	0.537
Liquid Products, lb/lb coal[h]	0.638	0.560
Net MAF Coal Hydrogasified, wt %[i]	29.2	36.0
Carbon Gasified, wt %	24.5	30.0
Steam Decomposed, lb/hr[j]	6.01	3.80
Steam Decomposed, % of steam fed	18.54	31.2
Steam Decomposed, % of total equivalent fed[k]	45.2	49.1
Overall Material Balance, %	97.6	96.4
Carbon Balance, %	99.9	100.2
Hydrogen Balance, %	95.1	87.4
Oxygen Balance, %	96.2	90.8

PRODUCT GAS PROPERTIES

	Feed	Product	Feed	Product
Gas Composition, mole %				
Nitrogen	--	30.3	--	41.5
Carbon Monoxide	37.3	12.4	37.2	9.0
Carbon Dioxide	5.4	16.5	5.4	15.7
Hydrogen	57.3	27.4	57.4	24.4
Methane	--	12.4	--	8.5
Ethane	--	0.5	--	0.4
Propane	--	0.1	--	0.4
Butane	--	--	--	--
Benzene	--	0.4	--	0.1
Hydrogen Sulfide	--	--	--	--
Total	100.0	100.0	100.0	100.0
Heating Value, Btu/SCF[m]	301	276	301	212
Specific Gravity (Air = 1.00)	0.483	0.772	0.482	0.808
Nitrogen Purge Rate, SCF/hr	343		248	

Table 7, Part 3. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF PRETREATED BITUMINOUS COAL AND DRIED SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR

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 feed gas 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 8. CHEMICAL AND SCREEN ANALYSES OF PRETREATED BITUMINOUS COAL AND DRIED SUBBITUMINOUS COAL FEED AND RESIDUE

Run No.	HT-242		HT-243	
Sample	Feed	Residue	Feed	Residue
Proximate Analysis, wt %				
Moisture	0.8	0.5	4.8	1.9
Volatile Matter	22.5	3.9	37.0	6.9
Fixed Carbon	62.1	73.7	51.2	77.8
Ash	14.6	21.9	7.0	13.4
Total	100.0	100.0	100.0	100.0
Ultimate Analysis (dry), wt %				
Carbon	67.4	73.6	68.3	80.3
Hydrogen	3.11	1.49	4.54	1.92
Nitrogen	1.03	0.78	0.96	0.82
Oxygen	10.26	0.59	18.13	2.84
Sulfur	3.23	1.57	0.74	0.48
Ash	14.70	21.97	7.33	13.64
Total	100.00	100.00	100.00	100.00
Screen Analysis, USS, wt %				
+20	24.3	24.5	3.0	1.2
+30	15.3	19.7	27.3	10.6
+40	15.5	17.5	25.1	27.5
+60	22.7	21.7	26.9	38.7
+80	12.8	9.2	11.8	13.9
+100	4.0	2.7	3.3	3.9
+200	4.8	4.0	2.2	3.5
+325	0.4	0.5	0.3	0.4
–325	0.2	0.2	0.1	0.3
Total	100.0	100.0	100.0	100.0

Table 9. COMPOSITION OF HYDROGASIFICATION LIQUID PRODUCTS

Run No.	HT-242	HT-243
Sample	Condenser	Condenser
Liquid Products,*		
lb/lb coal	0.638	0.560
Composition of Liquid Products, wt %		
Water	97.15	89.81
Oil	2.85	10.19
Total	100.00	100.00
Composition of Oil Fraction, wt %		
Carbon	85.00	84.60
Hydrogen	6.09	7.98
Total	91.09	92.58
Carbon in Oil Fraction		
lb/lb coal	0.01546	0.0483
wt % of carbon in coal	2.29	7.07

* Includes condensed, undecomposed steam.

Table 10. COMPARISON OF IRELAND MINE BITUMINOUS COAL HYDROGASIFICATION RESULTS WITH SYNTHESIS GAS AT 1000 AND 1500-psig SYSTEM PRESSURE

Run No.	HT-225	HT-169	HT-242
Feed Gas		Synthesis Gas	
Reactor Pressure, psig	1478	1026	1556
Coal Bed Temp Average, °F	1365	1615	1310
Coal Feed Rate, lb/hr	63.49	46.05	42.58
Synthesis Gas Rate, SCF/hr	626.5	512.6	674.0
Steam Feed Rate, lb/hr	33.06	25.06	32.41
Hydrogen/Coal Ratio, % of stoichiometric	13.0	15.4	24.8
Equivalent Hydrogen/Coal Ratio, % of stoichiometric	25.0	28.8	40.9
Steam Concentration in Feed Gas, mole %	52.6	50.7	50.3
Steam Decomposed, % of total equivalent steam fed	36.5	27.08	45.2
Carbon Gasified, %	17.4	23.8	24.5
MAF Coal Gasified, %	22.1	37.3	29.2
Hydrocarbon Yield, SCF/lb	2.69	4.12	3.46
CO + CO_2 Yield, SCF/lb	0.662	0.767	0.929
Product Gas Composition (nitrogen-free), mole %			
Carbon Monoxide	19.9	15.7	17.8
Carbon Dioxide	27.0	28.3	23.7
Hydrogen	30.5	26.0	39.3
Methane	21.0	28.5	17.8
Ethane	0.5	0.8	0.7
Propane	0.4	0.2	0.1
Benzene	0.4	0.3	0.6
Hydrogen Sulfide	0.3	0.2	0.0
Total	100.0	100.0	100.0
Product Gas Heating Value (nitrogen-free), Btu/SCF	404	447	396

a significantly lower hydrogen-to-coal ratio (15.4%) than in Run HT-242. The product gas of Run HT-169 has a larger methane concentration and a lower hydrogen concentration than either Run HT-225 or HT-242. The poorer results at a system pressure of 1500 psig are attributed to the relatively low average coal-bed temperatures, 1365°F in Run HT-225 and 1310°F in Run HT-242, compared to 1615°F in Run HT-169. The temperature level attained in Run HT-242 was limited by the heat input capacity of the reactor electrical heaters.

Gasification of Montana subbituminous coal in Run HT-243 at 500 psig with synthesis gas and steam resulted in a carbon gasification of 30.0% (Table 7). The hydrogen (methane) yield was 3.33 SCF/lb coal; the carbon oxides yield was 2.47 SCF/lb coal. An additional 7.07% of the carbon in the coal was converted to oil products (Table 9).

To show the effect of pressure on the gasification of Montana subbituminous coal with synthesis gas and steam, key results of Run HT-243 are compared with those of Run HT-214, conducted at 1111 psig, in Table 11. Results of Run HT-214 were originally reported in the November 1968 Project Status Report. The percentage of carbon gasification at both system pressures was about the same at 30%. However, the proportion of the gasified carbon going into hydrocarbons (methane) and carbon oxides was significantly different as indicated by the yields of these gaseous components per pound of coal fed. At 1111 psig these yields were 4.53 SCF/lb of hydrocarbons and 1.156 SCF/lb of carbon oxides. At 489 psig the hydrocarbon yield was 3.33 SCF/lb and the carbon oxides yield was 2.47 SCF/lb. These yields are also reflected in the product-gas compositions. The methane concentration was significantly higher and the carbon oxides concentration lower when gasifying at 1111 psig than when gasifying at 489 psig. The results of these two tests indicate that the reactivity of the Montana subbituminous coal as measured by the total carbon gasification at 500 psig is similar to that at 1000 psig. However, because of equilibrium effects methane production is lower at 500 psig.

Slurry Pumping

The nozzles acquired from Spraying Systems Corporation hold their designed spray angle and full-cone spray pattern with both water and coal plus water slurry feeds.

Table 11. COMPARISON OF MONTANA SUBBITUMINOUS COAL HYDROGASIFICATION RESULTS WITH SYNTHESIS GAS AT 500 AND 1000-psig SYSTEM PRESSURE

Run No.	HT-214	HT-243
Reactor Pressure, psig	1111	489
Coal Bed Temp Average, °F	1540	1550
Coal Feed Rate, lb/hr	58.64	16.66
Synthesis Gas Rate, SCF/hr	608.1	249.2
Steam Feed Rate, lb/hr	26.16	12.18
Hydrogen/Coal Ratio, % of stoichiometric	17.7	24.9
Equivalent Hydrogen/Coal Ratio, % of stoichiometric	28.9	41.0
Steam Concentration in Feed Gas, mole %	47.5	50.7
Steam Decomposed, % of total equivalent steam fed	50.6	49.1
Carbon Gasified, %	29.6	30.0
MAF Coal Gasified, %	35.2	36.0
Hydrocarbon Yield, SCF/lb	4.53	3.33
$CO + CO_2$ Yield, SCF/lb	1.156	2.47
Product-Gas Composition (nitrogen-free), mole %		
Carbon Monoxide	12.5	15.4
Carbon Dioxide	24.4	26.8
Hydrogen	31.7	41.7
Methane	28.6	14.5
Ethane	1.3	6.7
Propane	0.8	0.7
Benzene	0.3	0.2
Hydrogen Sulfide	0.4	--
Total	100.0	100.0
Product Gas Heating Value (nitrogen-free), Btu/SCF	479	362

A Polaroid camera was used to find the best photographic method to test the spray particle size. The fastest shutter speed, a closeup lens, and various lighting locations all failed to produce a sharp particle pattern. A strobe light was used and sharp pictures obtained. When these are developed and projected, it is hoped good measurements can be made.

Several minor problems had to be solved before smooth operation could be obtained with the Wilson-Snyder pump. First, the check valves failed to seat at medium flow rates when the pump was operated at atmospheric pressure. The pump piston packing and the check valve seats were replaced. The discharge system was modified to pump through an air-controlled valve. By maintaining a 1000-psi pressure on the pump, increasing flow was obtained up to one-half maximum speed.

Above half speed fluid "hammer" occurred. The slurry circulation pump was found to be the cause and replacement of the impeller stopped the knock. An air-controlled valve with a larger orifice had to be installed to operate the Wilson-Synder pump at high flow rates without exceeding 1000 psi pressure on the pump.

ELECTROTHERMAL GASIFICATION

Six tests were conducted in the electrothermal gasifier this month. All tests were made at 1000 psig pressure with the concentric electrode configuration and using FMC Project COED char as feed material. A silicon carbide tube was used as electrode material during two of the tests.

Run EG-49 was the first to use the 1.5-in.-OD silicon carbide tube as electrode material: A section 28 inches long was submerged in the fluidized bed. A sleeve and a metal pin served as connectors to the 316 stainless steel electrode rod above the bed. No changes were made on the 316 stainless steel outer electrode tube. The initial resistance during the heat-up period varied from 700 to 100 ohms; however, within several minutes the resistance decreased to 10 ohms, and for the remainder of the heat-up period the voltage-current relationship closely followed that observed with the stainless steel electrode. A noticeable increase in resistance occurred several minutes before steam introduction, causing termination of the test. Inspection of the reactor revealed that the silicon carbide electrode had been sheared off at the base of the connecting sleeve. A hot spot at that point probably caused the fracture (Figure 3).

A 5/8-in. stainless steel rod was inserted through the center of the silicon carbide tube to further stabilize the electrode for Run EG-50. The bottom of the electrode was capped off with refractory cement to prevent current transfer from the tip. The heat-up period was smooth and similar to that experienced with the stainless steel electrode. A malfunction of the steam orifice transmitter prevented introduction of steam, and heat-up to 1700°F was continued using nitrogen as a fluidizing medium. The resistance at 1700°F was approximately 1 ohm. Having brought the heat-up to operating temperature we terminated the test. Minor fluctuations in the resistance during the test were due to the fracture of the silicon carbide tube halfway from the connecting sleeve (Figure 4). A 1.5-in. OD solid silicon carbide rod which has been ordered is expected to have greater strength and will provide better results than the tube.

Meanwhile, a 316 stainless steel electrode was reinstalled for Run EG-51. A very smooth heat-up period was followed by a period of erratic steam flow. Fluidization was temporarily lost at one point, causing a current surge and termination of the test. Power to the unit was interrupted fast enough to avoid any serious damage. The only noticeable effect was a small burned spot at the tip of the electrode.

Run EG-52 was plagued by numerous mechanical problems. An early termination was forced when the superheater cycled off and attempts to restart it were unsuccessful. A faulty thermocouple was replaced in the superheater and some minor repairs on the unit were completed before Run EG-53.

Run EG-53 was conducted to provide two of our consultants with data on the electrical characteristics of our fluidized-bed operation. A Bell & Howell Type 5-134 recording oscillograph was installed to accumulate data on high- and low-frequency current fluctuations. Chart recordings were made when steady-state operating conditions were reached. Transient behavior of the bed following step changes in power input was recorded and will be used to define bed characteristics. The polarity of the electrodes was switched to observe differences in operation; however, the feedstock was used up before we could establish an adequate period of steady-state operation at reversed polarity. Another test will be made within several weeks.

The last test conducted this month was to collect additional data at 1900°F using FMC Project COED char. During Run EG-54 we discovered that water

Figure 3. SILICON CARBIDE ELECTRODE AT JUNCTION TO STAINLESS STEEL ROD AFTER RUN EG-49

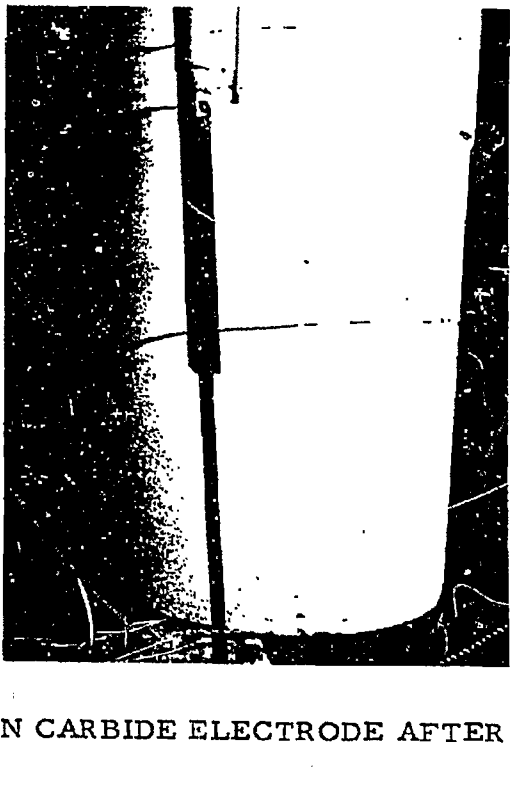


Figure 4. SILICON CARBIDE ELECTRODE AFTER RUN EG-50

had collected in the insulation as a result of the operational procedures of the last run: Apparently some condensation of steam occurred while the polarity of the electrodes was switched. Considerable time was spent drying out the insulation in order to establish good fluidization. High-resistance fluctuations prevented steady-state operation. The test was terminated after continuous slugging of solids in the reactor steadily decreased our operating temperature.

The completed operating data and results of Runs EG-46 and EG-47 appear in Table 12. The feed and residue analyses are given in Table 13. Results of Run EG-48 are now being processed. Note that the carbon content of the Run EG-47 feed material is much higher than previously experienced and the effect is reflected in the carbon conversion attained. We have never had a coal or char with such low-ash and high-carbon contents. Apparently, these analyses cannot be attributed to analytical techniques. We are continuing to try to find an explanation for this.

PILOT PLANT CONSTRUCTION

Engineering

Major engineering activity has been in the design and detailing of instrumentation, piping, and electrical additions. The emergency power system design has been finalized and specifications are being prepared. The total project's detailed design and drafting is 98% complete. The scale model will be completed during the next month.

Procurement

The insulation subcontract was let. The instrument/electrical subcontract quotation is being reviewed. Deliveries of the final shipments of shop-fabricated pipe slipped to April 20.

Construction

Major field activities during this report period were field shop pipe fabrication, area piping, erection of platforms and ladders, area paving, and electrical work. Piping is 40% complete.

We have experienced a total of 14 inclement weather days, 2 of which occurred in this report period. On these days, progress was significantly below normal.

Several views of the site are shown in Figures 5, 6, 7, and 8.

Table 12. OPERATING CONDITIONS AND RESULTS OF THE ELECTROTHERMAL GASIFICATION OF COAL CHAR

Run No.	EG-46	EG-47
Feed Char	HV Bituminous Hydrogasified Char	
Sieve Size, USS	−10+80	
Duration of Test, hr	4.65	5.28
Steady-State Operating Period, min	90	80
Electrode Material	316 SS	316 SS
OPERATING CONDITIONS		
Bed Height, ft	2.75	2.75
Reactor Pressure, psig	1008	1004
Reactor Temp, °F Inches From Bottom		
39	1930	1950
42	1805	1940
45	1815	1935
48	1945	1930
51	1870	1960
54	1890	1905
57	1950	1880
60	1910	1825
Average	1889	1916
Steam Feed Rate, lb/hr	104	140
Steam Residence Time, min*	0.22	0.17
Steam Superficial Velocity, ft/s†	0.20	0.28
Steam/Char Feed Ratio, lb/lb	0.78	1.22
Char Feed Rate (dry), lb/hr	132.6	114.4
Char Residence Time, min	5.3	6.1
Nitrogen Purge Rate, SCF/hr	682	558
Voltage, V	213	285
Current, A	344	221
Power Input, kW	73.3	63.0
Overall Resistance, ohms	0.62	1.3

Table 12, Cont. OPERATING CONDITIONS AND RESULTS OF THE ELECTROTHERMAL GASIFICATION OF COAL CHAR

Run No.	EG-46	EG-47
OPERATING RESULTS		
Product Gas Rate (dry), SCF/hr‡	2419	2294
Product Gas Yield (dry), SCF/lb char	18.2	20.1
Hydrogen Yield, SCF/lb	8.5	10.6
Hydrogen + Carbon Monoxide Yield, SCF/lb	14.4	15.5
Carbon Oxides Yield, SCF/lb	8.1	8.0
Char Gasified, wt %	34.0	33.7
Carbon Gasified, %	41.9	33.0
Liquid Products, lb/hr	36.2	70.4
Steam Decomposed, lb/hr	67.8	69.6
Steam Conversion, %	65.2	49.7
Overall Material Balance, %	97.7	95.5
Carbon Balance, %	101.0	98.0
Hydrogen Balance, %	100.0	100.0
Oxygen Balance, %	92.2	88.7
PRODUCT GAS PROPERTIES		
Composition, mole % ‡		
CO	32.6	24.3
CO_2	11.9	15.5
H_2	46.5	53.0
CH_4	9.0	7.2
H_2S	--	--
Total	100.0	100.0
Specific Gravity (Air = 1.00)	0.578	0.547

* Coal bed volume (top 2.75 ft)/SCF steam feed at average reactor temperature and pressure.

† CF/s steam at reactor temperature and pressure/cross-sectional area of reactor.

‡ Dry, nitrogen-free basis.

Table 13. CHEMICAL AND SCREEN ANALYSES OF ELECTROTHERMAL GASIFICATION FEEDS AND RESIDUES

Run No.	EG-46		EG-47	
Sample	Feed	Residue	Feed	Residue
Proximate Analysis, wt %				
Moisture	2.5	0.7	1.9	0.9
Volatile Matter	2.9	1.9	2.2	2.3
Fixed Carbon	73.5	65.4	91.4	90.0
Ash	21.1	32.0	4.5	6.8
Total	100.0	100.0	100.0	100.0
Ultimate Analysis, wt %				
Carbon	74.60	66.80	92.40	90.50
Hydrogen	1.42	0.56	0.81	0.81
Nitrogen	0.76	0.22	0.65	0.46
Oxygen	1.77	0.21	1.47	1.11
Sulfur	0.20	0.03	0.12	0.26
Ash	21.25	32.18	4.55	6.86
Total	100.00	100.00	100.00	100.00
Screen Analysis, USS, wt %				
+20	7.9	4.7	5.8	0.2
+30	23.0	13.4	19.3	0.4
+40	23.9	25.4	22.6	0.7
+60	28.2	35.9	32.4	4.6
+80	11.6	12.3	13.9	18.5
+100	4.3	3.1	3.5	13.3
+200	1.1	3.8	1.9	32.5
+325	--	0.7	0.2	9.0
−325	--	0.7	0.4	20.8
Total	100.0	100.0	100.0	100.0

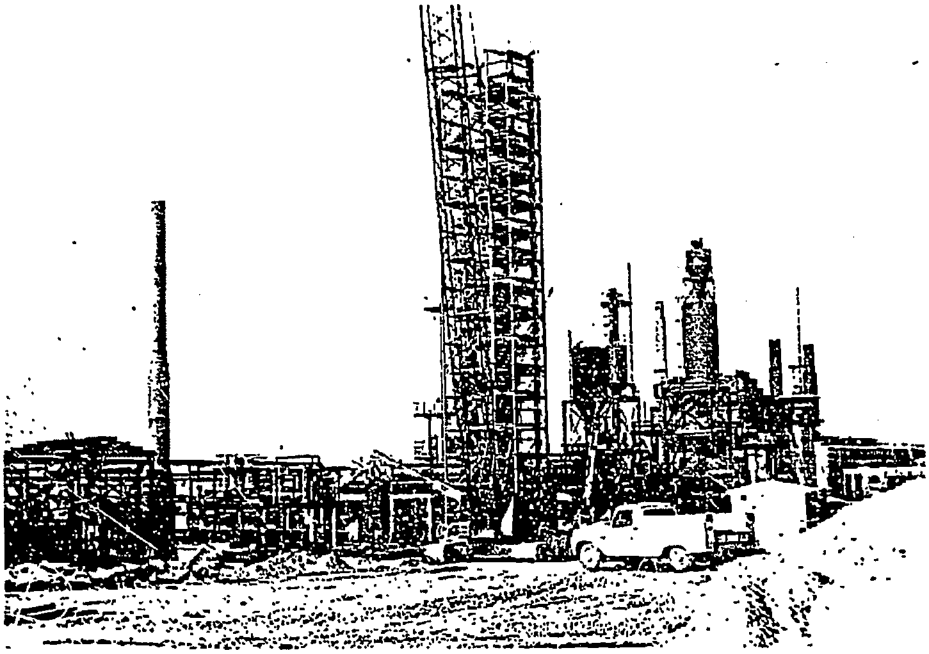


Figure 5. OVERALL PLANT LOOKING NORTHEAST

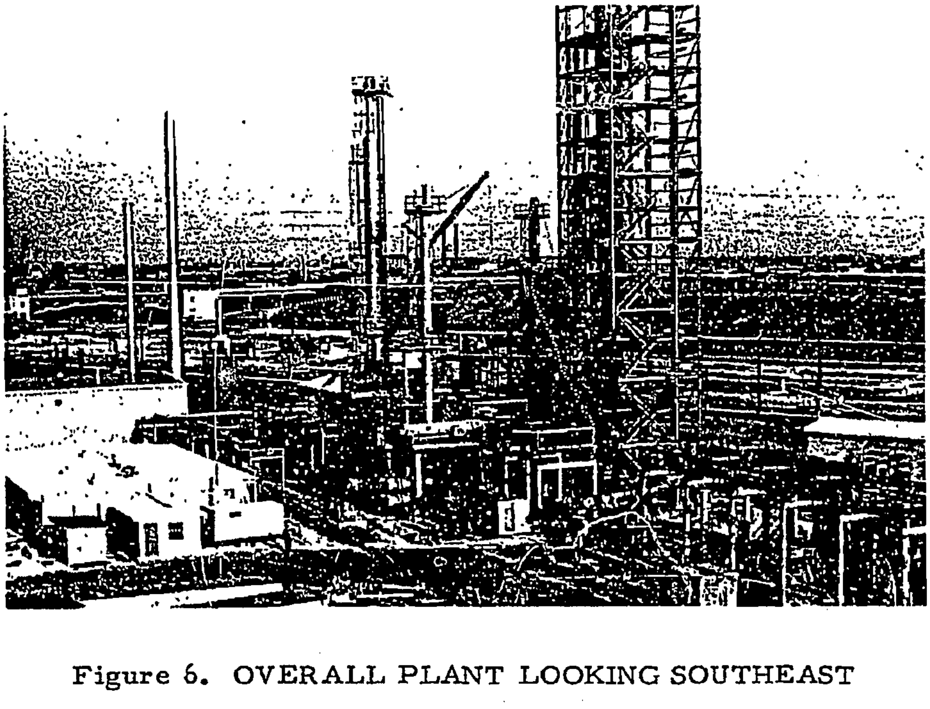


Figure 6. OVERALL PLANT LOOKING SOUTHEAST

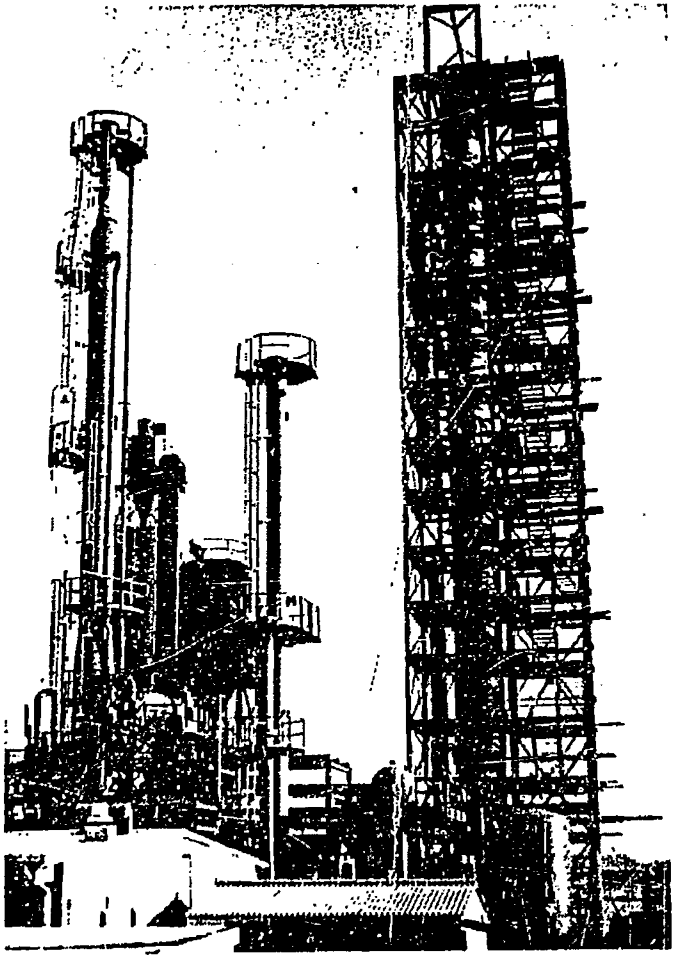


Figure 7. PROCESS AREA LOOKING SOUTH-SOUTHEAST

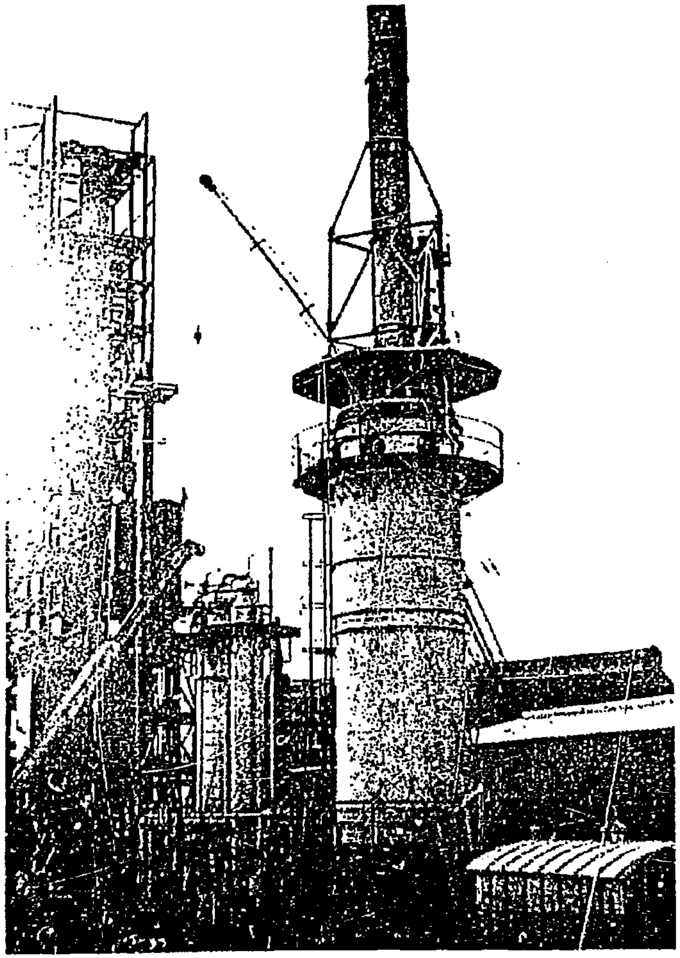
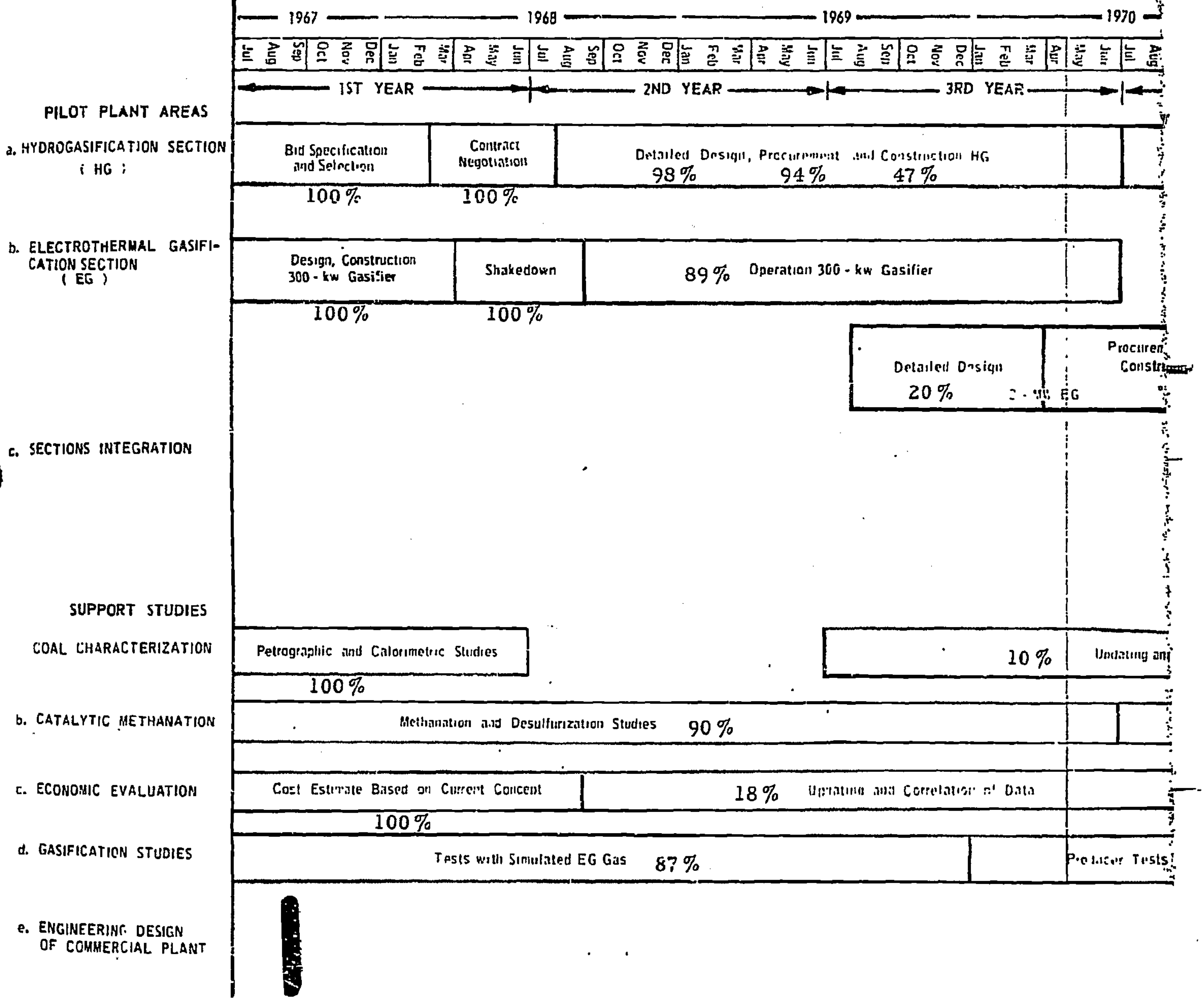


Figure 8. HYDROGEN PLANT LOOKING NORTHWEST

PILOT PLANT PROGRAM OF IGT HYDROGASIF

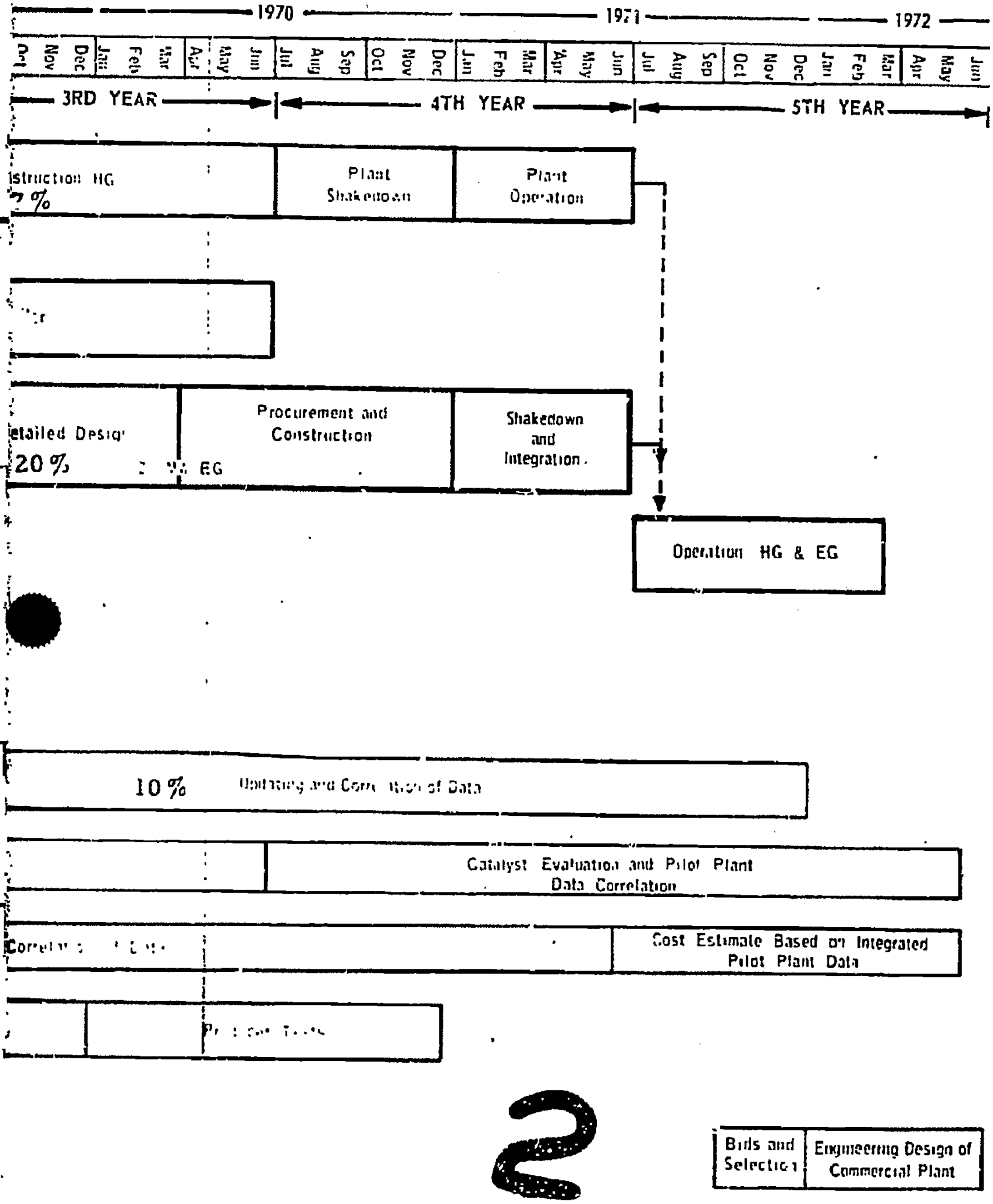
OCR CONTRACT No. 14-01-0001-381 (1)



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

F IGT HYDROGASIFICATION PROCESS

01-381(1) AGA: 1U-4-1



2

E - 1084

PROGRESS REPORT NO. 21

COAL HYDROGASIFICATION PILOT PLANT

FOR

INSTITUTE OF GAS TECHNOLOGY

CHICAGO, ILLINOIS

W-1784

TABLE OF CONTENTS

PROGRESS REPORT

PROJECT:	Institute of Gas Technology Chicago, Illinois Coal Hydrogasification Pilot Plant Procon Job No. W-1784
REPORT NO:	21
DATE:	April 15, 1970
PROCON PROJECT MANAGER:	T. A. Taylor

DISTRIBUTION

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Mr. C. J. Towle	1
Mr. T. A. Taylor	2
Field	1

W-1784

Coal Hydrogasification Pilot Plant

I. SUMMARY

Engineering	98%
Purchasing	94%
Material Receipt	90%
Construction	47%

A. ENGINEERING

Major engineering activity has been the design and detailing of instrumentation, piping and electrical additions. The emergency power system design has been finalized and specifications are being prepared.

Total project detailed design and drafting is 98 percent complete.

The scale model will be completed during the next report period.

B. PROCUREMENT

The insulation subcontract has been let. The instrument/electrical subcontract quotation is being reviewed.

Deliveries of the final shipments of shop fabricated pipe have slipped to April 20th.

C. CONSTRUCTION

Major field activities during this report period have been field shop pipe fabrication, area piping, erection of platforms and ladders, area paving and electrical work.

C. CONSTRUCTION (continued)

Piping is now 40 percent complete.

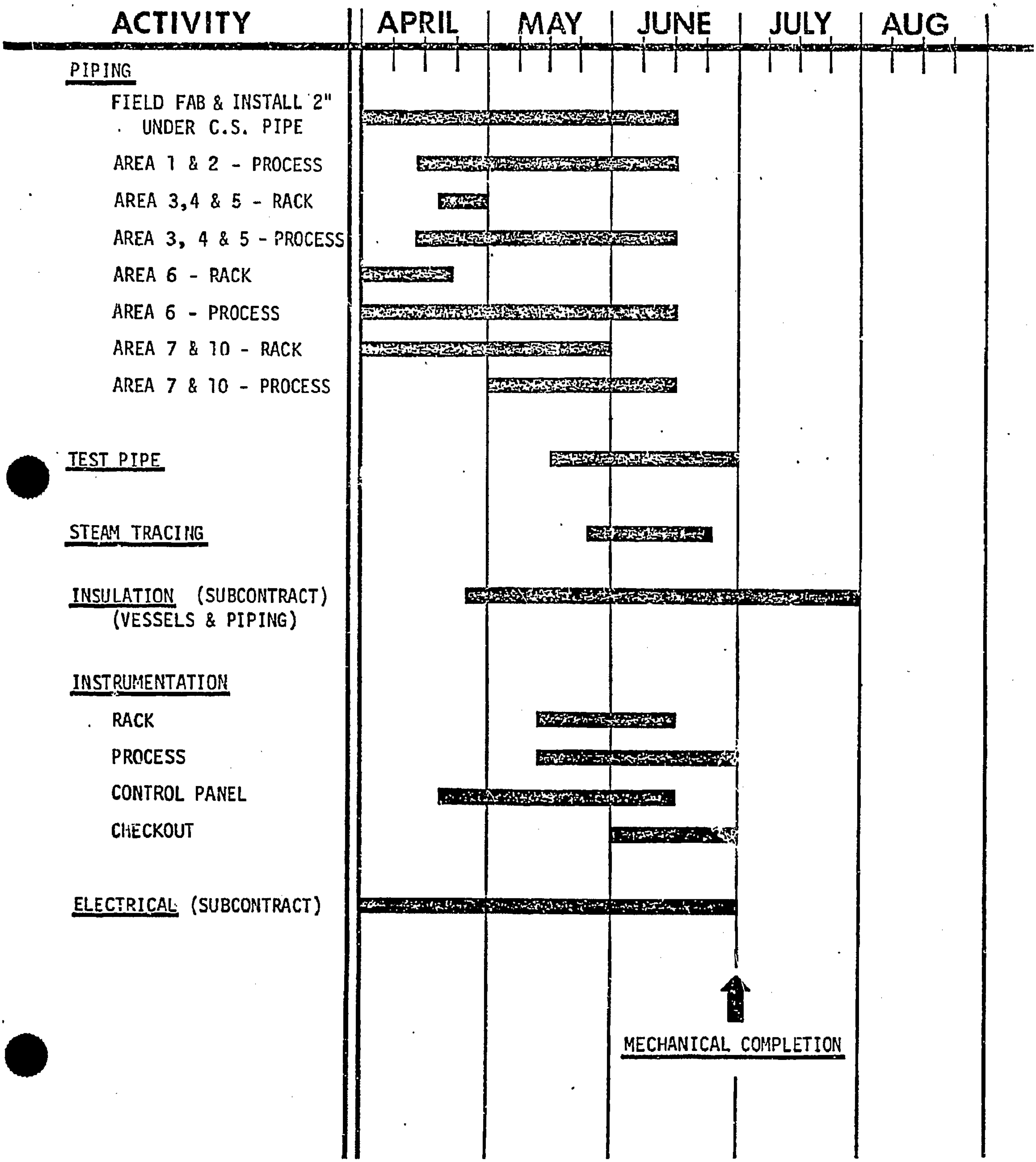
We have experienced a total of fourteen inclimate weather days, two of which occurred in this report period. On these days, progress was significantly below normal.

II. SCHEDULE AND S-CURVE REPORT

Review of the construction schedule by the field resulted in the bar chart schedule enclosed.

The S-Curve Report has been updated to show present progress.

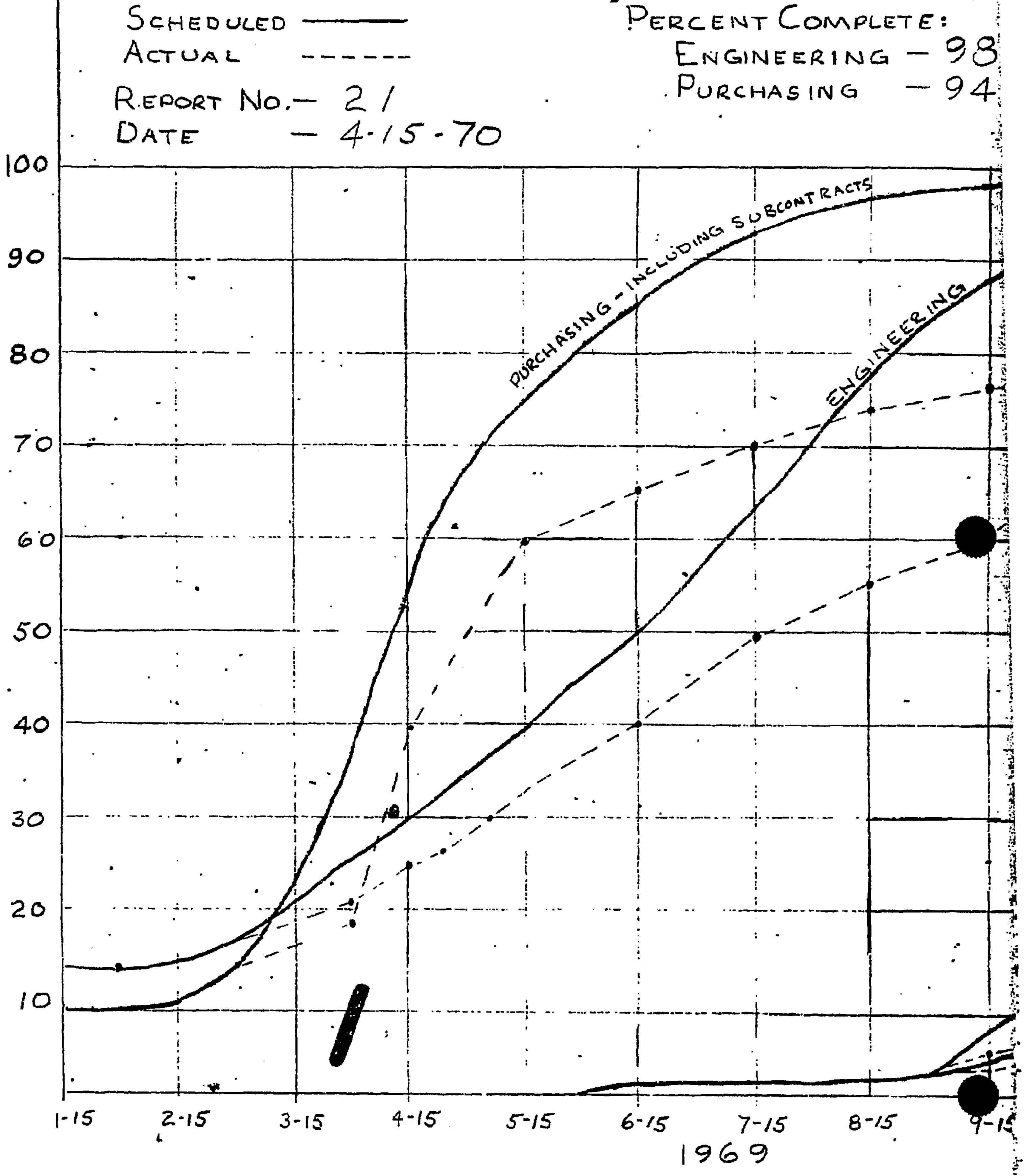
ARCHART SCHEDULE
FOR
IGT
HYDROGASIFICATION PLANT



INSTITUTE OF GAS

PERCENT COMPLETE:
ENGINEERING – 98
PURCHASING – 94

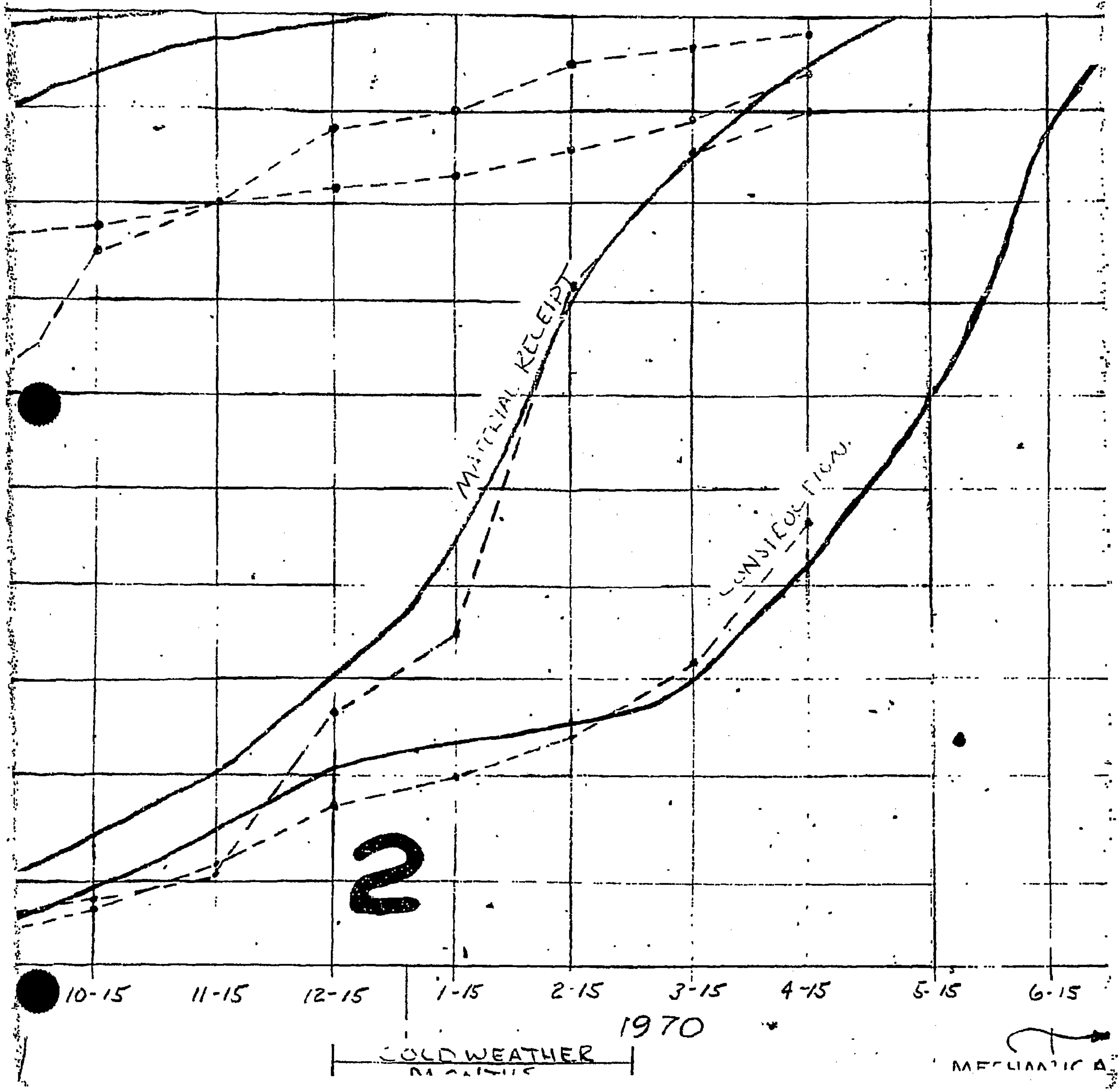
REPORT NO.– 21
DATE – 4-15-70



NOLOGY - 1784

MATERIAL RECEIVED - 90
CONSTRUCTION - 47

REV 0 - 3-15-69
REV. 1 - 4-15-69
REV. 2 - 7-15-69 CONSTRUCTION ONLY
REV. 3 - 8-15-69 " "
REV. 4 - 1-15-70 CONSTRUCTION & M
RECEIPT ONLY



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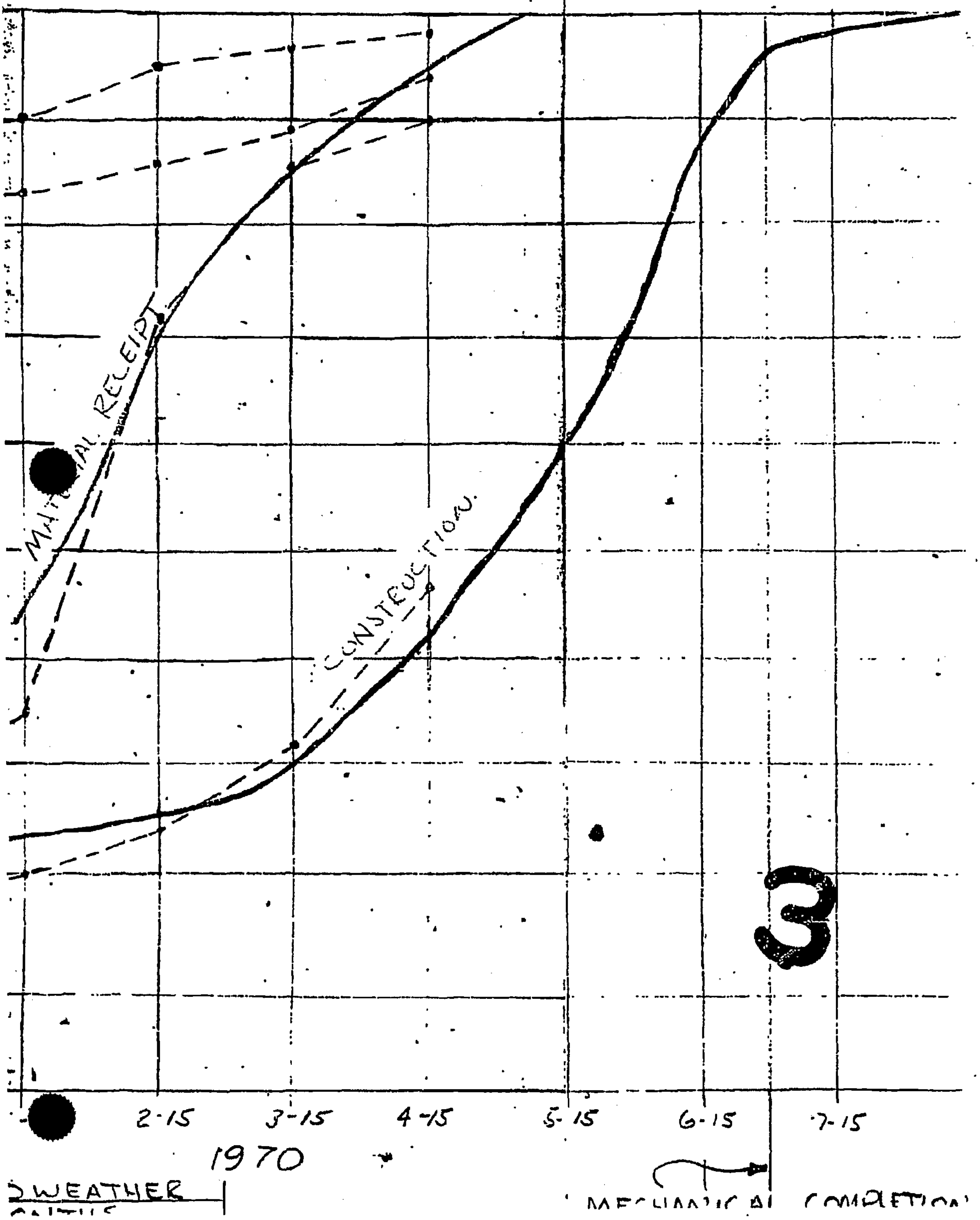
REV. 1 - 4-15-69
REV. 2 - 7-15-69 CONSTRUCTION ONLY
REV. 3 - 8-15-69 "
REV. 4 - 1-15-70 CONSTRUCTION & MATERIAL RECEIPT ONLY



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.

REPORT # 21

4/15/70

CONTRACT FINANCIAL REPORT (Dollars in thousands) (See instructions before preparation)	1 For Month Ended April 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.	Actual/ Estimated Current Month	Cumulative Actual/ Estimated To Date				
	a	b	c	a	b	c	d
Procon Incorporated W-1784	$ 4,400	$ 600	$ 5,000				
Less 10% Retention applicable	436	60	496				
	3,964	540	4,504				
(Above Includes 66 Thousand Dollars previously reported as W-1784-X-1)							

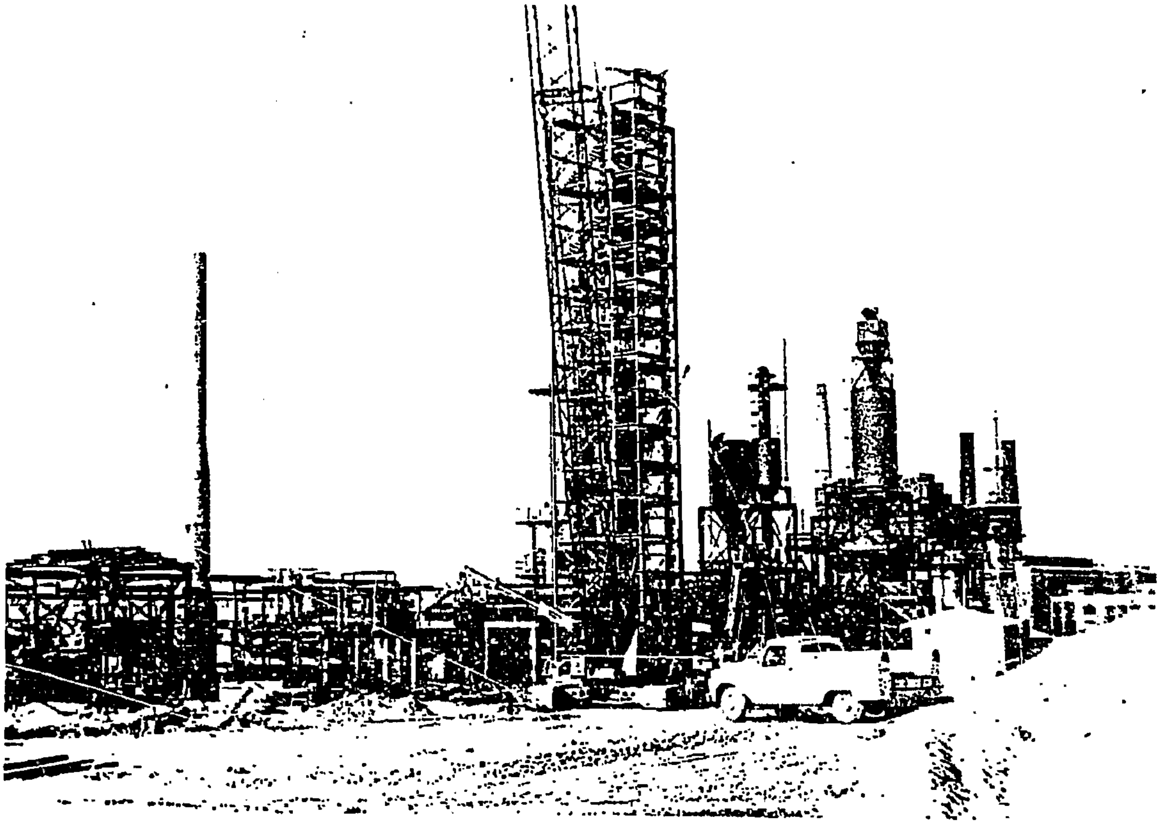
"The undersigned certifies that the amount is due and payable to Procon, in accordance with the terms of the contract up to the date of this Certificate and that the contractor has fully complied with the terms and conditions of the contract."

Progress Report No. 21, April 15, 1970

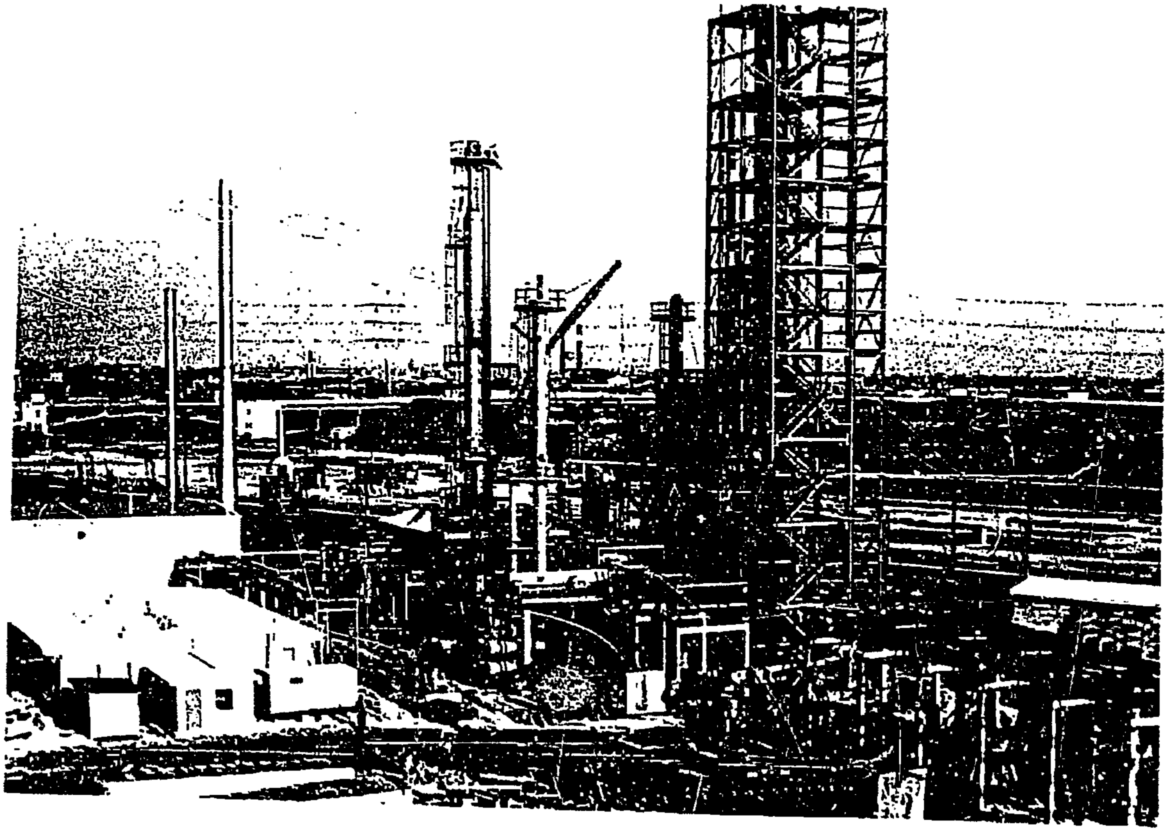
T A Taylor

T. A. Taylor

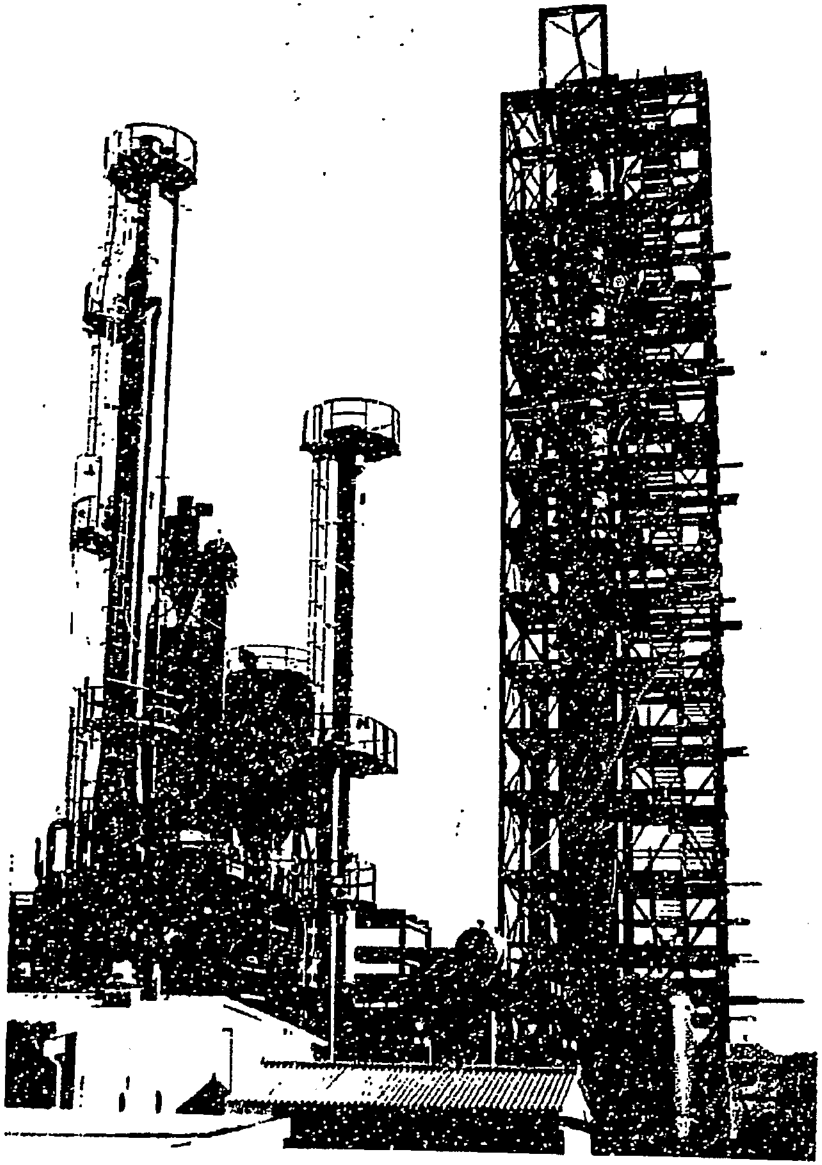
17 Total							



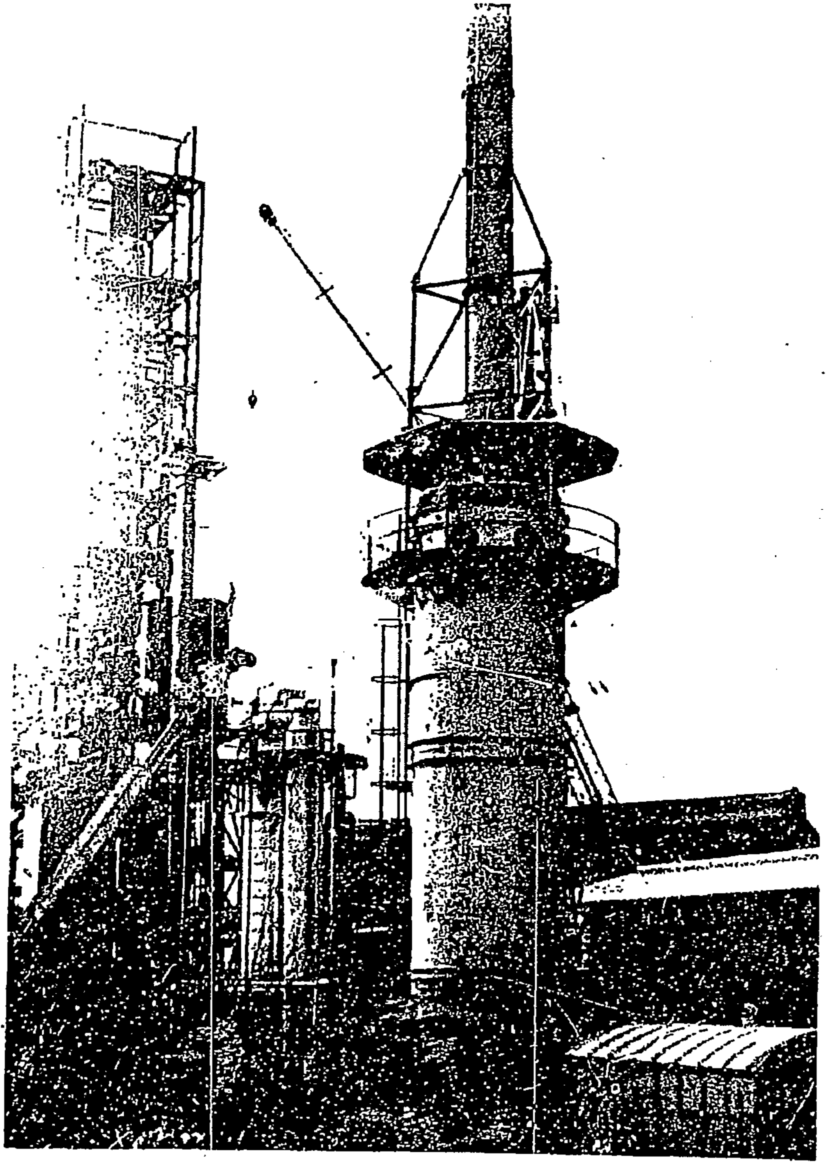
OVERALL PLANT - LOOKING NORTHEAST



OVERALL PLANT - LOOKING SOUTHEAST



PROCESS AREA - LOOKING SOUTH-SOUTHEAST



HYDROGEN PLANT - LOOKING NORTHWEST

IGT-MPR--5/70

IGT
EDUCATION RESEARCH

INSTITUTE OF GAS TECHNOLOGY · IIT CENTER · CHICAGO 60616

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

Report For May 1970
OCR Report No. 68

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

Microtumbler tests to date indicate that the attrition resistance of gasified residue does not change and, in fact, may increase slightly with conversion. It is possible that the expected decrease in strength caused by removal of material by gasification throughout the particle is compensated for by an increase in strength caused by coking or graphitization of the particle.

We are continuing to determine the distribution of minor products of hydrogasification. Ammonium and bicarbonate ions are the major ones in the effluent quench water, but phenol and cyanide have also been noticed.

HIGH-PRESSURE METHANATION

The brief study of ethylene hydrogenation is complete. The purpose of the study was to see if such a process can be adapted for starting up the pilot plant hydrogasifier. We tested both Ni-Mo catalyst and ammonia synthesis catalyst, the former being more active. Ethylene can be hydrogenated to ethane at room temperature over Ni-Mo catalyst.

DEVELOPMENT UNIT STUDIES

The study of Montana subbituminous coal was completed this month. Operating at 500 psi instead of 1000 psi with either hydrogen-steam or synthesis gas-steam mixtures definitely showed that 1000 psi is the desired pressure: Operation was smooth at 1000 psi, but erratic at 500 psi.

Plans are being made to examine the flow patterns and pressure balance in a model of the upper section of the HYGAS hydrogasifier. The model will permit preliminary study of any future modifications.

We are modifying the electrothermal gasifier to test a silicon carbide tube as the outer electrode and also to install a magnetic flip coil to suppress arcing and reduce current fluctuations. Work to date indicates that the 2-MW electrothermal gasifier will have a direct-current power supply and will employ the concentric electrode configuration.

PILOT PLANT CONSTRUCTION

Engineering is 98% complete, purchasing is 95% complete, material receipt is 94% complete, and construction is 62% complete. Due to the prolonged truck strike in the Chicago area, receipt of material needed for construction has been seriously delayed. The revised estimate of the mechanical completion date of the plant is now September 1. All efforts will be made to improve this date.

With Procon we have begun the design and construction of a 2-MW electrothermal gasifier system for the HYGAS pilot plant. The piping and instrument diagram was issued for review. The requisition for the reactor vessel was sent out for quotation. The reactor will be built aboveground instead of in a pit where solids transfer might be easier because the cost of underground construction was prohibitive. For safety purposes, the reactor will have a water jacket similar to the hydrogasifier's.

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)	$ 183,000

2. OCR Funding

A. Funds Expended This Month (estimated)	$ 540,000
B. Funds Expended Since Contract Amendment No. 1 (estimated)	$6,360,000

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
Jack Huebler
Vice-President

Appendix. Achievements in May

COAL CHARACTERIZATION

Microtumbler tests for attrition resistance were run on additional sieve fractions of several samples of hydrogasification residue. Table 1 gives these results and those of tests previously run on −20+30 sieve fractions and reported in the April 1970 Project Status Report. These tests were made with a 2-gram sample, 800 revolutions of the tumbler, and 12-5/16-inch-diameter polyethylene balls.

We consider the amount of −200-mesh fines produced in the test to be the most significant test result. Residue from lignite again shows the least attrition of the residues tested. Residues from FMC char and bituminous coal show more attrition, with the bituminous coal residue having the most. Tests on residues of bituminous coal gasified to different conversions (Runs HT-154, HT-210, EG-34) again show slightly decreasing attrition with increasing conversion. Residue from bituminous coal shows a slight increase in attrition with decrease in particle size. However, it is possible that neither of these trends is statistically significant.

Nevertheless, it is apparent that the constant or slight increase in attrition resistance with increase in conversion is characteristic of the residue as a whole and cannot be attributed to the unrepresentative behavior of the −20+30 sieve fraction. Perhaps the decrease in strength expected because of the removal of material by gasification throughout the particle is compensated for by an increase in strength because of a coking type of reaction.

We continued determination of minor components produced in the hydrogasification process. The yield of ammonia in Run HT-244 with Montana subbituminous coal as feed was 7.2 pounds NH_3 per ton of dry coal, corresponding to 32% of the nitrogen in the coal. The recalculated yield of ammonia in Run HT-243 with the same coal was 14.2 pounds NH_3 per ton of dry coal, corresponding to 64% of the nitrogen in the coal. These conversion values are based on an average nitrogen content of the coal.

The condensate water from Run HT-EG-7 with pretreated Ireland mine coal as feed contained 5 g/ℓ of phenol and 359 mg/ℓ of cyanide.

Work continued on the behavior of ammonia in the quench water system. The variation with temperature of the ionization constants of ammonia,

Table 1. MICROTUMBLER ATTRITION TESTS WITH POLYETHYLENE BALLS ON HYDROGASIFICATION RESIDUE AND RELATED SAMPLES

Sample	Carbon Conversion, wt %	Sieve Fraction, USS	Ash, wt %	Sieve Analysis of Tumbled Sample On USS 30	USS −30+100	USS −100+200	USS −200
				wt %			
Ireland Mine Coal	--	−20+30	--	82.9	14.9	0.6	1.6
Residue, Ireland Mine Coal, Run HT-154	24.6	Unsieved	18.1	--	--	--	--
		−20+30	--	19.8	52.6	12.6	15.0
		−40+60	18.9	--	65.0	18.7	16.3
		−80+100	--	--	52.9	30.0	17.1
Residue, Ireland Mine Coal, Run HT-210	41.7	Unsieved	19.4	--	--	--	--
		−20+30	23.3	31.6	43.9	13.5	11.0
		−40+60	21.8	--	70.3	15.2	14.5
		−80+100	27.4	--	50.6	31.1	18.3
Residue From Electrogasification, Run EG-34	37.2*	Unsieved	25.5	--	--	--	--
		−20+30	--	42.0	39.8	6.8	11.4
		−40+60	22.9	--	76.3	11.1	12.6
		−80+100	22.6	--	59.5	27.3	13.2
Residue From N.D. Lignite, Run HT-192	33.9	Unsieved	14.0	--	--	--	--
		−20+30	12.5	45.5	44.4	4.0	6.1
				46.2	42.9	3.6	7.3
		−40+60	13.9	--	88.2	6.3	5.5
		−80+100	16.4	--	69.8	23.8	6.4
FMC Char, Feed to EG-26	--	Unsieved	15.8	--	--	--	--
		−20+30	--	69.5	20.8	4.9	4.8
		−40+60	16.7	--	84.1	7.6	8.3
		−60+80	17.2	--	78.2	12.5	9.3
Residue, Indiana No. 6 Coal, Run HT-161	31.9	Unsieved	16.0	--	--	--	--
		−40+60	--	28.3	48.3	8.1	15.3
Pretreated Pocahontas No. 4	--	−40+60	--	27.6	53.5	9.7	9.2

* Electrogasification only. Feed was residue from Indiana No. 6 coal.

hydrogen sulfide, and carbonic acid are shown in Figure 1. Values for hydrogen sulfide have been extrapolated from 60°C up to 120°C. Henry's law constants calculated from data in the literature for the same components are shown in Figure 2. These values supersede those given in the First Quarter, 1970, Project Status Report.

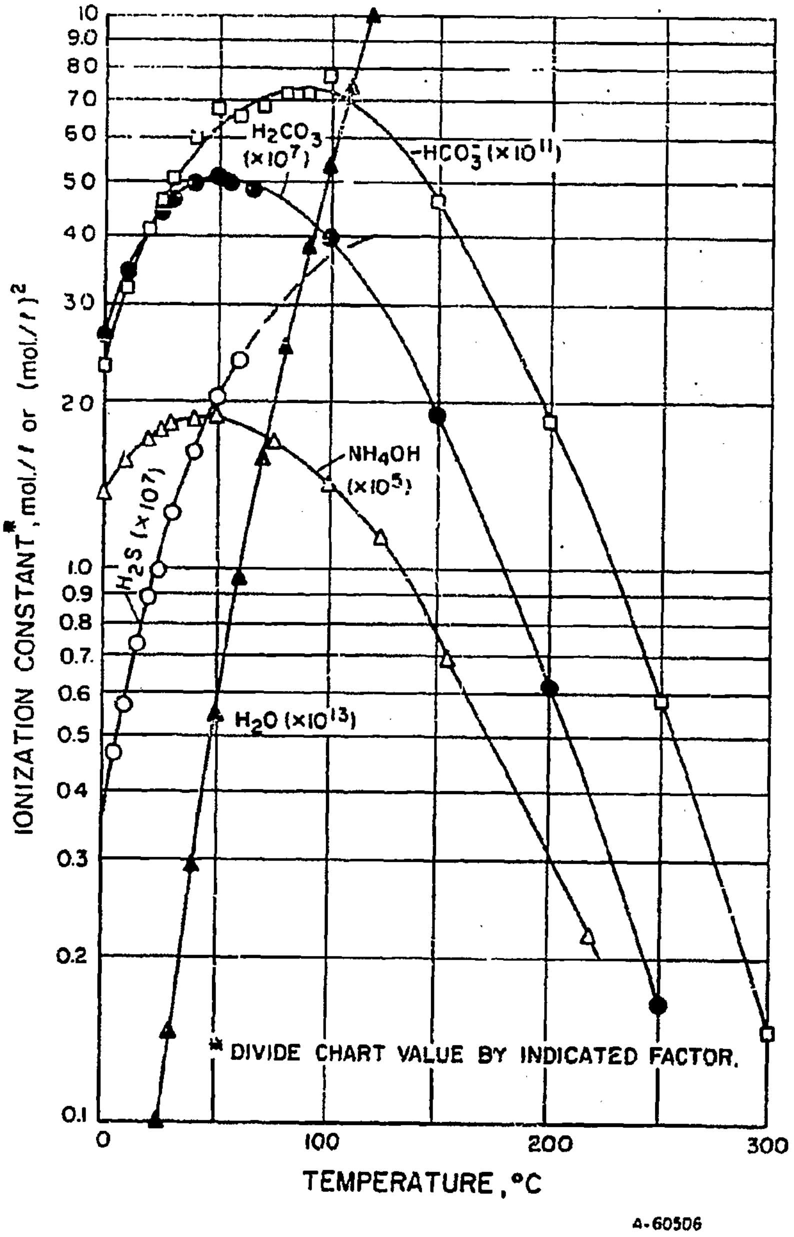


Figure 1. IONIZATION CONSTANTS

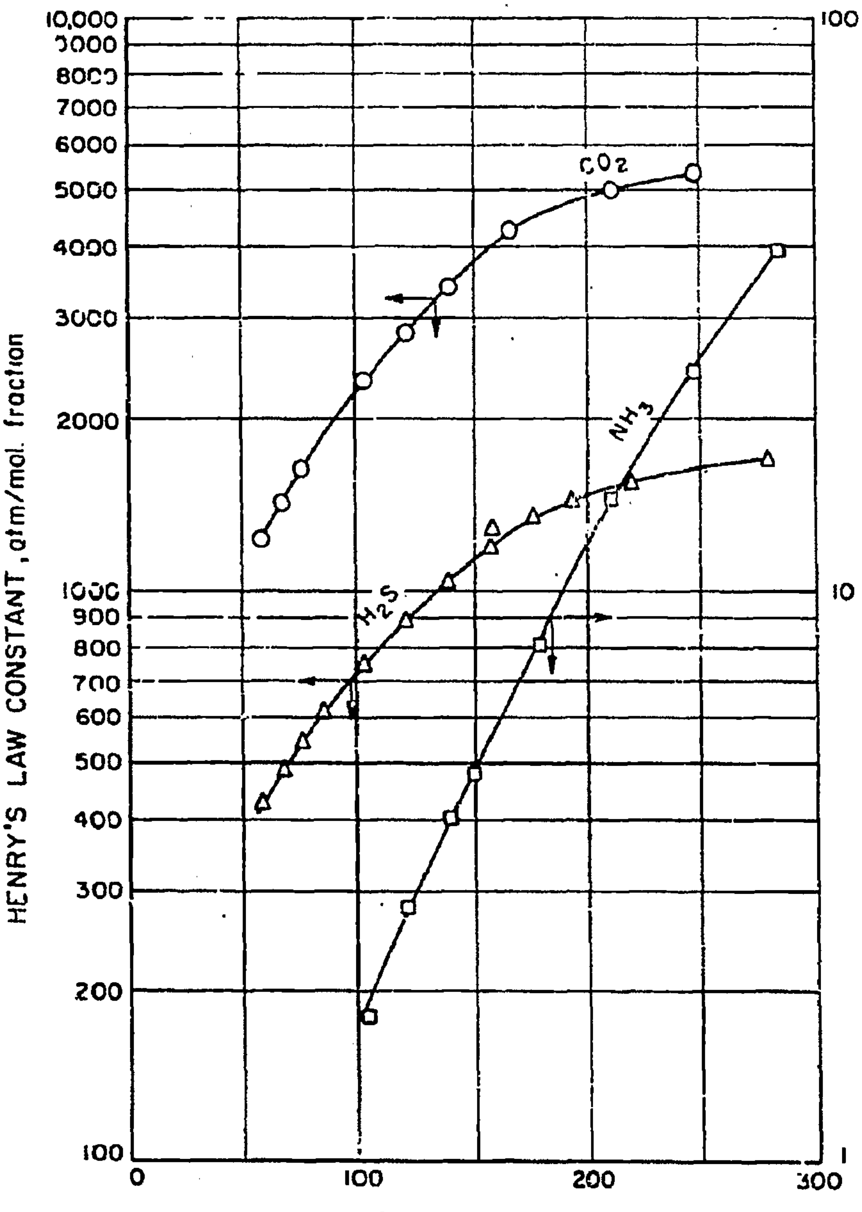


Figure 2. SOLUBILITY OF GASES IN WATER

With these constants the equilibrium concentrations of ammonia, carbon dioxide, and hydrogen sulfide in the quench water can be calculated with the assumption that the only ions (or compounds) formed in addition to the dissolved gases are NH_4^+, HCO_3^-, and HS^-. The second ionization constants of

H_2CO_3 and H_2S are small enough that the amounts of carbonate and sulfide ions are neglibible. However, partial pressure data for the system at temperatures from 68° to 140°F and pressures up to 1 atmosphere indicate that substantial amounts of other components or ions are present. Van Krevelen et al.[2] correlated their data in the form of apparent equilibrium constants, including one for the formation of carbamate ions:

$$K = \frac{(NH_2COO^-)}{(NH_3)(HCO_3^-)}$$

However, since this equilibrium "constant" changed tenfold over the range of ionic strength from 0 to 3 (normality), we hesitate to extrapolate this to our conditions of higher temperature and pressure, for which we have found no data in the literature.

As a first approximation, we have ignored such forms as carbonate in the calculation of concentrations of dissolved gases and NH_4^+, HCO_3^-, and HS^- ions in the quench water of the 500 billion Btu/day lignite plant of the study by Tsaros et al.[1] In this calculation (Table 2) we assumed that the quench water is cooled in a closed circuit so that the ammonia concentration in the quench water will build up until the ammonia in the wastewater equals the ammonia input, that all the ammonia in the gas is absorbed, and that the quench water leaving the tower at 250°F is in equilibrium with the partial pressure of carbon dioxide and hydrogen sulfide in the entering gas.

Table 2. COMPONENTS IN QUENCH WATER STREAMS

	In Quench Water From Tower				In Wastewater	
	Ammonia Rate, moles/hr					
	500		2000		500	2000
Component	Mole Fraction X 10^3	Moles/hr	Mole Fraction X 10^3	Moles/hr	Moles/hr	Moles/hr
NH_4^+	4.71	--	11.8	--	--	--
NH_3 Dissolved	3.15	--	19.6	--	--	--
NH_3 Total	7.86	5710	31.4	22,800	500	2000
HCO_3^-	4.71	--	11.8	--	--	--
CO_2 Dissolved	1.98	--	1.98	--	--	--
CO_2 Total	6.69	4860	13.8	10,000	426	878
HS^-	0.202	--	0.350	--	--	--
H_2S Dissolved	0.066	--	0.046	--	--	--
H_2S Total	0.268	195	0.396	288	17	25

References Cited

1. Tsaros, C. L., Arora, J. L., Lee, B. S., Pimental, L. S., Olson, D. P. and Schora, F. C., "Cost Estimate of a 500 Billion Btu/Day Pipeline Gas Plant Via Hydrogasification and Electrothermal Gasification of Lignite," R&D Rep. No. 22, Interim Rep. No. 4. Washington, D.C.: Office of Coal Research, 1968.

2. Van Krevelen, D. W., Hoftijzer, P. J. and Huntjens, F. J., "Composition and Vapour Pressures of Aqueous Solutions of Ammonia, Carbon Dioxide, and Hydrogen Sulfide," Rec. Trav. Chim. Pays-Bas 68, 191-216 (1949).

HIGH-PRESSURE METHANATION

Ethylene Hydrogenation

Experiments of ethylene hydrogenation with a catalyst were continued. The results are presented in Table 3. Runs EH-8 to EH-11 were made with the same batch of catalyst as those used for Runs EH-5 to EH-7, i.e., HT-100 (−100+200 mesh, 50% by weight) and sand (−100+200 mesh, 50% by weight). Runs EH-8 to EH-10 were made to study the temperature dependency and extent of ethylene hydrogenation because the heat of reaction depends on the product formed, as shown in Figure 3. Ethylene can be hydrogenated at 70°F (Run EH-10), although the product is mainly ethane. Figure 4 shows the effect of temperature on the degree of hydrogenation. Run EH-11 was conducted at a H_2/C_2H_4 ratio of 2, as shown in Figure 5. The effect on the rate of methane formation is small. When the bed was emptied after Run EH-11, carbon was found in abundance. The bed was then repacked with the same catalyst, which had been held at 1800°F for 48 hours. After the catalyst was reduced with H_2 at 650°F for 4 hours, Run EH-12 was made. The catalyst was still active, but not as active as fresh ones (compare to Run EH-5).

An ammonia synthesis catalyst was also tested. The results showed that it is not as active as Ni-Mo catalyst (Runs EH-13 and EH-14).

Assuming an Arrhenius equation for the methanation formation rate constant, we have —

$$r_i = k_o e^{-E/RT} h(x)$$

where, k_o = frequency factor

E = activation energy of the reaction

$h(x)$ = general function showing dependence on composition

Table 3. EXPERIMENTAL DATA FROM ETHYLENE HYDROGENATION

	Run No.						
Feed	EH-8	EH-9	EH-10	EH-11	EH-12	EH-13	EH-14
Press., psig	510	510	505	510	510	300	300
Temp, °F							
Furnace	450	250	70	600	650	600	390
9	110	75	70	130	520	140	105
10	430	235	70	690	725	650	400
11	455	205	70	710	725	--	--
Flow Rate, SCF/hr	16.9	17.0	17.1	14.6	16.6	11.2	11.8
Composition, mole %							
H_2	79	78	79	67	80	72	76
C_2H_4	21	22	21	33	20	28	24
Product							
Temp, °F							
Furnace	450	235	100	630	690	635	360
9	480	460	450	535	385	120	100
10	490	335	290	700	900	765	525
11	450	235	170	700	900	--	--
Flow Rate, SCF/hr	12.8	12.6	13.3	9.2	11.6	9.7	9.4
Composition, mole %							
H_2	67.0	69.6	71.2	41.7	64.7	66.8	57.8
CH_4	3.9	0.7	0.6	15.2	0.1	13.5	12.0
C_2H_6	28.7	29.2	27.8	41.9	32.9	16.3	22.8
C_3H_8	0.1	0.1	0.1	0.4	0.2	0.2	0.3
C_2H_4	0.3	0.4	0.3	0.7	1.2	2.9	7.0
C_3H_6	--	--	--	--	0.1	0.2	0.1
$\underline{n}$-C_4H_{10}	--	--	--	0.1	0.5	0.1	--

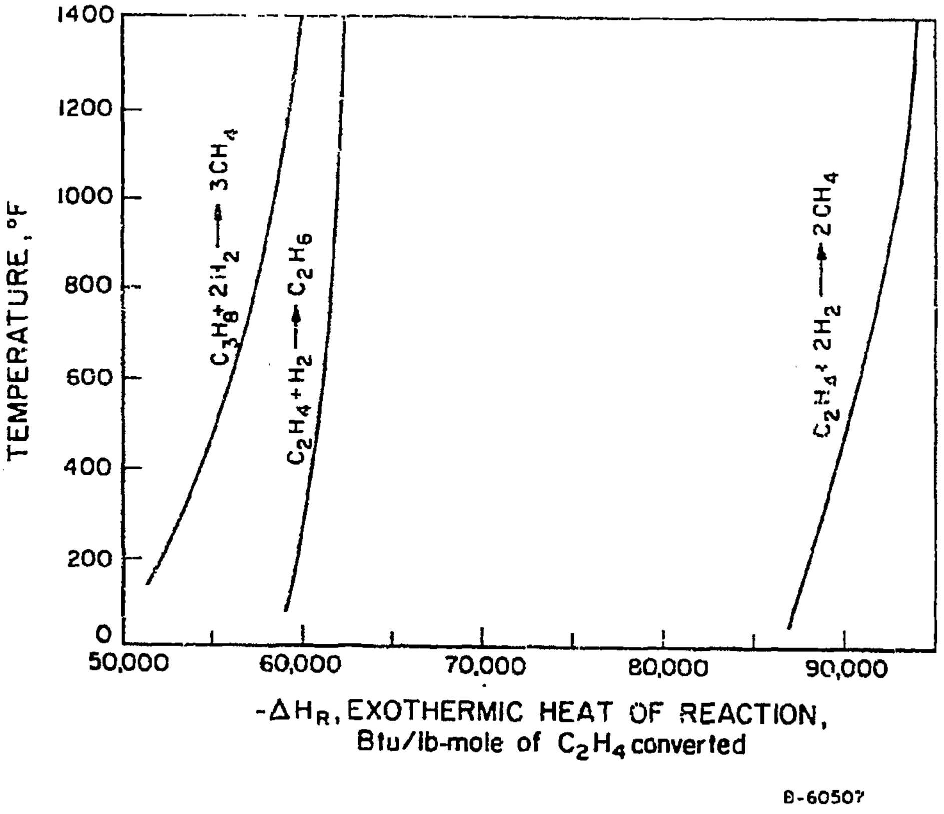


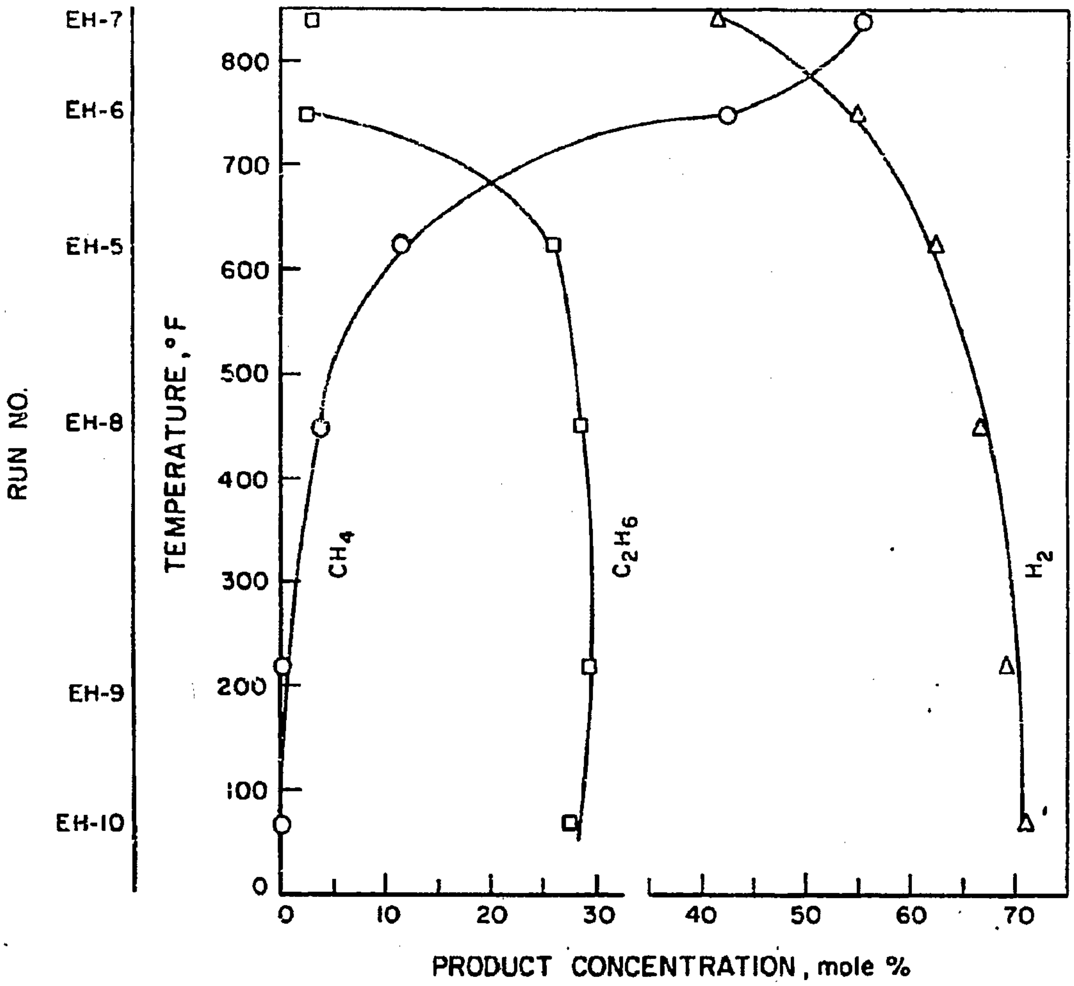
Figure 3. EFFECT OF TEMPERATURE ON THE HEAT OF REACTION

Thus, for compositions that vary in a narrow range, or for reactions that depend little on composition, a plot of $\ln r_i$ vs. $1/T$ should be linear. Figure 5 shows this temperature dependance.

DEVELOPMENT UNIT STUDIES

Hydrogasification Tests

We performed two hydrogasification tests this month in the high-temperature, balanced-pressure development unit. These two tests, Runs HT-247 and HT-248, conclude our current studies of the reactivity of a dried, but otherwise untreated, Montana subbituminous coal to hydrogasification at a system pressure of 500 psig. Reaction of the coal to gasification with hydrogen and steam was investigated in Run HT-247, and gasification with synthesis gas and steam was investigated in Run HT-248. Although both runs were comparatively short because of operating difficulties, adequate data were obtained for a meaningful evaluation of the tests.



A-60508

Figure 4. EFFECT OF TEMPERATURE ON DEGREE OF ETHYLENE HYDROGENATION

Significant features of the two tests are given in Table 4.

The operating conditions of Run HT-247 were similar to those of Run HT-244 (May 1970 Project Status Report). We repeated the earlier test to obtain a longer steady-state operating period. The nominal conditions were 51.6 lb/hr coal, 345 SCF/hr hydrogen (20% stoichiometric hydrogen/coal ratio), and 8.8 lb/hr steam (35 mole percent concentration in the feed gas). To minimize the coal agglomeration difficulties experienced in Run HT-244, the

Table 5, Part 1. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF PRETREATED BITUMINOUS COAL AND DRIED SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR

Coal	Montana Subbituminous Coal	Ireland Mine Bituminous Coal	Montana Subbituminous Coal
Source	Colstrip Mine	IGT Pretreated, FP-142	Colstrip Mine
Sieve Size, USS	—10+80		
Run No.	HT-244	HT-EG-7	HT-247
Duration of Test, hr	1-1/4	6-1/4	2-1/4
Steady-State Operating Period, min[a]	60-75	95-381	123-141
OPERATING CONDITIONS			
Bed Height, ft	3.5	3.5	3.5
Reactor Pressure, psig	488	1034	549
Reactor Temperature, °F[b] Inches From Bottom			
62-1/2	1020	1315	1125
67-3/4	1160	1535	1340
73	1325	1680	1515
78-1/4	1435	1510	1405
83-1/2	1495	1605	1560
89	1335	1525	1425
94-1/2	1420	1570	1515
100	1690	1680	1615
104	1570	1680	1625
Average	1385	1565	1460
Coal Rate, lb/hr[c]	69.45	64.77	54.60
Feed Gas Rate, SCF/hr	343.8	467.5	377.9
Steam Rate, lb/hr	9.22	25.08	8.90
Steam, mole % of hydrogen-steam mixture	36.0	53.0	33.1
Hydrogen/Coal Ratio, % of stoichiometric[d]	14.2	19.6	19.9
Hydrogen/Steam Ratio, mole/mole	1.77	0.887	2.02
Bed-Pressure Differential, in. wc	48.0	104.0	--
Coal Space Velocity, lb/cu ft-hr	224.5	118.9	176.5
Feed-Gas Residence Time, min[e]	0.332	0.341	0.341
Superficial Feed-Gas Velocity, ft/s[f]	0.176	0.171	0.173

Table 5, Part 2. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF PRETREATED BITUMINOUS COAL AND DRIED SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR

Run No.	HT-244	HT-EG-7	HT-247
OPERATING RESULTS			
Product Gas Rate, SCF/hr	714.8	993.9	889.5
Net Btu Recovery, 1000 Btu/lb	2.497	3.456	3.132
Product-Gas Yield, SCF/lb	10.29	15.34	16.29
Hydrocarbon Yield, SCF/lb	1.88	3.68	2.80
Carbon Oxide Yield, SCF/lb	2.72	2.22	2.22
Net Reacted Hydrogen, SCF/lb	1.153	2.66	1.708
Residue, lb/lb coal[g]	0.591	0.627	0.517
Liquid Products, lb/lb coal[h]	0.184	0.344	0.230
Net MAF Coal Hydrogasified, wt %[i]	34.7	36.2	36.2
Carbon Gasified, wt %	23.5	29.1	26.1
Steam Decomposed, lb/hr[j]	nil	3.22	nil
Steam Decomposed, % of steam fed	nil	12.8	nil
Steam Decomposed, % of total equivalent fed[k]	59.8	35.2	50.4
Overall Material Balance, %	94.1	95.9	94.0
Carbon Balance, %	97.3	98.4	92.6
Hydrogen Balance, %	82.6	94.9	88.7
Oxygen Balance, %	88.8	93.9	90.1
PRODUCT GAS PROPERTIES			
Gas Composition, mole %			
Nitrogen	18.1	31.6	36.9
Carbon Monoxide	20.2	8.3	8.5
Carbon Dioxide	6.2	6.2	5.1
Hydrogen	36.9	29.7	32.0
Methane	16.7	23.5	15.9
Ethane	1.0	0.4	0.9
Propane	0.6	0.1	0.4
Butane	--	--	--
Benzene	0.3	0.2	0.3
Hydrogen Sulfide	--	--	--
Total	100.0	100.0	100.0
Heating Value, Btu/SCF[m]	392	371	325
Specific Gravity (Air = 1.00)	0.612	0.644	0.653
Nitrogen Purge Rate, SCF/hr	129	314	328

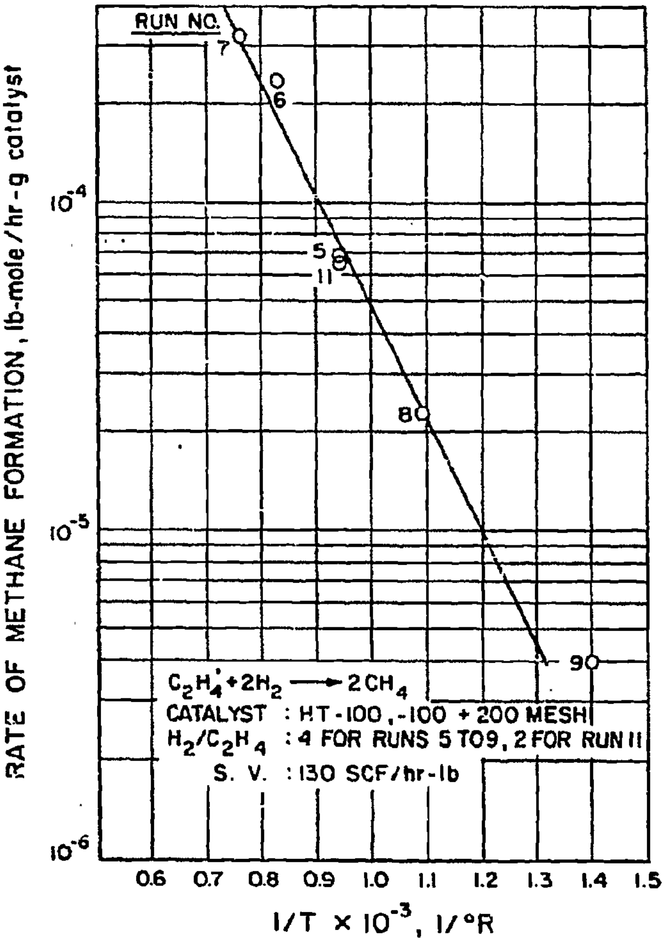


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Figure 5. TEMPERATURE DEPENDENCY OF ETHYLENE HYDROGENATION REACTION

test was started up with only hydrogen being fed to the bottom of the reactor. Steam feed was started only after the 3.5-foot fluidized coal bed was established. About 1-1/4 hours after coal feed was started, the coal feed screw stopped. Twelve minutes later the jamming was cleared and the feed screw was restarted. Feeding continued for 55 minutes before the screw stopped again. Light agglomeration of the coal at the top of the reactor was basically responsible for the plugging. A contributing cause was the loss of the stirrer in the coal injection tube. The stirrer had broken off at its connection to the

Table 4. FEATURES OF HYDROGASIFICATION TEST RESULTS FOR RUNS HT-247 AND HT-248

Run No.	Temperature, °F	Purpose of Run	Results
Feed Solids: Dried Montana Subbituminous Coal, Colstrip Mine			
HT-247	1300-1700	To study the hydrogasification reactivity of a subbituminous coal with hydrogen and steam at 500 psig	Successful, short duration
HT-248	1300-1700	Same as HT-247, except with synthesis gas and steam feed gas	Successful, short duration

drive shaft before the run was started. The duration of the test was 2-1/4 hours with a 20-minute steady-state period.

The evaluation of coal hydrogasification with synthesis gas and steam, Run HT-248, was made at conditions similar to those of Run HT-243 (April 1970 Project Status Report) to verify the results of the earlier test, which was of relatively short duration. Nominal feed rates were 21 lb/hr coal, 265 SCF/hr synthesis gas (54% hydrogen, 41% carbon monoxide, 5% carbon dioxide), and 12.6 lb/hr steam (50 mole percent concentration). The test lasted over 2-3/4 hours before we terminated it because of light agglomeration of the coal at the outlet of the coal feed tubes. A steady-state operating period of 3/4 hour was obtained. A failure of the temperature controls of Zone 6 of the reactor furnace, the lower heating zone, limited the average coal-bed temperature to 1295°F.

Complete hydrogasification results of Run HT-248 will be presented when analyses of this test are completed.

Operating conditions and results of Run HT-247, and those of Run HT-244 (March 1970 Project Status Report) and Run HT-EG-7 (April 1970 Project Status Report) are presented in Table 5. Run HT-244 was conducted with Montana subbituminous coal at conditions similar to those of Run HT-247. Run HT-EG-7 was conducted with lightly pretreated Pittsburgh seam bituminous coal from the Ireland mine to produce a partially hydrogasified char for use as feed in the electrothermal gasifier development unit studies. Compositions and screen analyses of the feeds and residues of these tests are given in Table 6. Liquid products and compositions are shown in Table 7.

Table 5, Part 3. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF PRETREATED BITUMINOUS COAL AND DRIED SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR

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 feed gas 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 6. CHEMICAL AND SCREEN ANALYSES OF PRETREATED BITUMINOUS COAL AND DRIED SUBBITUMINOUS COAL FEED AND RESIDUE

Run No.	HT-244		HT-EG-7		HT-247	
Sample	Feed	Residue	Feed	Residue	Feed	Residue
Proximate Analysis, wt %						
Moisture	3.7	1.0	1.6	0.9	1.6	0.9
Volatile Matter	36.5	6.9	22.4	2.4	37.1	3.1
Fixed Carbon	51.8	78.1	61.7	73.7	53.2	80.2
Ash	8.0	14.0	14.3	23.0	8.1	15.8
Total	100.0	100.0	100.0	100.0	100.0	100.0
Ultimate Analysis (dry), wt %						
Carbon	67.7	79.6	67.5	73.7	68.0	80.4
Hydrogen	4.22	1.88	3.07	0.92	4.34	1.08
Nitrogen	0.93	0.75	1.31	0.55	1.01	0.50
Oxygen	17.80	3.12	10.41	0.00	17.55	1.39
Sulfur	1.01	0.53	3.15	1.74	0.84	0.64
Ash	8.34	14.12	14.56	23.22	8.26	15.99
Total	100.00	100.00	100.00	100.13	100.00	110.00
Screen Analysis, USS, wt %						
+20	3.3	8.2	23.1	16.0	3.0	11.3
+30	22.4	16.2	20.9	20.1	23.5	13.3
+40	24.5	22.5	19.0	16.6	26.5	24.3
+60	33.0	29.2	24.5	27.5	30.0	31.7
+80	11.8	13.1	10.0	12.8	11.6	11.9
+100	2.9	4.5	1.7	3.2	2.6	3.4
+200	1.7	5.1	0.6	2.7	2.5	3.5
+325	0.2	0.6	0.2	0.5	0.2	0.4
–325	0.2	0.6	0.0	0.6	0.1	0.2
Total	100.0	100.0	100.0	100.0	100.0	100.0

Table 7. COMPOSITION OF HYDROGASIFICATION LIQUID PRODUCTS

Run No.	HT-244	HT-EG-7	HT-247
Sample	Condenser	Condenser	Condenser
Liquid Products, lb/lb coal	0.1840	0.344	0.230
Composition of Liquid Products, wt %			
Water	81.31	98.08	81.31
Oil	18.69	1.92	18.69
Total	100.00	100.00	100.00
Composition of Oil Fraction, wt %			
Carbon	85.30	87.00	85.90
Hydrogen	7.76	6.65	7.79
Total	93.06	93.65	93.69
Carbon in Oil Fraction,			
lb/lb coal	0.0293	0.00575	0.0369
wt % of carbon in coal	4.33	0.85	5.43

The results of Runs HT-244 and HT-247 are similar, indicating good reproducibility. Carbon gasification was 24% in Run HT-244 and 26% in Run HT-247. Carbon conversion to oils was only somewhat larger in Run HT-247, with 5.43% of the carbon in the coal converted to oil, compared to 4.33% for Run HT-244 (Table 7). Partially responsible for the somewhat larger carbon conversions of Run HT-244 are the higher average coal-bed temperature (1460°F compared to 1385°F) and the larger hydrogen-to-coal ratio (19.9% of stoichiometric versus 14.2% of stoichiometric). (See Table 5.)

The effect of pressure on the Montana subbituminous coal hydrogasification results is shown by a comparison of the key results of Runs HT-244 and HT-247 with those of Run HT-216, conducted earlier at similar conditions except for 1000 psig (Table 8). The carbon gasification, hydrocarbon yield, and product-gas heating value are significantly greater at 1000 psig than at 500 psig.

The results of Run HT-EG-7 are similar to those of other tests with the Ireland mine bituminous coal conducted at similar hydrogasification conditions.

Table 8. COMPARISON OF MONTANA SUBBITUMINOUS COAL HYDROGASIFICATION RESULTS WITH HYDROGEN AND STEAM AT 500 AND 1000-psig SYSTEM PRESSURE

Run No.	HT-216	HT-244	HT-247
Reactor Pressure, psig	1052	488	549
Coal-Bed Temp Average, °F	1615	1385	1460
Coal Feed Rate, lb/hr	53.74	69.45	54.60
Hydrogen Rate, SCF/hr	461.3	343.8	377.9
Steam Feed Rate, lb/hr	11.80	9.22	8.90
Hydrogen/Coal Ratio, % of stoichiometric	24.80	14.2	19.9
Steam Concentration in Feed Gas, mole %	34.9	36.0	33.1
Steam Decomposed, % of total equivalent steam fed	63.1	59.8	50.4
Carbon Gasified, %	43.1	23.5	26.1
MAF Coal Gasified, %	56.3	34.7	36.2
Hydrocarbon Yield, SCF/lb	5.12	1.88	2.80
CO + CO_2 Yield, SCF/lb	3.37	2.72	2.22
Product-Gas Rate (nitrogen-free), SCF/hr	664.2	585.4	561.3
Product-Gas Composition (nitrogen-free), mole %			
Carbon Monoxide	16.4	24.7	13.5
Carbon Dioxide	10.9	7.6	8.1
Hydrogen	30.7	45.0	50.7
Methane	39.0	20.4	25.2
Ethane	1.8	1.2	1.4
Propane	0.6	0.7	0.6
Benzene	0.6	0.4	0.5
Hydrogen Sulfide	--	--	--
Total	100.0	100.0	100.0
Product-Gas Heating Value (nitrogen-free), Btu/SCF	608	479	514

During the month we began the pretreatment of a Pittsburgh No. 8 seam, Ireland mine bituminous coal. The coal was pretreated with air and nitrogen in a fluidized bed at 750°-800°F (Run FP-143). The pretreated coal will be used in hydrogasification tests to produce a char for use in the electrothermal gasifier development unit.

Plans are being made to examine the solids flow pattern and pressure balance in a model of the upper section of the hydrogasifier designed for the HYGAS pilot plant. Char will be used for circulation at ambient conditions. This model will also permit us to check any proposed flow modifications before incorporating them into the pilot plant.

ELECTROTHERMAL GASIFICATION

During the month, we conducted one test in the electrothermal gasifier at 1900°F and 100 psig using FMC Project COED char as the feed material. Some major modifications are being made on the unit to suppress current fluctuations to a more desirable magnitude.

The purpose of Run EG-55 was to obtain additional data under conditions leading to a high carbon conversion. Abnormally high resistances were observed after briefly reaching operating conditions. Unable to maintain a 1900°F temperature in the fluidized bed, we switched to nitrogen as a fluidizing medium in an attempt to lower the resistance. A plug developed in the exit-gas line, causing a pressure rise throughout the unit. When efforts to clear the exit line were unsuccessful, we terminated the run. During the operating period several thermocouples became inoperable, reducing the number of thermocouples in the bed region to two.

At this point we decided to completely disassemble the reactor for the following reasons:

1. Replacement of thermocouples
2. Installation of a 6.0-inch-ID silicon carbide reactor tube
3. Installation of a magnetic "flip coil"

The first item is self-explanatory. The installation of a silicon carbide tube to act as one of the electrodes in a concentric configuration will provide us with more information on the electrical characteristics of this material in our environment. The last reason requires a more detailed explanation.

The application of a "flip coil" is a direct result of the studies conducted by our two electrical consultants. The goal of these studies was to identify the power supply requirements in conjunction with the electrical characteristics of a 2-MW electrothermal gasifier. In a report submitted to us on May 6, 1970, they presented a summary of the results from tests made during Run EG-53 and a number of conclusions.

Analysis of the data on current oscillations and transient behavior of the system led to the following observations:

1. Current oscillations are in the 2-10 cps range, most objectionable for a flicker.
2. The ratio of peak-to-trough current observed is in excess of 4.
3. No high-frequency disturbances are noted.
4. The gasifier itself should produce no radio frequency interference.
5. Reverse polarity experiments did not confirm that symmetrical operation would be obtained on alternating current.

The report pointed out that the important considerations for the design of the electrothermal gasifier are the power density at which the unit can be operated (impedance and current density) and protection against arcing and flashover, which could damage the apparatus. Similarly important design criteria for the power supply are —

1. Voltage-current (V-I) characteristics of the load
2. Voltage-current linearity
3. Voltage-current symmetry for alternating current
4. Flicker (both 3 Φ and 1 Φ) arising from load instability or fluctuations
5. Phase unknowns

Nonlinearity, phase imbalance, and load asymmetry are not factors if a d-c gasifier is employed, assuming that current fluctuations occur at a frequency of much less than 60 cps, as is our case, and that flicker is still held to an acceptable level. The consultants, therefore, recommended that the 2-MW pilot plant unit utilize the present coaxial configuration with a d-c power supply because —

1. The estimated power supply cost is, at worst, no greater than that for the a-c supply and may be considerably less.

2. The experimental data and other considerations indicate that technical and/or cost factors are much more uncertain with the a-c system.

3. The existing coaxial configuration cannot be used on alternating current because the 3 Φ — 1 Φ conversion at high power level that would be required is impractical.

4. The 2 Φ coaxial and 3 Φ "electric arc furnace" alternative configurations that might be considered will require development work with uncertain results.

The remaining problem is the existence of flicker, which could either be the result of arcing in the bed or other types of disturbances such as "bubbles." If the cause of the flicker is arcing, the use of the magnetic "flip coil" should reduce the fluctuations considerably. In essence the "flip coil" is simply a coil placed inside the reactor near the outside wall of the pressure vessel. Current passing through the coil will induce a magnetic field, which should suppress electrical transients in the gasifier. Three beneficial effects will result:

1. Reduced power supply costs due to reduction of flicker-suppression requirements

2. Increased bed life and reduced likelihood of destructive arcing

3. Increased mean bed impedance leading to higher power density of a given electrode current density

Current fluctuations observed during the experiments were precisely at the worst possible flicker frequencies (2-6 cps) and were at magnitudes comparable to those observed with arc furnaces where flicker alone can add $8-$25/kW to the power supply cost.

A more detailed description of the coil and the silicon carbide tube will follow as soon as the installation is completed. Table 9 presents the operating conditions and results of Run EG-48. Chemical and screen analyses of the feed and residue materials of that same run are in Table 10.

PILOT PLANT CONSTRUCTION

Engineering

Under the original scope of the guaranteed maximum price, the electrical details of the control panel and motor control center interconnecting wiring

Table 9. OPERATING CONDITIONS AND RESULTS OF THE ELECTROTHERMAL GASIFICATION OF COAL CHAR

Run No.	EG-48
Feed Char	HV Bituminous Hydrogasified Char
Sieve Size, USS	−10+80
Duration of Test, hr	5.62
Steady-State Operating Period, min	90
Electrode Material	316 SS
OPERATING CONDITIONS	
Bed Height, ft	2.75
Reactor Pressure, psig	1010
Reactor Temp, °F Inches From Bottom	
39	2035
42	1850
45	1845
48	1900
51	1920
54	1900
57	1980
63	1880
72	1915
Average	1914
Steam Feed Rate, lb/hr	121
Steam Residence Time, min*	0.19
Steam Superficial Velocity, ft/s†	0.24
Steam/Char Feed Ratio, lb/lb	1.45
Char Feed Rate (dry), lb/hr	83.2
Char Residence Time, min	8.8
Nitrogen Purge Rate, SCF/hr	856
Voltage, V	235
Current, A	286
Power Input, kW	67.2
Overall Resistance, ohm	0.82

Table 9, Cont. OPERATING CONDITIONS AND RESULTS OF THE ELECTROTHERMAL GASIFICATION OF COAL CHAR

Run No.	EG-48
OPERATING RESULTS	
Product-Gas Rate (dry), SCF/hr‡	2138
Product-Gas Yield (dry), SCF/lb char	25.7
Hydrogen Yield, SCF/lb	13.1
Hydrogen + Carbon Monoxide Yield, SCF/lb	20.1
Carbon Oxides Yield, SCF/lb	9.7
Char Gasified, wt %	34.5
Carbon Gasified, wt %	47.0
Liquid Products, lb/hr	57.0
Steam Decomposed, lb/hr	64.0
Steam Conversion, wt %	52.9
Overall Material Balance, %	98.8
Carbon Balance, %	109.3
Hydrogen Balance, %	100.0
Oxygen Balance, %	92.6
PRODUCT-GAS PROPERTIES	
Composition, mole %‡	
CO	27.3
CO_2	13.9
H_2	51.1
CH_4	7.7
H_2S	--
Total	100.0
Specific Gravity (Air = 1.00)	0.553

* Coal-bed volume (top 2.75 ft)/SCF steam feed at average reactor temperature and pressure.

† CF/s steam at reactor temperature and pressure/cross-sectional area of reactor.

‡ Dry, nitrogen-free basis.

Table 10. CHEMICAL AND SCREEN ANALYSES OF ELECTROTHERMAL GASIFICATION FEEDS AND RESIDUES

Run No.	EG-48	
Sample	Feed	Residue
Proximate Analysis, wt %		
Moisture	1.2	7.4
Volatile Matter	2.5	1.5
Fixed Carbon	85.2	75.2
Ash	11.1	15.9
Total	100.0	100.0
Ultimate Analysis, wt %		
Carbon	85.90	81.70
Hydrogen	1.02	0.93
Nitrogen	0.69	0.40
Oxygen	0.57	--
Sulfur	0.57	0.11
Ash	11.25	17.17
Total	100.00	100.31
Screen Analysis, USS, wt %		
+20	6.6	2.7
+30	16.3	17.1
+40	17.3	20.8
+60	35.5	32.3
+80	16.8	15.7
+100	4.1	4.9
+200	2.9	5.1
+325	0.1	0.7
--325	0.4	0.7
Total	100.0	100.0

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and the painting specification remain to be completed. Engineering work is under way on several recent additions to the project. The scale model has been completed. Field changes and corrections are being recorded, and "as built" drawings will be made after completion of construction.

Procurement

The instrument/electrical subcontract is being reviewed and will be let shortly. The fireproofing and refractory subcontracts have already been let. The truck strike has caused serious delays in receipt of outstanding material.

Construction

Major field activities during this report period have been field shop pipe fabrication, area piping, final setting and aligning of equipment, coal unloading facilities work, vessel insulation, and electrical work. Piping is approximately 60% complete.

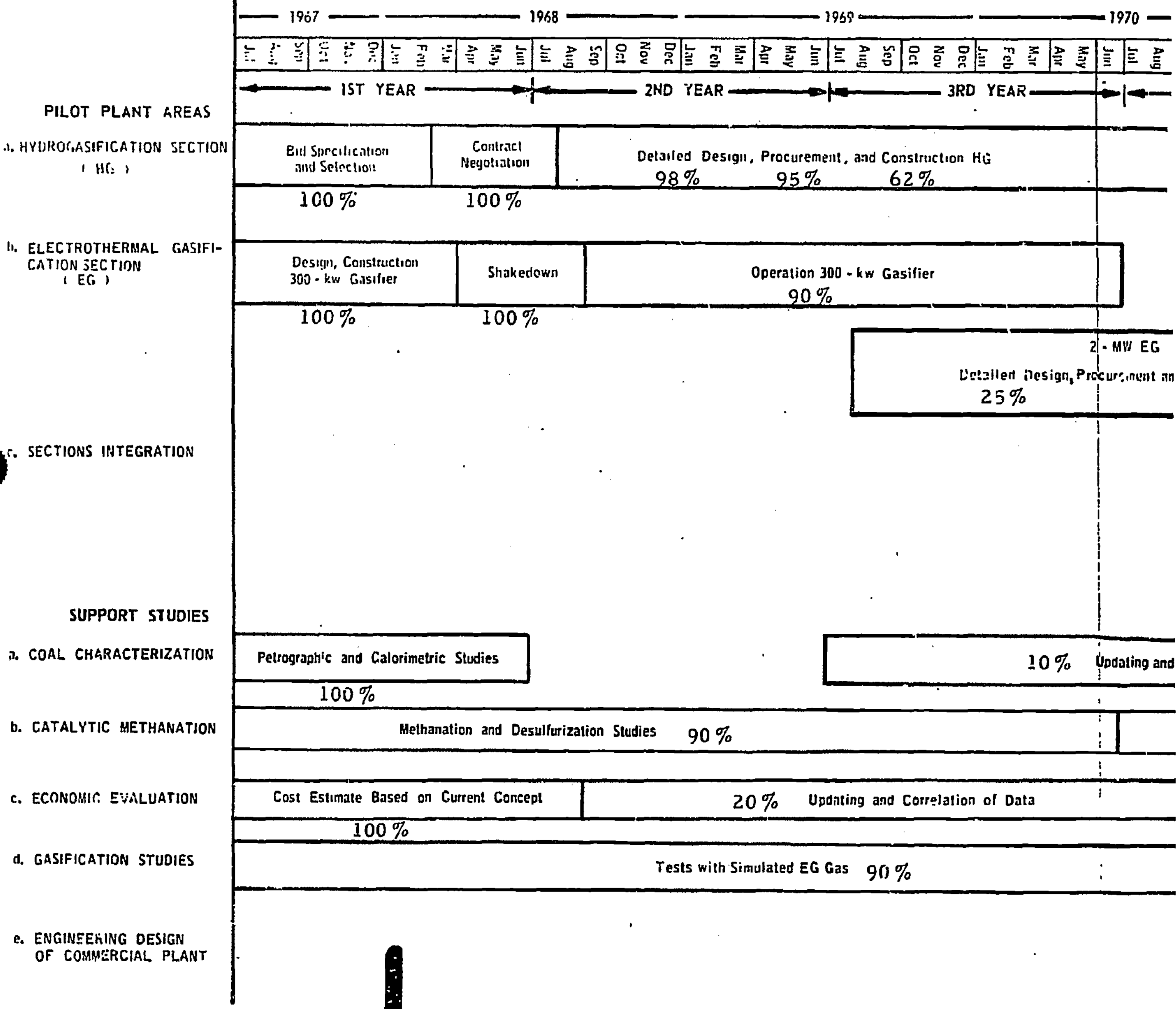
We have experienced a total of 21 inclement weather days, 7 of which occurred in this report period. On these days progress was negligible.

Schedule

The project schedule has been reviewed; mechanical completion is now scheduled for September 1, 1970.

PILOT PLANT PROGRAM OF IGT HYDROGASIFI

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



OF IGT HYDROGASIFICATION PROCESS

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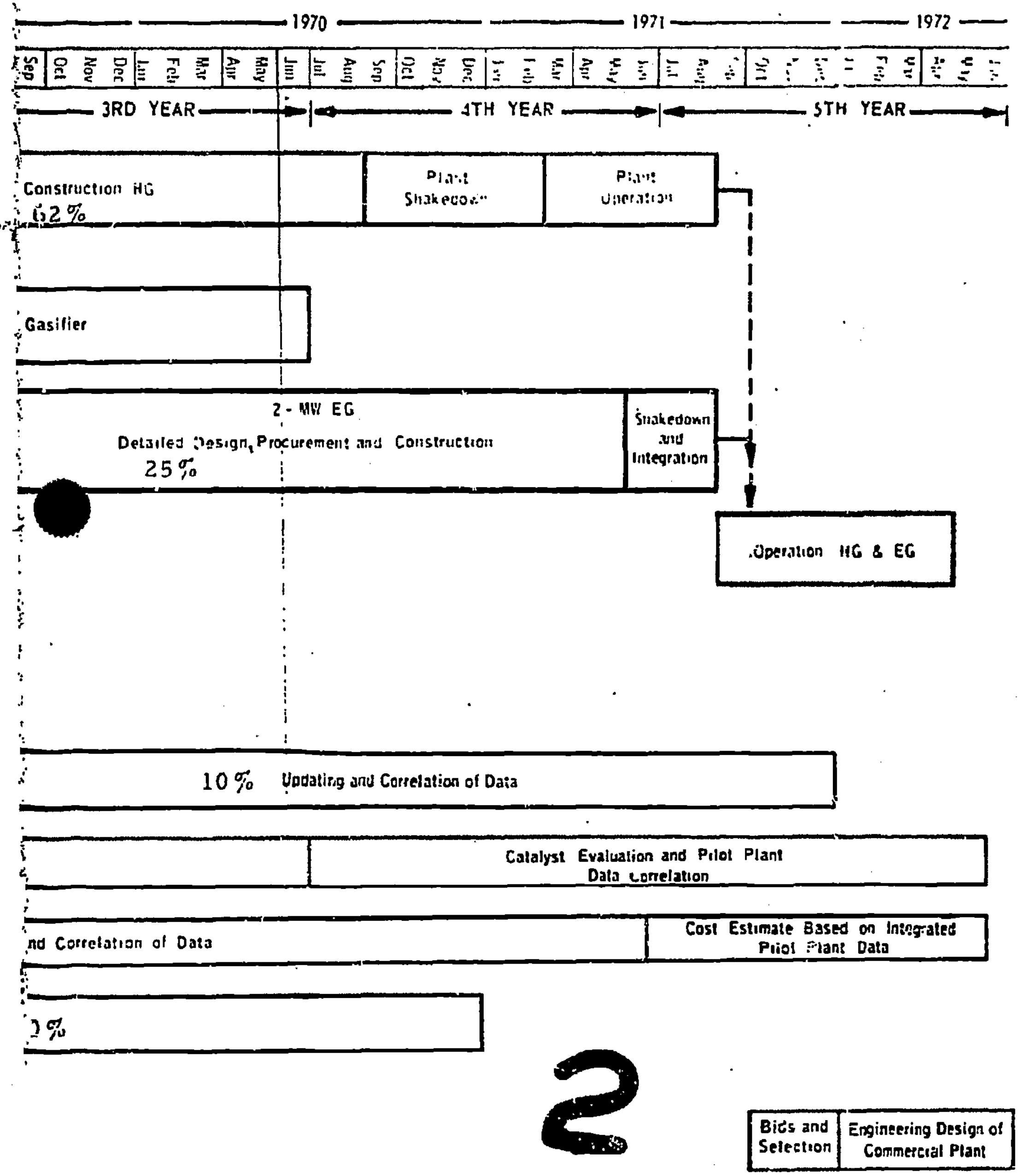

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Project Status Report
For
OFFICE OF COAL RESEARCH
and
AMERICAN GAS ASSOCIATION

Report For Second Quarter, 1970
OCR Report No. 69

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

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. The yield of ammonia in Run HT-248 with Montana subbituminous coal as feed was 14.5 lb NH_3/ton of dry coal, corresponding to 64.9% of the nitrogen in the coal. Since over 75% of the ammonia appears in the recycle quench water, if a cooling tower is used, most of the ammonia is released to the atmosphere. We are studying means of recovering the ammonia and avoiding atmospheric pollution. Phenol and cyanide were also found in the effluent.

To date microtumbler tests indicate that the attrition resistance of gasified residue does not change and, in fact, may increase slightly with conversion. It is possible that the expected decrease in strength caused by removal of material by gasification throughout the particle is compensated for by an increase in strength caused by coking or graphitization of the particle.

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. The temperature dependence of the reaction rate is also being studied through tests at 630° and 865°F.

The laboratory unit to measure the sulfur tolerance of methanation catalysts for use in the pilot plant has been completed. The two pilot plant process gas chromatographs were tied-in with this unit to allow the analyzers to be checked out and tried with gases similar to those that will be encountered in the pilot plant.

Ethylene hydrogenation was briefly studied to see if this type of process can be adapted for starting up the pilot plant hydrogasifier. We tested both Ni-Mo and ammonia synthesis catalysts. Ethylene can be hydrogenated to ethane at room temperature over the more active Ni-Mo catalyst. Without a catalyst the reaction does not occur below 800°F; however, hydrogenation occurring at 1100°F was accompanied by the formation of carbon and tar.

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.

The economics of lock hopper and slurry systems for feeding pretreated char to the hydrogasifier were compared using recent data. 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.

The use of partial or total air cooling for a pipeline gas plant was examined in detail. For a lignite based plant total air cooling or air cooling to 140° F, followed by water cooling to 100° F, shows that the plant makeup water requirement can be reduced by 82-88% from a totally water-cooled plant. The capital investment for an air-cooled plant is less than for a water-cooled plant; substantial savings in power consumption should also result. The work was done in cooperation with Hudson Products Corporation, a major supplier of air coolers.

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.

The study of Montana subbituminous coal was completed. Operating at 500 instead of 1000 psi with either hydrogen-steam or synthesis gas-steam mixtures definitely showed that the latter is the desired pressure: Operation was smooth at 1000 psi, but erratic at 500 psi. Lower methane and higher carbon oxides yields were also obtained at 500 psi.

During this month over 2700 lbs of Ireland mine coal were pretreated, and will later be hydrogasified to supply the electrothermal gasifier.

Designs are being completed to examine the flow patterns and pressure balance in a model of the upper section of the HYGAS hydrogasifier. The model will permit preliminary study of any future modifications.

After a number of successful runs at 1900° F and 1000 psi using IGT's hydrogasified char, a silicon carbide tube was tried as the central electrode. It failed due to brittleness. A solid rod was obtained and will be tried next.

A run aimed at defining the electrical characteristics of the bed yielded much data. High- and low-frequency current and voltage fluctuations, along with transient response of the bed to step-changes in power input, should provide the necessary data for designing the power package of the 2-MW EG unit. Work to date indicates that the 2-MW electrothermal gasifier will have a direct-current power supply and a concentric electrode configuration.

The gasifier was disassembled to install a magnetic flip coil to suppress arcing and reduce current fluctuations. A 6-inch-diameter silicon carbide tube was installed to serve as the outer electrode. The first test this month showed that modifications are required to overcome the high resistance of silicon carbide relative to a metal. A metal tube is being reinstalled to check out the flip coil.

The nozzles from Spraying Systems Corp. appear to show no wear when dispersing coal-water slurries. Photographs were taken that should permit measurement of spray distribution.

NEW PROCESS STUDIES

A fuel cell engineering study to supply power to the electrothermal gasifier was completed, including details of power plant configuration and cost calculations. 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 $99 and $123 /kW for cell power densities of 300 and 150 watts/sq ft.

PILOT PLANT CONSTRUCTION

Engineering is 99% complete, purchasing is 97% complete, material receipt is 96% complete, and construction is 67% complete. Due to the prolonged truck strike in the Chicago area, receipt of material needed for construction has been seriously delayed. The revised estimate of the mechanical completion date of the plant is now September 1. All efforts will be made to improve this date. The instrument/electrical subcontract has been let.

Together with Procon we began the design and construction of a 2-MW electrothermal gasifier system for the HYGAS pilot plant. The piping and instrument diagram was issued for review. The requisition for the reactor vessel was sent out for quotation. The reactor will be built aboveground instead of in a pit, where solids transfer might be easier, because the cost of underground construction was prohibitive. For safety purposes, the reactor will have a water jacket similar to the hydrogasifier's.

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)	$ 25,000
C.	Funds Expended to Date (estimated)	$ 208,000

2. OCR Funding

A.	Funds Expended This Month (estimated)	$ 281,000
B.	Funds Expended Since Contract Amendment No. 1 (estimated)	$6,740,000

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.
Frank C. Schora, Jr.
Director

Signed
Bernard S. Lee
Manager

Jack Huebler
Jack Huebler
Vice-President

Appendix. Achievements in June

HIGH-PRESSURE METHANATION

Kinetics of the methanation reaction were studied at 630° and 865°F. The results are presented in Table 1. These data were analyzed for the methanation reaction's temperature dependence; the initial results are presented as an Arrhenius plot in Figure 1.

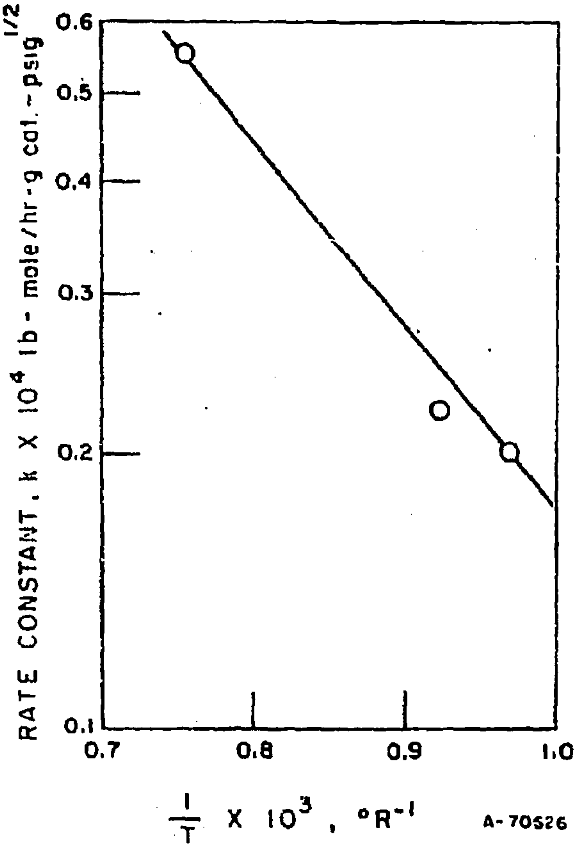

Figure 1. ARRHENIUS PLOT FOR METHANATION

Sulfur Resistance Studies

The laboratory unit to measure the sulfur resistance of commercial methanation catalysts for use in the pilot plant was completed. Gas mixtures of hydrogen sulfide in hydrogen have been prepared for initial tests ; other sulfur compounds can be tested later.

Table 1. METHANATION RUN DATA

Catalyst: Harshaw Ni0104T; 1/4-in. pellets in CSTR

Run No.	298	299	300	301	302	303	304	305	306
Catalyst Wt, g	15.3838	15.3838	15.3838	15.3838	15.3838	15.3838	11.5559	11.5559	11.5559
Feed Gas Rate, SCF/hr	1.8848	1.8217	1.8230	1.6456	3.8197	3.5872	3.7751	2.5607	2.5259
Feed Gas Composition, mole %									
Hydrogen	68.2	35.5	11.6	6.5	92.4	15.4	91.6	90.64	89.3
Carbon Monoxide	28.3	14.5	4.45	3.35	7.5	5.8	8.3	9.3	10.7
Carbon Dioxide	0.13	0.0	0.0	0.25	0.1	0.3	0.1	0.06	0.0
Methane	0.44	50.0	83.95	89.90	0.0	78.5	0.0	0.0	0.0
Total	97.07	100.0	100.00	100.00	100.0	100.0	100.0	100.00	100.0
Reactor Temperature, °F	634	632	573	575	858	850	897	867	622
Reactor Pressure, psig	300.0	302.5	304.0	305.0	605.0	610.0	300.0	95.0	97.5
Product Gas Rate, SCF/hr	0.5454	1.0686	1.7291	1.6032	2.8297	2.9739	3.0197	1.9431	1.8801
Product Gas Composition, mole % (dry basis)									
Hydrogen	3.5	3.0	1.0	0.5	91.3	4.3	90.4	87.78	85.5
Carbon Monoxide	0.1	0.15	0.1	0.1	0.3	0.6	0.8	1.4	2.4
Carbon Dioxide	--	0.35	0.2	1.13	0.1	1.6	0.1	0.12	0.1
Methane	96.4	96.50	98.7	98.27	8.3	93.5	8.6	10.7	12.0
Total	100.0	100.00	100.0	100.00	100.0	100.0	99.9	100.00	100.0
Feed H_2/CO Ratio	2.08	2.45	2.58	1.91	12.1	2.53	11.0	9.7	8.3
Water Produced, lb/hr X 10^4	251	110	186	17	51	91	53	81	51
Rate of Methane Formation, (lb-mole/hr-g catalyst) X 10^4	0.90	0.44	0.13	0.07	0.48	0.26	0.65	0.47	0.51

This unit was installed in a laboratory with the two pilot plant gas chromatographs so that they can be tested as well as provide analytical data on the laboratory unit feed and exit gases. This will allow any modifications in these analyzers to be made now, minimizing the number of problems encountered once plant operation begins. It will also give the IGT data collection staff an opportunity to test computer software as it is being developed for pilot plant data logging.

ENGINEERING ECONOMICS STUDIES

Use of Air Cooling Instead of Water Cooling in Pipeline Gas Plants

We completed the first part of the study of air cooling in the pipeline gas-from-lignite plant design.[1] In the original design, water cooling was used for cooling process streams and in the condensing turbines of the combined MHD-steam power cycle. To obtain estimates for air coolers, cooling requirements were submitted to Hudson Products Corporation. We considered cooling completely by air and also air cooling to 140°F from higher temperatures followed by water cooling to 100°F where required by the process. We are currently working out systems to compare air and water cooling where all the power is generated by conventional steam turbine systems.

Table 2 summarizes the results of this study to date. There are some variations in the applications of air cooling to the quench system, which is used to condense the light oil from the hydrogasifier effluent and return it to the slurry feed system (Figures 2, 3, and 4).

Depending on the degree of air cooling, plant makeup water requirements can be reduced 81.5-88% from the original design. Makeup water for cooling is 3.5% of the circulating cooling water; process water represents about 27% of the total net makeup in the original design. When air coolers are used, additional water can be obtained by cooling the lignite dryer flue gas to 140° and 100°F, and collecting the condensate for use in the plant, thus permitting a reduction greater than 73%. Collecting the condensate is not practical with water cooling because the heat of condensation must be dissipated by evaporating an equivalent amount of water in the plant cooling tower.

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[1] Tsaros, C. L., Arora, J. L., Lee, B. S., Pimental, L. S., Olson, D. P. and Schora, F. C., "Cost Estimate of a 500 Billion Btu/Day Pipeline Gas Plant Via Hydrogasification and Electrothermal Gasification of Lignite," R&D Rep. No. 22, Interim Rep. No. 4. Washington, D.C.: Office of Coal Research, 1968.

Table 2. SUMMARY OF WATER VS. AIR COOLING, PIPELINE GAS FROM LIGNIT

	Original Design[a]	Cooling b Coolers[b] (T- Water Cooling
Makeup Water		
Process Water, gpm	4,252	4,
Coolers, gpm	8,023	5
Quench Tower, gpm	3,640	1
Total, gpm	15,915	5,
Recovered From Lignite Dryer, gpm	--	2,
Plant Net Makeup, gpm	--	2,
Plant Makeup, gal/day	22,903,200	4,250,
Change From Original, gal/day	--	18,652,
Bare Equipment Cost, $		
Coolers	4,189,000	12,321,
Cooling Towers and Pumps	4,344,000	316,
Quench System	13,262,400	4,[illegible]
Makeup Water System	2,300,000	426,
Total Cost	24,095,400	17,209,
Net Change in Investment Over Original Design	--	6,886,
Power Required		
Cooling Water System, hp (kW)	14,400 (10,740)	1,070
Process Air Coolers,[e] hp (kW)	--	10,690 (7
Quench System, hp (kW)	15,200 (11,335)	3,640 (2
Total For Cooling, hp (kW)	29,600 (22,075)	15,400 (11
Net Change, hp (kW)	--	14,200 (10
Annual Consumption, kWhr	156,972,000	51,142,
Net Change in By-product Power, kWhr	--	105,829,
Value at $0.003/kWhr	--	317,
20-yr Avg Gas Price, ¢/million Btu	32.90	31

a. Water cooling power from MHD and steam turbine condensers.

b. Air cooling of process streams to 140°F; water cooling to 100°F. Quench-tower water is cooled by a condensers are air cooled.

c. Air cooling of process streams to 140°F (Figure 3); water cooling to 100°F. Quench-tower water is Turbine steam condensers are air cooled.

d. All air cooling of process streams and steam turbine condensers.

e. Power for air coolers for quench included in quench system total.

NG, 500 BILLION Btu/day
GNITE

oling by Air s[b] (T-140°F); Cooling to 100°F	Cooling by Air Coolers[c] (T-140°F); Water Cooling to 100°F	Cooling Entirely by Air Coolers[d]
4,252	4,252	4,252
594	785	12
191	--	--
5,027	5,037	4,264
2,075	2,075	2,342
2,952	2,952	1,912
4,250,880	4,250,880	2,753,280
18,652,320	18,652,320	20,149,920
12,321,100	12,558,300	13,668,000
316,800	419,100	8,700
144,500	4,020,600	4,020,600
426,800	426,800	276,400
17,209,200	17,424,800	17,973,700
6,886,200	6,670,600	6,121,700
,070 (800)	1,410 (1,055)	20 (15)
,690 (7,971)	10,690 (7,971)	12,710 (9,478)
,640 (2,714)	3,320 (2,475)	3,710 (2,766)
,400 (11,485)	15,420 (11,501)	16,440 (12,259)
,200 (10,590)	14,180 (10,574)	13,160 (9,816)
51,142,900	51,808,900	49,448,000
5,829,100	105,163,100	107,523,700
317,500	315,500	322,570
31.83	31.86	31.92

ed by a cooling tower from 130° to 90°F (Figure 2); turbine steam

ter is cooled by a shell-and-tube exchanger from 130° to 90°F.

2

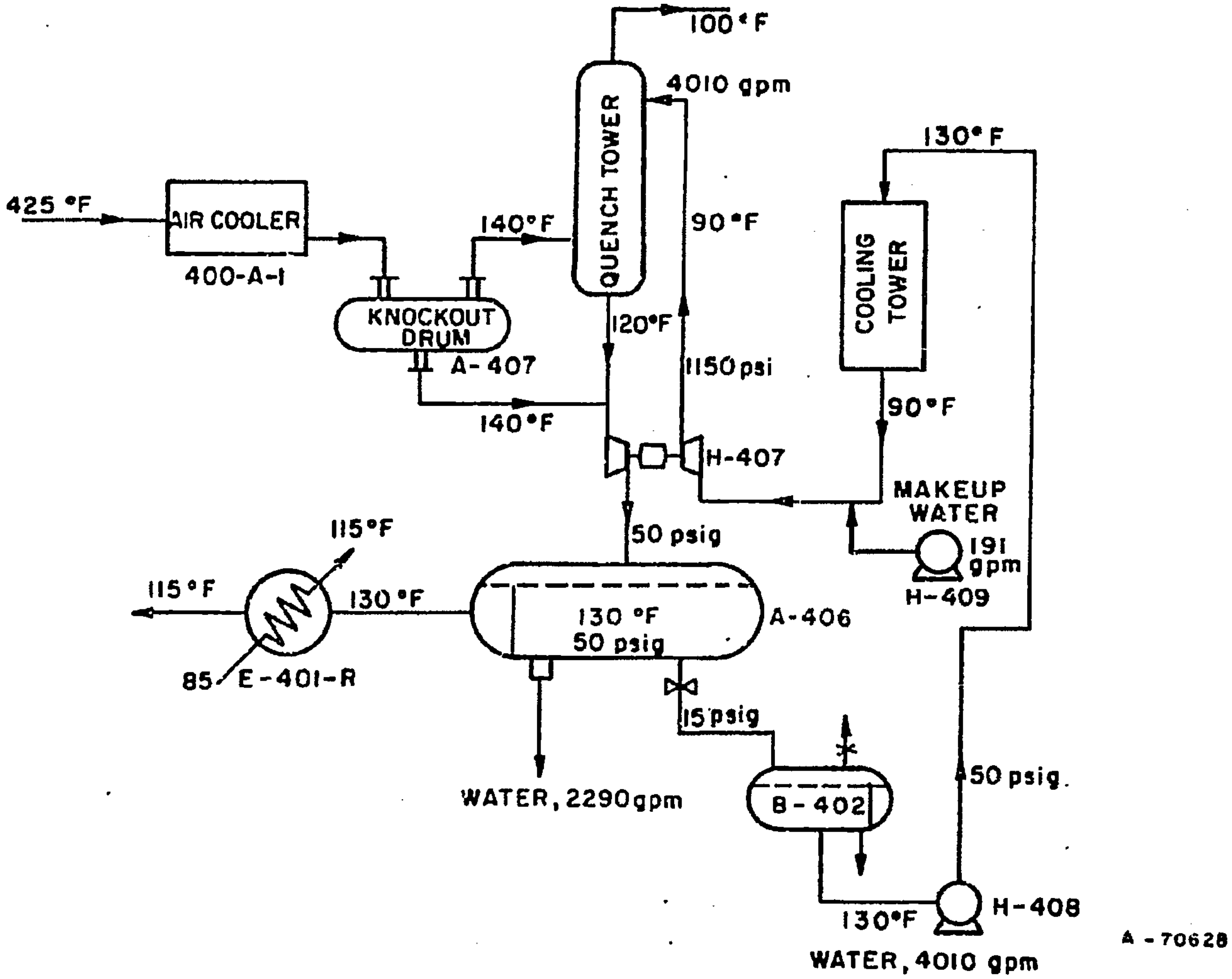

Figure 2. DIAGRAM OF QUENCH SYSTEM USING AIR COOLERS, A SHELL-AND-TUBE EXCHANGER, AND A COOLING TOWER FOR QUENCH WATER COOLING TO 90°F

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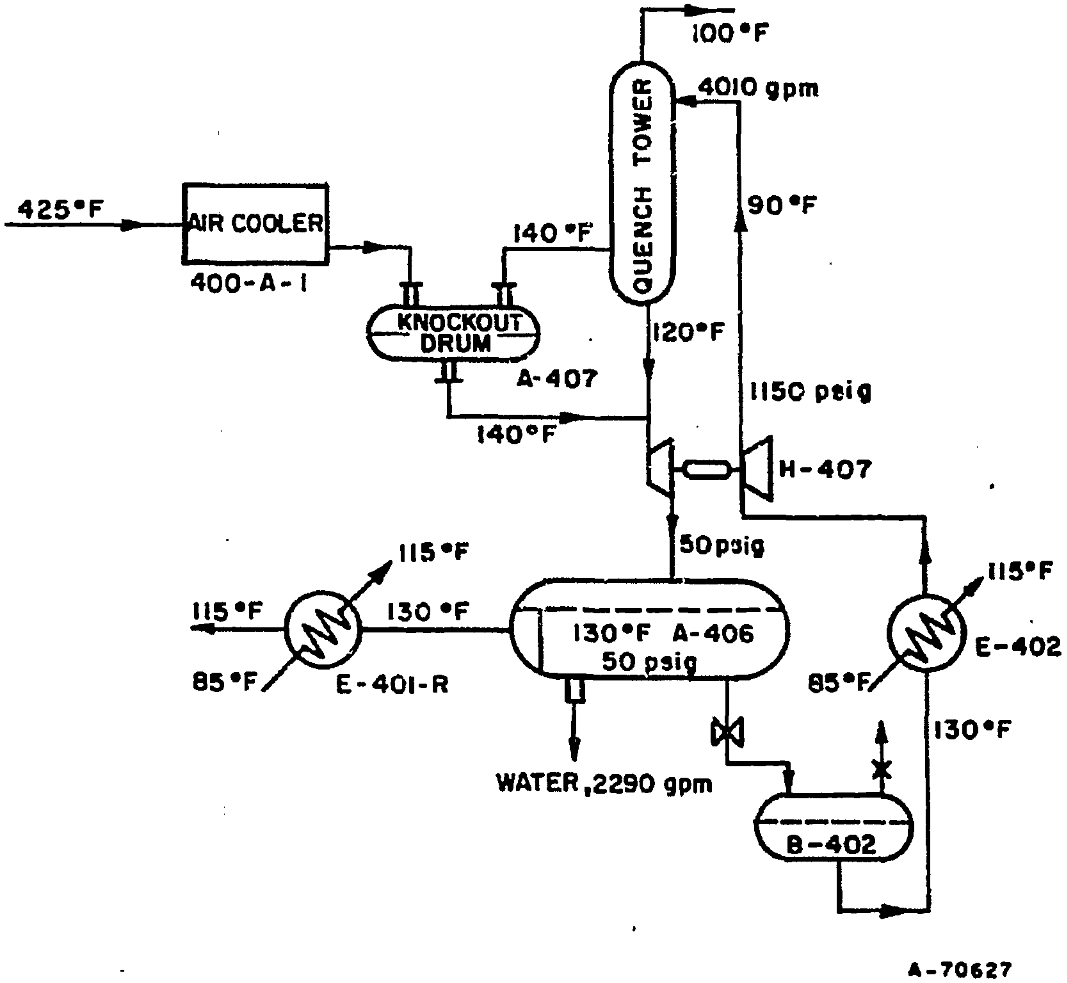

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Figure 3. DIAGRAM OF QUENCH SYSTEM USING SHELL-AND-TUBE EXCHANGERS FOR COOLING QUENCH WATER TO 90°F

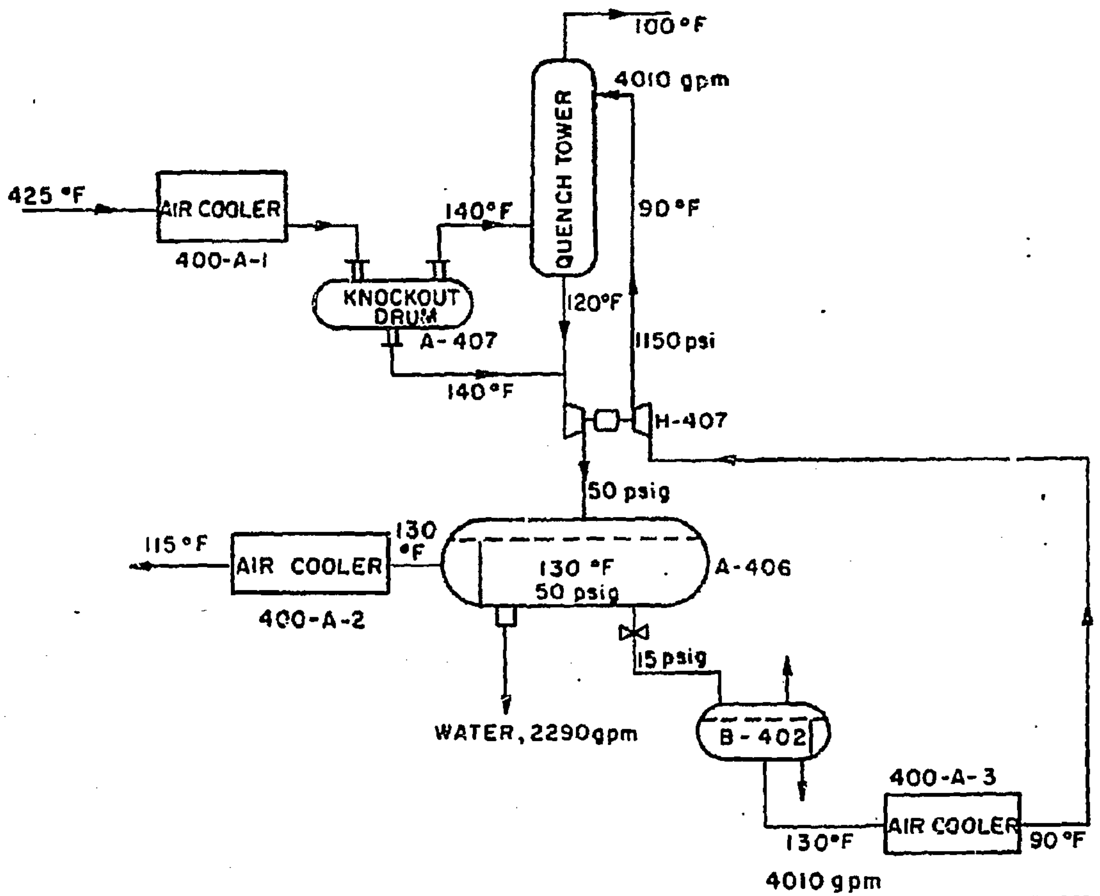

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Figure 4. DIAGRAM OF QUENCH-SYSTEM USING ONLY AIR COOLERS

Process air coolers cost $8.1-$9.5 million more than shell-and-tube exchangers. Net equipment costs are reduced by $6.1-$6.9 million because of the savings in the expensive quench system equipment by substituting air coolers.

If coal were fed dry, and not in a light oil slurry, the heat load on the quench system would be reduced by about 20%. Elimination of the slurry feed would not greatly alter the relative cooling requirement between the air and water systems.

When air cooling is used for the bulk of the cooling total power required for the plant decreases, mainly because of reductions in the power used to pump water in the high-pressure quench system. With the bulk of the heat transferred to air, water is mainly used for scrubbing, with the flow rate less than 20% of the original design (Figures 2, 3, and 4). Annual power requirements for process air coolers average about one-third of the design horsepower, according to Hudson Products. Design ambient air temperature is 85°F, but as the air gets cooler less air is required and the power is reduced by the third power of the air rate. With water cooling annual power consumption will be closer to design than for air coolers because the water flow rate to the exchangers is usually kept high, despite changes in ambient temperature, in order to avoid scale deposition. An example is given below for calculating the annual power requirements (pp. 19 and 22).

Since annual power consumption for air cooling is less than for water cooling, there is more by-product power credit. Lower investment and more by-product power result in lowering the gas price by about 1¢/million Btu. Thus, for the particular system used in the lignite design, air cooling shows a great saving in makeup water and a reduction in overall cost.

Comparison of Cases With and Without Air Cooling

In the original estimate, the process cooling is done entirely by cooling water. With the proposed replacement by air coolers, Table 3 compares the cooling water requirements. Three cases are shown:

Case I. Entire cooling done by water (original design)

Case II. Cooling done by air coolers to 140°F and the rest of the cooling is by shell-and-tube heat exchangers using cooling water

Case III. Entire cooling done by air coolers

Table 3, Part 1. COMPARISON OF BASIC REQUIREMENTS FOR AIR AND WATER COOLING OF PROCESS STREAMS AND TURBINE CONDENSER IN A PLANT FOR PRODUCING 5 BILLION Btu/day PIPELINE GAS FROM NORTH DAKOTA LIGNITE

	Case I (Original Design)				
	Shell and Tube Heat Exchanger				
Service	Equipment No.	Cost of Equipment (f.o.b.), $	Cooling Range, °F	Duty, 10^6 Btu/hr	Cooling Water Required, gpm
Condenser for Dryer Flue Gases	Used with air cooling only to recover makeup water				
Cooling Recycle Light Oil	E-401	25,300	250-115	109.8	7,320
Condenser for Hot Carbonate I	E-501	648,000	228-100	867.0	57,800
Cooling CO Shift Effluent	E-605	10,600	256-240	50.3	3,152
Condenser for Hot Carbonate II	E-701	399,000	229-100	587.2	39,146
Cooling Methanation Feed	E-702	75,600	240-100	150.1	10,008
Condenser for Activated Carbon Regeneration	E-702 (condenser)	66,000	285-100	132.7	8,840
Cooling Activated Carbon Recycle Compressor *	--	--	550-100	5.0	335
Cooling Methanation Effluent	E-804	126,000	200-100	213.3	14,220
Subtotal, Process Cooling					140,821
MHD Turbine Steam Condensation	--	744,000	--	--	88,400
Total, Cooling Water					229,221
Makeup Water					
Amount, gpm					8,023
Investment Cost, $					1,484,600
Total Exchanger Cost, $		2,094,500	--	--	--

* The activated carbon cycle compressor comes with intermediate water coolers; therefore, there is no estimated price on them. No air cooling is estimated here.

Table 3, Part 2. COMPARISON OF BASIC REQUIREMENTS FOR AIR AND
PROCESS STREAMS AND TURBINE CONDENSER IN A PLANT FO
5 BILLION Btu/day PIPELINE GAS FROM NORTH DAKOTA

	Case II (Air Cooling Down to 140°F and Final Cooling						
	Air Cooler				Shell and Tub		
Service	Equipment No.	Cost of Equipment (f. o. b.), $	Cooling Range, °F	Duty, 10^6 Btu/hr	Equipment No.	Cost of Equipment (f. o. b.), $	Co Rang
Condenser for Dryer Flue Gases	200-A-1	1, 200, 000	200-140	1099. 7	--	--	
Cooling Recycle Light Oil	--	--	--	--	E-401-R	9, 800	13
Condenser for Hot Carbonate I	500-A-1	678, 600	228-140	771. 7	E-501-R	97, 000	14
Cooling CO Shift Effluent	600-A-1	28, 700	256-240	50. 3	--	--	
Condenser for Hot Carbonate II	700-A-1	511, 000	229-140	542. 2	E-701-R	46, 600	140.
Cooling Methanation Feed	700-A-2	148, 000	240-140	115. 8	E-702-R	34, 900	[illegible]
Condenser for Activated Carbon Regeneration	700-A-3	109, 000	285-140	130. 3	E-702-R (condenser)	3, 700	140-
Cooling Activated Carbon Recycle Compressor *	--	--	--	--	--	--	550-
Cooling Methanation Effluent	800-A-1	230, 000	200-140	151. 4	E-804-R	123, 000	140-
Subtotal, Process Cooling							
MHD Turbine Steam Condensation	1200-A-1	4, 600, 000	123-103	1324. 0	--	--	
Total, Cooling Water		[redacted]					
Makeup Water		[redacted]					
Amount, gpm							
Investment Cost, $							
Total Exchanger Cost, $	--	7, 505, 300	--	--	--	315, 000	

[7, 505, 300 + 315, 000] = 7, 820, 300

* The activated carbon cycle compressor comes with intermediate water coolers; therefore, there is no estimated price on them. No air cooling is estimated here.

UIREMENTS FOR AIR AND WATER COOLING OF ONDENSER IN A PLANT FOR PRODUCING AS FROM NORTH DAKOTA LIGNITE

to 140°F and Final Cooling by Water)

	Shell and Tube Heat Exchanger			
quipment No.	Cost of Equipment (f. o. b.), $	Cooling Range, °F	Duty, 10^6 Btu/hr	Cooling Water Required, gpm
--	--	--	--	--
E-401-R	9,800	130-115	10.4	690
E-501-R	97,000	140-100	95.3	6,353
--	--	--	--	--
E-701-R	46,600	140-100	45.0	3,000
E-702-R	34,900	140-100	34.3	2,287
E-702-R (condenser)	3,700	140-100	2.4	167
--	--	550-100	5.0	335
E-804-R	123,000	140-100	61.9	4,128
				16,960
--	--	--	--	--
				16,960
				594
				109,900
--	315,000	--	--	--

315,000] = 7,820,300

Table 3, Part 3. COMPARISON OF BASIC REQUIREMENTS FOR AIR AND WATER COOLING OF PROCESS STREAMS AND TURBINE CONDENSER IN A PLANT FOR PRODUCING 5 BILLION Btu/day PIPELINE GAS FROM NORTH DAKOTA LIGNITE

	Case III (Entire Cooling by Air Coolers)				
	Air Cooler				
Service	Equipment No.	Cost of Equipment (f. o. b.), $	Cooling Range, °F	Duty, 10^6 Btu/hr	Cooling Water Required, gpm
Condenser for Dryer Flue Gases	200-A-1 (alt)	1, 712, 000	200-100	1295. 3	--
Cooling Recycle Light Oil	400-A-2	25, 500	130-115	10. 4	--
Condenser for Hot Carbonate I	500-A-1 (alt)	970, 800	228-100	867. 0	--
Cooling CO Shift Effluent	600-A-1	28, 700	256-140	50. 9	--
Condenser for Hot Carbonate II	700-A-1 (alt)	616, 400	229-100	587. 2	--
Cooling Methanation Feed	700-A-2 (alt)	230, 000	240-140	150. 1	--
Condenser for Activated Carbon Regeneration	700-A-3 (alt)	183, 600	285-100	132. 7	--
Cooling Activated Carbon Recycle Compressor *	--	--	--	--	335
Cooling Methanation Effluent	800-A-1 (alt)	362, 000	200-100	213. 3	--
Subtotal, Process Cooling					335
MHD Turbine Steam Condensation	1200-A-1	4, 600, 000	123-103	1324. 0	--
Total, Cooling Water					335
Makeup Water					
Amount, gpm					12
Investment Cost, $					2300
Total Exchanger Cost, $	--	8, 729, 000	--	--	--

---

* The activated carbon cycle compressor comes with intermediate water coolers; therefore, there is no estimated price on them. No air cooling is estimated here.

In the original design 229,221 gpm of cooling water is needed using only shell-and-tube exchangers. In addition to the original heat exchanger's needs, an air cooler is proposed for condensing water in the dryer flue gases to 140°F in Case II and to 100°F in Case III. By cooling the dryer flue gases, 2075 and 2342 gpm of water can be condensed at temperatures of 140° and 100°F. The cooling water requirement for Case II is 16,960 gpm.

In Case III, by cooling process fluids to 100°F entirely by air coolers, the amount of cooling water needed drops to 335 gpm for the activated-carbon recycle compressor. No air cooling procedure is proposed for this, as the cost for the three-stage compressor was assumed to include intermediate coolers.

Quench System

In the original estimate,[1] the light-oil vaporizer effluent at 425°F is cooled to 100°F by direct quenching in the quench tower; the quench water carries oil at 250°F. This oil is separated and cooled to 115°F in a shell-and-tube heat exchanger by cooling water. The water is cooled in the cooling tower to 90°F and then pumped back into the quench tower. With the proposed use of air coolers the vaporizer effluent is cooled to 140°F and then sent to the quench tower. The oil-water mixture at 130°F is then separated.

The oil is cooled either by air cooler (Figure 4) or by conventional shell-and-tube heat exchanger (Figures 2 and 3) to 115°F. The water for the quench system can be cooled to 90°F by the following ways: 1) a cooling tower, 2) a shell-and-tube exchanger, or 3) an air cooler. These systems are shown in Figures 2, 3, and 4; a comparative study on the quench system is shown in Table 4.

The quench water in the proposed system is 4010 gpm. When a cooling tower is used for cooling the quench water from 130° to 90°F, the loss due to evaporation is replaced by 191 gpm of makeup water.

The cooling tower for the quench water can be avoided if a shell-and-tube exchanger is used for cooling the quench water to 90°F with 85°F cooling water. The additional 5467 gpm of cooling is needed for this; the incremental cost in the offsite cooling tower investment is $130,000.

Complete air cooling is also suggested where the quench water is cooled to 90°F by an air cooler (400-A-3) designed for 85°F ambient air, the same

Table 4. SUMMARY OF REVISED EQUIPMENT IN

Equipment	Equipment No.	Description
Vessels:		
	Air Cooling Cases (All cases of air cooling and combinations h	
Knockout Drum	A-407	8.5-ft OD x 25.5-ft width x 3.5-in. thickness, 1100 psi, 140° F
Oil-Water Separator	A-406	18-ft OD x 54-ft width x 0.5-in. thick 1/2-hr residence time, 130° F, 50 ps 12-ft liquid space
Recycle-Water Settling Tank	B-402	8-ft OD x 20-ft width x 0.25-in. thick 130° F, 15 psi
Pumps:		
Quench Tower Cooling-Water Feed Pump (Used in all cases of air cooling and combinations.)	H-407	4100 gpm quench water from 0 to 110 90° F, 3300 hp driven by hydraulic tur generating 1320 hp at full load plus e motor sized for full pumping load.
Quench-Water Cooling Tower Feed Pump (Used only when a cooling tower is used.)	H-408	4264 gpm of water from 15 to 50 psi, 130° F, 150-hp motor-driven centrifug pump
Quench Tower Makeup-Water Pump (Used only when a cooling tower is used.)	H-409	191 gpm of water to 15 psi, 90° F, 5 hp
Quench Tower:		
Quench Tower (Used in all cases of air cooling and combinations.)	A-405	14-ft, 3-in. ID x 135-ft tan-tan x 6-in. wall thickness containing 120-ft packe bed of 3-1/2 in. plastic pall rings; 2,050,000 lb/hr water flow rate, 1100 gas cooled from 140° to 100° F, water heated from 90° to 120° F
Quench-Water Cooling Tower (Used only when a cooling tower is used.)	D-401	Cools 2,131,844 lb/hr of quench water from 130° to 90° F

* Includes $53,000 for packing.

PMENT IN QUENCH SYSTEM

	No. Required	Cost/Unit, $	Total Cost, $
nbinations have common vessels.)			
x 3.5-in. F	4	32,150	128,600
.5-in. thickness, 30°F, 50 psi,	4	60,200	240,800
.25-in. thickness,	4	3,575	14,300
or [illegible]o 1100 psi, ydraulic turbine load plus electric ing load.	1 + 1 spare	82,000	164,000
.5 to 50 psi, ven centrifugal	1 + 1 spare	6,200	12,400
i, 90°F, 5 hp	1 + 1 spare	600	1,200
-tan x 6-in. 120-ft packed all rings; rate, 1100 psi, 00°F, water	1	1,253,600	1,306,600*
uench water	1	85,800	85,800

as the cooling water on hot days (5% of the time) when water cooling would be needed.

Because of the cost of air coolers, the heat exchangers' total costs are higher than the original case shown in Table 5. The bare cost for a shell-and-tube exchanger is twice the f.o.b. cost; the bare cost for air coolers is four-thirds the f.o.b. cost. The pumps and the vessels involved in the quench systems are lower than the original case. Total equipment costs are less with air cooling.

Table 6 shows the comparison of the utility requirements for plant cooling water systems. By using complete air cooling the cooling water requirement is 335 gpm. When air cooling is used to 140°F and then subsequent cooling is done by shell-and-tube exchangers, the process cooling water requirement increases. If a quench water cooling tower is used for cooling the quench water to 90°F, the process cooling water requirement is 16,960 gpm. If the quench water is cooled by a shell-and-tube exchanger to 90°F, the process cooling water requirement is 22,427 gpm. This subsequently raises the off-site cooling tower investment by $130,000. The annual power requirement for plant cooling drops and the by-product power credit increases.

The following are sample calculations of power requirements when air coolers are used to cool the process fluids:

Total Horsepower for Air Coolers

Air Cooler No.	Temperature Range, °F	Design hp
200-A-1	200-100	1622
400-A-1	425-140	2000
400-A-2	130-115	29
400-A-3	130-90	357
500-A-1	228-100	2500
600-A-1	256-240	40
700-A-1	229-100	400
700-A-2	240-100	243
700-A-3	285-100	192
800-A-1	221-100	511
		7894 (5886 kW)

Steam Turbine Air Cooler Requirement, hp	5976
MHD Condenser Operation at 2 in. of Hg, hr/yr (Calculated from temp vs. hr/yr data for Bismark, N. D.)	5500
MHD Condenser Operation Between 2 and 4 in. of Hg, hr/yr	(365 X 24 X 0.9) – 5500 = 2400

Table 5. COMPARISON OF WATER AND AIR COOLING FOR QUENCH SYSTEN PIPELINE GAS FROM NORTH DAKOTA LIGNITE PLANT PRODUCING 500 BILLIC

	Original Design	Air Cooling to 140°F, but Quench Water is Cooled to 90°F by Quench-Water Cooling Tower	Air Cooling to 1 and-Tube Exchan to Cool the Que to 90°I
Quench Water Requirement, lb/hr	11,941,000	2,050,000	2,050,00
Quench Tower Cost, $	3,660,000	1,306,600	1,306,60
Bare Cost of Quench Tower, $	7,492,000	2,660,500	2,660,50
Cooling Tower for Quench Water, $	500,000	85,800	--
Bare Cost of Cooling Tower, $	500,000	85,800	--
Increment in the Offsite Cooling Tower Investment for Process Cooling Water to Cool Quench Water, $	--	--	129,90
Quench System Makeup Water			
gpm	3,640	191	19
Facilities, $	526,400	27,600	27,60
Heat Exchangers			
Light-Oil Cooler (E-401), $	25,300	--	--
Air Cooler for Vaporizer Effluent (400-A-1), $	--	1,263,000	1,263,00
Light-Oil Air Cooler (400-A-2), $	--	--	--
Quench Water Air Cooler (400-A-3), $	--	--	--
Light-Oil Cooler (E-401-R), $	--	9,800	9,80
Quench Water Cooler (E-402), $	--	--	118,60
Bare Cost of Exchangers,[‡] $	50,600	1,703,600	1,940,80
Equipment			
Vessels:			
Knockout Drum (A-407), $	--	128,600	128,60
Oil-Water Separator (A-406), $	908,000	240,800	240,80
Recycle Water Settling Tank (B-402), $	132,000	14,300	14,30
Bare Cost of Vessels, $	2,257,600	900,900	900,90
Pumps:			
Quench-Water Cooling Tower Recycle Pump (H-406), $	86,000	--	--
Quench Tower Cooling-Water Feed Pump (H-407), $	735,000	164,000	164,00
Quench-Water Cooling Tower Feed Pump (H-408), $	52,500	12,400	--
Quench Tower Makeup-Water Pump (H-409), $	202,500	1,200	--
Bare Cost of Pumps, $	3,012,800	497,300	459,20
Total Bare Cost, $	13,839,400	5,875,700	5,961,40
Power for Quench System			
Pumps, hp (kW)	14,000 (10,440)	1,430 (1,065)	1,320 (
Cooling Tower Fan, hp (kW)	1,200 (895)	210 (160)	--
Air Coolers, hp (kW)	--	2,000 (1,490)	2,000 (1,
Total, hp (kW)	15,200 (11,335)	3,640 (2,715)	3,320 (2,
Annual Consumption, 1000 kWhr	84,662	12,732	11,68

[*] The quench-water cooling is done by a shell-and-tube exchanger; the 191-gpm makeup water investment is included in the offsite cooling tower.

[†] The quench-water cooling is done by an air cooler.

[‡] Shell and tube exchangers bare cost is taken as twice the f.o.b. cost. Air coolers are taken at 4/3 the f.o.b. cost.

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COOLING FOR QUENCH SYSTEM FOR E PLANT PRODUCING 500 BILLION Btu/hr

ir Cooling to 140°F, but Quench Water is Cooled to 90°F by Quench-Water Cooling Tower	Air Cooling to 140°F; Shell-and-Tube Exchanger is used to Cool the Quench Water to 90°F	Entire Cooling by Air Coolers
2,050,000	2,050,000	2,050,000
1,306,600	1,306,600	1,306,600
2,660,500	2,660,500	2,660,500
85,800	--	--
85,800	--	--
--	129,900	--
191	191*	--†
27,600	27,600	--
--	--	--
1,263,000	1,263,000	1,263,000
--	--	25,500
--	--	259,000
9,800	9,800	--
--	118,600	--
1,703,600	1,940,800	2,065,300
128,600	128,600	128,600
240,800	240,800	240,800
14,300	14,300	14,300
900,900	900,900	900,900
--	--	--
164,000	164,000	164,000
12,400	--	--
1,200	--	--
497,300	459,200	459,200
5,875,700	5,961,400	6,083,900
1,430 (1,065)	1,320 (985)	1,320 (985)
210 (160)	--	--
2,000 (1,490)	2,000 (1,490)	2,390 (1,780)
3,640 (2,715)	3,320 (2,475)	3,710 (2,765)
12,732	11,680	12,440

water investment is

2

Table 6. COMPARISON OF UTILITY REQUIREMENTS – PLANT COOLING-WATER SYSTEM

	Original Design Cooling Entirely By Water (T-100°F)	Cooling By Air Coolers (T-140°F); Water Cooling (140°-100°, 90°F)*	Cooling By Air Coolers (T-140°F); Water Cooling (140°-100°, 90°F)†	Cooling Entirely By Air Cooler (T-100°F)*
Cooling Water, gpm	229,221	16,960	22,427	335
Cooling Tower Cost (including piping), $	3,200,000	236,800	313,100	4,700
Cooling Tower Circulating Pumps				
Horsepower	12,000	930	1,230	20
Cost, $	1,144,000	80,000	106,000	4,000
Cooling Water Makeup				
gpm	8,028	594	785	12
Installed Cost, $	1,160,200	85,900	113,500	1,700
Offsite Cooling Tower Investment, $	5,504,200	402,700	532,600	9,400
Total Power for Cooling ‡				
Design hp (kW)	14,400 (10,740)	11,660 (8,695)	12,100 (9,023)*	12,730 (9,492)
Avg Annual Consumption, MWhr	72,310	38,411	40,128	36,999
Change in Cooling Water, gpm	--	(212,261)	(206,794)	(228,686)
Change in Makeup Water, gpm	--	(7,429)	(7,238)	(8,012)

* The quench water in the quench system is cooled to 90°F by a quench water-cooling tower system.

† The quench water is cooled by a shell-and-tube exchanger to 90°F. The extra cooling water, amounting to 5467 gpm, is included here for comparison.

‡ Power for cooling in quench system is included in the Quench System, Table 4.

Air Coolers, design hp 7894 + 5976 = 13,870 (10,340 kW)

Original design for steam turbine is for 2 in. of Hg. For air cooling at 85°F ambient, design is for 4 in. of Hg.

Annual Average Consumption for Air Coolers, kWhr = 5886 X (1/3) X 365 X 24 X 0.9 + 5976 X 0.7457 X 2400 + 5976 X (1/3) X 0.7457 X 5500 = 34,340 MWhr

Pump Requirements

Cooling Water System, hp	20
Quench Tower System, hp	1320

Steam turbine condenser in MHD unit has two 1250-hp pumps for recirculating water between direct condenser and air cooler. With power recovery and turbine consumption at 50% recovery —

Horsepower Need for Two Pumps	1250
Pumps' Annual Consumption, kWhr	15,108,300
Total Consumption on Annual Basis, MWhr	34,340 + 15,108 = 49,448

Increase in By-Product Power, Original Design

Cooling Water System, Design hp	
Fan	3,200
Pumps	11,200
Total	14,400
Quench System, Design hp	
Pumps	14,000
Quench Water Cooling Tower Fan	1,200
Total	15,200

Annual Average Consumption, hp

Cooling Water System $= \frac{3200}{3} + 11,200 =$ 12,300

Quench System $= \frac{1200}{3} + 14,000 =$ 14,400

Total 26,700

Annual Average Consumption (90% stream factor), MWhr 156,972

Hence, increase in by-product power due to air coolers is—

156,972 – 49,448 = 107,524 MWhr

In the original report[1] on electrothermal gasification of lignite (p. 58) there is a mistake in tabulating power requirements for the quench tower cooling water system and an error in calculating the carbonate-steam feedwater pump horsepower. The correct values are 15,200 and 750 hp. Table 7 shows the revised power requirements.

Table 7. POWER REQUIREMENTS

Process Unit	Horsepower Requirement
Electrogasifier	894,286 (666,880 kW)
Lignite Storage, Recovery, Conveying	2,680
Lignite Crushers	900
Lignite Grinding-Drying	18,300
Slurry Preparation	1,440
Slurry Feed Pumps	7,560
Carbonate Circulating Pumps	25,900
Quench Tower Cooling Water System	15,200
Plant Cooling Water System	14,400
Reaction Steam Feedwater Pump	5,210
Light-Oil Recycle Pump	200
Carbonate-Steam Feedwater Pump	750
Miscellaneous	4,784
Activated-Carbon Tower Compressor	2,000
Total hp	993,610
Total kW	740,935
Electrogasifier, kW	666,880
Plant Motors, kW	74,054

DEVELOPMENT UNIT STUDIES

Hydrogasification Tests

Complete hydrogasification results of last month's Run HT-248 are presented. The purpose of this test was to evaluate the hydrogasification performance of a dried, but otherwise untreated, Montana subbituminous coal to gasification with a synthesis gas and steam feed gas at a system pressure of 500 psig. Results of this test, as well as operating conditions, are presented in Table 8. A carbon gasification of 27.4% was obtained, as a product gas of 418 Btu/SCF (nitrogen-free basis) was produced. The lower section of the reactor furnace became inoperative during the run. This is a critical heating zone for the 3.5-ft fluidized coal bed, and its failure accounts for the unusually low average coal-bed temperature of 1295°F. Compositions and screen analyses of the coal feed and residue of this test are given in Table 9. As determined by an ash balance, the residue produced was 0.671 lb per lb of coal fed. Liquid products and their composition are shown in Table 10. Carbon converted to oils represents 3.83% of the carbon in the coal.

For comparative purposes, key results of Run HT-248 are tabulated in Table 11 along with those of three other hydrogasification tests (Runs HT-243, HT-244, and HT-247) performed with the same Montana subbituminous coal, at a system pressure in the range of 488-594 psig. In Runs HT-244 and HT-247 the coal was gasified with hydrogen and steam. In Run HT-243, as in Run HT-248, the gasification medium was synthesis gas and steam. Coal conversion of Run HT-248, as measured by the carbon gasified and the moisture-, ash-free coal gasified, is somewhat lower than that of Run HT-243. This is explained primarily by the lower coal-bed temperatures of Run HT-248, and also by the lower hydrogen-to-coal ratios. The reduced gaseous hydrocarbon and carbon oxides yields per lb of coal also reflect the lower coal conversion of Run HT-248. The higher product-gas heating value of Run HT-248, 418 Btu/SCF, compared to the 362 Btu/SCF of Run HT-243, is explained by the lower unreacted hydrogen concentration in the product gas resulting from the lower feed hydrogen-to-coal ratio, and by the higher methane concentration resulting from the lower average coal-bed temperature. A comparison of the syntheiss gas and steam feed tests with the hydrogen and steam feed tests shows a favorable level of coal conversion with synthesis-gas operation. However, it should also be observed that

Table 8, Part 1. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF MONTANA SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR FOR RUN HT-248

Coal	Montana Subbituminous Coal
Source	Colstrip Mine
Sieve Size, USS	−10+80
Run No.	HT-248
Duration of Test, hr	2-3/4
Steady-State Operating Period, min[a]	127-168
OPERATING CONDITIONS	
Bed Height, ft	3.5
Reactor Pressure, psig	594
Reactor Temperature, °F[b] Inches From Bottom	
62-1/2	645
67-3/4	940
73	1330
78-1/4	1310
83-1/2	1490
89	1360
94-1/4	1415
100	1555
104	1595
Average	1295
Coal Rate, lb/hr[c]	28.96
Feed Gas Rate, SCF/hr	249.2
Steam Rate, lb/hr	12.31
Steam, mole % of hydrogen-steam mixture	50.9
Hydrogen/Coal Ratio, % of stoichiometric[d]	15.6
Hydrogen/Steam Ratio, mole/mole	0.596
Bed Pressure Differential, in. wc	126.0
Coal Space Velocity, lb/cu ft-hr	93.61
Feed Gas Residence Time, min[e]	0.447
Superficial Feed Gas Velocity, ft/s[f]	0.130

Table 8, Part 2. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF MONTANA SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR FOR RUN HT-248

Run No.	HT-248
OPERATING RESULTS	
Product Gas Rate, SCF/hr	598.5
Net Btu Recovery, 1000 Btu/lb	2.964
Product Gas Yield, SCF/lb	20.66
Hydrocarbon Yield, SCF/lb	3.00
Carbon Oxides Yield, SCF/lb	2.01
Net Reacted Hydrogen, SCF/lb	0.24
Residue, lb/lb coal[g]	0.671
Liquid Products, lb/lb coal[h]	0.356
Net MAF Coal Hydrogasified, wt %[i]	33.7
Carbon Gasified, wt %	27.4
Steam Decomposed, lb/hr[j]	2.89
Steam Decomposed, % of steam fed	23.4
Steam Decomposed, % of total equivalent fed[k]	49.8
Overall Material Balance, %	99.5
Carbon Balance, %	106.6
Hydrogen Balance, %	89.6
Oxygen Balance, %	92.7

PRODUCT GAS PROPERTIES

Gas Composition, mole %	Feed	Product
Nitrogen	--	35.2
Carbon Monoxide	36.6	11.7
Carbon Dioxide	1.5	13.9
Hydrogen	61.9	24.6
Methane	--	13.7
Ethane	--	0.6
Propane	--	0.2
Butane	--	--
Benzene	--	0.1
Hydrogen Sulfide	--	--
Total	100.0	100.0
Heating Value, Btu/SCF[m]	313	271
Specific Gravity (Air = 1.0)	0.420	0.773

Nitrogen Purge Rate, SCF/hr	211

Table 8, Part 3. OPERATING CONDITIONS AND RESULTS OF THE HYDROGASIFICATION OF MONTANA SUBBITUMINOUS COAL IN HIGH-TEMPERATURE ADIABATIC REACTOR FOR RUN HT-248

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 feed gas 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 9. CHEMICAL AND SCREEN ANALYSES OF MONTANA SUBBITUMINOUS COAL FEED AND RESIDUE

Run No.	HT-248	
Sample	Feed	Residue
Proximate Analysis, wt %		
Moisture	2.1	1.7
Volatile Matter	36.8	9.9
Fixed Carbon	52.9	76.2
Ash	8.2	12.2
Total	100.0	100.0
Ultimate Analysis (dry), wt %		
Carbon	67.3	78.9
Hydrogen	4.43	2.19
Nitrogen	1.00	1.00
Oxygen	17.89	4.89
Sulfur	1.03	0.57
Ash	8.35	12.45
Total	100.00	100.00
Screen Analysis, USS, wt %		
+20	1.6	10.0
+30	18.0	12.3
+40	26.0	23.2
+60	33.0	34.6
+80	13.8	12.3
+100	3.6	3.4
+200	3.3	3.5
+325	0.6	0.4
−325	0.1	0.3
Total	100.0	100.0

Table 10. COMPOSITION OF HYDROGASIFICATION LIQUID PRODUCTS

Run No.	HT-248
Sample	Condenser
Liquid Products,*	
lb/lb coal	0.356
Composition of Liquid Products, wt %	
Water	91.49
Oil	8.51
Total	100.00
Composition of Oil Fraction, wt %	
Carbon	85.10
Hydrogen	8.46
Total	93.56
Carbon in Oil Fraction,	
lb/lb coal	0.0258
wt % of carbon in coal	3.83

* Includes condensed, undecomposed steam.

Table 11. COMPARISON OF MONTANA SUBBITUMINOUS COAL HYDROGASIFICATION RESULTS AT 488-594 psig USING HYDROGEN-STEAM AND SYNTHESIS GAS-STEAM FEED GASES

Run No.	HT-244	HT-247	HT-243	HT-248
Feed Gas	Hydrogen-Steam		Synthesis Gas-Steam	
Reactor Pressure, psig	488	549	489	594
Coal Bed Temp Average, °F	1385	1460	1550	1295
Coal Feed Rate, lb/hr	69.45	54.60	16.66	28.96
Feed Gas Rate, SCF/hr				
Hydrogen	343.8	377.9	--	--
Synthesis Gas	--	--	249.2	249.2
Steam Feed Rate, lb/hr	9.22	8.90	12.18	12.31
Hydrogen/Coal Ratio, % of stoichiometric	14.2	19.9	24.9	15.6
Equivalent Hydrogen/Coal Ratio, % of stoichiometric*	14.2	19.9	41.0	24.9
Steam Concentration in Feed Gas, mole %	36.0	33.1	50.7	50.9
Steam Decomposed, % of total equivalent fed	59.8	50.4	49.1	49.8
Carbon Gasified, %	23.5	26.0	30.0	27.4
MAF Coal Gasified, %	34.7	36.2	36.0	33.7
Hydrocarbon Yield, SCF/lb coal	1.88	2.80	3.33	3.00
$CO + CO_2$ Yield, SCF/lb coal	2.72	2.22	2.47	2.01
Carbon in Oil Products, % of carbon in coal	4.33	5.43	7.07	3.83
Product-Gas Rate (nitrogen-free), SCF/hr	585.5	561.3	349.1	387.8
Product-Gas Composition (nitrogen-free), mole %				
Carbon Monoxide	24.7	13.5	15.4	18.1
Carbon Dioxide	7.6	8.1	26.8	21.5
Hydrogen	45.0	50.7	41.7	38.0
Methane	20.4	25.2	14.5	21.1
Ethane	1.2	1.4	0.7	0.9
Propane	0.7	0.6	0.7	0.3
Benzene	0.4	0.5	0.2	0.1
Total	100.0	100.0	100.0	100.0
Product-Gas Heating Value (nitrogen-free), Btu/SCF	479	514	362	418

* Includes carbon monoxide equivalent of hydrogen.

this generally satisfactory conversion with synthesis gas was obtained with significantly larger equivalent hydrogen-to-coal ratios than with hydrogen operation and at substantially lower coal feed rates. The ratios of carbon monoxide to carbon dioxide in the product gases differ substantially in the two sets of operations. With a synthesis gas and steam feed, this ratio is less than 1, reflecting the considerable amount of carbon monoxide shifting with steam to form carbon dioxide. With a hydrogen and steam feed, the ratio of carbon monoxide to carbon dioxide is in the range of 1.5-3; the carbon dioxide concentration in the product gas is of the order of 8 mole percent.

Coal Pretreatment

The pretreatment of a Pittsburgh No. 8 seam, Ireland mine bituminous coal, started last month, was continued. In the first pretreatment stage, Run FP-143, approximately 2700 lb of the raw coal was treated with air and nitrogen at 750°-800°F in a fluidized bed. Laboratory agglomeration tests of the treated coal showed a slight tendency for agglomeration. The coal was treated in a second-stage operation, Run FP-144, to render it suitable for hydrogasification use. Treatment conditions were similar to those of Run FP-143. The pretreated coal will be hydrogasified to produce a char suitable for use in the electrogasifier development unit.

ELECTROTHERMAL GASIFICATION

We completed major modifications and equipment changes during the month. One run was conducted in the electrothermal gasifier at 1800°F and 1000 psig pressure using FMC project COED char as the feed material. It was the first run during which we used a 6.0-in.-ID silicon carbide tube as one of the electrodes in a concentric configuration. The magnetic flip coil was in place for the run; however, the power for it was not connected.

The installation of the 21-3/8-in.-OD magnetic coil required the complete disassembly of the reactor vessel. The coil has a total of 78 turns of 1/2-in. OD x 0.060-in. wall copper tubing and reaches 45 in. in length (Figure 5). A thin layer of high-temperature epoxy insulates and protects the coil. Additional electrical insulation is provided by a sheet of Mylar paper between the reactor and the coil. Cooling water passes through the coil at the rate of 2 gpm to prevent the insulation from deteriorating during a run. Two Milton-Roy high-pressure pumps were overhauled to provide the necessary flow rate.

Figure 5. FLIP COIL BEING LOWERED INTO REACTOR VESSEL. HIGH-TEMPERATURE W. R. EPOXY COATING PROVIDES INSULATION AND ADDS TO MECHANICAL STRENGTH OF COIL.

The coil will operate at a low frequency, typically 1 cps, and has a power requirement up to 6 kW. A generator set designed by our consultants has been incorporated into our system to supply this power to the coil. The magnetic field established by the coil will attempt to suppress any gross arcing in the fluidized bed, as explained in more detail in the May 1970 Project Status Report.

After the coil was installed, the reactor was repacked with Fiberfrax insulation and the silicon carbide tubes were substituted for the stainless steel tube (Figure 6). In the concentric configuration these tubes act as the outer electrode. The inner electrode was a 1.5-in.-OD stainless steel rod. The two sections of the 6.0-in. ID x 1/2-in. wall x 36-in.-long tubes were sealed together with refractory cement and topped off by a 28-in.-long precast refractory tube. Stainless steel thermowells were placed along the outside of the tubes to monitor the temperatures. Electrical connection was provided on the bottom by a stainless steel cone.

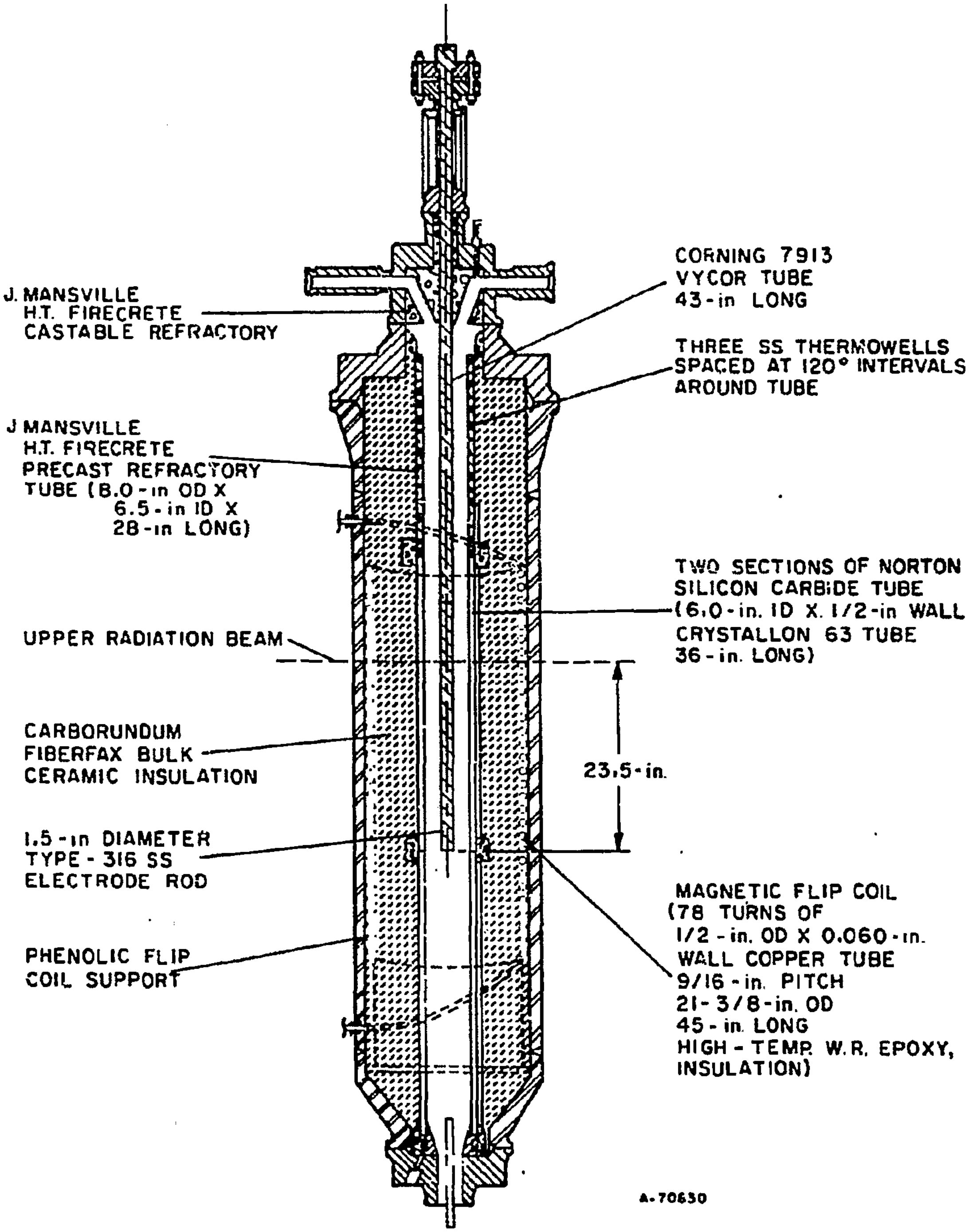


Figure 6. CUTAWAY VIEW OF REACTOR VESSEL IN A CONCENTRIC CONFIGURATION FOR RUN EG-56

The main purpose of Run EG-56 was to test the new electrode arrangement. In addition, a close check was made on the temperature of the cooling water for the magnetic flip coil. Power was not applied to the coil. The initial resistance of the fluidized bed was approximately 1500 ohms, and then dropped rapidly to less than 60 ohms. During the run the resistance varied erratically between 60 and 5 ohms. Steam was introduced when temperatures in the fluidized bed reached 1450°F. Fifteen minutes later a bundle of six thermocouples burned out and the run was terminated.

A thorough inspection of the reactor revealed several small burned spots at the junction of the two silicon carbide tubes exactly opposite the thermowells and a 2-in. burn at the tip of the electrode, also directly opposite the junction. No deterioration of the silicon carbide was visible, although a small discoloration was noticeable on the upper silicon carbide tube where a thermowell had touched the wall. A small hole had been burned in the thermowell.

The electrical characteristics of silicon carbide might explain these observations. In our operating region the resistance of silicon carbide decreases with increasing temperature. We have a temperature gradient between the fluidized bed and the bottom cone amounting to several hundred degrees Fahrenheit; as a result the least path of resistance may be along the thermowells.

We have to eliminate this path and obtain a 5-ft section of silicon carbide tube which will eliminate the problem of a junction. A silicon carbide cement is being investigated as a possibility of providing a good conductive junction between silicon carbide bricks. A solid 1.5-in.-OD silicon carbide rod is also ready for testing in the coming weeks.

A 6.0-in.-ID stainless steel tube will be reinstalled for the next run to allow our electrical consultants to evaluate the effectiveness of the magnetic flip coil. We then expect to continue investigating the properties of silicon carbide.

A major equipment change was initiated when Procon requested that the Wilson-Snyder high-pressure pump, which we had been using for our quench water requirements, be made available for installation in the HYGAS pilot plant. A Kerr Mustang-3250 high-pressure pump was installed in its place.

PILOT PLANT CONSTRUCTION

Engineering

Engineering work is continuing on recent additions to the project. Several of the most recent additions concern tie-ins with the electrogasifier unit. Field changes and corrections are being recorded and "as built" drawings will be made after construction is complete.

Procurement

The instrument/electrical subcontract has been let and the painting specification is out for bids. The truck strike has caused serious delays in receipt of outstanding material; Several critical items have been freighted. The remaining material to be delivered is under continuing review to determine the appropriate action to be taken.

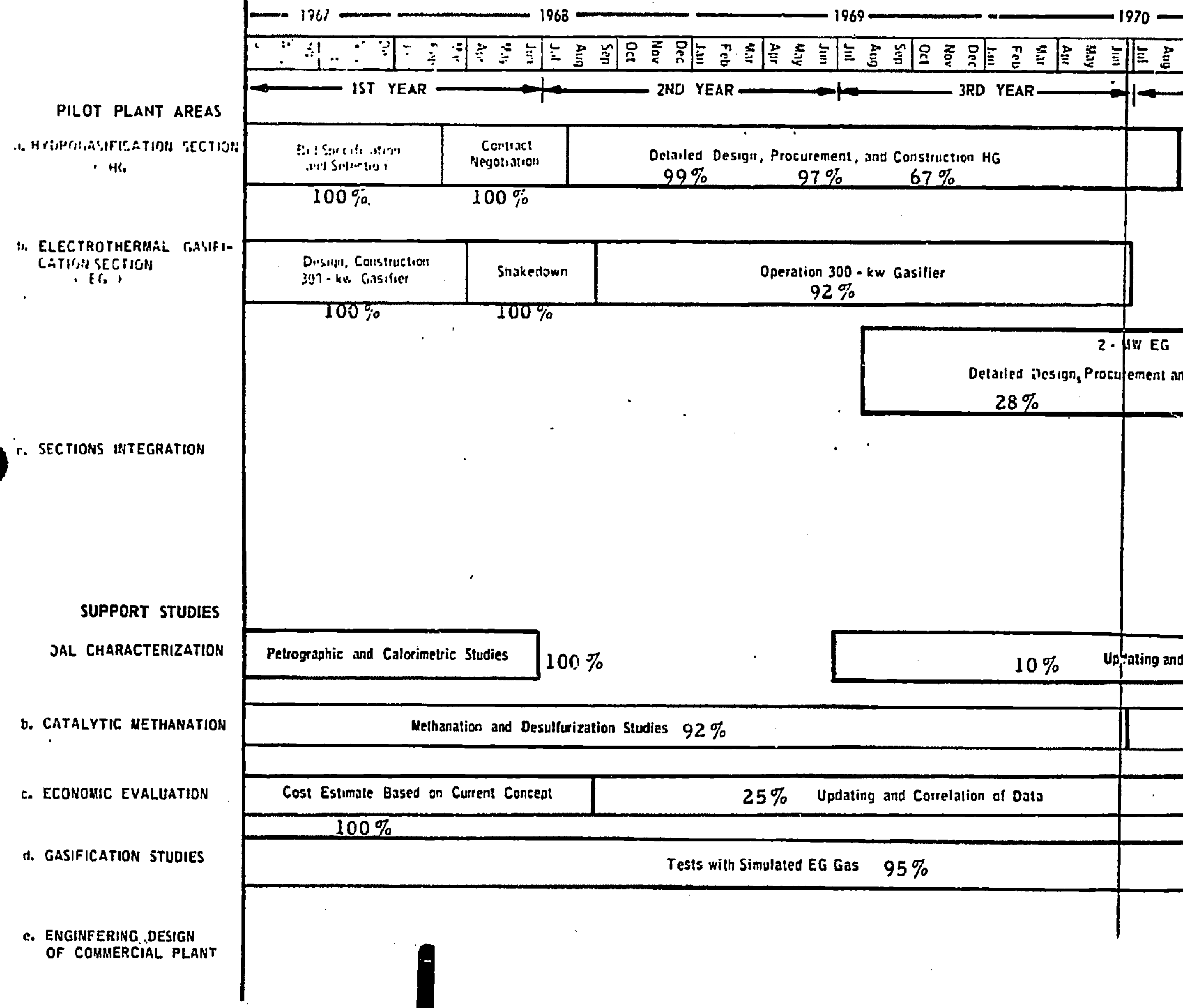
Construction

Major field activities during this report period have been area piping, coal unloading facilities work, vessel insulation, electrical work, and instrumentation. Piping is approximately 77% complete.

We experienced a total of 24 inclement weather days; 3 occurred during this report period. On these days progress was negligible.

PILOT PLANT PROGRAM OF ICT HYDROGASIFI

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



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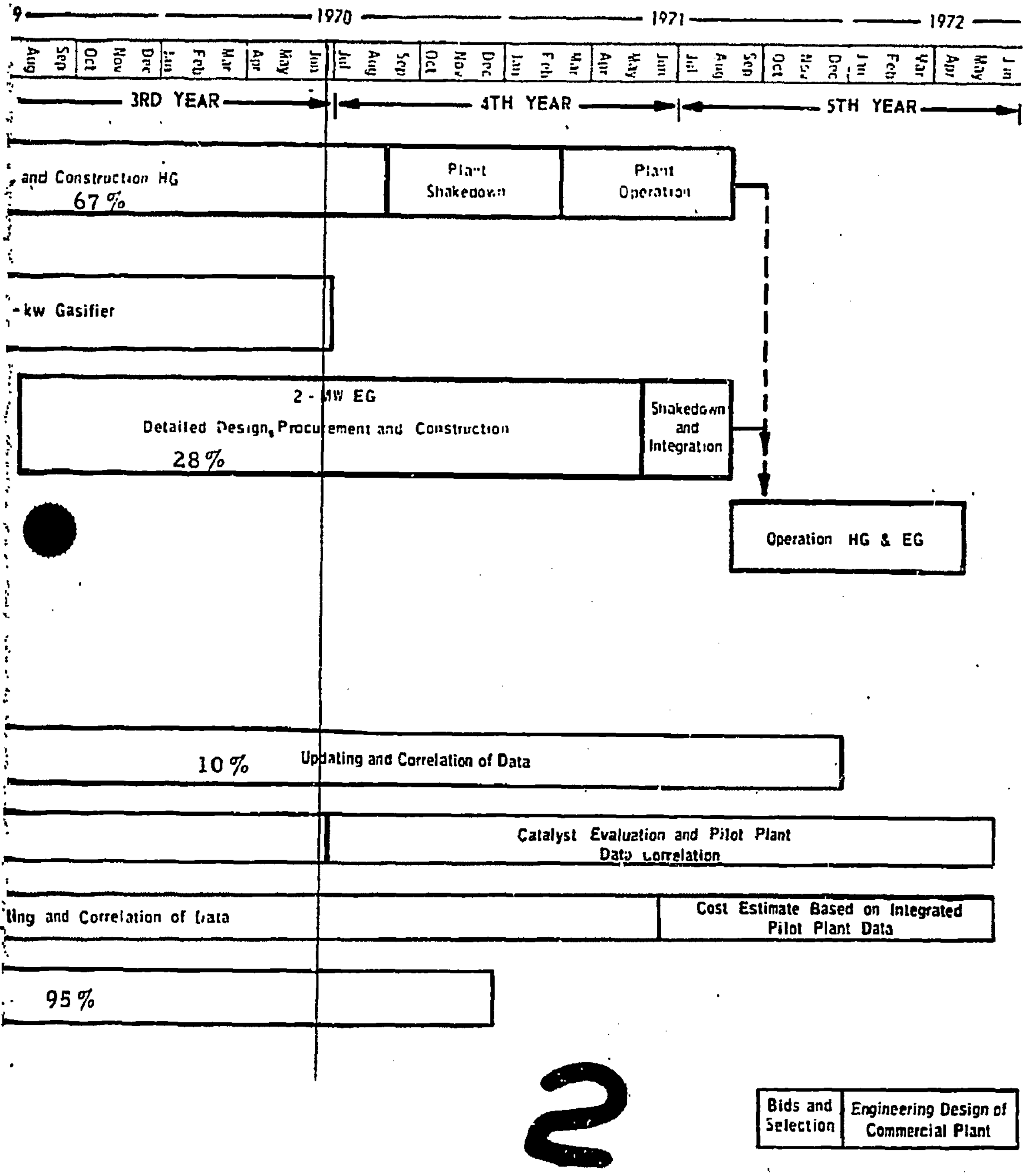

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