FE-2434-18

PIPELINE GAS FROM COAL-HYDROGENATION (IGT HYDROGASIFICATION PROCESS)

Project 9000 Monthly Status Report

For the Period August 1 through August 31, 1977

Prepared by

Institute of Gas Technology

IIT Center, 3424 S. State Street

Chicago, Illinois 60616

Date Published — October 1977

Prepared for the

UNITED STATES ENERGY RESEARCH
AND
DEVELOPMENT ADMINISTRATION

Under Contract No. EF-77-C-01-2434

NOTICE

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REA

IGT
EDUCATION · RESEARCH

INSTITUTE OF GAS TECHNOLOGY · IIT CENTER · CHICAGO 60616

Project Status Report
for
ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION
AND
AMERICAN GAS ASSOCIATION

Report for
August 1977

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

ERDA Contract No. EF-77-C-01-2434

I. PROJECT OBJECTIVE

The objective of this project is to perform the necessary pilot plant operations and related support studies to acquire data for a commercial/demonstration plant design based on the HYGAS® Process.

Tasks 1 through 6, which concerned demonstrating the feasibility of the HYGAS pilot plant using lignite, bituminous, and subbituminous coal feed-stocks, were completed under ERDA Contract No. EF-77-C-01-2434 (July 1, 1976, through June 30, 1977).

The extension of this contract began July 1, 1977, and involves completing Tasks 7 through 9, which are detailed in the body of this report. Since the beginning of this contract extension, we have used our annual plant turnaround period for routine maintenance, repairs, and installation of equipment. All of these are necessary for successful plant operation, and therefore, are a significant part of advancing pilot plant studies.

After plant turnaround, which was completed early in August, we conducted Test 64. We are now trying to acquire data for a commercial/demonstration plant by conducting an extended run with Peabody No. 10 mine bituminous coal at high carbon conversions. Eight and one-half days of self-sustained operation at coal conversions ranging from 72% to 97% occurred during Test 64. This was a major step toward achieving our immediate objective.

II. SUMMARY

The first part of August was spent on completing the annual plant turnaround and on preparing for Test 64. This test, conducted during the last half of August, was our longest test to date with Peabody No. 10 mine bituminous coal. Eight and one-half days of self-sustained operation were achieved, with coal conversions ranging from 72% to 97%.

A cold-flow model was assembled to simulate the second-stage gasifier and the steam-oxygen gasifier. Results of tests made with this model are presented in this report.

Representatives of ERDA, Procon, and IGT held their initial kick-off meeting this month on the Procon design of the commercial/demonstration plant based on the HYGAS Process.

III. ACHIEVEMENTS

Task 7. Pilot Plant Experimental Operation

Work during our annual plant turnaround and preparations for Test 64 were completed early in August. The new ductwork was installed around the coal mill-wet scrubber area, and the ducts were insulated. Prior to beginning Test 64, tests of the coal mill showed that the secondary fan vibrated excessively and that its shaft was bent. We replaced the shaft and balanced the impeller. The coal preparation section was then readied for Test 64.

In previous tests, the pretreater section had operated very well, and no modifications were contemplated for Test 64. A 50:50 mixture of nitrogen and air was planned to be used to fluidize the char in the char cooler. The pretreater was readied for operation.

The slurry preparation section was also readied for Test 64. A 2-foot-long spool piece for instrumentation, requested by Argonne National Laboratory (ANL) was installed. ANL is developing a new acoustic slurry-flow measurement technique.

Gray Serv's technicians completed their machine work and inspected all of the Grayloc nozzles on the reactor and other high-pressure vessels in the plant. Several nozzles were remachined, and repairs were completed on the line 321 expansion joint. The reactor was reassembled and pressure tested. ANL inspected the high-pressure cyclone and noted some wear on the wall of the cyclone and additional wear in the solids discharge line of the cyclone. Wall thickness in both areas is well within the original design limits. However, we installed a replaceable sleeve in the solids discharge line to preserve the parent metal's wall thickness in future operations. A new conductivity level detector was installed on the cyclone slurry pot, and a new eductor assembly was also added to the cyclone slurry pot to improve slurry mixing for better handling. The reactor section was then ready for Test 64.

The quench section was modified so that it could be better monitored. A splash plate was installed at the gas-exit nozzle of the quench tower. The heat exchangers in the quench system were cleaned and buttoned up. Pressure testing of the quench section revealed a leak at the Grayloc fitting at the bottom of the quench tower which was then repaired.

The purification section was completely cleaned. Packing in the absorber tower and regenerator tower was replaced, and the vessels were headed up in preparation for Test 64.

No changes were made in the IGT fixed-bed catalyst methanation unit. The Chem Systems liquid-phase methanation unit was modified.

The light-oil and solids recovery sections were both prepared for operation. The new high-capacity incinerator was installed. The utilities were ready for operation at the start of August.

The hydrogen plant was started up, and a leak was discovered in the diglycolamine heat exchanger that had been repaired during turnaround. The hydrogen plant was then shut down and the diglycolamine cooler opened and inspected. Gasket surfaces in its internals were polished up, and the unit was reassembled and started up for Test 64.

Light-off for Test 64 occurred at 0525 hours on August 14. The objective of Test 64 was to operate the HYGAS pilot plant with Peabody No. 10 mine pretreated coal for a 7-day steady-state period at high coal conversions. We began coal feed to the pretreater on August 18 at 0350 hours. The pretreater operated satisfactorily during Test 64, except for two periods when coal feed was stopped when a hole developed in a gasket between the

60-ton hopper and the bin vibrator in the bottom of the raw-coal storage hopper and again when the Sweco oversize screener broke down. These interruptions were not detrimental to the unit's overall operation. A total of 437 tons of raw coal was fed to the pretreater reactor over a period of 197 hours, averaging 2.2 tons/hr. IGT agglomeration boat tests showed that the pretreated char was free flowing. Throughout the entire test the pretreater was operated at temperatures ranging from 750^{o} to $770^{o}F$.

Pretreated char was introduced to the reactor at 2100 hours on August 18, and the reactor operation was self-sustained at 0500 hours on August 19, when smooth solids flow in the reactor was established. Test 64 was terminated at 1600 hours on August 27 when we were unable to withdraw solids from the steam-oxygen gasification zone and had difficulty in moving solids from the second-stage gasifier to the steam-oxygen gasifier. All attempts to reestablish solids flow were unsuccessful, and the test was terminated.

Test 64 operated in a self-sustained manner for 203 hours, and a total of 396 tons of pretreated char was processed through the gasifier over a slurry feeding period of 205 hours. Coal conversion by fast ash analysis ranged from 72% to 97%. The operating temperatures of the steam-oxygen gasifier ranged from 1700^{o} to $1825^{o}F$. Reactor operations for Test 64 were exceptionally smooth and troublefree. No periodic upsets of solids flow were experienced during this test.

Preliminary test results for Test 64 are given in Table 1. It shows the degree of coal conversion, operating conditions, oxygen-to-char feed ratios, steam-to-char feed ratios, and operating temperatures in the steam-oxygen gasifier.

The operation of the product gas cyclone dust slurry pot appeared to be very satisfactory. The quench system operation was troublefree. The purification section was put on stream at 1015 hours on August 22. There were some initial high pressure drop problems in the absorber tower; however, the section was eventually lined out. Purified gas was delivered to the liquid-phase methanation section from 1730 hours on August 24 until Test 64 was terminated. Sixty-nine hours of purified gas feed to the liquid-phase methanation pilot unit were logged. The effluent cleanup section operated satisfactorily during Test 64. The utility section in the hydrogen plant operated as required during the test.

Table 1. PRELIMINARY ASH RESULTS FROM SPENT CHAR SAMPLES* FROM TEST 64
(Quick Ash From State Street)

Date (Time)	Char Feed Rate,† lb/hr	Ash in Spent Char	Ash in Pretreated Char	Char Gasified	Carbon Gasified by Gas Analyses	lb O_2/lb Char Feed	lb Steam/ lb Char Feed	SOG Bed Temp, °F
		%						
8/20 (0600)	1802	50.8	12.7‡	86	--	0.38	4.2	1565
8/21 (0600)	3318	55.7	12.7‡	88	75	0.27	2.3	1675
8/22 (0630)	3714	74.7	12.7	95	86	0.29	2.1	1730
8/22 (1030)	3620	84.4	12.7‡	97	90	0.31	2.2	1749
8/22 (1630)	4367	56.0	12.7‡	89	80	0.26	1.8	1737
8/23 (0630)	4454	39.3	14.1	75	78	0.23	1.7	1788
8/23 (1400)	4296	35.9	13.7	72	61	0.21	1.8	1750
8/24 (0630)	4610	40.4	12.7	79	75	0.21	1.7	1811
8/25 (0630)	4553	41.9	12.7	80	65	0.21	1.7	1808
8/25 (1030)	4336	36.2	12.7‡	74	68	0.22	1.8	1799
8/25 (1400)	4284	38.9	12.3	78	72	0.22	1.8	1803
8/26 (0630)	4717	34.6	12.5	73	68	0.20	1.6	1811
8/26 (1000)	4033	36.5	13.3	73	76	0.24	1.9	1798
8/26 (1400)	4359	43.7	13.1	81	77	0.23	1.8	1821
8/27 (0630)	4492	50.5	12.9	85	86	0.23	1.7	1818
8/27 (1000)	4521	41.6	13.4	78	73	0.22	1.7	1765

* All samples are grab samples.

† Char feed rates on a wet basis.

‡ Assumed.

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During Test 64, 8-1/2 days of self-sustained operation were achieved in the reactor, with coal conversions ranging from 72% to 97%. Solids flow in and out of the reactor and throughout it was also very smooth. There were no periodic upsets of solids flow from the second-stage gasifier to the steam-oxygen gasifier as had been seen in previous tests. We cooled down and depressurized the reactor, and early next month we will inspect and clean the entire plant in preparation for Test 65.

Several long, steady-state periods of Test 64 were picked for study. Detailed results will be presented when they are available.

The results of Test 63, conducted in June, are presented in Tables 2 through 4 and Figures 1 through 18. Three periods for the pretreater section were selected and their results are presented here. The oxygen balance is again forced because the quench water effluent stream from the pretreater quench tower was not measured. Samples were collected for that stream to determine fines, tars, and oils flows. Five periods were chosen for reactor study. Their results are presented here. Due to the periodic operation of the light-oil recovery unit, the water and oil balances cannot be obtained.

The mechanical status of the HYGAS pilot plant for August is shown in Figure 19.

The American Gas Association project advisors met at IGT on August 10, 1977, to review the work done on the HYGAS program since May 1977.

The Office of Program Planning and Assessment, Fossil Energy, U. S. ERDA, has contracted for Scientific Design Company, Inc., to evaluate the HYGAS program. An initial meeting between representatives of ERDA-CCU, and ERDA-OPPA, IGT, and Scientific Design was held in July to establish the program and set the basis for interaction between ERDA, Scientific Design, and IGT. Since August 15, a resident engineer from Scientific Design Company, Inc., has been onsite monitoring all HYGAS operations.

Task 8. Demonstration Plant Design Support

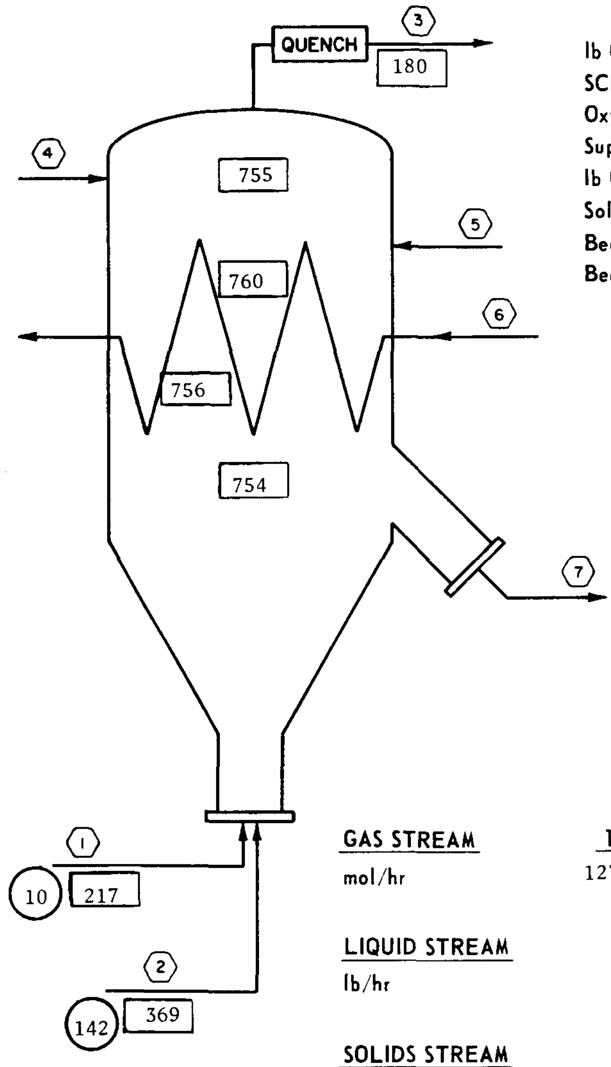
Representatives of ERDA, Procon, and IGT met at the HYGAS pilot plant on August 11 for the initial kick-off meeting on the Procon design of the commercial/demonstration plant based on the HYGAS Process. General

Table 2. MATERIAL BALANCE SUMMARY FOR THE PRETREATER SECTION FOR TEST 63 FROM 6/21/77 (1700 Hours) TO 6/22/77 (0600 Hours)

Basis = 1 hr. All units in lbs unless otherwise noted.

INPUT		C	H	O	N	S	Ar	Ash	Other	Total
Coal Feed	Wt % (Dry)	68.07	4.88	9.65	1.21	4.71		11.48		100
	Coal (Dry)	3167	227	449	56	219		534		4652
	Moisture		19	148						167
Streams to Pretreater	Air			852	2771		50			3673
	Steam		182	1452						1634
Nitrogen from purges					40					40
Air from purges				9	29					38
H_2O to venturi scrubber			2006	16,051						18,057
H_2O to quench tower			653	5225						5878
Air to char cooler				97	313		6			416
Cooling water to char cooler			98	783						881
TOTAL INPUT		3167	3185	25,066	3209	219	56	534		35,436
OUTPUT										
Pretreated Char to Gasifier	Wt.% (Dry)	67.83	3.49	8.72	1.36	4.30		14.30		100
	Char (Dry)	2596	134	334	52	165		547		3828
	Moisture		10	76						86
Slurry Waste from Quench	Wt.% (Dry)	56.60	2.61	11.27	1.45	4.10		23.97		100
	Solids (Dry)	118	5	24	3	9		50		209
	Tars & Oils	106	10	8	1	4				129
	H_2O & Dis. materials	13	2719	21,745	2	42				24,521
Quench Tower Off-Gas	Total	151	284	2879	3314		56			6684
	Components: H_2		1							1
	CO_2	107		287						394
	C_2H_6	3	1							4
	N_2				3314					3314
	CH_4	9	3							12
	CO	32		43						75
	O_2			321						321
	Ar						56			56
	H_2O		279	2228						2507
TOTAL OUTPUT		2984	3162	25,066	3372	220	56	597		35,457
Net (Output – Input)		–183	–23	0	163	1	0	63		21
% Balance (Output/Input)		94	99	100	105	100	100	112		100

STREAM No.
PRESSURE, psig
TEMPERATURE, °F



lb Char/lb Coal =	0.78
SCF O_2/lb Coal =	2.2
Oxygen Breakthrough, % fed =	26.3
Superficial Velocity, ft/s =	0.91
lb O_2 Reacted/lb Coal =	0.14
Solids Residence Time, hr =	1.8
Bed Density, lb/cu ft =	20
Bed Height, ft =	10

GAS STREAM	3
COMPONENTS, mol %	
H_2	0.14
CO_2	3.17
C_2H_6	0.05
O_2	4.17
N_2	41.93
Ar	--
CH_4	0.26
CO	0.94
H_2O	49.34
mol/hr	282
mol wt	24

GAS STREAM	1	2
mol/hr	127	91

LIQUID STREAM	5	6
lb/hr	0	1636

SOLIDS STREAM	4	7
lb/hr	4652	3828

1. AIR TO PRETREATER
2. STEAM TO PRETREATER
3. PRETREATER OFF-GAS AFTER QUENCH
4. RAW COAL TO PRETREATER
5. DIRECT WATER QUENCH
6. COOLING WATER TO COOLING COIL IN PRETREATER
7. PRETREATER CHAR TO CHAR COOLER

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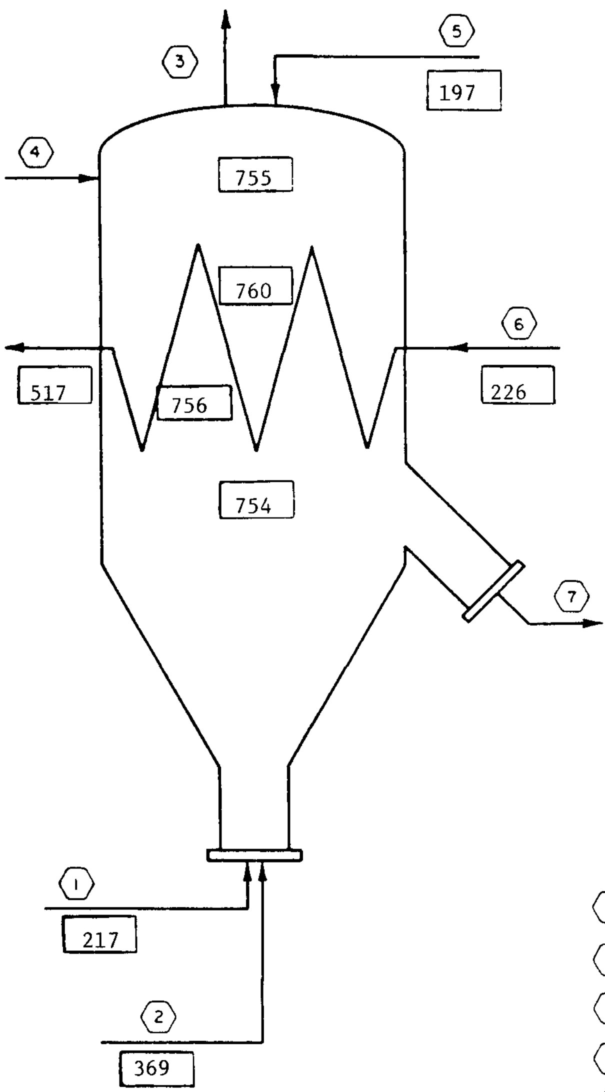
Figure 1. PRETREATMENT DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/21/77 (1700 Hours) TO 6/22/77 (0600 Hours)

Stream No. Temperature, °F

Basis: 1 hour

Datum Condition: 77°F, 1 atm, material in standard state.



INPUT	Btu
Sensible Heat (Streams 1, 2, 4, 5, 6)	630,728
Heat of Combustion (Stream 4)	57,168,428
Steam Enthalpy (Streams 2 & 5)	2,803,496
Total	60,602,652
OUTPUT	
Sensible Heat (Streams 3 & 7)	3,204,197
Heat of Combustion (Streams 3 & 7)	50,192,711
Steam Enthalpy (Streams 3 & 6)	4,659,354
Total	58,056,262
% Balance	96

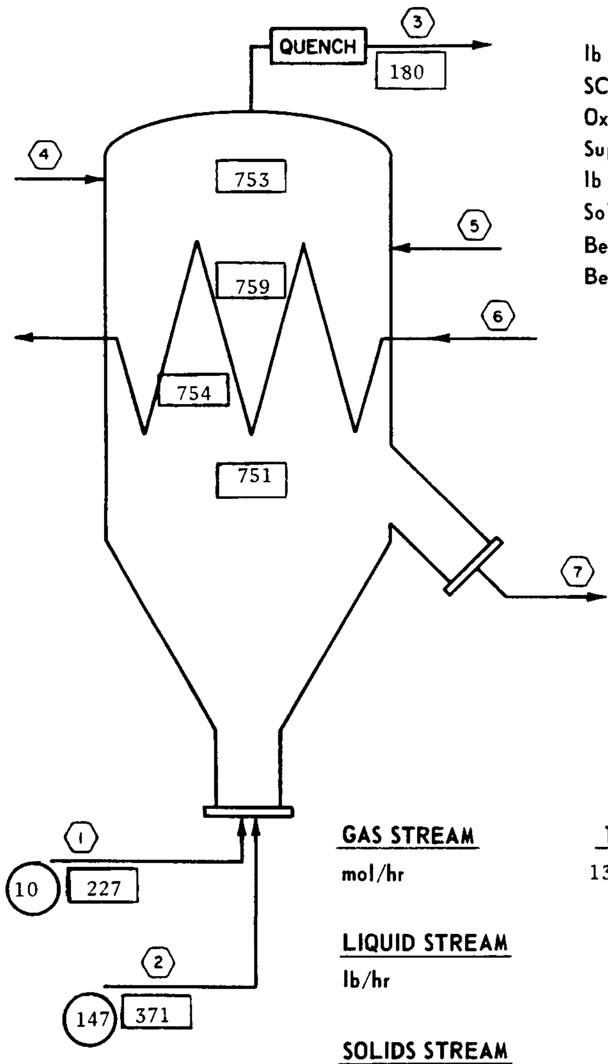
1. Air to Pretreater
2. Steam to Pretreater
3. Pretreater Overhead
4. Raw Coal to Pretreater
5. Gas From Char Cooler
6. Cooling Water to Cooling Coil in Pretreater
7. Pretreated Char to Char Cooler

Figure 2. PRETREATER HEAT BALANCE DATA SHEET FOR TEST 63 FOR STEADY PERIOD FROM 6/21/77 (1700 Hours) TO 6/22/77 (0600 Hours)

Table 3. MATERIAL BALANCE SUMMARY FOR THE PRETREATER SECTION FOR TEST 63 FROM 6/25/77 (0100 Hours) TO 6/25/77 (1200 Hours)

Basis = 1 hr. All units in lbs unless otherwise noted.

INPUT		C	H	O	N	S	Ar	Ash	Other	Total
Coal Feed	Wt % (Dry)	67.50	4.95	9.63	1.26	4.89		11.77		100
	Coal (Dry)	3312	243	472	62	240		578		4907
	Moisture		19	154						173
Streams to Pretreater	Air			879	2856		52			3787
	Steam		198	1582						1780
Nitrogen from purges					32					32
Air from purges				9	29					38
H_2O to venturi scrubber			2006	16,050						18,056
H_2O to quench tower			653	5221						5874
Air to char cooler				130	424		8			562
Cooling water to char cooler			95	756						851
TOTAL INPUT		3312	3214	25,253	3403	240	60	578		36,060
OUTPUT										
Pretreated Char to Gasifier	Wt.% (Dry)	67.37	3.53	8.44	1.39	4.35		14.92		100
	Char (Dry)	2562	134	321	53	165		567		3802
	Moisture		6	48						54
Slurry Waste from Quench	Wt.% (Dry)	57.00	2.56	11.62	1.49	4.33		23.00		100
	Solids (Dry)	137	6	28	4	10		55		240
	Tars & Oils	117	12	11	1	4				145
	H_2O & Dis. materials	25	2731	21,847	3	49				24,655
Quench Tower Off-Gas	Total	291	291	2998	3392		58			7030
	Components: H_2		1							1
	CO_2	156		416						572
	C_2H_6	3	1							4
	N_2				3392					3392
	CH_4	5	2							7
	CO	127		169						296
	O_2			120						120
	Ar						58			58
	H_2O		287	2293						2580
TOTAL OUTPUT		3132	3180	25,253	3453	228	58	622		35,926
Net (Output – Input)		–180	–34	0	50	–12	–2	44		–134
% Balance (Output/Input)		95	99	100	101	95	97	108		100



STREAM No.
PRESSURE, psig
TEMPERATURE, °F

lb Char/lb Coal =	0.76
SCF O_2/lb Coal=	2.1
Oxygen Breakthrough, % fed=	0
Superficial Velocity, ft/s =	0.95
lb O_2 Reacted/lb Coal=	0.18
Solids Residence Time, hr=	1.5
Bed Density, lb/cu ft =	19
Bed Height, ft =	9

GAS STREAM	3
COMPONENTS, mol %	
H_2	0.05
CO_2	4.42
C_2H_6	0.04
O_2	1.28
N_2	41.22
Ar	0.49
CH_4	0.14
CO	3.59
H_2O	48.77
mol/hr	294
mol wt	24

GAS STREAM	1	2
mol/hr	131	99

LIQUID STREAM	5	6
lb/hr	0	1906

SOLIDS STREAM	4	7
lb/hr	4907	3802

1 AIR TO PRETREATER
2 STEAM TO PRETREATER
3 PRETREATER OFF-GAS AFTER QUENCH
4 RAW COAL TO PRETREATER
5 DIRECT WATER QUENCH
6 COOLING WATER TO COOLING COIL IN PRETREATER
7 PRETREATER CHAR TO CHAR COOLER

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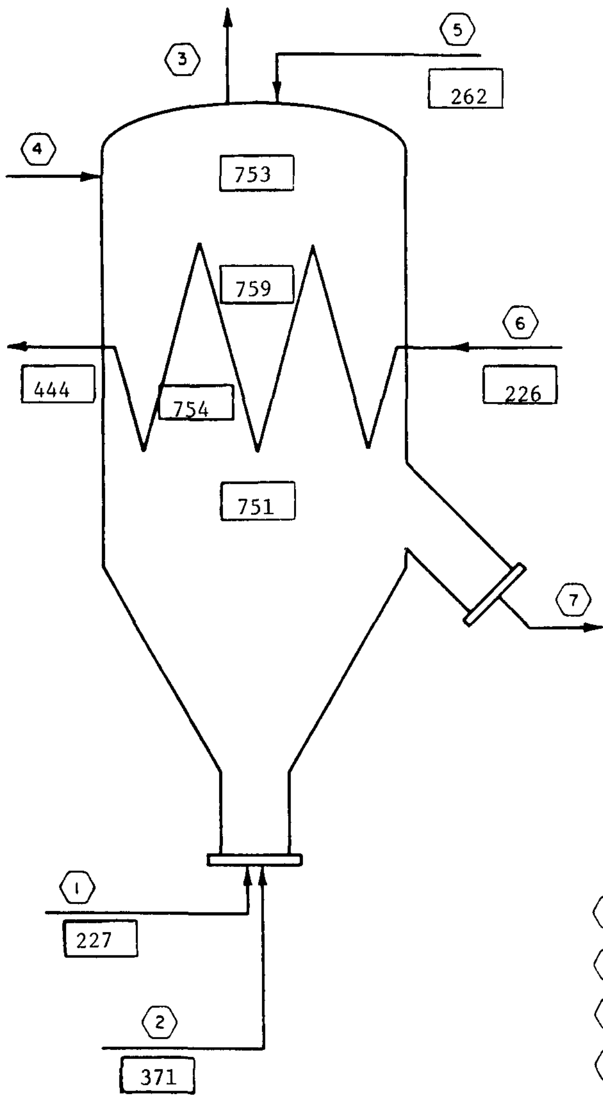
Figure 3. PRETREATMENT DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/25/77 (0100 Hours) TO 6/25/77 (1200 Hours)

Stream No. Temperature, °F

Basis: 1 hour

Datum Condition: 77°F, 1 atm, material in standard state.



INPUT	Btu
Sensible Heat (Streams 1, 2, 4, 5, 6)	741,150
Heat of Combustion (Stream 4)	59,948,819
Steam Enthalpy (Streams 2 & 5)	3,006,410
Total	63,696,379
OUTPUT	
Sensible Heat (Streams 3 & 7)	3,310,893
Heat of Combustion (Streams 3 & 7)	51,006,009
Steam Enthalpy (Streams 3 & 6)	5,002,884
Total	59,339,786
% Balance 93	

1 Air to Pretreater
2 Steam to Pretreater
3 Pretreater Overhead
4 Raw Coal to Pretreater
5 Gas From Char Cooler
6 Cooling Water to Cooling Coil in Pretreater
7 Pretreated Char to Char Cooler

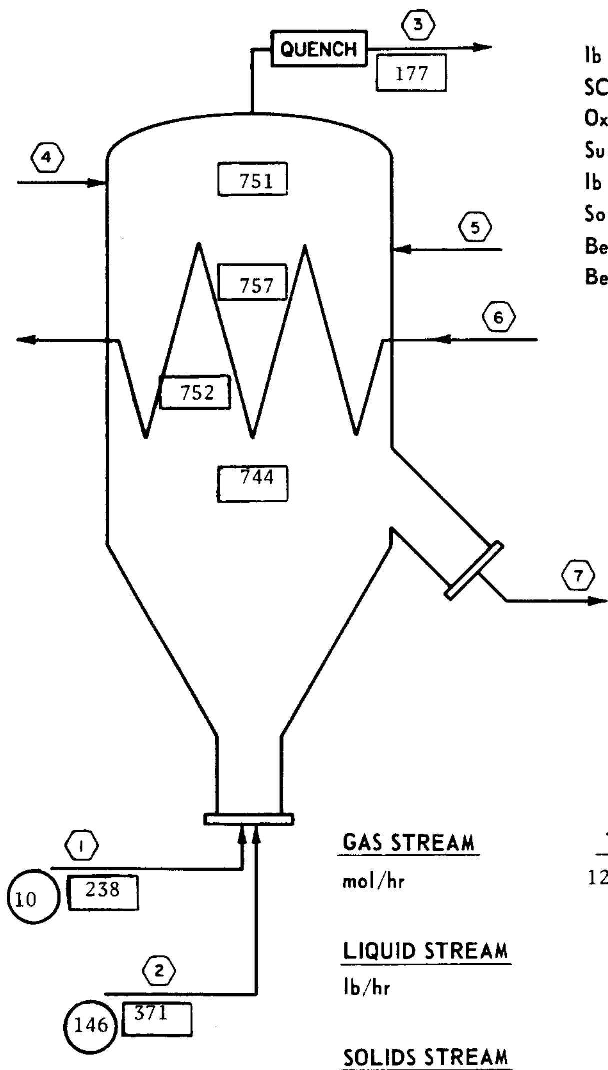
Figure 4. PRETREATER HEAT BALANCE DATA SHEET FOR TEST 63 FOR STEADY PERIOD FROM 6/25/77 (0100 Hours) TO 6/25/77 (0600 Hours)

Table 4. MATERIAL BALANCE SUMMARY FOR THE PRETREATER SECTION FOR TEST 63 FROM 6/25/77 (1400 Hours) TO 6/25/77 (1700 Hours)

Basis = 1 hr. All units in lbs unless otherwise noted.

INPUT		C	H	O	N	S	Ar	Ash	Other	Total
Coal Feed	Wt % (Dry)	67.50	4.82	10.04	1.25	4.72		11.67		100
	Coal (Dry)	3166	226	471	59	221		547		4690
	Moisture		18	147						165
Streams to Pretreater	Air			837	2719		49			3605
	Steam		195	1563						1758
Nitrogen from purges					33					33
Air from purges				9	29					38
H_2O to venturi scrubber			2006	16,048						18,054
H_2O to quench tower			653	5225						5878
Air to char cooler				136	442		8			586
Cooling water to char cooler			71	565						636
TOTAL INPUT		3166	3169	25,001	3282	221	57	547		35,443
OUTPUT										
Pretreated Char to Gasifier	Wt.% (Dry)	66.80	3.43	9.17	1.53	4.33		14.74		100
	Char (Dry)	2505	129	344	57	162		553		3750
	Moisture		9	75						84
Slurry Waste from Quench	Wt.% (Dry)	56.90	2.41	11.91	1.32	4.41		23.05		100
	Solids (Dry)	91	4	19	2	7		37		160
	Tars & Oils	102	11	10	1	3				127
	H_2O & Dis materials	25	2750	21,991	3	50				24,819
Quench Tower Off-Gas	Total	285	238	2562	3327		57			6469
	Components: H_2		1							1
	CO_2	153		407						560
	C_2H_6	3	1							4
	N_2				3327					3327
	CH_4	5	2							7
	CO	124		166						290
	O_2			118						118
	Ar						57			57
	H_2O		234	1871						2105
TOTAL OUTPUT		2008	3141	25,001	3390	222	57	590		35,409
Net (Output – Input)		–158	–28	0	108	–1	0	43		–34
% Balance (Output/Input)		95	99	100	103	100	100	108		100

STREAM No.
PRESSURE, psig
TEMPERATURE, °F



lb Char/lb Coal =	0.76
SCF O_2/lb Coal=	2.1
Oxygen Breakthrough, % fed=	0
Superficial Velocity, ft/s =	0.92
lb O_2 Reacted/lb Coal=	0.18
Solids Residence Time, hr =	1.3
Bed Density, lb/cu ft =	21
Bed Height, ft =	7

GAS STREAM	3
COMPONENTS, mol %	
H_2	0.06
CO_2	4.81
C_2H_6	0.04
O_2	1.39
N_2	44.90
Ar	0.54
CH_4	0.16
CO	3.91
H_2O	44.19
mol/hr	265
mol wt	24

GAS STREAM	1	2
mol/hr	125	98

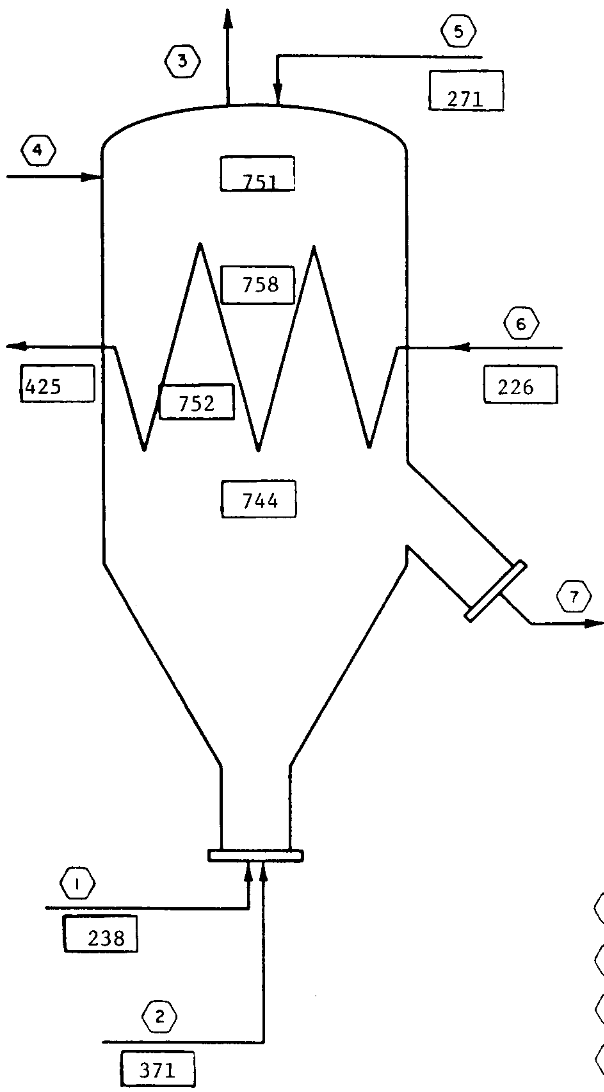
LIQUID STREAM	5	6
lb/hr	0	1546

SOLIDS STREAM	4	7
lb/hr	4690	3750

1 AIR TO PRETREATER
2 STEAM TO PRETREATER
3 PRETREATER OFF-GAS AFTER QUENCH
4 RAW COAL TO PRETREATER
5 DIRECT WATER QUENCH
6 COOLING WATER TO COOLING COIL IN PRETREATER
7 PRETREATER CHAR TO CHAR COOLER

A77051073

Figure 5. PRETREATMENT DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/25/77 (1400 Hours) TO 6/25/77 (1700 Hours)



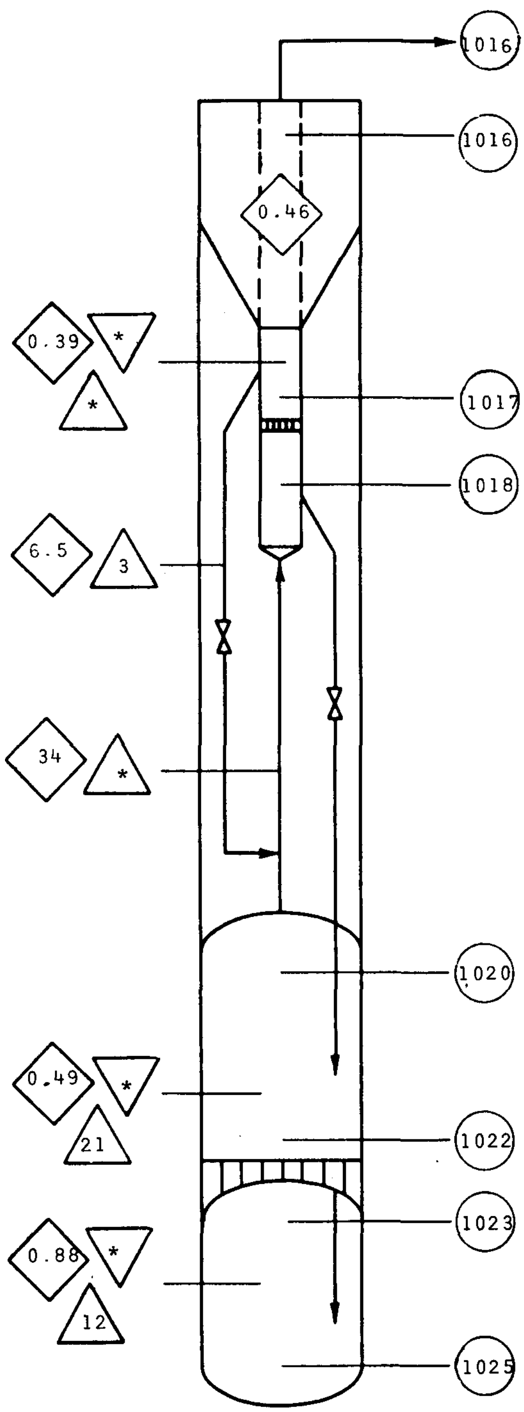
Basis: 1 hour

Datum Condition: 77°F, 1 atm, material in standard state.

INPUT	Btu
Sensible Heat (Streams 1, 2, 4, 5, 6)	669,747
Heat of Combustion (Stream 4)	57,391,530
Steam Enthalpy (Streams 2 & 5)	2,728,176
Total	60,789,453
OUTPUT	
Sensible Heat (Streams 3 & 7)	3,080,344
Heat of Combustion (Streams 3 & 7)	48,970,249
Steam Enthalpy (Streams 3 & 6)	4,087,094
Total	56,137,687
% Balance 92	

1. Air to Pretreater
2. Steam to Pretreater
3. Pretreater Overhead
4. Raw Coal to Pretreater
5. Gas From Char Cooler
6. Cooling Water to Cooling Coil in Pretreater
7. Pretreated Char to Char Cooler

Figure 6. PRETREATER HEAT BALANCE DATA SHEET FOR TEST 63 FOR STEADY PERIOD FROM 6/25/77 (1400 Hours) TO 6/25/77 (1700 Hours)



○ PRESSURE, psig
△ DENSITY, lb/cu ft
◇ VELOCITY, ft/s
▽ MEAN RESIDENCE TIME, min
* NOT AVAILABLE

Product Gas – dry, nitrogen- and acid-gas-free basis
Coal Fed – dry basis
Carbon (net) = total carbon in – carbon in overhead

lb Oxygen/ lb Carbon (net) =	0.35
lb Steam/ lb Carbon (net) =	3.6
lb Oxygen/lb Coal Fed=	0.20
lb Steam/lb Coal Fed=	2.1
lb Hydrogen/lb Coal Fed=	0
lb Coal Fed/1000 SCF Product Gas=	135

By Ash Balance

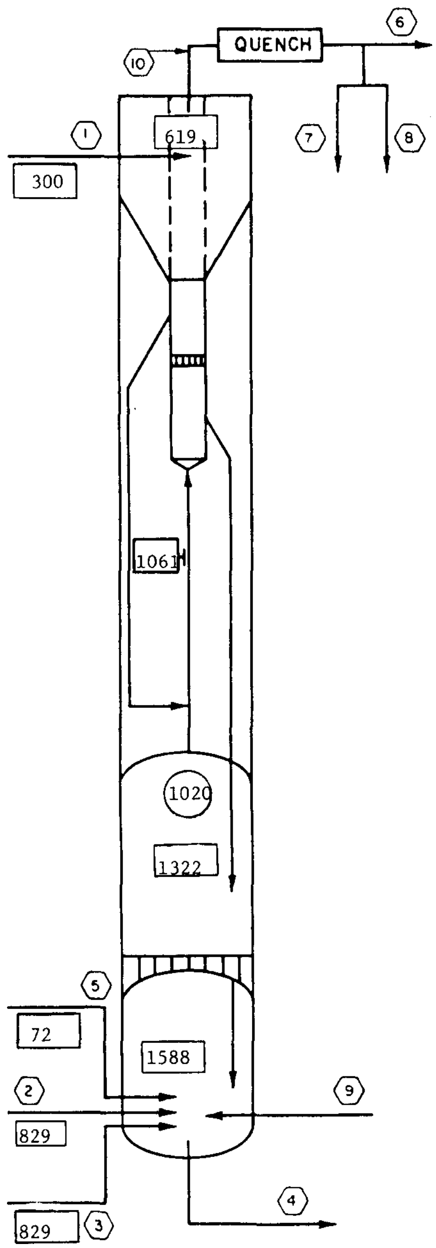
Coal Gasified, %= 52
Carbon Gasified, %= 42

Methane Yield SCF/ lb coal fed= 3.0
Equivalent Methane Yield, SCF/ lb coal fed = 4.3

Bed Height, ft

Slurry Dryer= *
HTR= 14
SOG= 22

Figure 7. HYGAS REACTOR ENGINEERING DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/21/77 (1700 Hours) TO 6/21/77 (2300 Hours)



MATERIAL BALANCE, %
C
H
O
ASH

⬡ STREAM No.
◯ PRESSURE, psig
▭ TEMPERATURE, °F

GAS STREAMS	6	10
mol % (dry)		
H_2	28.77	24.23
CO_2	28.94	36.63
C_2H_6	0.59	0.70
Ar	--	--
N_2	13.42	11.87
H_2S	1.24	1.59
CH_4	21.54	20.30
CO	5.50	4.68
mol/hr (dry)	124	152

SOLIDS STREAMS	1	10	4
wt % (dry)			
C	67.00	77.10	69.20
H	3.36	2.60	1.28
N	1.29	1.17	0.55
Cl	--	--	--
S	4.52	2.50	1.39
ASH	14.98	12.59	26.70
O	8.85	4.04	0.88
lb/hr (dry)	3889		

MOISTURE, wt % 2.0

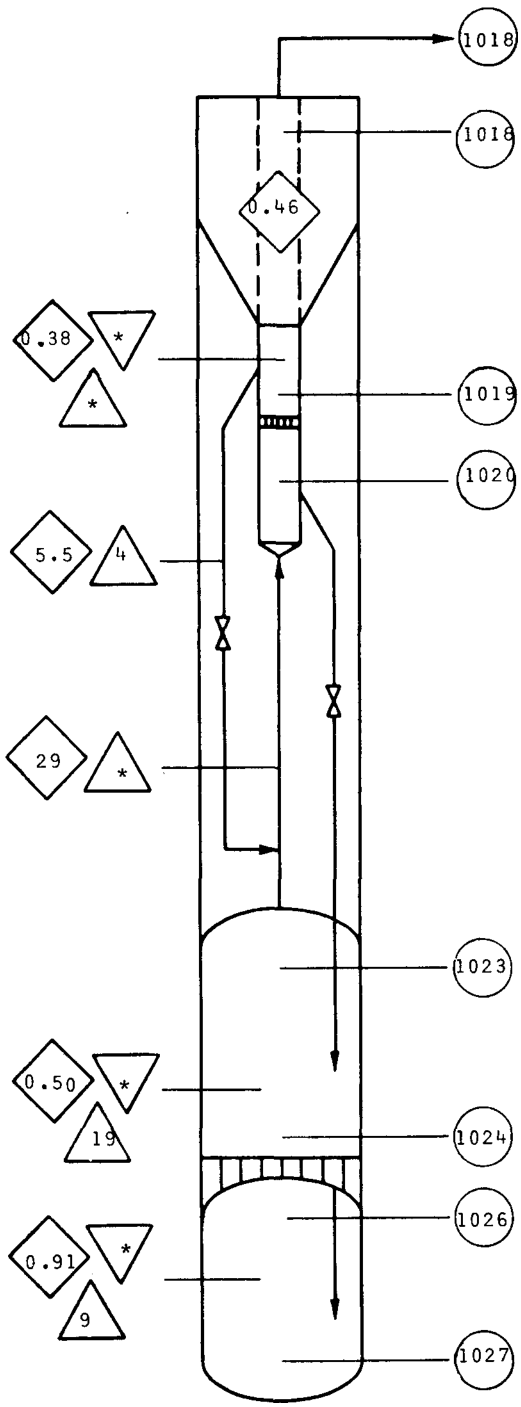
SLURRY CONCN, wt % 29

LIQUID STREAMS	7	8	1
	lb/hr		
H_2O	7898	--	--
LIGHT OIL	--	--	9954

GAS FEED STREAMS	2	3	5	9
	mol/hr			
	366	83	24	0

1 FEED SLURRY
2 HIGH-PRESSURE STEAM TO STEAM-OXYGEN SPARGER
3 HIGH-PRESSURE STEAM TO STRIPPING RING
4 SPENT CHAR
5 HIGH-PRESSURE OXYGEN TO STEAM-OXYGEN SPARGER
6 GASIFIER EFFLUENT AFTER QUENCH
7 WATER MADE
8 LIGHT OIL TO RECYCLE
9 GASES FROM EXTERNAL HEATER
10 REACTOR OVERHEAD

Figure 8. HYGAS REACTOR DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/21/77 (1700 Hours) TO 6/21/77 (2300 Hours)



○ PRESSURE, psig
△ DENSITY, lb/cu ft
◇ VELOCITY, ft/s
▽ MEAN RESIDENCE TIME, min
* NOT AVAILABLE

Product Gas – dry, nitrogen- and acid-gas-free basis
Coal Fed – dry basis
Carbon (net) = total carbon in – carbon in overhead

lb Oxygen/lb Carbon (net) =	0.36
lb Steam/lb Carbon (net) =	3.5
lb Oxygen/lb Coal Fed =	0.23
lb Steam/lb Coal Fed =	2.2
lb Hydrogen/lb Coal Fed =	0
lb Coal Fed/1000 SCF Product Gas =	115

By Ash Balance

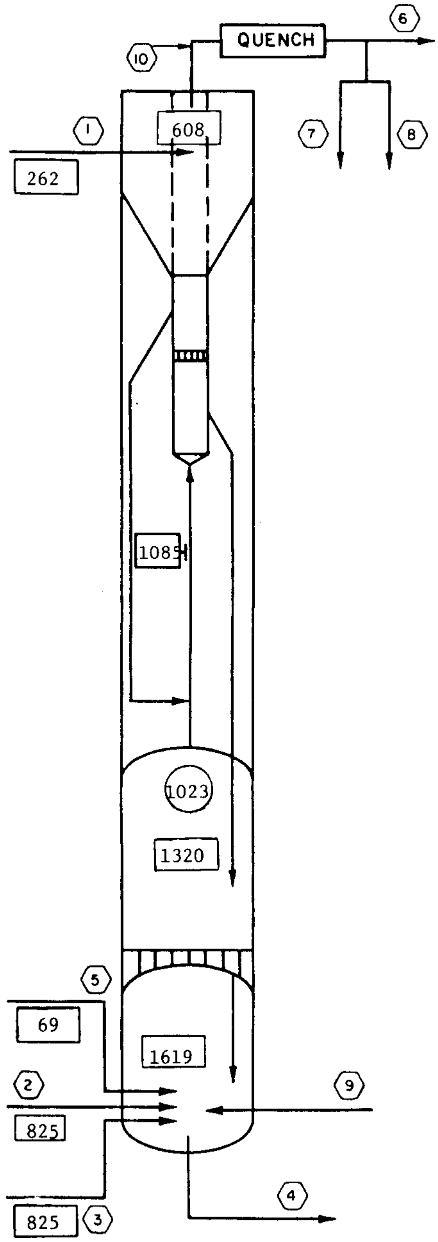
Coal Gasified, % =	64
Carbon Gasified, % =	57

Methane Yield SCF/lb coal fed = 3.4
Equivalent Methane Yield, SCF/lb coal fed = 4.9

Bed Height, ft

Slurry Dryer = *
HTR = 16
SOG = 27

Figure 9. HYGAS REACTOR ENGINEERING DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/22/77 (0200 Hours) TO 6/22/77 (0800 Hours)



MATERIAL BALANCE, %
C
H
O
ASH

⬡ STREAM No.
◯ PRESSURE, psig
▭ TEMPERATURE, °F

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	29.63	26.59
CO_2	30.76	35.59
C_2H_6	0.63	0.69
Ar	--	--
N_2	10.04	9.60
H_2S	1.06	1.26
CH_4	21.69	20.78
CO	6.19	5.49
mol/hr (dry)	136	157

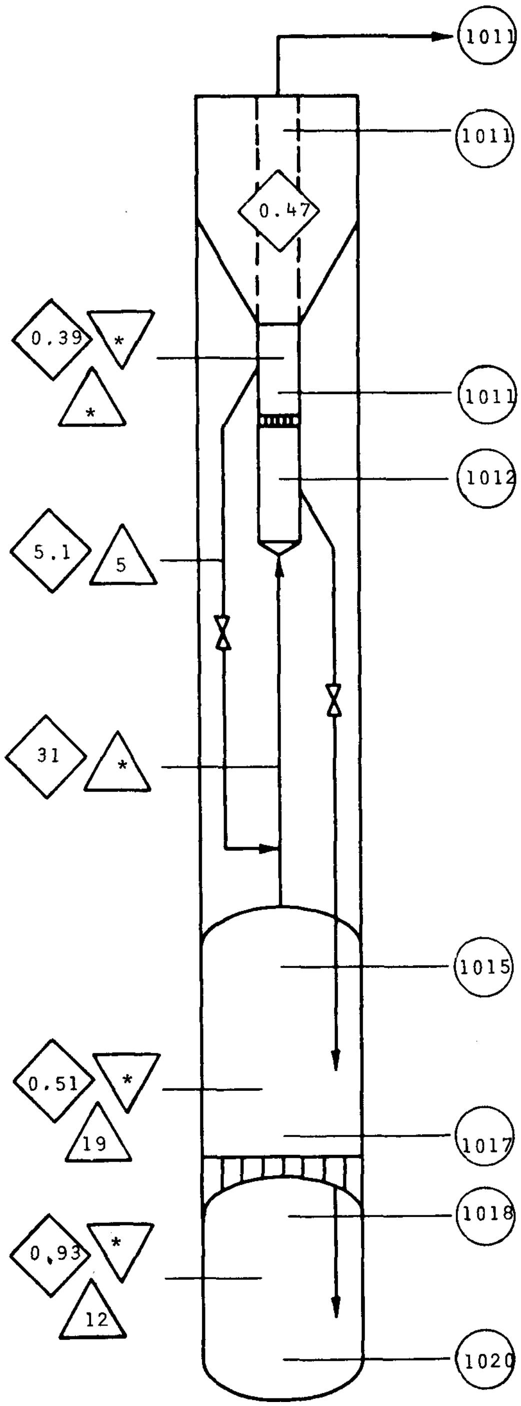
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	67.70	76.20	64.60
H	3.51	3.37	1.16
N	1.37	1.15	0.42
Cl	--	--	--
S	4.26	2.98	1.04
ASH	14.49	13.23	32.08
O	8.67	3.07	0.70
lb/hr (dry)	3664		
MOISTURE, wt %	2.0		
SLURRY CONCN, wt %	26		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	7754	--	--
LIGHT OIL	--	--	10,855

GAS FEED STREAMS			
2	3	5	9
mol/hr			
370	83	26	0

1. FEED SLURRY
2. HIGH-PRESSURE STEAM TO STEAM-OXYGEN SPARGER
3. HIGH-PRESSURE STEAM TO STRIPPING RING
4. SPENT CHAR
5. HIGH-PRESSURE OXYGEN TO STEAM-OXYGEN SPARGER
6. GASIFIER EFFLUENT AFTER QUENCH
7. WATER MADE
8. LIGHT OIL TO RECYCLE
9. GASES FROM EXTERNAL HEATER
10. REACTOR OVERHEAD

Figure 10. HYGAS REACTOR DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/22/77 (0200 Hours) TO 6/22/77 (0800 Hours)



○ PRESSURE, psig
△ DENSITY, lb/cu ft
◇ VELOCITY, ft/s
▽ MEAN RESIDENCE TIME, min
* NOT AVAILABLE

<u>Product Gas</u> – dry, nitrogen- and acid-gas-free basis
<u>Coal Fed</u> – dry basis
<u>Carbon (net)</u> = total carbon in – carbon in overhead

lb Oxygen/lb Carbon (net) =	0.41
lb Steam/lb Carbon (net) =	3.5
lb Oxygen/lb Coal Fed =	0.23
lb Steam/lb Coal Fed =	2.0
lb Hydrogen/lb Coal Fed =	0
lb Coal Fed/1000 SCF Product Gas =	89

<u>By Ash Balance</u>

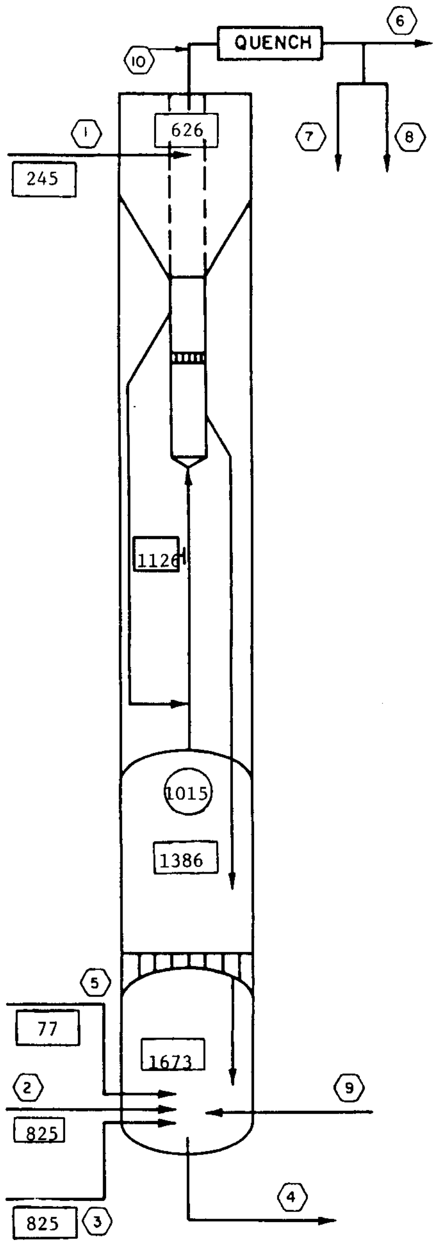
Coal Gasified, % =	64
Carbon Gasified, % =	57

Methane Yield SCF/lb coal fed = 4.2
Equivalent Methane Yield, SCF/lb coal fed = 6.2

<u>Bed Height, ft</u>

Slurry Dryer = *
HTR = 15
SOG = 24

Figure 11. HYGAS REACTOR ENGINEERING DATA FOR TEST 63 STEADY PERIOD FROM 6/23/77 (1800 Hours) TO 6/24/77 (1300 Hours)



MATERIAL BALANCE, %	
C	
H	
O	
ASH	

⬡ STREAM No.
◯ PRESSURE, psig
▭ TEMPERATURE, °F

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	28.92	25.92
CO_2	31.37	36.19
C_2H_6	0.59	0.68
Ar	--	--
N_2	8.54	8.18
H_2S	1.11	1.40
CH_4	21.40	20.52
CO	8.07	7.11
mol/hr (dry)	189	218

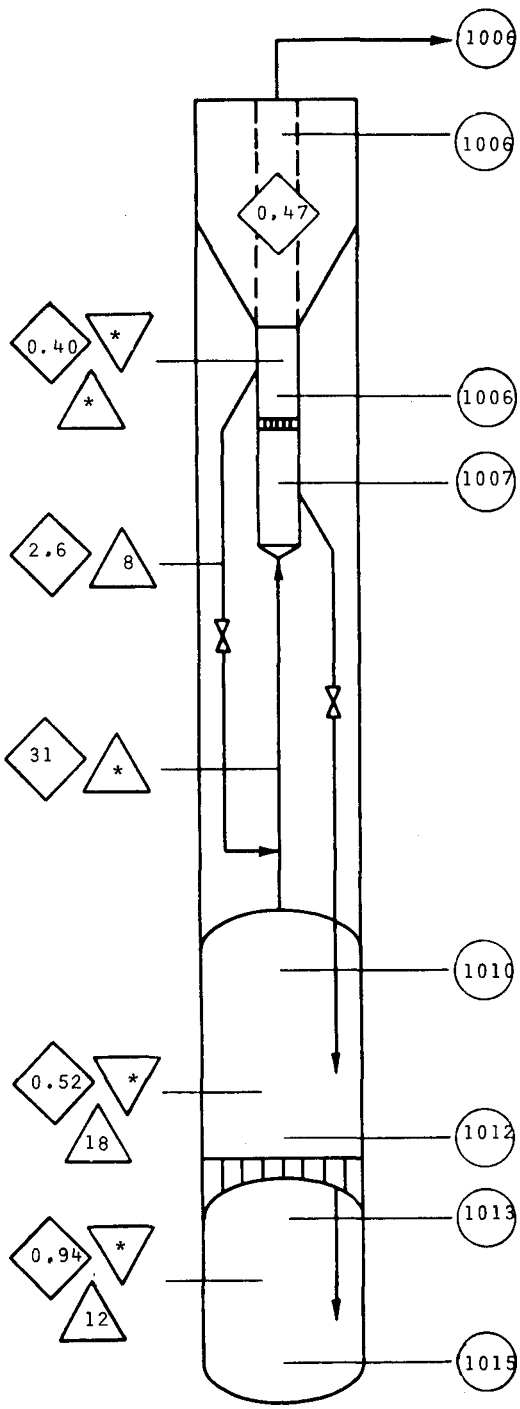
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	66.87	76.50	62.66
H	3.37	2.59	1.06
N	1.45	1.18	0.38
Cl	--	--	--
S	4.29	2.79	1.02
ASH	15.56	13.67	34.13
O	8.46	3.27	0.75
lb/hr (dry)	3999		
MOISTURE, wt %	1.7		
SLURRY CONCN, wt %	30		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	6778	--	--
LIGHT OIL	--	--	9737

GAS FEED STREAMS			
2	3	5	9
mol/hr			
359	85	29	0

(1) FEED SLURRY
(2) HIGH-PRESSURE STEAM TO STEAM-OXYGEN SPARGER
(3) HIGH-PRESSURE STEAM TO STRIPPING RING
(4) SPENT CHAR
(5) HIGH-PRESSURE OXYGEN TO STEAM-OXYGEN SPARGER
(6) GASIFIER EFFLUENT AFTER QUENCH
(7) WATER MADE
(8) LIGHT OIL TO RECYCLE
(9) GASES FROM EXTERNAL HEATER
(10) REACTOR OVERHEAD

Figure 12. HYGAS REACTOR DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/23/77 (1800 Hours) TO 6/24/77 (1300 Hours)



○ PRESSURE, psig
△ DENSITY, lb/cu ft
◇ VELOCITY, ft/s
▽ MEAN RESIDENCE TIME, min
* NOT AVAILABLE

Product Gas – dry, nitrogen- and acid-gas-free basis
Coal Fed – dry basis
Carbon (net) = total carbon in – carbon in overhead

lb Oxygen/lb Carbon (net) =	0.46
lb Steam/lb Carbon (net) =	4.0
lb Oxygen/lb Coal Fed =	0.25
lb Steam/lb Coal Fed =	2.1
lb Hydrogen/lb Coal Fed =	0
lb Coal Fed/1000 SCF Product Gas =	87

By Ash Balance

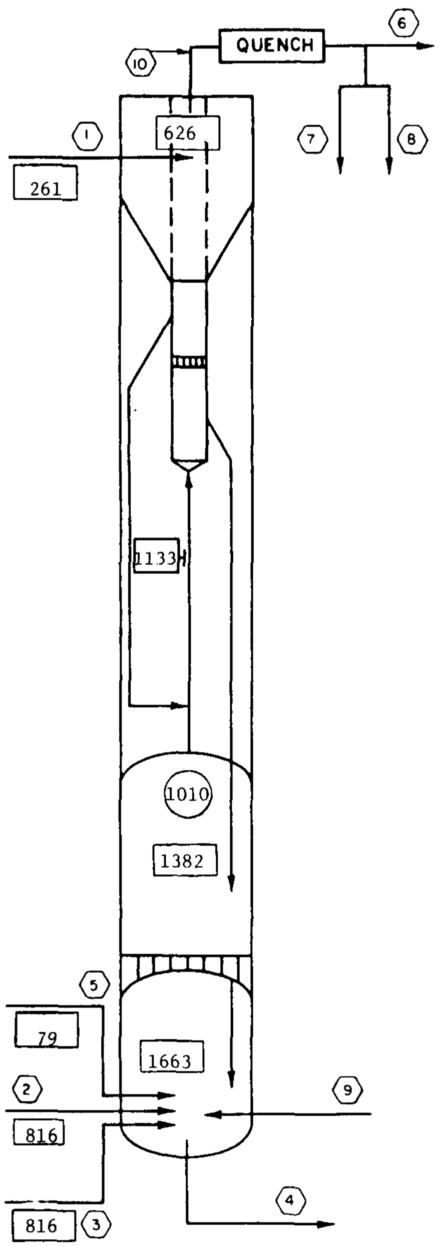
Coal Gasified, % =	66
Carbon Gasified, % =	62

Methane Yield SCF/lb coal fed = 4.2
Equivalent Methane Yield, SCF/lb coal fed = 6.3

Bed Height, ft

Slurry Dryer = *
HTR = 16
SOG = 23

Figure 13. HYGAS REACTOR ENGINEERING DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/23/77 (1800 Hours) TO 6/25/77 (1700 Hours)



MATERIAL BALANCE, %

C

H

O

ASH

◯ STREAM No.
◯ PRESSURE, psig
▭ TEMPERATURE, °F

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	29.69	26.51
CO_2	30.80	35.70
C_2H_6	0.59	0.67
Ar	--	--
N_2	8.81	8.52
H_2S	1.26	1.50
CH_4	20.80	19.95
CO	8.05	7.15
mol/hr (dry)	184	213

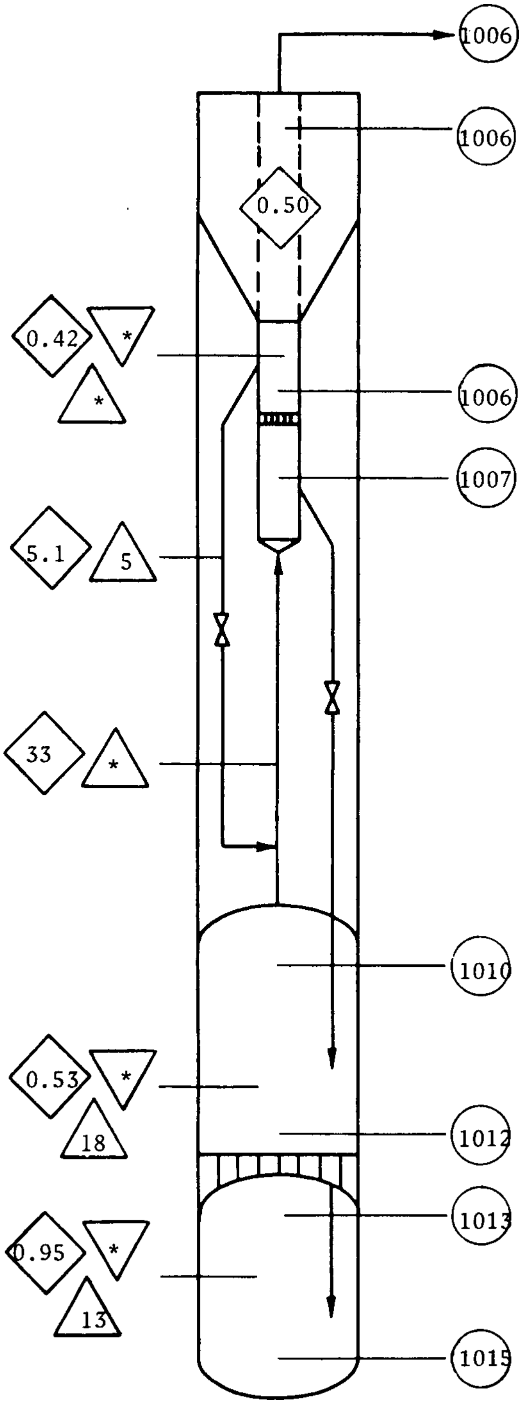
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	68.34	75.29	59.32
H	3.49	2.65	1.01
N	1.37	1.15	0.35
Cl	--	--	--
S	4.53	2.82	1.06
ASH	14.71	14.48	33.92
O	7.56	3.61	4.34
lb/hr (dry)	3810		
MOISTURE, wt %	2.1		
SLURRY CONCN, wt %	27		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	6946	--	--
LIGHT OIL	--	--	10,347

GAS FEED STREAMS			
2	3	5	9
mol/hr			
365	86	30	0

(1) FEED SLURRY
(2) HIGH-PRESSURE STEAM TO STEAM-OXYGEN SPARGER
(3) HIGH-PRESSURE STEAM TO STRIPPING RING
(4) SPENT CHAR
(5) HIGH-PRESSURE OXYGEN TO STEAM-OXYGEN SPARGER
(6) GASIFIER EFFLUENT AFTER QUENCH
(7) WATER MADE
(8) LIGHT OIL TO RECYCLE
(9) GASES FROM EXTERNAL HEATER
(10) REACTOR OVERHEAD

Figure 14. HYGAS REACTOR DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/23/77 (1800 Hours) TO 6/25/77 (1700 Hours)



○ PRESSURE, psig
△ DENSITY, lb/cu ft
◇ VELOCITY, ft/s
▽ MEAN RESIDENCE TIME, min
* NOT AVAILABLE

<u>Product Gas</u> – dry, nitrogen- and acid-gas-free basis
<u>Coal Fed</u> – dry basis
<u>Carbon (net)</u> = total carbon in – carbon in overhead

lb Oxygen/lb Carbon (net) = 0.49
lb Steam/lb Carbon (net) = 4.1
lb Oxygen/lb Coal Fed = 0.25
lb Steam/lb Coal Fed = 2.1
lb Hydrogen/lb Coal Fed = 0
lb Coal Fed/1000 SCF Product Gas = 86

<u>By Ash Balance</u>

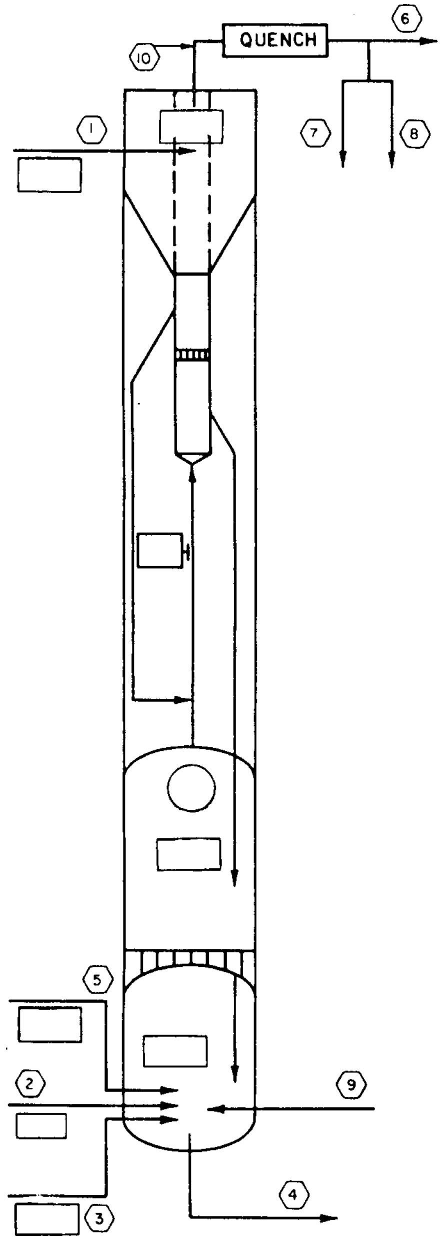
Coal Gasified, % = 75
Carbon Gasified, % = 71

Methane Yield SCF/lb coal fed = 4.2
Equivalent Methane Yield, SCF/lb coal fed = 6.2

<u>Bed Height, ft</u>

Slurry Dryer = *
HTR = 16
SOG = 22

Figure 15. HYGAS REACTOR ENGINEERING DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/24/77 (1400 Hours) TO 6/25/77 (1000 Hours)



STREAM No.
PRESSURE, psig
TEMPERATURE, °F

MATERIAL BALANCE, %
C
H
O
ASH

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	30.06	26.87
CO_2	30.74	26.87
C_2H_6	0.58	0.64
Ar	--	--
N_2	8.68	8.18
H_2S	1.37	1.50
CH_4	20.49	19.74
CO	8.08	7.16
mol/hr (dry)	186	215

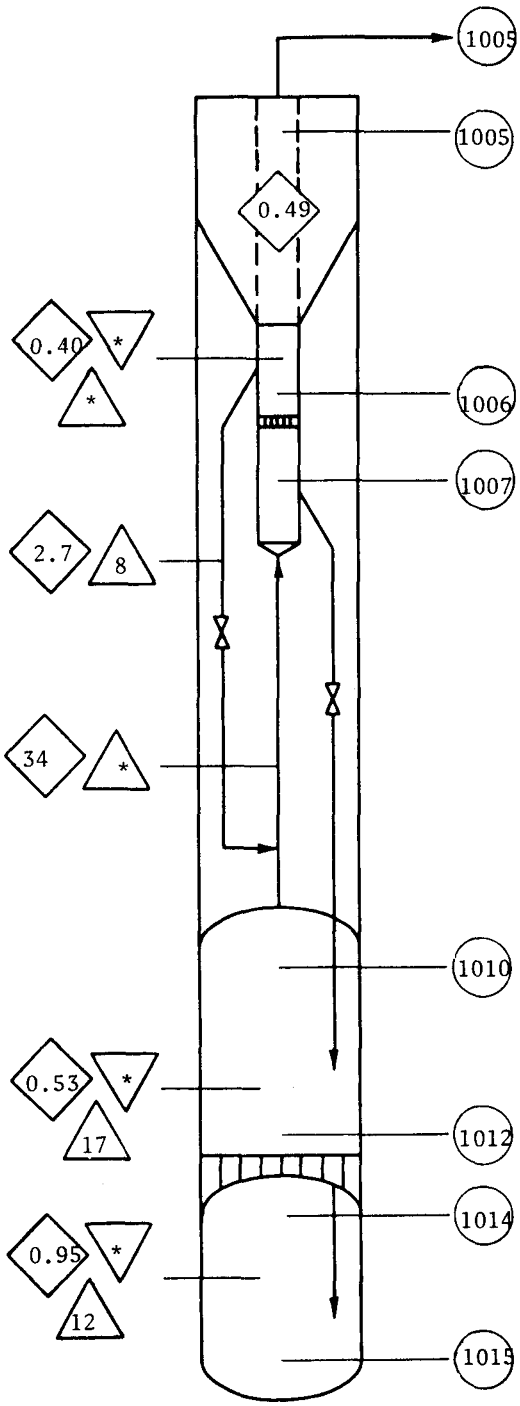
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	68.36	75.20	56.10
H	3.51	2.62	0.96
N	1.37	1.14	0.34
Cl	--	--	--
S	4.53	2.82	1.00
ASH	14.78	14.60	41.35
O	7.45	3.62	0.25
lb/hr (dry)	3834		
MOISTURE, wt %	2.1		
SLURRY CONCN, wt %	27		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	7300	--	--
LIGHT OIL	--	--	10,688

GAS FEED STREAMS			
2	3	5	9
mol/hr			
369	86	30	0

1 FEED SLURRY
2 HIGH-PRESSURE STEAM TO STEAM-OXYGEN SPARGER
3 HIGH-PRESSURE STEAM TO STRIPPING RING
4 SPENT CHAR
5 HIGH-PRESSURE OXYGEN TO STEAM-OXYGEN SPARGER
6 GASIFIER EFFLUENT AFTER QUENCH
7 WATER MADE
8 LIGHT OIL TO RECYCLE
9 GASES FROM EXTERNAL HEATER
10 REACTOR OVERHEAD

Figure 16. HYGAS REACTOR DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/24/77 (1400 Hours) TO 6/25/77 (1000 Hours)



○ PRESSURE, psig
△ DENSITY, lb/cu ft
◇ VELOCITY, ft/s
▽ MEAN RESIDENCE TIME, min
* NOT AVAILABLE

Product Gas – dry, nitrogen- and acid-gas-free basis
Coal Fed – dry basis
Carbon (net) = total carbon in – carbon in overhead

lb Oxygen/lb Carbon (net) =	0.50
lb Steam/lb Carbon (net) =	4.1
lb Oxygen/lb Coal Fed =	0.27
lb Steam/lb Coal Fed =	2.2
lb Hydrogen/lb Coal Fed =	0
lb Coal Fed/1000 SCF Product Gas =	83

By Ash Balance

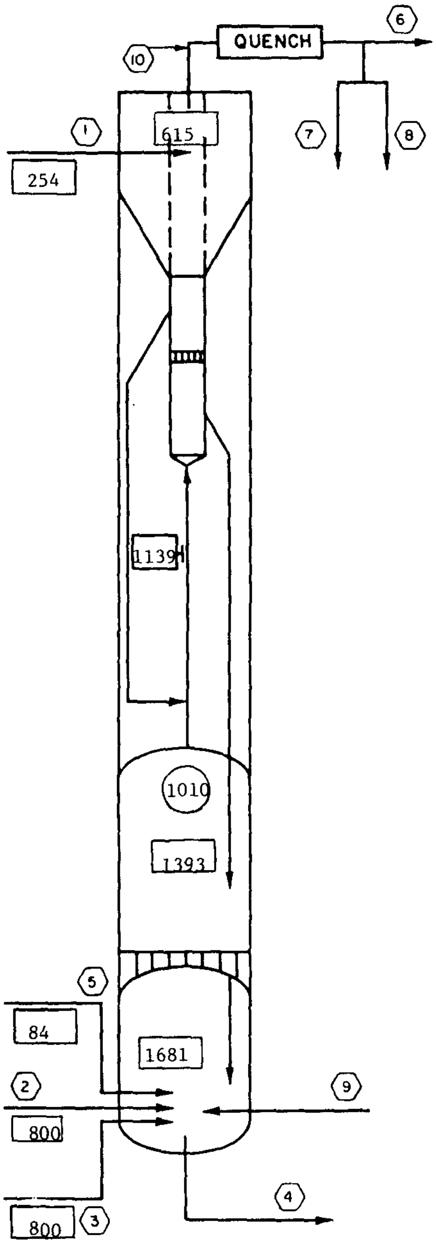
Coal Gasified, % =	74
Carbon Gasified, % =	69

Methane Yield SCF/lb coal fed = 4.4
Equivalent Methane Yield, SCF/lb coal fed = 6.5

Bed Height, ft

Slurry Dryer = *
HTR = 17
SOG = 23

Figure 17. HYGAS REACTOR ENGINEERING DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/25/77 (1200 Hours) TO 6/25/77 (1700 Hours)



STREAM No. (hexagon)
PRESSURE, psig (circle)
TEMPERATURE, °F (box)

MATERIAL BALANCE, %

C	
H	
O	
ASH	

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	30.63	26.91
CO_2	30.16	36.11
C_2H_6	0.61	0.66
Ar	--	--
N_2	8.39	7.87
H_2S	1.22	1.38
CH_4	20.68	19.79
CO	8.31	7.28
mol/hr (dry)	182	215

	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	68.70	74.80	57.85
H	3.47	2.70	0.93
N	1.30	1.14	0.31
Cl	--	--	--
S	4.69	2.83	0.75
ASH	14.35	14.79	39.57
O	7.49	3.74	0.59
lb/hr (dry)	3702		
MOISTURE, wt %	2.4		
SLURRY CONCN, wt %	24		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	7042	--	--
LIGHT OIL	--	--	11,828

GAS FEED STREAMS			
2	3	5	9
mol/hr			
367	86	31	0

1. FEED SLURRY
2. HIGH-PRESSURE STEAM TO STEAM-OXYGEN SPARGER
3. HIGH-PRESSURE STEAM TO STRIPPING RING
4. SPENT CHAR
5. HIGH-PRESSURE OXYGEN TO STEAM-OXYGEN SPARGER
6. GASIFIER EFFLUENT AFTER QUENCH
7. WATER MADE
8. LIGHT OIL TO RECYCLE
9. GASES FROM EXTERNAL HEATER
10. REACTOR OVERHEAD

Figure 18. HYGAS REACTOR DATA FOR TEST 63 FOR STEADY PERIOD FROM 6/25/77 (1200 Hours) TO 6/25/77 (1700 Hours)

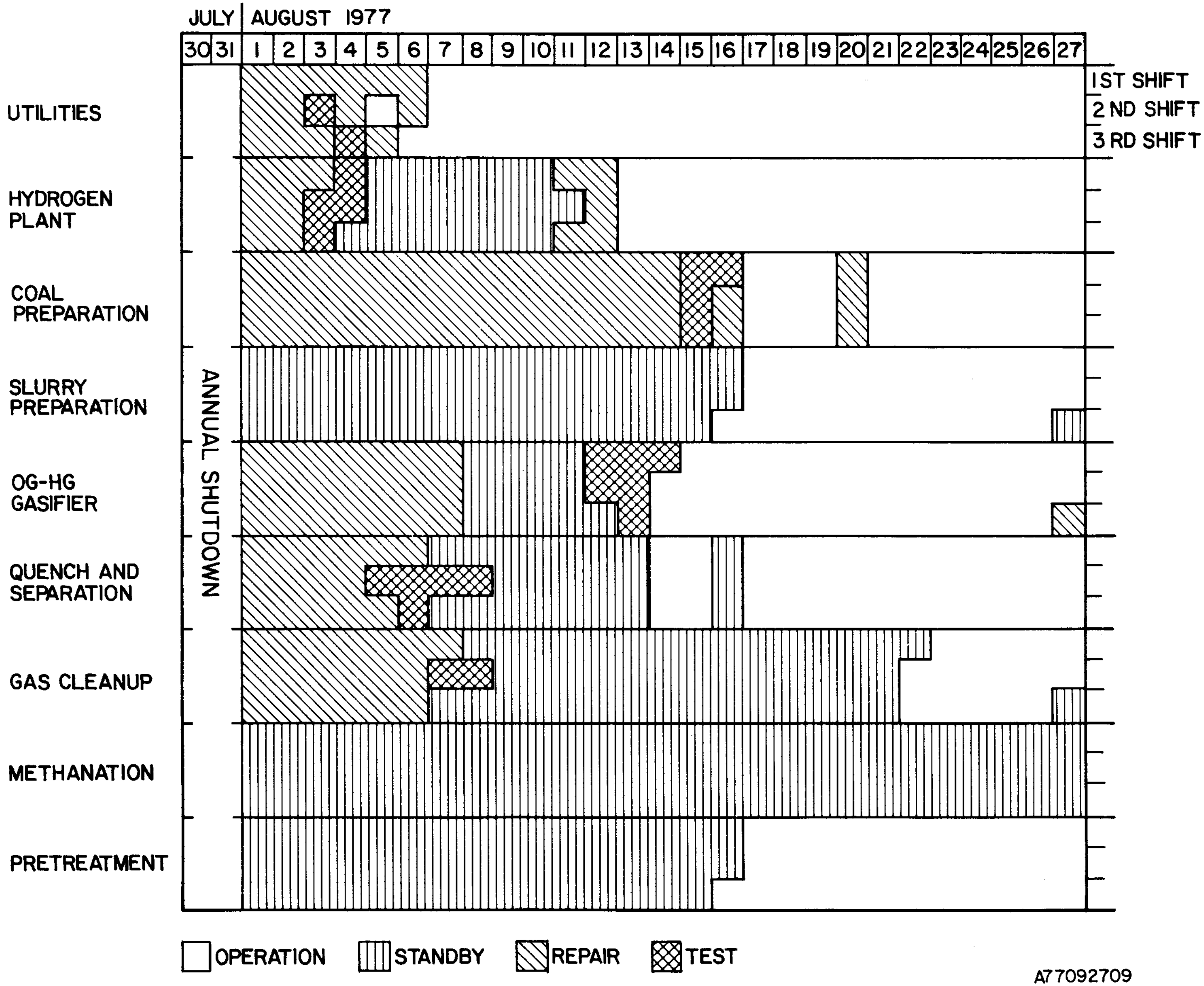


Figure 19. MECHANICAL STATUS OF THE HYGAS PLANT FOR AUGUST 1977

discussions were held on the data requirements that Procon must have for the initiation of its design study.

The objective of the work effort under this task is to provide ERDA-Major Facilities Program Management (MFPM) and Procon with engineering assistance in the design of a commercial/demonstration plant based on the HYGAS Steam-Oxygen Process.

Initially, IGT supplied Procon with typical analyses of lignite, subbituminous, and mid-continent bituminous coals, as possible candidates for the design coal. In the meeting with ERDA-Major Facilities Program Management (MFPM) and Procon, the coal selection was narrowed to lignite, mid-continent run-of-mine, and mid-continent washed bituminous coals for a starting point for material balance calculations for the plant design. Basic material balances, product yields, and feed requirements will be calculated with each coal type and at pressure levels of 800, 1000, and 1200 psig in the gasifier. This will allow Procon to start on the selection process for the commercial plant design case and allow trade-off studies to be initiated on the effect of pressure on plant optimization. Details of these calculations will be reported as they become available.

The cold-flow modeling program will be initiated as soon as ERDA, Procon, and IGT can make a selection on the reactor configuration.

Task 9. Operation Support Studies

A. Plant Effluent Processing

The light-oil recovery unit and the solids recovery unit both worked satisfactorily for Test 64. The new surge pot was not put in service because of a leaking packing on the mechanical stirrer. Equipment and instrumentation for a new high-capacity incinerator are being installed. Shakedown of the new incinerator is expected in mid-September.

B. Test Methanation Systems and Catalysts

Purified product gas from the HYGAS reactor was fed to Chem Systems liquid-phase methanation (LPM) pilot unit starting at 1730 hours on August 24. The LPM unit operated for 69 continuous hours prior to the shutdown of Test 64. Carbon monoxide conversion levels in the pilot unit ranged from 45% to 98%. This test in the LPM pilot unit is very important because it is the first time that this unit has received purified product gas from the HYGAS plant for an extended period of time. Prior to Test 64, screeners were installed in the suction side of the oil circulation pumps in the LPM unit because some ceramic inerts were found in these pumps during a hot-oil circulation test.

The IGT fixed-bed catalyst, cold-gas recycle methanation system was on standby for Test 64. However, bench-scale catalyst evaluation studies were continued by IGT.

Recent Methanation Catalyst Evaluation Studies

We continued our evaluation studies of the methanation catalyst, LDI X-826. The same batch of catalyst, which was used in experimental Runs 476 through 487 (ERDA Report No. FE-2434-9, January 1977), was tested again. The reactor was sealed off in a hydrogen atmosphere after Run 487 at 52 psig. When we introduced a feed gas, similar to that used in Run 480, to the catalyst bed after 2000 hours of inactivity to determine if the catalyst was deactivated by aging, we found that the catalyst had not been deactivated. This finding was expected, because the catalyst had not been subjected to unfavorable conditions such as high or low temperatures or to poisons.

Steam, which consisted of about 20 mole percent of the feed, was added to ensure that the catalyst would perform adequately in a recycle methanator, which might accumulate a higher concentration of steam than that produced stoichiometrically by the methanation reaction. The results are presented in Figure 20 and Table 5. Again, as expected, the activity of the catalyst was not affected by this amount of steam. These experiments, consisting of Runs 476 through 492, concluded the first evaluation step of our standard program in which the activity of the catalyst is established within an ideal temperature range and with a feed composition simulating the gas effluent from the coal gasification reactor.

C. Investigate Hot-Oil Quench System

Initial design work was started with a study of equilibrium between light oil, gas, and water to determine what condensate compositions could be expected.

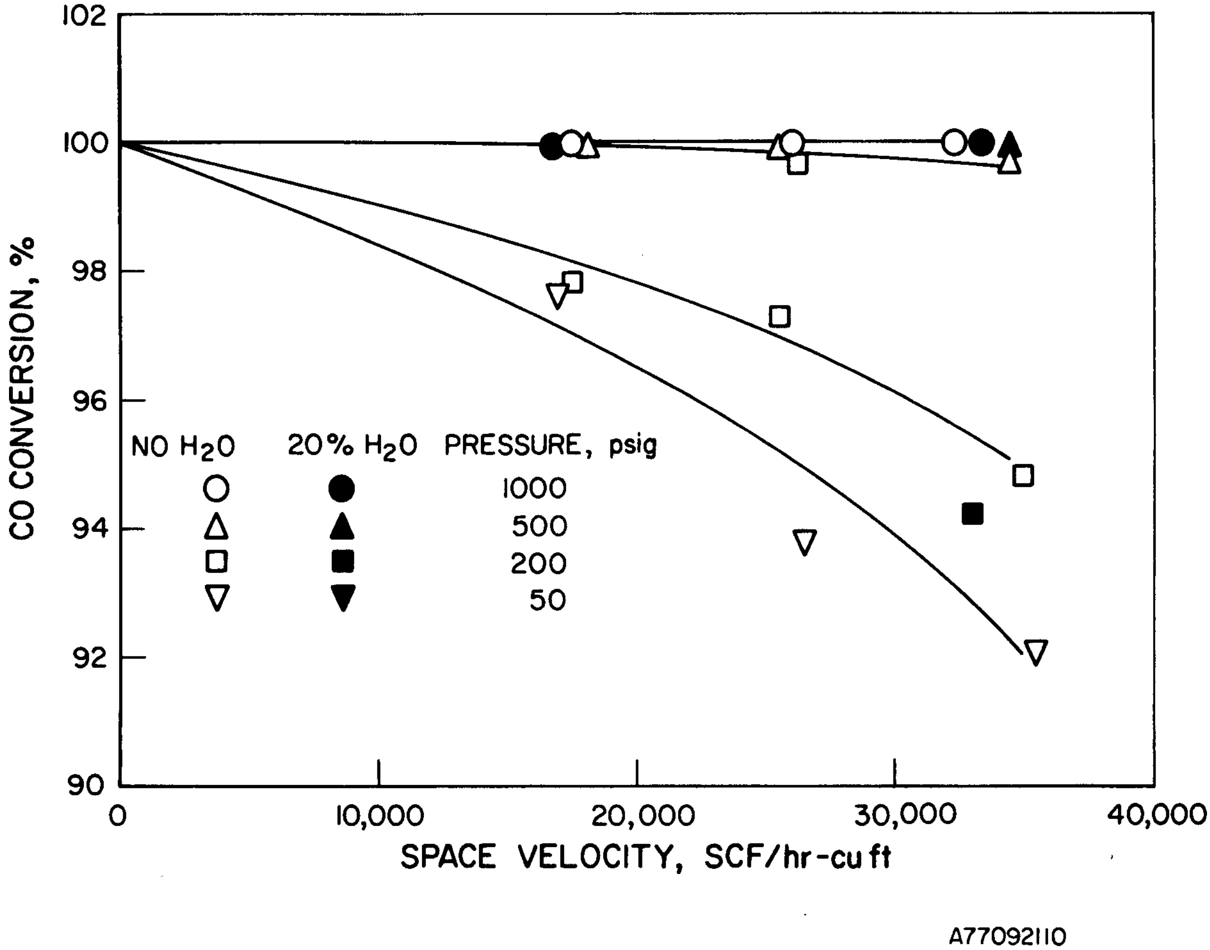


Figure 20. EFFECT OF SPACE VELOCITY AND STEAM ON THE CONVERSION OF CARBON MONOXIDE AT TEMPERATURES RANGING FROM 550° TO 680°F (LDI X-826 Catalyst, 1/8-Inch Cylinders)

D. Materials Testing

Exposure for MPC corrosion and erosion test coupons was carried out during Test 64. Nondestructive testing was performed on slurry lines and high-pressure lines in the HYGAS pilot plant prior to Test 64. Results of this test will be reported as soon as all evaluations have been made.

E. Engineering Services

A cold-model simulation study of the operation of line 339 between the second-stage, high-temperature reactor (HTR) and the steam-oxygen beds was undertaken to explain why solids flow could not be started between the HTR and the steam-oxygen beds. An existing low-pressure solids recirculation unit (used for nonmechanical valve testing) was modified to include two Plexiglas fluidized beds as shown in Figures 21 and 22.

In operation, solids from the solids receiver at the top of the unit were transferred through an L-valve into the upper fluidized bed (simulating the HTR). Solids from the upper fluidized bed were transferred to the lower fluidized bed (simulating the steam-oxygen bed) through a 3-inch overflow pipe fitted with a flapper valve, similar to the one at the bottom of line 339 (Figure 23). The solids from the lower bed were then transferred into the lift line, which transported the solids back into the solids receiver.

Air was used for the fluidizing medium and for the lift gas. Provision was made for fluidizing the two fluid beds independently, and also for fluidizing the upper bed with the fluidizing gas from the lower bed. Most of the runs were made using the latter technique. After passing through the beds, the gas was exhausted to the atmosphere.

Initial tests were made with –40+120 mesh size Rosebud subbituminous coal. After Test 64 was initiated, a supply of pretreated material was used in all of the remaining runs. No significant difference between the two types of solids was observed in these tests.

Solids were successfully passed down the simulated 339 line using two different techniques. With the first technique, the 339 flapper valve was first closed, and gas flow was started through the lower bed. Solids were then charged to the upper (HTR) bed through the L-valve, and a bed level higher than the overflow pipe was established. After line 339 was filled with solids (determined by visual inspection), the flapper valve was cracked open and solids were passed into the lower (steam-oxygen) bed. With this technique, the solids level in the upper bed was always higher than the overflow pipe height. The solids in line 339 were in packed-bed (stick-slip) flow, and the solids flow rate into the lower bed was controlled by the opening of the flapper valve. Increasing the flapper valve opening increased the solids flow. If the flapper valve position remained constant, solids flow could also be controlled by the aeration gas rate fed to line 339

Table 5 . METHANATION CATALYSIS — EVALUATION OF LDI CATALYST CO. LDI X-826 CATALYST
(1/8-Inch, 22.29-Gram Cylinders)

Run No.	488		489		490		491		492	
Time, hr	599		631		679		799		871	
Basis for Analysis	Dry	Wet	Dry	Wet	Dry	Wet	Dry	Wet	Dry	Wet
Pressure, psig	200	200	1000	1000	1000	1000	200	200	500	500
Reactor Temperature, °F										
Inlet	483	483	510	510	342	342	310	310	315	315
Quarter Bed	578	578	595	595	542	542	543	543	546	546
Middle Bed	568	568	609	609	538	538	540	540	540	540
Outlet	662	662	611	611	659	659	649	649	640	640
Furnace Temperature, °F										
Top Zone	478	478	478	478	478	478	472	472	475	475
Bottom Zone	568	568	570	570	592	592	592	592	549	549
Flow Rate, lb-mol/hr										
Feed	0.04866	0.04866	0.03312	0.04195	0.06469	0.08277	0.06529	0.08332	00.06628	0.08417
H_2O	0	0	0	0.00884	0	0.01808	0	0.01803	0	0.01789
Feed Composition, mol %										
H_2	10.29	10.29	14.15	11.16	12.3	9.6	10.2	8.0	10.2	8.03
N_2	2.4	2.4	2.17	1.71	2.3	1.8	2.5	1.9	2.6	2.04
CH_4	81.08	81.08	79.4	62.71	80.7	63.1	82.23	64.4	82.33	64.8
C_2H_6	0.27	0.27	0.19	0.14	.2	0.19	0.29	0.23	0.27	0.21
C_3H_8	0	0	0	0	0.0	0.04	0	0	0	0
C_4H_{10}'s	0	0	0	0	0	0	0	0	0	0
CO_2	2.4	2.4	1.31	1.03	1.4	1.09	1.4	1.1	1.4	1.1
CO	3.4	3.4	2.63	2.07	2.6	2.0	3.2	2.5	3.2	2.52
He	0.16	0.16	0.15	0.12	0.4	0.31	0.18	0.14	0	0
H_2O	0	0	0	21.06	0	21.87	0	21.73	0	0
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Flow Rate, lb-mol/hr										
Product	0.04423	0.04602	0.02913	0.03969	0.05903	0.07994	0.06106	0.08055	0.06234	0.08212
H_2O in Product	0	0.001785	0	0.01056	0	0.02091	0	0.01949	0	0.01978
Product Composition, mol %										
H_2	0.27	0.26	6.3	4.7	4.3	3.2	3.1	2.4	4.4	3.3
N_2	2.7	2.6	2.4	1.8	2.5	1.8	2.7	2.0	2.6	1.97
CH_4	95.13	91.4	89.74	66.7	91.09	67.4	92.96	70.5	90.67	68.4
C_2H_6	0.02	0.02	0	0	0.07	0.05	0.02	0.01	0.03	0.02
C_3H_8	0	0	0	0	0	0	0	0	0	0
C_4H_{10}'s	0	0	0	0	0	0	0	0	0	0
CO_2	1.6	1.53	1.3	0.93	1.6	1.2	0.82	0.62	2.3	1.74
CO	0.1	0.1	0	0	0	0	0.2	0.15	0	0
He	0.18	0.17	0.26	0.12	0.44	0.33	0.2	0.15	0	0
H_2O	0	3.91	0	25.75	0	26.02	0	24.17	0	24.57
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
CO Consumed, lb-mol/hr	0.001607	--	0.000868	--	0.001682	--	0.001967	--	0.002121	--
CO_2 Changed, lb-mol/hr	−0.000469	--	−0.000052	--	+0.000059	--	−0.000415	--	+0.000501	--
H_2 Consumed, lb-mol/hr	0.004885	--	0.00276	--	0.005383	--	0.00473	--	0.00405	--
H_2O Produced, lb-mol/hr	0.001785	--	0.001728	--	0.002830	--	0.00146	--	0.001890	--
CH_4 Produced, lb-mol/hr	o.002233	--	0.000986	--	0.001999	--	0.00308	--	0.001622	--
C_2H_6 Consumed, lb-mol/hr	0.000122	--	0.000059	--	0.000122	--	0.000181	--	0.000162	--
C_3H_8 Consumed, lb-mol/hr	0	--	0	--	0.000032	--	--	--	--	--
Space Velocity, SCF/hr-cu ft	25,313	--	17,238	--	33,671	--	33,754	--	34,500	--

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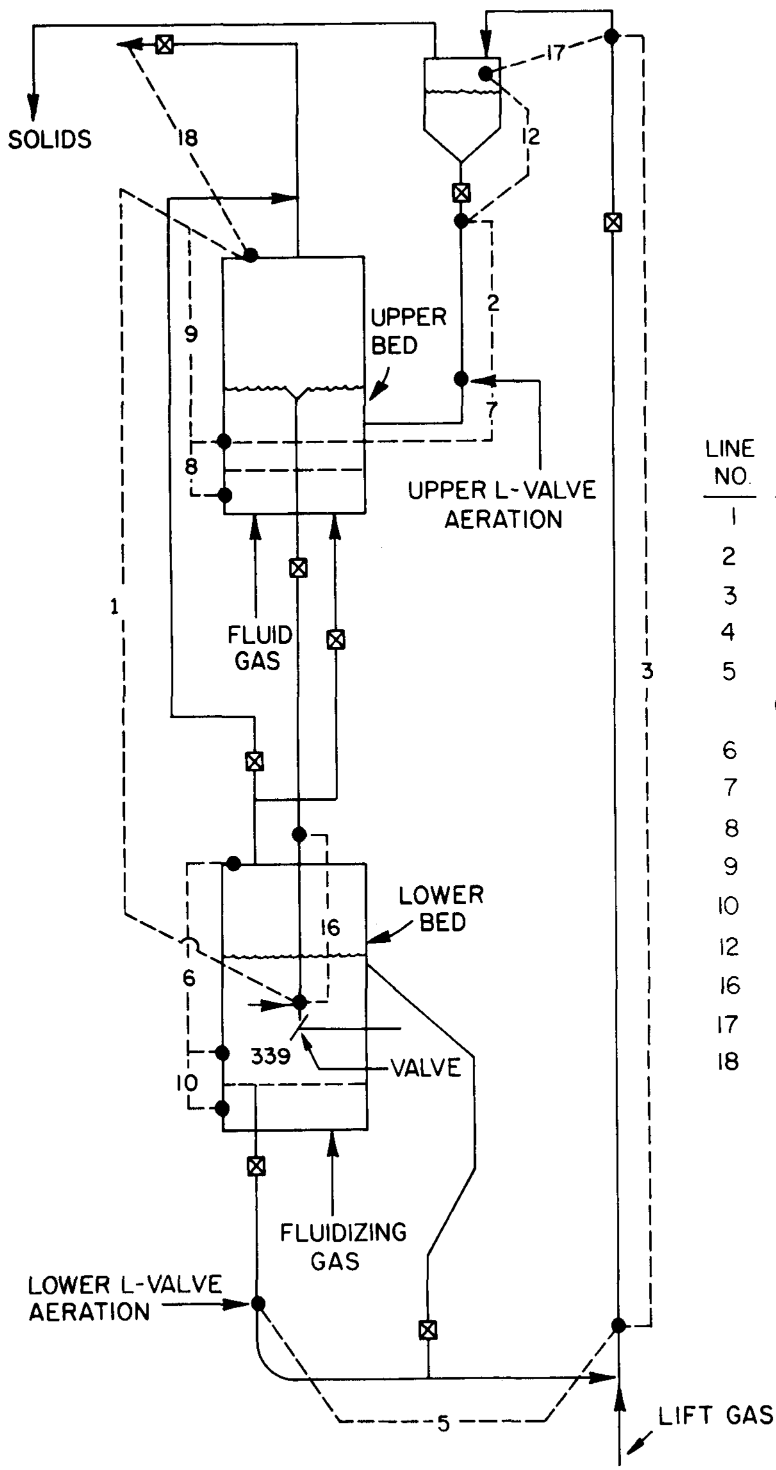

LINE NO.	Δp CELL	PRESS. RANGE, in. H_2O
1	BETWEEN BED DOWNCOMER	0-50
2	UPPER BED L-VALVE DOWNCOMER	0-100
3	TOTAL LIFT	0-100
4	ORIFICE	0-100
5	LOWER BED L-VALVE DOWNCOMER OR L-HORIZONTAL	0-50
6	BED Δp (Lower Bed)	0-25
7	UPPER BED L-HORIZONTAL	0-25
8	UPPER BED DISTRIBUTOR	0-25
9	BED Δp (Upper Bed)	0-25
10	LOWER BED DISTRIBUTOR	0-25
12	SOLIDS RECEIVER	0-50
16	DOWNCOMER	0-25
17	BEND	0-50
18	OUTLET GAS	0-25

● = PRESSURE TAP

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Figure 21. INSTRUMENTATION SCHEMATIC FOR THE COLD-MODEL UNIT

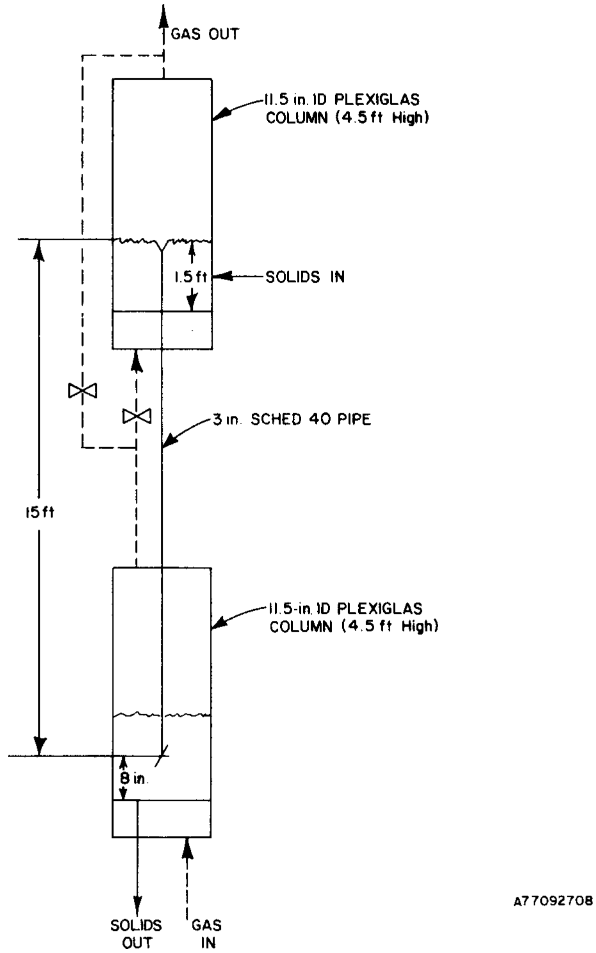


Figure 22. COLD-MODEL FLOWSHEET USED TO SIMULATE THE OPERATION OF LINE 339

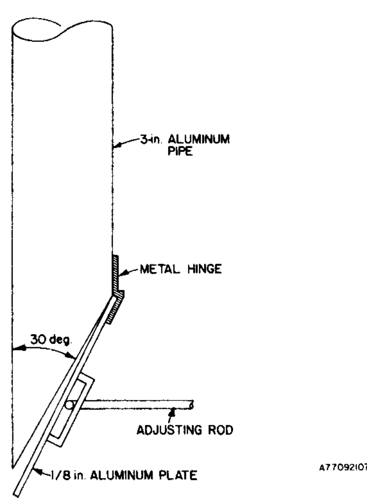


Figure 23. CROSS SECTION OF SIMULATED LINE 339 FLAPPER VALVE

at a location approximately 4 inches above the flapper valve. Using this technique, the solids flow rates to the upper bed, through the flapper valve, and out of the lower bed had to be matched to achieve constant bed levels.

With the second technique, the flapper valve was closed and gas flow started through the lower bed. Solids were charged to the upper bed, and a bed level higher than the overflow pipe was established. After line 339 was filled with solids, the flapper valve was used to control the solids flow rate into the lower bed until the solids covered the flapper valve. The flapper valve was then opened all the way letting line 339 function as a dipleg. The upper bed then operated as an overflow bed. Line 339 developed a fluidized-bed seal leg high enough (approximately 3 to 4 feet) to overcome the pressure drop thorugh the lower bed above the flapper valve, the upper bed distributor, and the upper bed. The upper part of line 339 had solids streaming down in a dilute-phase flow in this model. Using this technique, the solids flow rate into the upper bed had only to be matched by the solids flow rate out of the lower bed to keep the bed levels constant.

Both modes of flow could be established if line 339 could be filled with solids; however, it could not be filled with solids if the gas velocity through the lower bed was too high. Flapper valve 339, although fully closed, would let a substantial quantity of leakage gas pass up line 339, thus preventing the solids from filling the line.

The gas rate that would prevent the pretreated solids from filling line 339 was determined in the following manner. The gas rate to the lower bed was set very high, and the upper bed was filled to a level above the overflow pipe. Valve 339 was closed. At this high gas rate, the solids would not fill line 339. The gas rate to the lower bed was then decreased until the solids were seen falling down line 339. At this point, the bed pressure drop (1.25 inches H_2O) across the upper distributor was noted. A 3-inch, full-port ball valve in line 339 was closed routing all of the gas to the lower bed through the upper bed distributor. The upper bed distributor pressure drop was again noted (9.5 inches H_2O), and the amount of gas passing through the distributor in both instances was calculated using the orifice equation. (See the section on calculations.)

The difference in gas flow was calculated, and we found that a gas velocity of 6.6 ft/s in line 339 was the gas velocity at which the solids would start to fill the line. The amount of gas passing up line 339 in the unit was over 60% of the gas fed to the lower bed when the flapper valve was closed.

The terminal velocity of the average particle size of the pretreated solids was also calculated and found to be 6.8 ft/s. (See the section on calculations.) This agrees excellently with the experimental results.

Apparently, the reason solids will not pass down line 339 when trying to fill the steam-oxygen bed is because of excessive gas flow through the "closed" flapper valve and up line 339. This gas flow exceeds the terminal velocity of the particles passing line 339, and they cannot fill the line.

Because the HYGAS pilot plant operates with higher gas densities than the cold model, the gas flow up line 339 in the plant would not need to be 6 to 7 ft/s to prevent solids from filling the line. The calculated terminal velocity of the solids in the pilot plant is approximately 3.1 ft/s; thus, the gas rate through the closed flapper valve could be much less than in the cold model and still prevent solids from filling the pipe. The approximate volumetric flow rate in the steam-oxygen bed is $(\frac{\pi}{4})(2)^2(0.9)(60) =$ 170 CF/min. At 3.1 ft/s the volumetric flow rate up line 339 is approximately $(3.1)(0.05)(60) = 9.3$ CF/min, or only 5.5% of the total gas flow.

At the start of Test 64 in the HYGAS pilot plant, the steam-oxygen bed was successfully filled using a technique similar to that used to fill line 339 in the cold model. The gas rate to the steam-oxygen bed was decreased to as low a value as possible, and with no initial bed in the HTR, line 339 filled. The flapper valve was then cracked open, and solids flow to the steam-oxygen bed was started. The steam-oxygen bed was then filled satisfactorily. However, sometime later because of a faulty instrument reading the steam-oxygen bed was lost. Attempts to restart the solids flow failed, i.e., line 339 could not be filled with solids. This failure is probably due to the fact that there was a full bed in the HTR (when upon initial start-up there was none) which caused more gas to pass up line 339 than before, thus exceeding the terminal velocity in the line and preventing solids from filling it.

Several things could be done to solve the gas bypassing problem, but each solution has advantages and disadvantages. If a positive shutoff valve (such as a ball valve) could be installed in line 339 it would solve the problem. However, there are no such valves that can stand up under the conditions found in the reactor, and such a valve would also be difficult to install.

A curved nonmechanical valve could be installed at the bottom of line 339. With this valve if the steam-oxygen bed were lost for any reason, the valve would retain solids (much like a drain trap in a sink) and give a greater resistance to gas flow than a leaky flapper valve. Thus, line 339 could be filled and solids flow started again. However, this type of valve would have to have some means of closing off its end for initial start-up. It would also be difficult to install.

Another approach would be to convert the overflow line 399 to an underflow line. This would have the effect of decreasing the amount of gas flow up the line (because of the effect of the bed above the line) if the steam-oxygen bed were lost, and would enable the line to be filled with solids using the present flapper valve. However, the convenience of operating an overflow HTR bed would be lost, and the solids flow rate out of flapper valve 339 would have to be matched to the solids flow rate into the HTR bed and to the solids flow rate out of the steam-oxygen bed.

Calculations

The upper distributor area is equivalent to fifty-seven 1/8-inch diameter holes.

Area of 1/8-inch diameter hole : 0.00008522 sq ft

Distributor open area: 0.004857 sq ft

3-inch, Schedule 40 pipe area (line 339 area): 0.0513 sq ft

Condition 1

Δp across upper distributor when flapper valve 339 is closed and ball valve is open: 1.25 inches H_2O.

Condition 2

Δp across upper distributor when flapper valve 339 is closed and ball valve is closed: 9.5 inches H_2O.

The gas flow rate through distributor holes for Condition 1 is —

$$U_{or} = C_d\sqrt{\frac{2g_c\ \Delta P}{\rho_g}} = 0.6\sqrt{\frac{2(32.2)\ 1.25\ (144)}{0.095\ (27.7)}}$$

$$= 39.8 \text{ ft/s} \rightarrow 0.19 \text{ CF/s} = 11.6 \text{ CF/min}$$

The gas flow rate through distributor holes for Condition 2 is —

$$U_{or} = 109.8 \text{ ft/s} \rightarrow 0.533 \text{ CF/s} = 32.0 \text{ CF/min}$$

Flow difference: 32.0 — 11.6 = 20.4 CF/min

Flow rate up line 339: $\frac{20.4}{(60)\ (0.0513)} = 6.6$ ft/s

Percent flow up line 339: $\frac{20.4}{32}\ (100) = 63.8\%$

The terminal velocity of the pretreated solids (cold-flow unit) is —

$$U_t = \left[\frac{4}{225}\ \frac{(\rho_s-\rho_g)^2\ g^2)}{\rho_g\ \mu}\right]^{1/3} D_p$$

$$= \left[\frac{4}{225}\ \frac{(60 - 0.095)^2\ 32^2}{(0.095)\ (1 \times 10^{-5})}\right]^{1/3} 0.0017 \simeq 6.8 \text{ ft/s}$$

The terminal velocity of the pretreated solids in the HYGAS gasifier (HTR) is —

$$U_t = \left[\frac{4}{225}\ \frac{(60 - 1)^2\ 32^2}{(1)\ 1 \times 10^{-5}}\right]^{1/3} 0.0017 \simeq 3.1 \text{ ft/s}$$

IV. PROBLEMS

No critical problems were encountered during August.

V. PATENT STATUS

The work performed during August, as reported herein, is not considered patentable.

VI. WORK PLAN AND SCHEDULE

A work schedule for the HYGAS Process is presented in Figure 24.

Approved B. S. Lee — Signed Wilford G. Bair

Bernard S. Lee
Vice President

Wilford G. Bair
Director

TASK 7. PILOT PLANT EXPERIMENTAL OPERATION

A. PEABODY NO. 10 MINE WASHED BITUMINOUS COAL*

1. INVESTIGATE HIGH CARBON CONVERSION

2. INVESTIGATE MAXIMUM PILOT PLANT CAPACITY

B. INVESTIGATE MINIMUM PRETREATMENT CONDITIONS

C. INVESTIGATE FINES UTILIZATION*

1. TEST DOUBLE SCREENED FEED

2. TEST FINES REINJECTION

D. RUN-OF-MINE BITUMINOUS COAL

1. INVESTIGATE HIGH CARBON CONVERSION

2. INVESTIGATE MAXIMUM PILOT PLANT CAPACITY

TASK 8. DEMONSTRATION PLANT DESIGN SUPPORT

A. EXPERIMENTAL PROCESS STUDIES AND COLD FLOW MODELING OF REACTOR

B. ASSIST ERDA-MFPM IN PLANT DESIGN TASKS

TASK 9. OPERATIONS SUPPORT STUDIES

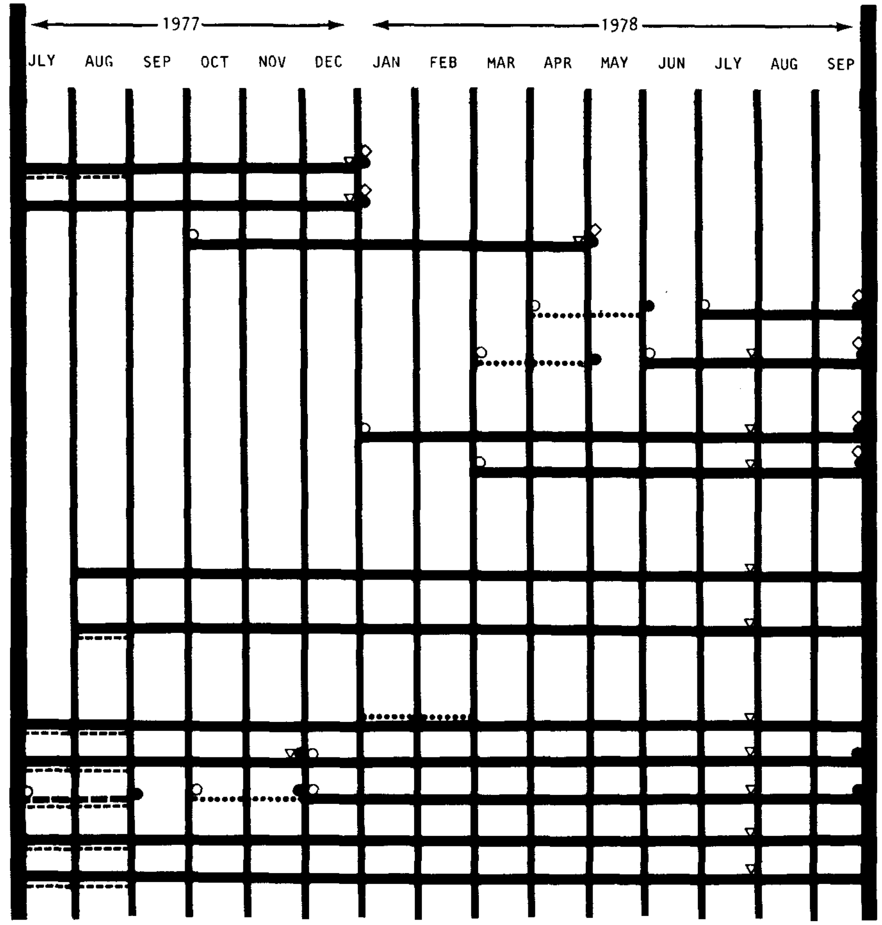
A. PLANT EFFLUENT PROCESSING*

B. TEST METHANATION SYSTEMS AND CATALYSTS*

C. INVESTIGATE HOT OIL QUENCH SYSTEM

D. MATERIALS TESTING*

E. ENGINEERING SERVICES*



*THIS TASK INITIATED UNDER PRIOR CONTRACT

LEGEND

O INITIATE
● COMPLETE
◇ MILESTONE
▽ DECISION POINT
▬▬ DESIGN EFFORT
████ TEST OPERATION
••••••• INSTALL EQUIPMENT
------ ACTUAL PROGRESS

Figure 24. WORK SCHEDULE FOR HYGAS PROCESS FOR AUGUST 1977