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FOSSIL ENERGY

PIPELINE GAS FROM COAL-HYDROGENATION (IGT HYDROGASIFICATION PROCESS)

Quarterly Report No. 5, July 1—September 30, 1977

February 1978
Date Published

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

Institute of Gas Technology
Chicago, Illinois

MASTER

U. S. DEPARTMENT OF ENERGY

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For the Period July 1 Through September 30, 1977

Prepared by

Institute of Gas Technology

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Chicago, Illinois 60616

Date Published — February 1978

Prepared for the

UNITED STATES DEPARTMENT OF ENERGY

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TABLE OF CONTENTS

TABLE OF CONTENTS, Cont.

LIST OF FIGURES

LIST OF FIGURES, Cont.

LIST OF TABLES

SUMMARY

This quarterly report covers work conducted between July 1 and September 30, 1977. Early in July, a post-run inspection and cleanup were conducted after Test 63. Details are reported together with results of Test 63.

The annual plant turnaround was conducted during July. This turnaround period is required for routine maintenance, repairs, and installation of new equipment, all of which are necessary for successful plant operation. As such, the annual plant turnaround is an integral part of advancing pilot plant studies to provide data for a commercial/demonstration plant design.

Test 64 was conducted after the plant turnaround was completed early in August. Current tests are aimed at acquiring 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 the immediate contract objective.

Test 65 was run during September. An electrical power failure, caused by an equipment malfunction in Commonwealth Edison's main power supply to the plant, forced the termination of this test after only 9 tons of coal had been processed through the gasifier.

A cold-flow model was designed and constructed to simulate the second-stage gasifier and steam-oxygen gasifier. This model was used to investigate how the fluidized bed is established in the steam-oxygen gasifier and the stability of the bed to process upsets. Results of tests made with this model are presented in this report.

Engineering assistance was provided to ERDA-Major Facilities Program Management (MFPM) and Procon, Inc., in their design of a commercial/demonstration plant based on the HYGAS® Process. This report contains the computer-generated material balance data presented to them.

A detailed comparison of the actual HYGAS reactor results vs. those predicted by a computer model simulation is presented.

INTRODUCTION

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 feedstocks, 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.

The annual plant turnaround and two hydrogasification tests were conducted during this quarter, including Test 64 — the most significant test to date with Peabody No. 10 Mine bituminous coal.

PROGRESS REPORT

Task 7. Pilot Plant Experimental Operation

Test 63

Test 63 was conducted late in June and was terminated on June 28. Details are given in Project 9000 Quarterly Report No. 4, June 1977, for this contract. A post-run inspection of the plant after Test 63 indicated that the coal-handling area was in good shape. The coal mill was clean, and the Sweco screener was intact. However, after the 60-ton, raw-coal storage hopper was emptied, dry caked coal was found near the top of the live bottom section covering 25% of the open area. This deposit was cleaned.

The pretreater reactor was in excellent condition. A few small pieces of agglomerated coal were found on the gas distributor grid, although a check of the gas distributor nozzles showed that they were all clear. Dust that appeared to contain a large amount of ash was also found along the southeast wall of the pretreater. The cyclone diplegs and the hot-char transfer lines to the char cooler were clean. The cooling coil in the pretreater was pressure-tested and found to be in good condition.

A hard clinker was found in the bottom of the char cooler below the solids discharge standpipe. To eliminate clinker formation, changes were made to mix equal amounts of nitrogen and air for fluidizing gas to the char cooler in future tests. The rest of the char cooler was in satisfactory condition. The pretreater quench tower and the venturi scrubber were in good condition, with the usual amount of coal and tar accumulations, which were cleaned.

An inspection showed that the slurry preparation section was in good condition. When the reactor was opened, the slurry dryer area was clean, containing only the usual amount of dust. Line 321 from the slurry dryer area to the first-stage gasifier and the lift-line area were clear; however, the solids downcomer from the spouting bed to the second-stage gasifier was plugged downward from the expansion joint. This plug was believed to have formed during shutdown, because there had been no indication of plugging during Test 63. The plug, about 29 feet long, was cleared by rodding and blasting with nitrogen. The second-stage gasifier was clean except for the two large pieces of refractory found lying on the grid area. These were approximately 12 x 8 x 2 inches and 10 x 6 x 3 inches. This spalled refractory fell from the freeboard area of the gasifier and could have caused poor gas distribution in the second-stage gasifier. When it fell cannot be deduced from the operating data. A clinker had formed above the gas sparger in the steam-oxygen gasifier. Two major clinker areas were found along the southeast and southwest walls of the steam-oxygen gasifier. The one clinker extended up to and around the solids transfer line 339 valve. The upper clinker was porous and spanned the entire steam-oxygen gasifier cross-sectional area. (The steam-oxygen sparger was clear.) Exactly when this clinker formation began is not known. It could have formed during upset conditions experienced in the steam-oxygen gasifier during the test.

The reactor high-pressure cyclone and its slurry pot were clean when examined. When the quench system was inspected, a higher than normal solids buildup was found in the vessels. The prequench tower top tray was full of solids. The lines in this section were all taken apart and cleaned.

The purification section was opened up. The absorber was clean, except for the demister at the top of the tower, which was cleaned. An inspection of the regenerator revealed that the packing was in poor condition.

The results of Test 63 are presented in Tables 1 through 3 and Figures 1 through 18. Three periods were selected for the pretreater section, and the results are presented here. The oxygen balance is 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 oil flows. Results of five periods, chosen for reactor study, are also included. Due to the periodic operation of the light-oil recovery unit, water and oil balances could not be obtained.

Plant Turnaround

Following Test 63, the annual plant turnaround was conducted. Modifications and maintenance work on the plant took about 1 month. Details of this work are discussed in the following subsections.

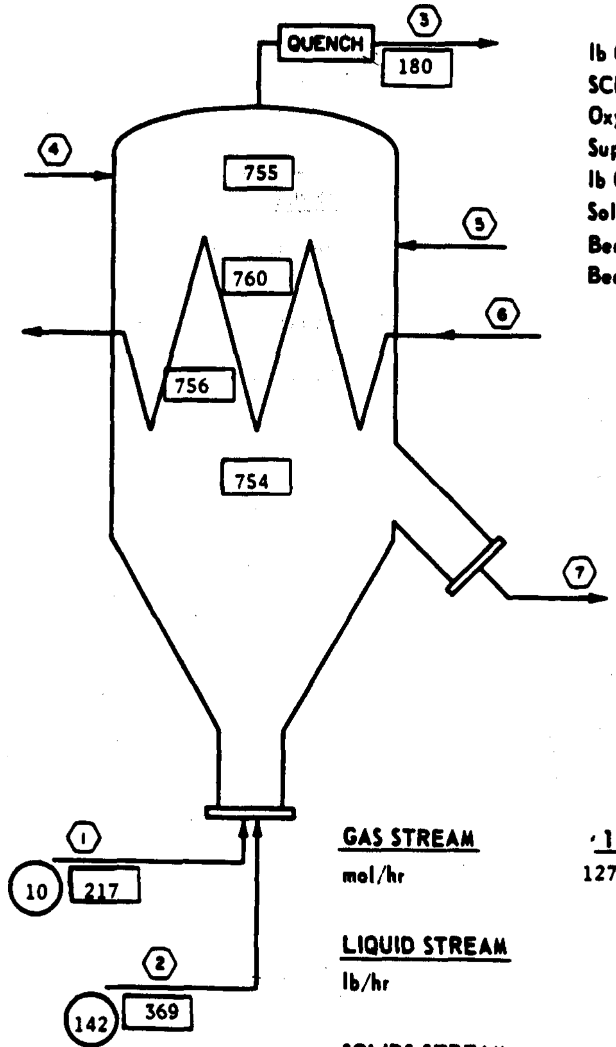
Coal Preparation

The coal mill was in good shape. The ductwork around the wet scrubber was replaced by ductwork with an epoxy-coated internal surface to prevent corrosion. The walls of the wet scrubber were sand-blasted, and the wet scrubber internals were resurfaced. Tests of the coal mill showed that

Table 1. 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

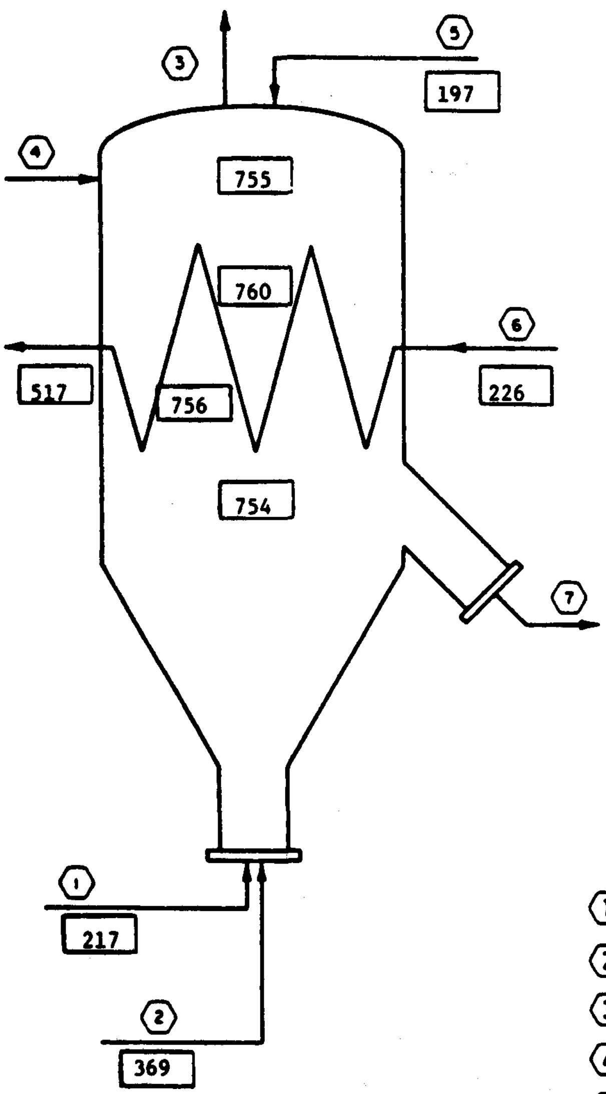
1 AIR TO PRETREATER
2 STEAM TO PRETREATER
3 PRETREATER OFF-GAS AFTER QUENCH
4 RAW COAL TO PRETREATER
5 DIRECT WATER QUENCH
6 COOLING WATER TO COOLING COIL IN PRETREATER
7 PRETREATER CHAR TO CHAR COOLER

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

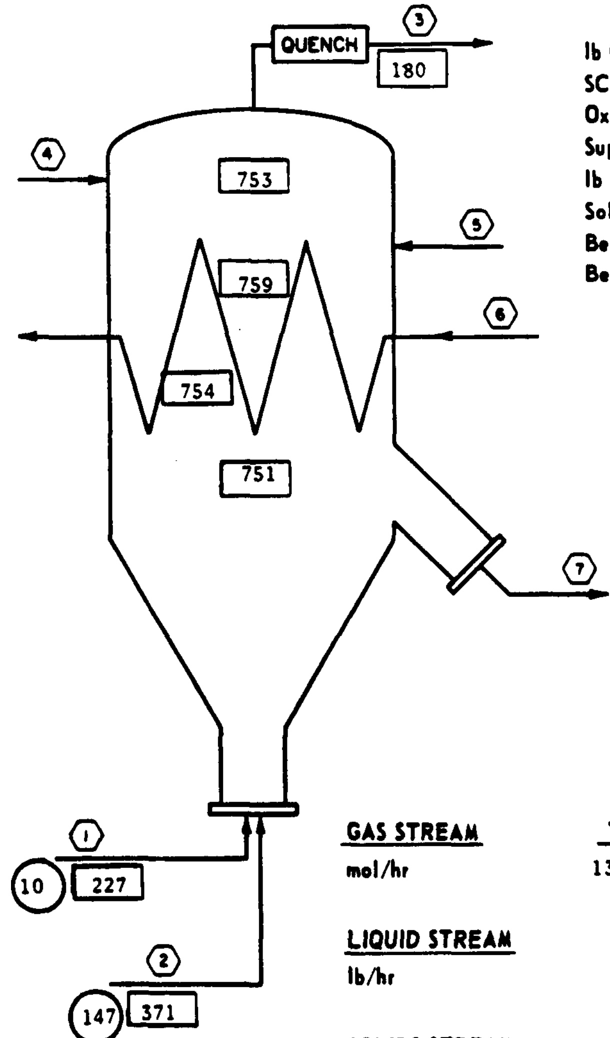
(7) Pretreated Char to Char Cooler

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 2. 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

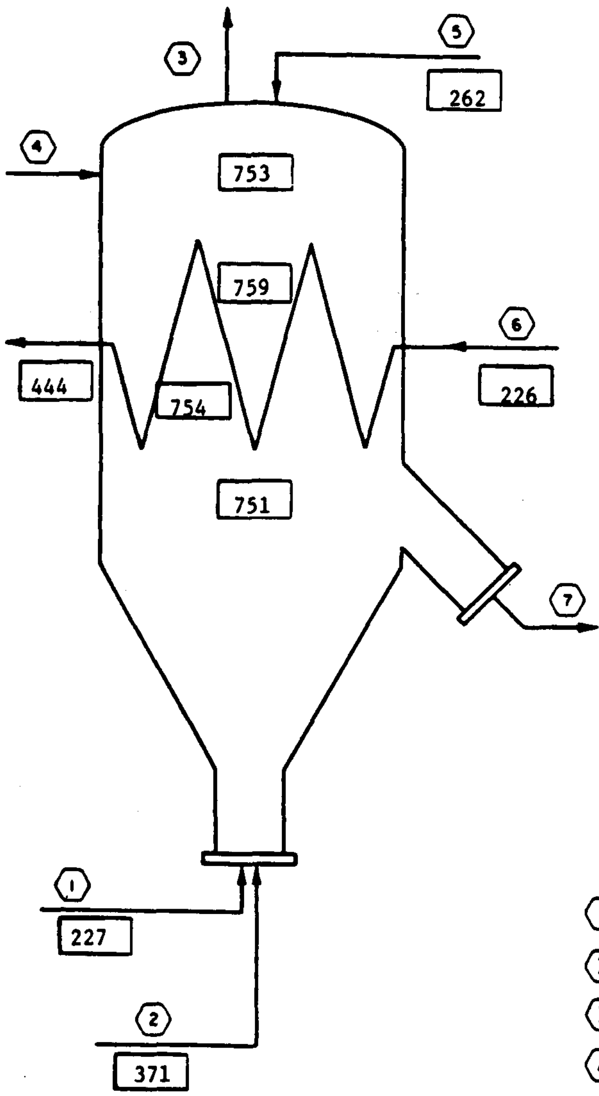
1 AIR TO PRETREATER
2 STEAM TO PRETREATER
3 PRETREATER OFF-GAS AFTER QUENCH
4 RAW COAL TO PRETREATER
5 DIRECT WATER QUENCH
6 COOLING WATER TO COOLING COIL IN PRETREATER
7 PRETREATER CHAR TO CHAR COOLER

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

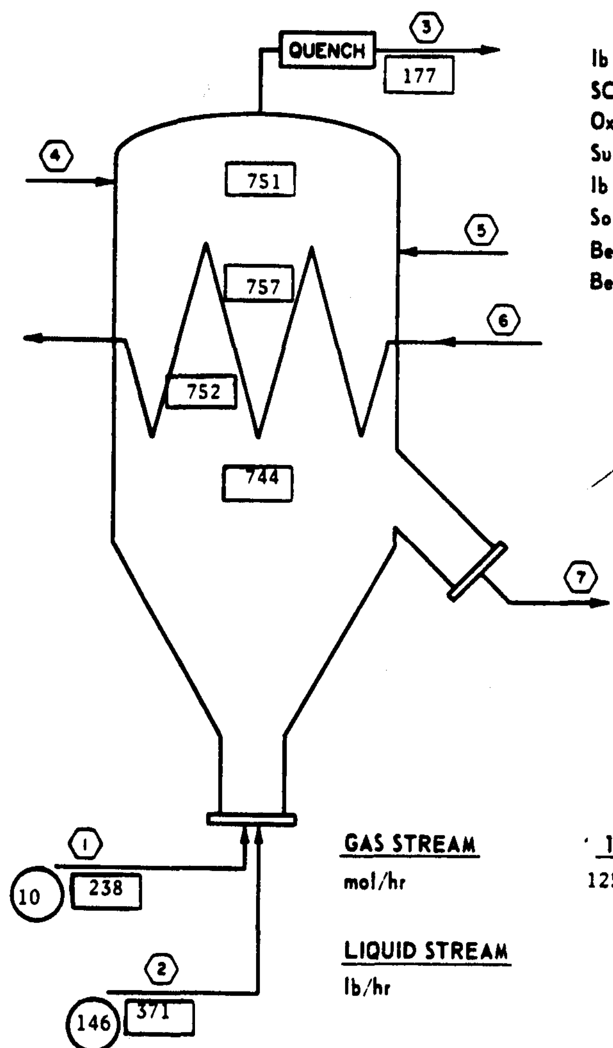
Figure 4. PRETREATER HEAT BALANCE DATA SHEET FOR TEST 63 FOR STEADY PERIOD FROM 6/25/77 (0100 Hours) TO 6/25/77 (0600 Hours)

Table 3. 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

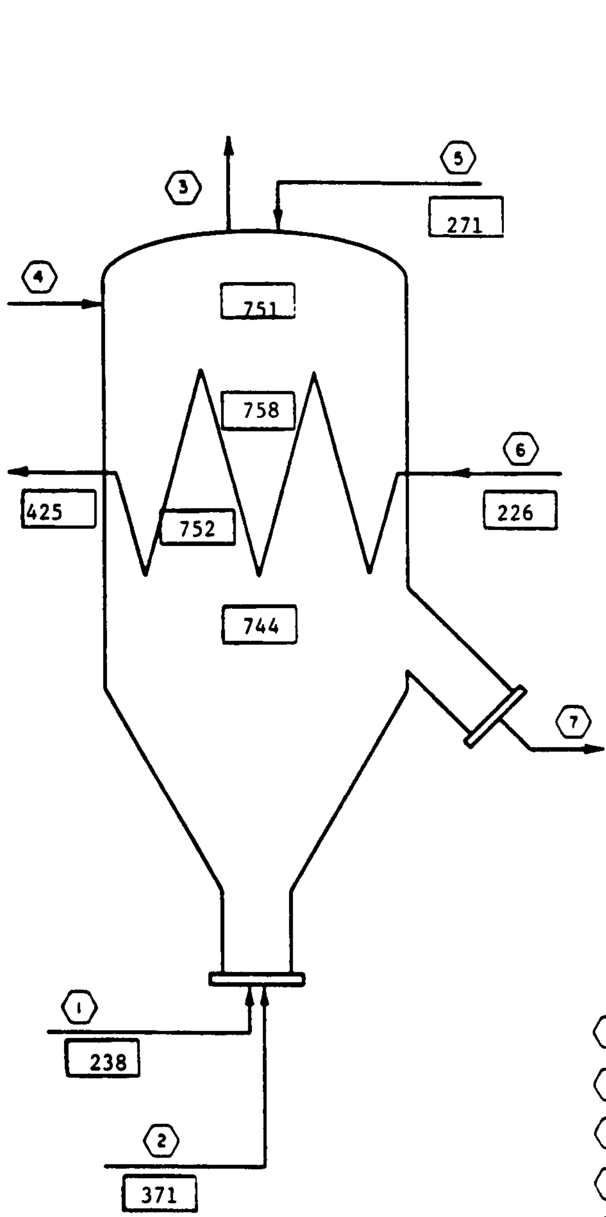
GAS STREAM	1	2
mol/hr	125	98

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

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



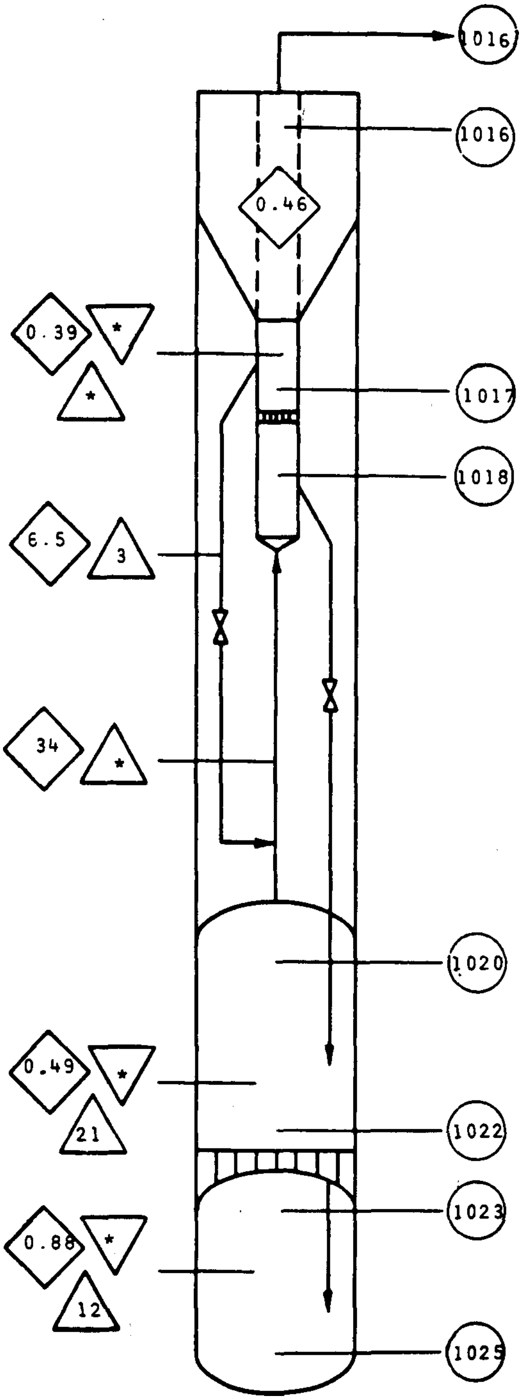
Basis: 1 hour

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

<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) = 0.35
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

<u>By Ash Balance</u>

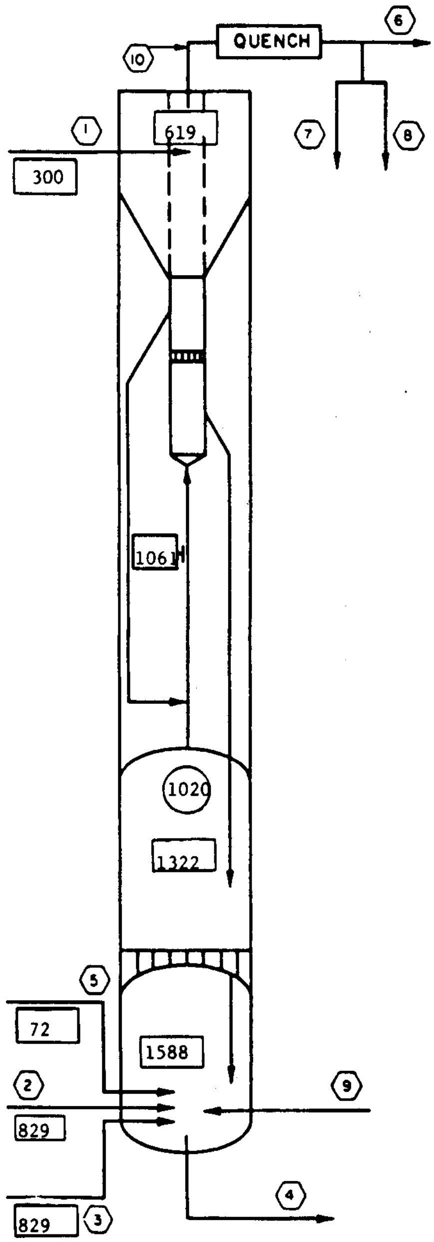
Coal Gasified, % = 52
Carbon Gasified, % = 42

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

<u>Bed Height, ft</u>

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

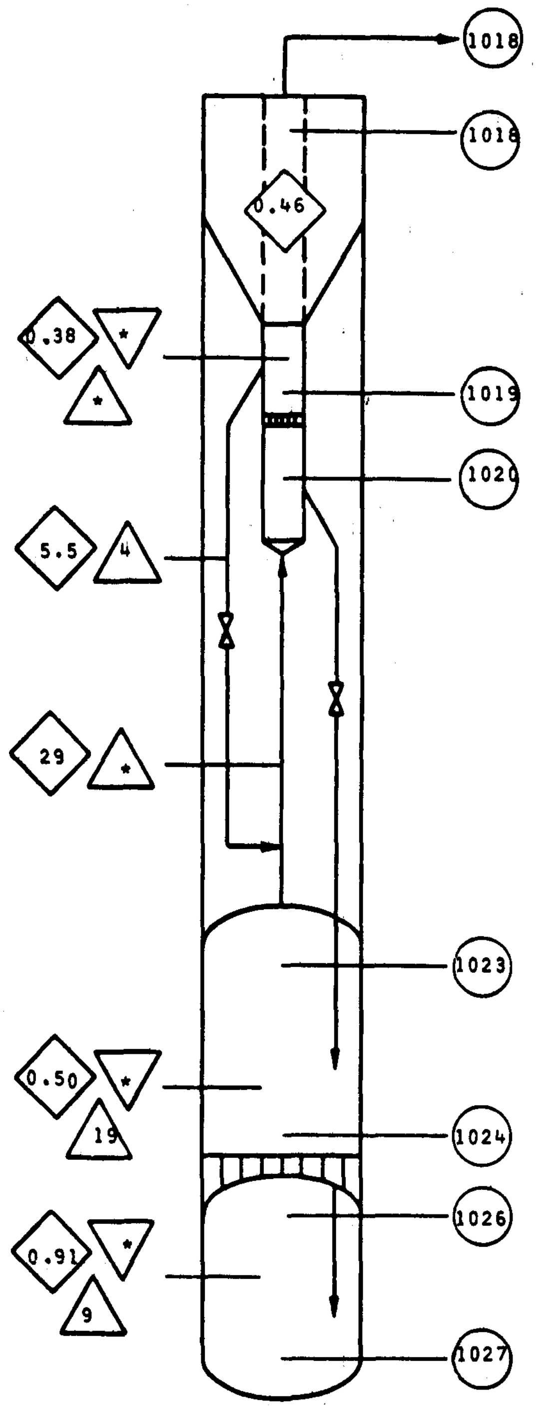
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	67.00	77.10	69.20
H	3.36	2.60	1.28
N	1.29	1.17	0.55
Cl	--	--	--
S	4.52	2.50	1.39
ASH	14.98	12.59	26.70
O	8.85	4.04	0.88
lb/hr (dry)	3889		
MOISTURE, wt %	2.0		
SLURRY CONCN, wt %	29		

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

GAS FEED STREAMS			
2	3		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

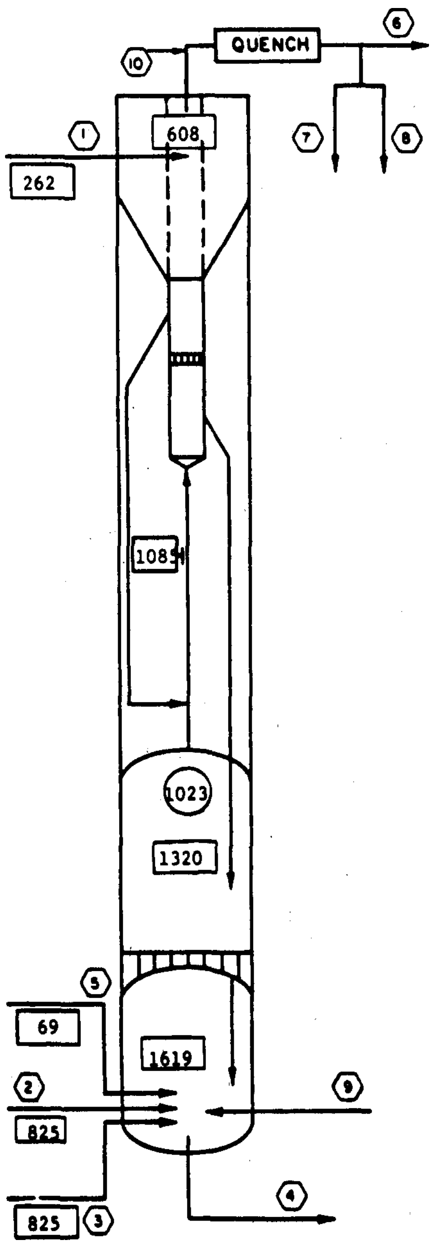
Coal Gasified, % = 64
Carbon Gasified, % = 57

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

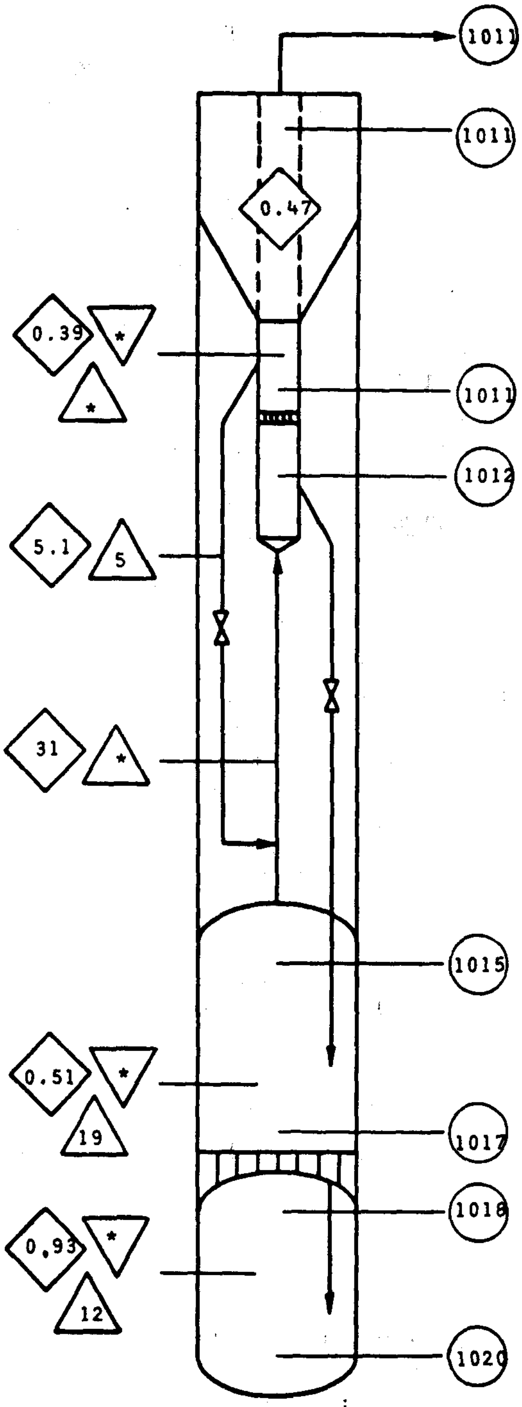
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	67.70	76.20	64.60
H	3.51	3.37	1.16
N	1.37	1.15	0.42
Cl	--	--	--
S	4.26	2.98	1.04
ASH	14.49	13.23	32.08
O	8.67	3.07	0.70
lb/hr (dry)	3664		
MOISTURE, wt %	2.0		
SLURRY CONCN, wt %	26		

	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

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) =	0.41
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

By Ash Balance

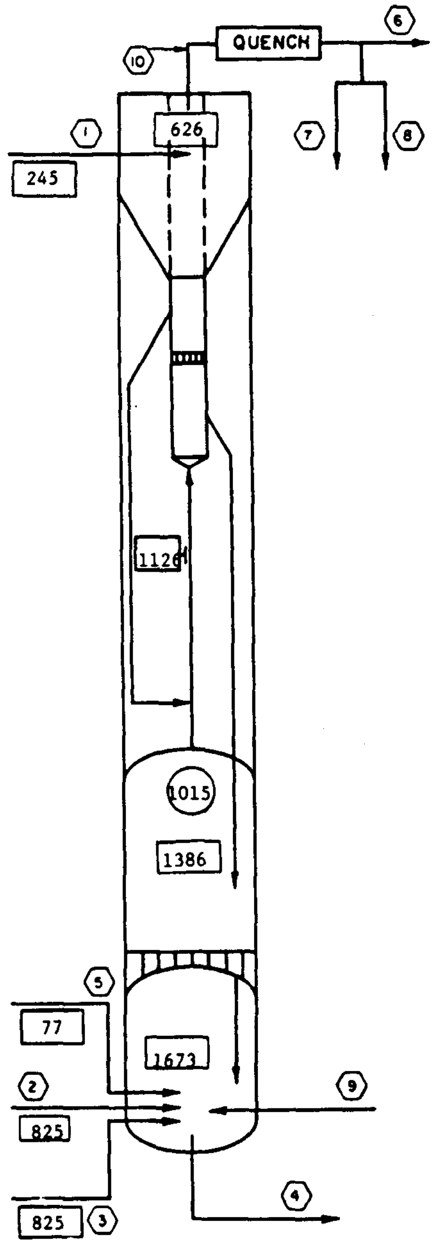
Coal Gasified, % =	64
Carbon Gasified, % =	57

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

Bed Height, ft

Slurry Dryer = *
HTR = 15
SOG = 24

Figure 11. HYGAS REACTOR ENGINEERING DATA FOR TEST 63 FOR 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

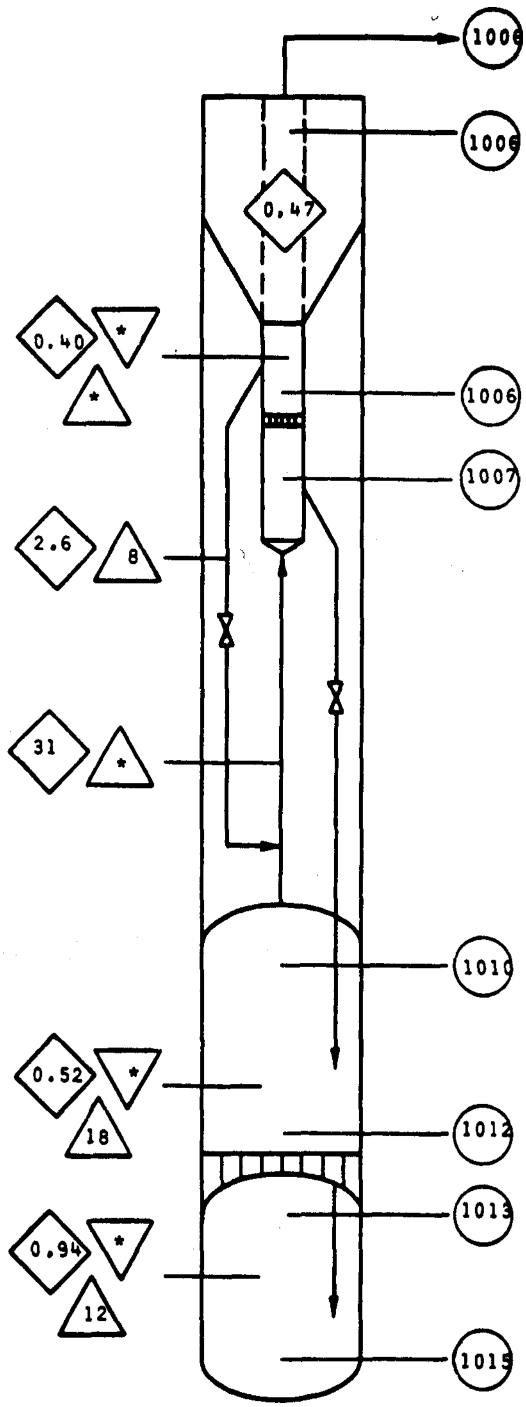
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	66.87	76.50	62.66
H	3.37	2.59	1.06
N	1.45	1.18	0.38
Cl	--	--	--
S	4.29	2.79	1.02
ASH	15.56	13.67	34.13
O	8.46	3.27	0.75
lb/hr (dry)	3999		
MOISTURE, wt %	1.7		
SLURRY CONCN, wt %	30		

	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

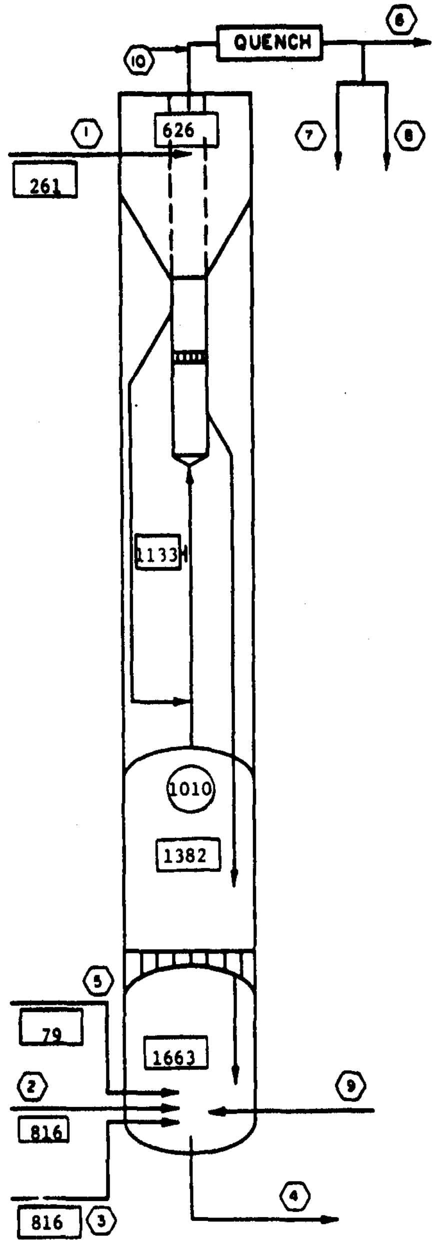
Coal Gasified, % = 66
Carbon Gasified, % = 62

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)



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

MATERIAL BALANCE, %

C
H
O
ASH

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

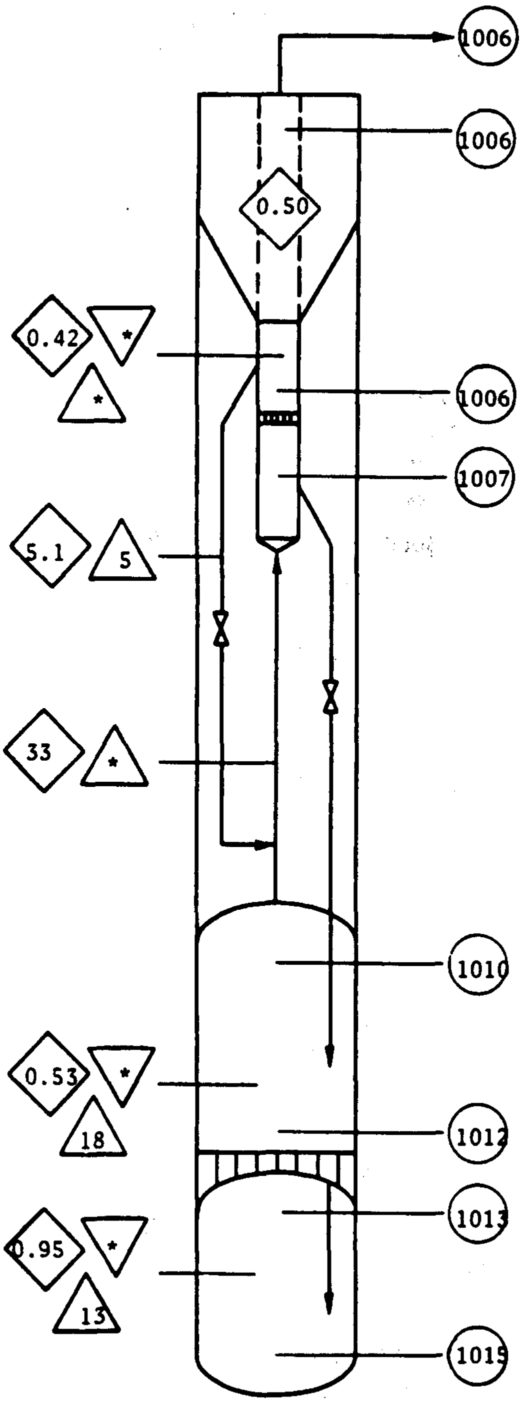
SOLIDS STREAMS	1	10	4
wt % (dry)			
C	68.34	75.29	59.32
H	3.49	2.65	1.01
N	1.37	1.15	0.35
Cl	--	--	--
S	4.53	2.82	1.06
ASH	14.71	14.48	33.92
O	7.56	3.61	4.34
lb/hr (dry)	3810		
MOISTURE, wt %	2.1		
SLURRY CONCN, wt %	27		

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

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) =	0.49
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

By Ash Balance

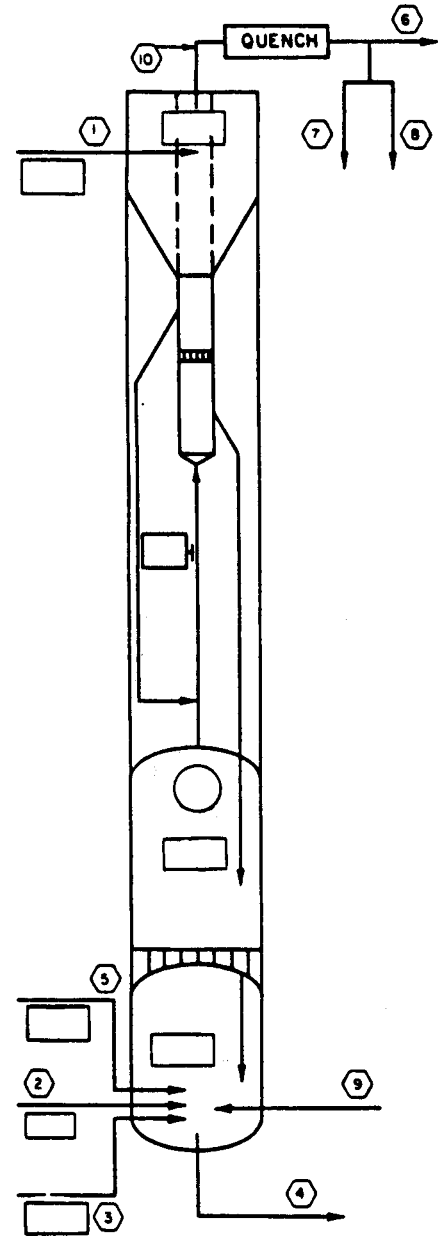
Coal Gasified, % =	75
Carbon Gasified, % =	71

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

Bed Height, ft

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)



MATERIAL BALANCE, %

C

H

O

ASH

STREAM No.

PRESSURE, psig

TEMPERATURE, °F

	GAS STREAMS 6	GAS STREAMS 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	SOLIDS STREAMS 10	SOLIDS STREAMS 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	LIQUID STREAMS 8	LIQUID STREAMS 1
	lb/hr		
H_2O	7300	--	--
LIGHT OIL	--	--	10,688

GAS FEED STREAMS 2	GAS FEED STREAMS 3	GAS FEED STREAMS 5	GAS FEED STREAMS 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

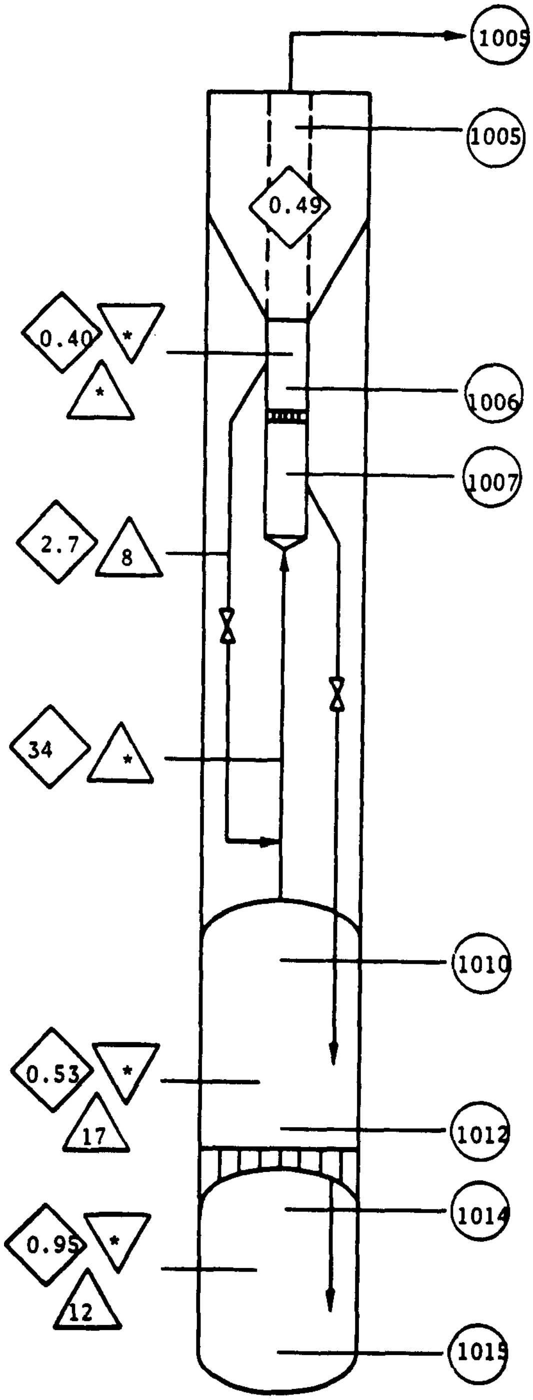
7 WATER MADE

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

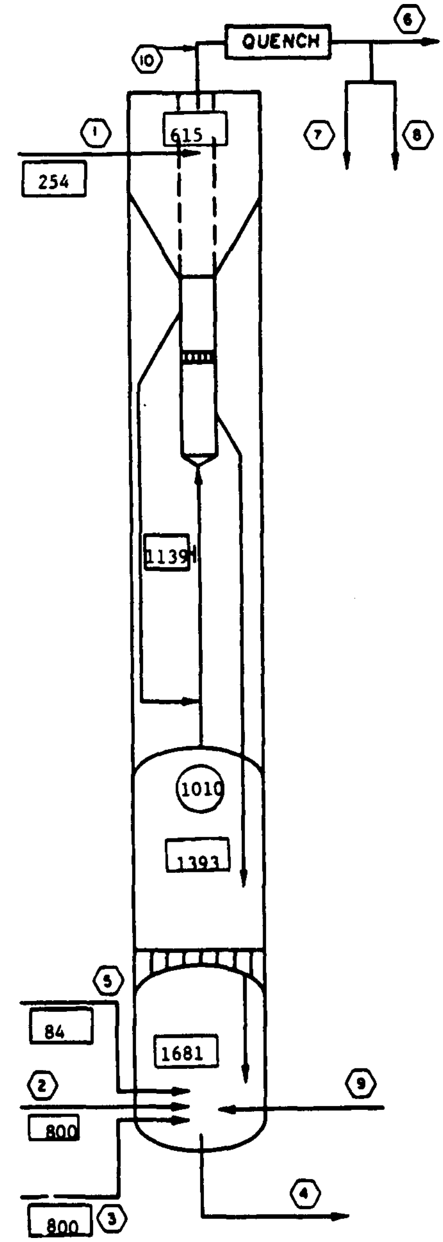
Bed Height, ft

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.
PRESSURE, psig
TEMPERATURE, °F

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)

the secondary fan vibrated excessively and that its shaft was bent. The shaft was replaced and the impeller balanced.

Pretreater

The pretreater section operated very well this past year. The only change made was a modification of the char cooler so that 50% nitrogen could be mixed with air in future tests to form the fluidizing gas in the char cooler to reduce the tendency of clinker formation. Regular maintenance work was performed on the equipment in this section.

Slurry Preparation

This section was inspected and maintenance work was done on it. All piping and vessels were in good condition. A 2-foot-long spool-piece section on the low-pressure coal slurry line was installed to house a test unit from Argonne National Laboratory to test the acoustic flow-metering of the low-pressure slurry.

Reactor

The reactor was inspected and cleaned. The plug and the valve body of the spent-char discharge valve (340) were badly eroded, which could have accounted for past erratic level control in the steam-oxygen gasifier. A replacement valve was installed. The reactor second-stage gasifier refractory above the manway 3 area was patched because pieces of refractory had fallen off the wall during Test 63. Line 339 seal on the second-stage refractory grid was in poor condition and was repacked.

The reactor high-pressure cyclone was sent to Argonne National Laboratory for inspection. The cyclone was found to be in satisfactory condition; however, a wear sleeve was installed on the solids exit line from the cyclone to prevent excessive wear in the future. A new conductivity level detector was installed on the cyclone slurry pot to replace the old float-type level controller, which had numerous operational problems. Also, a new liquid eductor assembly was added to the cyclone slurry pot to help mix the slurry for better handling.

Gray Serv technicians completed machine work and inspection of all 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. Regular maintenance work was performed in the reactor area. The reactor was reassembled and pressure tested.

Quench Section

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. Vessels, lines, and all quench section heat exchangers were cleaned. To eliminate the vertical expansion loop, the equalizing line between the quench tower and the quench separator was modified by installing a horizontal expansion loop. A gas-flow orifice meter was added to the product-gas line from the prequench tower to 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.

Purification Section

The purification section was completely cleaned. The regenerator tower packing was in poor condition after Test 63, so the entire charge was replaced. The absorber vessel was emptied, and the packing was cleaned and replaced. The heat exchanger in the reboiler was cleaned and reinstalled.

Methanation

No changes were made in the IGT fixed-bed catalyst methanation unit. The Chem Systems liquid-phase methanation unit was modified, as described under Task 9.

Effluent Cleanup

The light-oil and solids recovery sections were cleaned. The surge pot was readied for Test 64. Routine maintenance was performed on the Edens and the Alar filters. Installation of the new high-capacity incinerator was begun. Details are reported under Task 9.

Utilities

The utility section was shut down for the turnaround period. Annual inspections of the boilers were completed, and they were found to be satisfactory. Minor refractory patchwork was required around the burner block in the low-pressure boiler. A new, spare, high-pressure boiler feedwater pump and a spare, high-pressure process water pump were installed. A leak was discovered on the cation vessel in the demineralizer for the boiler feedwater. The resin was removed from the vessel, the leak was fixed, and the rubber lining was repaired. Other routine maintenance work was also performed.

Hydrogen Plant

The hydrogen plant reformer stack refractory was repaired, and the diglycolamine heat exchanger was retubed; however, when the hydrogen plant was started up, a leak was discovered in the diglycolamine heat exchanger. The hydrogen plant was shut down, and the diglycolamine cooler was opened and inspected. Its gasket surfaces and internals were polished, and the unit was reassembled.

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. Coal feed to the pretreater was begun on August 18 at 0350 hours. The pretreater operated satisfactorily during Test 64, except for two periods, one 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° to 770°F.

Pretreated char was introduced to the reactor at 2100 hours on August 18, and 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 solids could not be withdrawn from the steam-oxygen gasification zone and difficulty was experienced in moving them 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 (8-1/2 days) 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° to 1825°F. Reactor operations for Test 64 were exceptionally smooth and troublefree. No periodic solids flow upsets were experienced during this test.

Three heat and material balance periods for Test 64 were completed and are presented in Tables 4 through 6 and Figures 19 through 27.

The product-gas cyclone dust slurry pot operated very satisfactorily, and 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.

After Test 64, a post-run inspection of the plant showed that the pre-treater reactor was in excellent condition, following a long test period of 197 hours, during which 437 tons of raw coal were processed. When a clinker formation was found in the bottom of the char cooler despite the 50:50 mixture of nitrogen and air used during Test 64, a decision was made to use 100% nitrogen gas to fluidize the char-cooler solids in future tests. The coal mill, coal preparation section, and slurry preparation section were in good condition after Test 64.

Inspection of the reactor indicated that the slurry dryer bed was clean except for 2 feet of wet solids left behind after the sudden shutdown of Test 64. Lines 321 and 322 were clean, but wet coal was found in the lift line. This was readily cleared by blowing the line out with air.

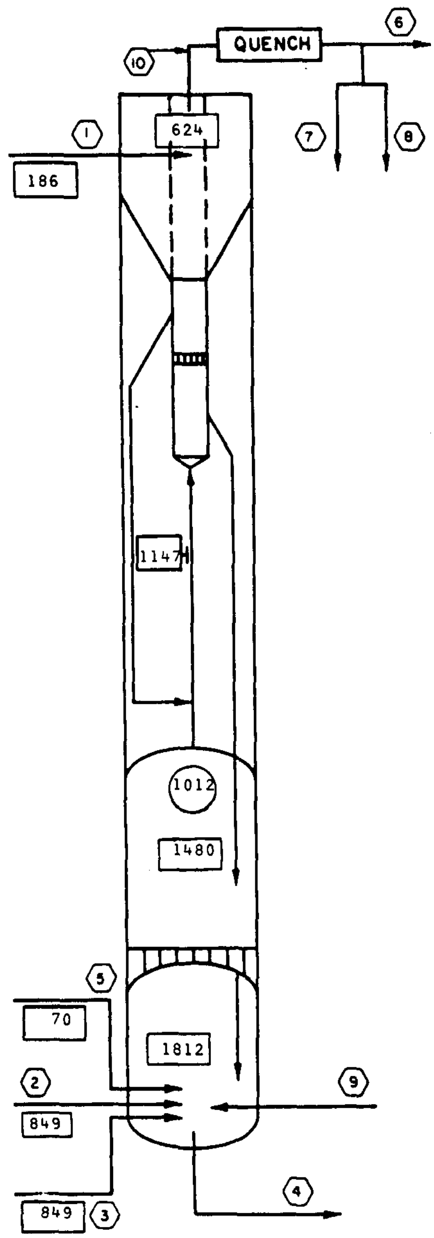
The second-stage reactor and line 339 were also clean. A clinker was found in the steam-oxygen gasifier. Its formation was probably triggered when trouble occurred while discharging solids through the spent-char slurry discharge chokes. This problem, in turn, upset ash withdrawal from the steam-oxygen gasifier. The clinker's consistency changed from a hard mass to soft, red sinter. Samples of these clinkers were sent to the laboratory for analysis. The steam-oxygen sparger ring was undamaged: All of the clinkers had formed

Table 4. MATERIAL BALANCE SUMMARY FOR THE HYGAS GASIFIER FOR TEST 64 FROM 8/24/77 (1500 Hours) TO 8/26/77 (0700 Hours)

Basis = 1 hour. All units in pounds unless noted otherwise.

INPUT		C	H	O	N	S	Cl	ASH	OTHER	TOTAL
Coal Feed	Wt % (Dry)	69.49	3.46	8.80	1.41	4.22	--	12.62		100
	Coal (Dry)	2978	148	377	61	181	--	541		4286
	Moisture		12	97						109
Sparger	Oxygen			960						960
	Steam		695	5564						6259
Burner	Oxygen			0						0
	Steam		0	0						0
	Hydrogen		0							0
Stripping Ring	Steam		166	1331						1497
Nitrogen From Purges					525					525
Pump Seal Flush			74	593						667
Cooling Water Spray			0	0						0
Water to Cyclone Pot			354	2835						3189
Light Oil In		7896	755							8651
TOTAL INPUT		10,874	2204	11,757	586	181	--	541		26,144
OUTPUT										
Reactor Overhead	Wt % (Dry)	74.85	2.59	4.50	1.19	2.83	--	14.04		100
	Dust (Dry)	434	15	26	7	16	--	81		579
Spent Char	Wt % (Dry)	59.20	0.87	0.38	0.28	0.84	--	38.43		100
	Char (Dry)	660	10	4	3	9	--	429		1115
Product Gas After Quench	Total (Dry)	1559	314	2354	506	92				4825
	Components H_2		134							134
	CO_2	738		1967						2705
	C_2H_5	29	7							36
	H_2S		6			92				98
	N_2				506					506
	CH_4	502	167							669
	CO	290		387						677
Water Out + Dissolved Materials		33	1164	9259	25	37				10,518
Toluene Storage Tank Vent Gases		217	15	465	16	5				718
Stripper Vent Gas		83	11	102	39	1				236
Light Oil Out		7888	755							8643
TOTAL OUTPUT		10,874	2284	12,210	596	160	--	510		26,634
Net (Output - Input)		0	80	453	10	-21	--	-31		490
% Balance (Output/Input)		100	104	104	102	88	--	94		102

STREAM No.
PRESSURE, psig
TEMPERATURE, °F



MATERIAL BALANCE, %

C	100
H	104
O	104
ASH	94

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	30.88	28.46
CO_2	28.40	32.33
C_2H_6	0.55	0.58
Ar	--	--
N_2	8.35	8.25
H_2S	1.32	1.25
CH_4	19.33	18.79
CO	11.17	10.34
mol/hr (dry)	216	243

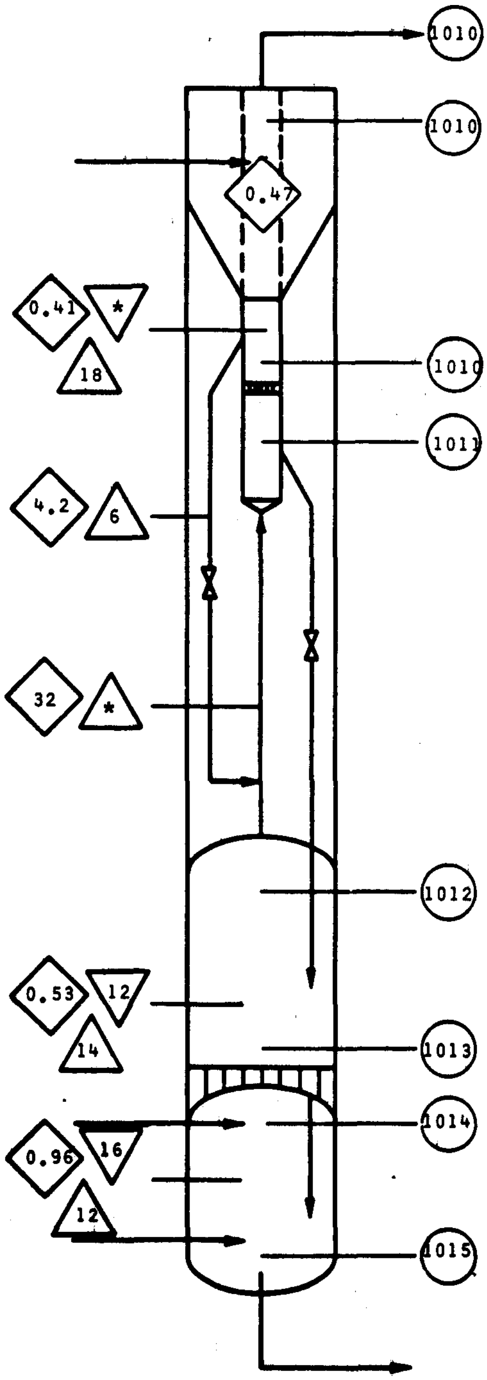
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	69.49	74.85	59.20
H	3.46	2.59	0.87
N	1.41	1.19	0.28
Cl	--	--	--
S	4.22	2.83	0.84
ASH	12.62	14.04	38.43
O	8.80	4.50	0.38
lb/hr (dry)	4286	579	1115
MOISTURE, wt %	2.5		
SLURRY CONC'N, wt %	34		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	6662	--	--
LIGHT OIL	--	8643	8651

GAS FEED STREAMS			
2	3	5	9
mol/hr			
348	83	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 19. HYGAS REACTOR DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/24/77 (1500 Hours) TO 8/26/77 (0700 Hours)



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

REACTOR PRODUCT GAS - dry, nitrogen- and acid-gas-free basis
COAL FED - dry basis
CARBON (net) = total carbon in char feed - carbon in overhead solids

lb OXYGEN/ lb CARBON (net) = 0.38
lb STEAM/ lb CARBON (net) = 3.05
lb OXYGEN / lb COAL FED = 0.22
lb STEAM / lb COAL FED = 1.81
lb COAL FED / 1000 SCF REACTOR PRODUCT GAS = 80

BY ASH BALANCE

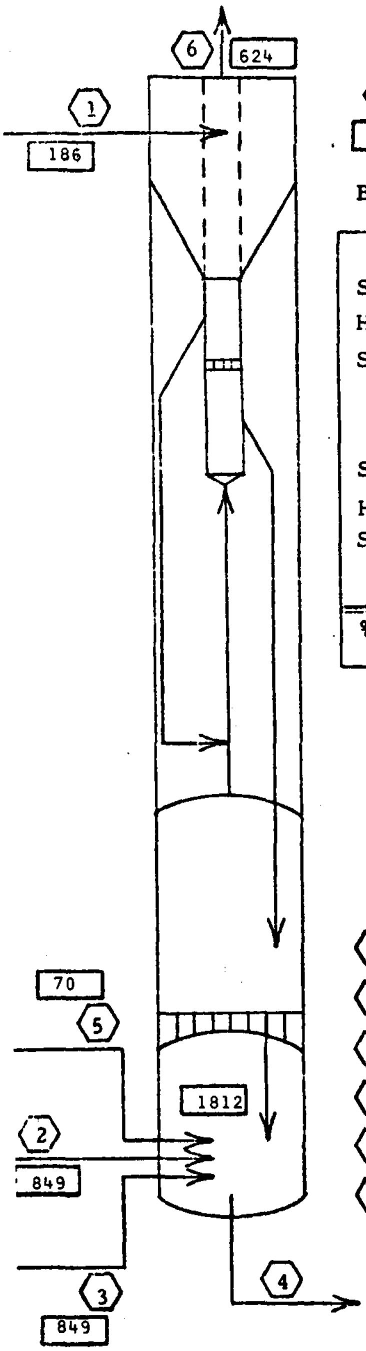
MAF† COAL GASIFIED, % = 77
CARBON GASIFIED, % = 72
METHANE YIELD SCF/lb COAL FED = 4.0
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 6.3

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 10
SOG = 14

†MOISTURE ASH FREE.

Figure 20. HYGAS REACTOR ENGINEERING DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/24/77 (1500 Hours) TO 8/26/77 (0700 Hours)



Stream No.

Temperature, $^{\circ}F$

Basis: 1 hour; Datum condition: $77^{\circ}F$, 1 atm, material in standard state.

INPUT	Btu
Sensible Heat (Streams 1 & 5)	600,220
Heat of Combustion* (Stream 1)	209,611,825
Steam Enthalpy (Streams 2 & 3)	10,659,846
Total	220,871,891
OUTPUT	
Sensible Heat (Streams 4 & 6)	5,314,212
Heat of Combustion* (Streams 4 & 6)	207,324,700
Steam Enthalpy + Light Oil Latent Heat (Stream 6)	9,203,896
Total	221,842,808
% Balance	100

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 Reactor Overhead

* High heating value.

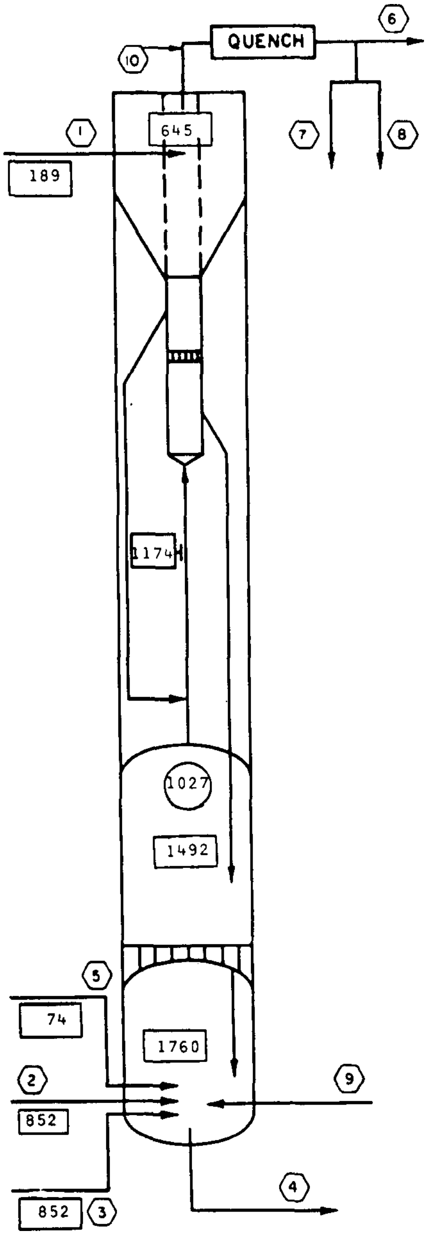
Figure 21. HYGAS REACTOR HEAT BALANCE DATA SHEET FOR TEST 64 FOR STEADY PERIOD FROM 8/24/77 (1500 Hours) TO 8/26/77 (0700 Hours)

Table 5. MATERIAL BALANCE SUMMARY FOR THE HYGAS GASIFIER FOR TEST 64 FROM 8/22/77 (1500 Hours) TO 8/23/77 (0700 Hours)

Basis = 1 hour. All units in pounds unless noted otherwise.

INPUT		C	H	O	N	S	Cl	ASH	OTHER	TOTAL
Coal Feed	Wt % (Dry)	69.00	3.58	7.72	1.44	4.77	--	13.49		100
	Coal (Dry)	2942	153	329	61	203	--	575		4263
	Moisture		11	85						96
Sparger	Oxygen			1044						1044
	Steam		698	5584						6282
Burner	Oxygen			0						0
	Steam		0	0						0
	Hydrogen		0							0
Stripping Ring	Steam		167	1333						1500
Nitrogen From Purges					452					452
Pump Seal Flush			74	593						667
Cooling Water Spray			0	0						0
Water to Cyclone Pot			361	2886						3247
Light Oil In		7970	762							8732
TOTAL INPUT		10,912	2226	11,854	513	203	--	575		26,283
OUTPUT										
Reactor Overhead	Wt % (Dry)	74.78	2.69	4.33	1.20	2.96	--	14.04		100
	Dust (Dry)	673	24	39	11	27	--	126		900
Spent Char	Wt % (Dry)	45.65	0.69	0.34	0.19	0.63	--	52.50		100
	Char (Dry)	370	6	3	2	5	--	426		812
Product Gas After Quench	Total (Dry)	1744	343	2625	442	95				5249
	Components H_2		145							145
	CO_2	809		2157						2966
	C_2H_5	32	8							40
	H_2S		6			95				101
	N_2				442					442
	CH_4	552	184							736
	CO	351		468						819
Water Out + Dissolved Materials		29	1113	8840	35	31				10,048
Toluene Storage Tank Vent Gases		218	15	472	16	5				726
Stripper Vent Gas		59	9	97	29					194
Light Oil Out		7800	745							8545
TOTAL OUTPUT		10,893	2255	12,076	535	163	--	552		26,474
Net (Output - Input)		-19	29	222	22	-40	--	-23		191
% Balance (Output/Input)		100	101	102	104	80	--	96		101

STREAM No.
PRESSURE, psig
TEMPERATURE, °F



MATERIAL BALANCE, %

C	100
H	101
O	102
ASH	96

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	30.74	28.66
CO_2	28.69	32.33
C_2H_6	0.56	0.59
Ar	--	--
N_2	6.72	6.65
H_2S	1.27	1.21
CH_4	19.57	19.00
CO	12.45	11.56
mol/hr (dry)	235	262

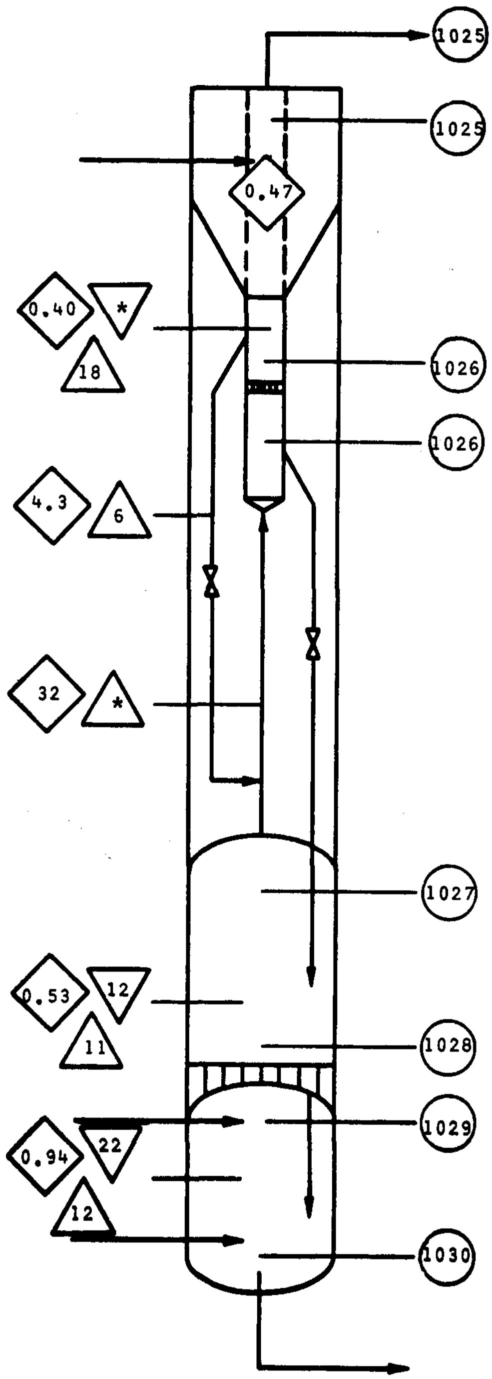
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	69.00	74.78	45.65
H	3.58	2.69	0.69
N	1.44	1.20	0.19
Cl	--	--	--
S	4.77	2.96	0.63
ASH	13.49	14.04	52.50
O	7.72	4.33	0.34
lb/hr (dry)	4263	900	812
MOISTURE, wt %	2.2		
SLURRY CONCN, wt %	33		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	6134	--	--
LIGHT OIL	--	8545	8732

GAS FEED STREAMS			
2	3	5	9
mol/hr			
349	83	33	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 22. HYGAS REACTOR DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/22/77 (1500 Hours) TO 8/23/77 (0700 Hours)



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

REACTOR PRODUCT GAS - dry, nitrogen- and acid-gas-free basis
COAL FED - dry basis
CARBON (net) = total carbon in char feed - carbon in overhead solids

lb OXYGEN/ lb CARBON (net) = 0.46
lb STEAM/ lb CARBON (net) = 3.4
lb OXYGEN/ lb COAL FED = 0.24
lb STEAM/ lb COAL FED = 1.8
lb COAL FED / 1000 SCF REACTOR PRODUCT GAS = 72

BY ASH BALANCE

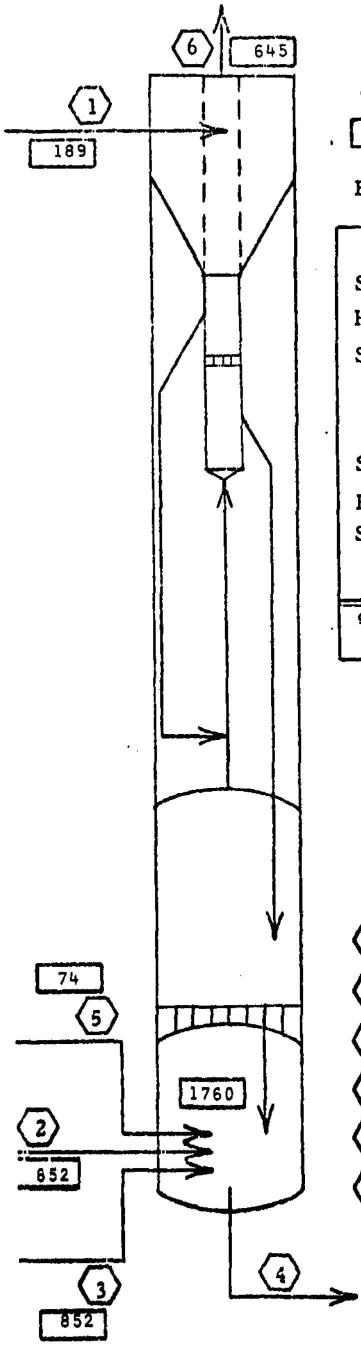
MAF† COAL GASIFIED, % = 86
CARBON GASIFIED, % = 84
METHANE YIELD SCF/lb COAL FED = 4.4
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 7.0

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 10
SOG = 18

†MOISTURE ASH FREE.

Figure 23. HYGAS REACTOR ENGINEERING DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/22/77 (1500 Hours) TO 8/23/77 (0700 Hours)



Stream No.

Temperature, °F

Basis: 1 hour; Datum condition: 77°F, 1 atm, material in standard state.

INPUT	Btu
Sensible Heat (Streams 1 & 5)	617,906
Heat of Combustion* (Stream 1)	209,623,003
Steam Enthalpy (Streams 2 & 3)	10,676,904
Total	220,917,813
OUTPUT	
Sensible Heat (Streams 4 & 6)	5,012,498
Heat of Combustion* (Streams 4 & 6)	208,103,689
Steam Enthalpy + Light Oil Latent Heat (Stream 6)	8,567,664
Total	221,683,851
% Balance 100	

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 Reactor Overhead

* High heating value.

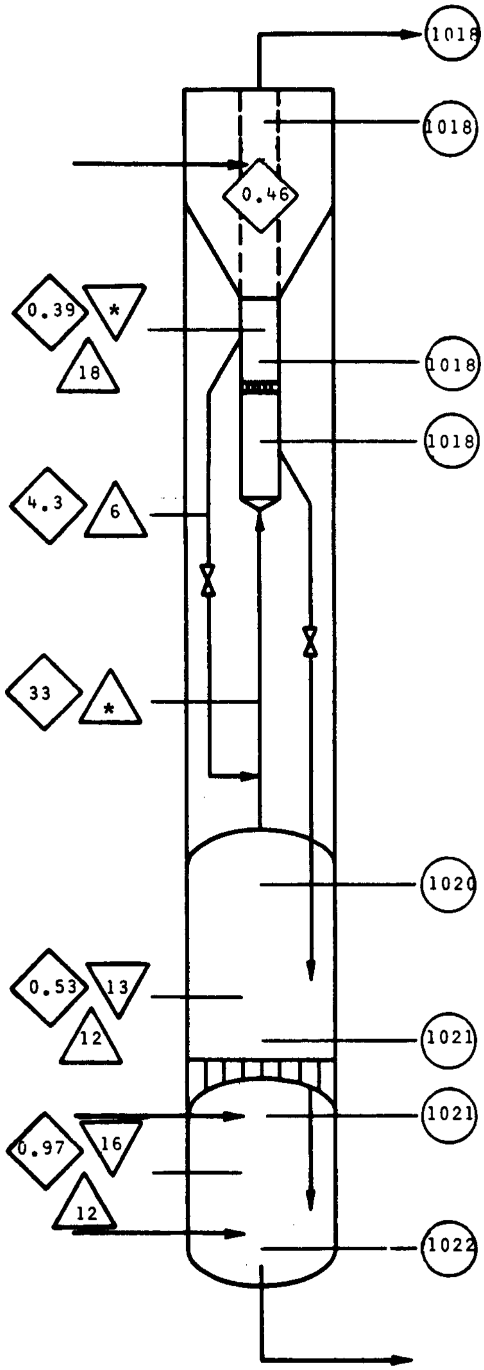
Figure 24. HYGAS REACTOR HEAT BALANCE DATA SHEET FOR TEST 64 FOR STEADY PERIOD FROM 8/22/77 (1500 Hours) TO 8/23/77 (0700 Hours)

Table 6. MATERIAL BALANCE SUMMARY FOR THE HYGAS GASIFIER FOR TEST 64 FROM 8/23/77 (1600 Hours) TO 8/24/77 (0500 Hours)

Basis = 1 hour. All units in pounds unless noted otherwise.

INPUT		C	H	O	N	S	Cl	ASH	OTHER	TOTAL
Coal Feed*	Wt % (Dry)	69.49	3.46	8.80	1.41	4.22	--	12.62		100
	Coal (Dry)	3154	157	399	64	192	--	573		4539
	Moisture		13	103						116
Sparger	Oxygen			967						967
	Steam		704	5631						6335
Burner	Oxygen			0						0
	Steam		0	0						0
	Hydrogen		0							0
Stripping Ring	Steam		168	1340						1508
Nitrogen From Purges					474					474
Pump Seal Flush			74	593						667
Cooling Water Spray			0	0						0
Water to Cyclone Pot			335	2683						3018
Light Oil In		7569	713							8282
TOTAL INPUT		10,723	2164	11,716	538	192	--	573		25,906
OUTPUT										
Reactor Overhead	Wt % (Dry)	76.63	2.75	4.26	1.19	2.88	--	12.29		100
	Dust (Dry)	638	23	36	10	24	--	102		833
Spent Char	Wt % (Dry)	58.44	0.83	0.57	0.26	0.93	--	38.97		100
	Char (Dry)	640	9	6	3	10	--	427		1095
Product Gas After Quench	Total (Dry)	1538	310	2310	496	90				4744
	Components H_2		128							128
	CO_2	732		1952						2684
	C_2H_5	28	7							35
	H_2S		6			90				96
	N_2				496					496
	CH_4	509	170							678
	CO	269		358						627
Water Out + Dissolved Materials		30	1126	8946	36	30				10,168
Toluene Storage Tank Vent Gases		218	15	464	16	0				713
Stripper Vent Gas		64	9	88	29	0				190
Light Oil Out		7314	689							8003
TOTAL OUTPUT		10,442	2181	11,850	590	154	--	529		25,746
Net (Output - Input)		-281	17	134	52	-38	--	-44		-160
% Balance (Output/Input)		97	101	101	110	80	--	92		99

*Coal feed analysis taken from a previous time period.



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

REACTOR PRODUCT GAS - dry, nitrogen- and acid-gas-free basis
COAL FED - dry basis
CARBON (net) = total carbon in char feed - carbon in overhead solids

lb OXYGEN/ lb CARBON (net) = 0.38
lb STEAM/ lb CARBON (net) = 3.12
lb OXYGEN / lb COAL FED = 0.21
lb STEAM / lb COAL FED = 1.73
lb COAL FED / 1000 SCF REACTOR PRODUCT GAS = 87

BY ASH BALANCE

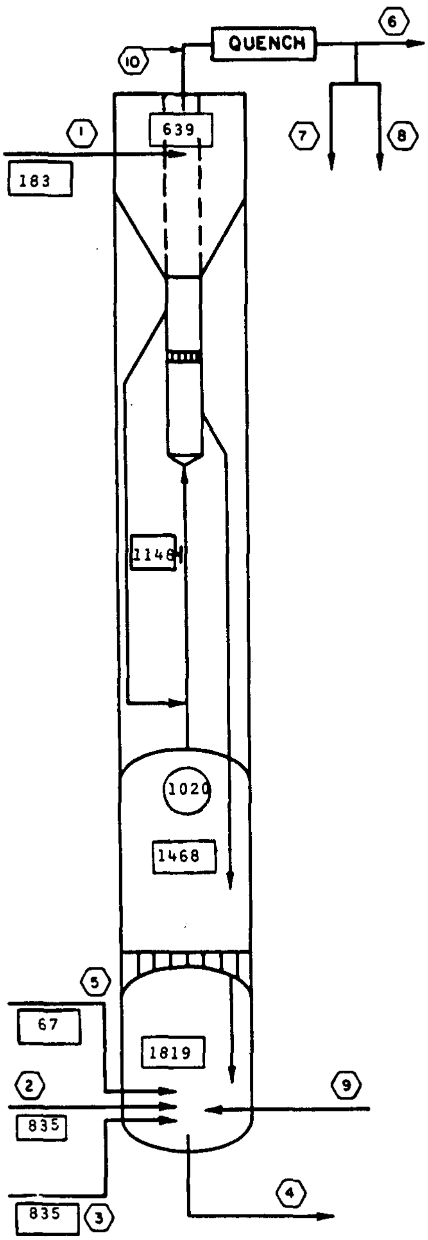
MAF† COAL GASIFIED, % = 77
CARBON GASIFIED, % = 73
METHANE YIELD SCF/lb COAL FED = 3.9
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 5.9

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 11
SOG = 14

† MOISTURE ASH FREE.

Figure 25. HYGAS REACTOR ENGINEERING DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/23/77 (1600 Hours) TO 8/24/77 (0500 Hours)



STREAM No.
PRESSURE, psig
TEMPERATURE, °F

MATERIAL BALANCE, %	
C	97
H	101
O	101
ASH	92

	GAS STREAMS	
	6	10
mol % (dry)	30.26	28.08
H_2	28.85	32.78
CO_2	0.56	0.59
C_2H_6	--	--
Ar	8.38	8.11
N_2	1.33	1.18
H_2S	20.04	19.49
CH_4	10.59	9.77
CO	211	237
mol/hr (dry)		

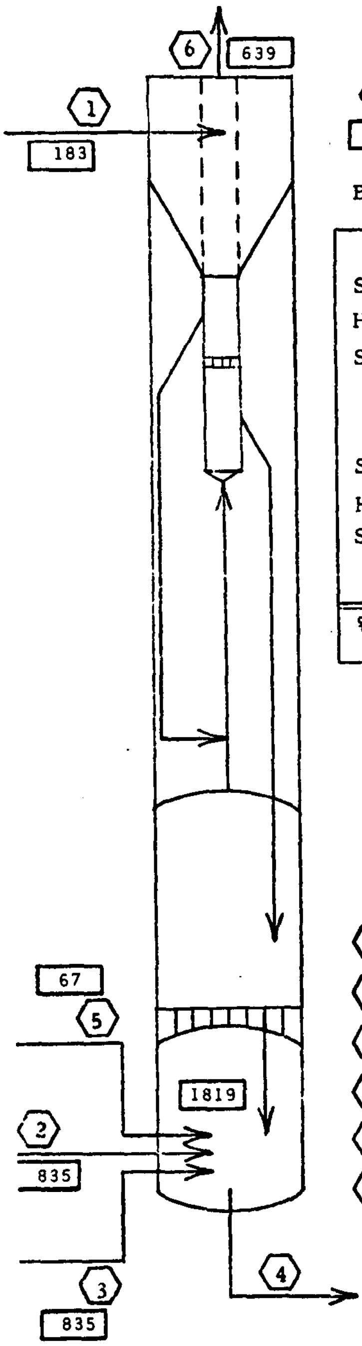
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	69.49	76.63	58.44
H	3.46	2.75	0.83
N	1.41	1.19	0.26
Cl	--	--	--
S	4.22	2.88	0.93
ASH	12.62	12.29	38.97
O	8.80	4.26	0.57
lb/hr (dry)	4539	833	1095
MOISTURE, wt %	1.8		
SLURRY CONCN, wt %	36		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	6433	--	--
LIGHT OIL	--	8003	8282

GAS FEED STREAMS			
2	3	5	9
mol/hr			
352	84	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 26. HYGAS REACTOR DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/23/77 (1600 Hours) TO 8/24/77 (0500 Hours)



⬡ Stream No.

▭ Temperature, °F

Basis: 1 hour; Datum condition: 77°F, 1 atm, material in standard state.

INPUT	Btu
Sensible Heat (Streams 1 & 5)	570,020
Heat of Combustion* (Stream 1)	204,892,240
Steam Enthalpy (Streams 2 & 3)	10,697,852
Total	216,160,112
OUTPUT	
Sensible Heat (Streams 4 & 6)	5,341,964
Heat of Combustion* (Streams 4 & 6)	198,159,841
Steam Enthalpy + Light Oil Latent Heat (Stream 6)	8,819,191
Total	212,320,996
% Balance 98	

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 Reactor Overhead

* High heating value.

Figure 27. HYGAS REACTOR HEAT BALANCE DATA SHEET FOR TEST 64 FOR STEADY PERIOD FROM 8/23/77 (1600 Hours) TO 8/24/77 (0500 Hours)

above the sparger. A minor adjustment in the spacing of the steam-oxygen sparger nozzles was made, principally to prevent direct impingement of the steam-oxygen jets onto the reactor wall, which is believed to be the starting point of clinker formation. These adjustments will give the hottest solids in the bed (which exist in the gas bubbles or jets) a chance to dissipate their heat to the surrounding solids before they contact the reactor wall.

An inspection of valve 340 indicated some wear on the valve body, but it was still in good enough condition to stay in service. The quench section was clean, although a normal solids accumulation was found in the quench separator. All vessels and lines in the quench section were hydraulically cleaned for Test 65.

During Test 64, a high pressure drop was observed in the absorber tower in the purification section. Antifoam agents were used, but were ineffective. The absorber tower was taken apart after the test, and a 5-foot section of the bottom of each packed section was replaced with 2-inch, stainless-steel pall rings, instead of the 1-inch rings normally used for the rest of the packing, to prevent physical holdup of the amine solution in the tower. Routine maintenance and minor mechanical modifications were carried out on the liquid-phase methanation pilot unit.

Test 65

Light-off for Test 65 occurred at 1400 hours on September 17. Coal feed to the pretreater was started at 0430 hours on September 19, and slurry feed to the reactor was begun at 1600 hours on the same day. An electric power failure occurred at 1330 hours on September 20 and forced the termination of Test 65. Solids flow was being established throughout the reactor when a main fuse on the incoming 12.8-kV line from Commonwealth Edison failed. Nine tons of coal were processed through the gasifier, and 30 tons of coal were processed through the pretreater for Test 65.

The power outage was later discovered to have resulted from a mechanical failure of the fuse holder on the 12.8-kVA line. The fuse holder overheated when an accumulation of coal dust and dirt on the equipment caused a high resistance in this element. Commonwealth Edison replaced the entire fuse-holder assembly and restored power to the pilot plant after a 2-hour outage on September 20. The pilot plant emergency electrical generator system operated satisfactorily during the power outage in Test 65, and the pilot plant was brought to a safe condition following the power failure. Commonwealth Edison instituted a regular inspection procedure to avoid future problems of this nature. No heat and material balance data were obtained for Test 65.

Scientific Design

The Office of Program Planning and Assessment, Fossil Energy, U.S. ERDA, contracted for Scientific Design Company, Inc., to evaluate the HYGAS program. An initial meeting between representatives of ERDA-CCU, 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. As of August 15, a resident engineer from Scientific Design Company, Inc., has been onsite monitoring all HYGAS operations.

Task 8. Demonstration Plant Design Support

The objective of the work done under this task is to provide engineering assistance to ERDA-MFPM and Procon, Inc., in their design of a commercial/demonstration plant based upon the HYGAS Steam-Oxygen Process. At this stage, IGT's functions are to —

a. Assist in the selection of basic design parameters, i.e., coal feedstock.

b. Assist in the identification of potential process trade-off studies.

c. Provide basic design data for the coal gasifier based upon the operation of the HYGAS pilot plant and on the use of the IGT kinetic model and mathematical gasifier analysis.

d. Share general knowledge acquired by IGT in its analysis of alternative coal gasification processing schemes that have been developed.

As the program proceeds, additional types of work are anticipated, including —

a. Acquiring new data desired by Procon for its design effort

b. Cold-flow modeling portions of the design gasifier internal mechanical configuration.

To begin work under this task, 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 discussions of the data requirements that Procon must have for the initiation of its design study were held. After this, a series of discussions, both formal and informal, were held with Procon and ERDA-MFPM. Major areas discussed included data requirements from IGT, the factors underlying the IGT computer model for coal gasification, alternative processing configurations, and mechanical design criteria for the gasifier.

Early in the program, IGT supplied Procon and ERDA-MFPM with copies of monthly, quarterly, interim, and final reports prepared under the HYGAS program. These reports give the general background on the process and on the operation of the pilot plant. IGT also supplied a typical run variability analysis, indicating the variation in the major process parameters during steady-state operating periods.

The first coal analyses supplied were typical compositions of the three major coal types, lignite, subbituminous, and bituminous coals, found in the United States. After discussions with Procon and ERDA-MFPM, it was decided that the primary design for the facility should be based on either an Eastern or a Western coal and that a delta design, with the same detail as the primary design, should be prepared for the alternative coal. The Western coal selected for this analysis was a lignite, because of the extent of IGT's operating experience in the pilot plant with this type of coal. The Eastern bituminous coal was selected from the Mid-Continent Basin for the same reason — IGT's extensive operating experience with this type of coal. Actually, two Eastern coals, washed and run-of-mine, were selected so that Procon could make a process trade-off study on the relative merits of washing the coal feed. Analyses of the selected coals are presented in Table 7. These analyses are

Table 7. ANALYSES OF COALS PROPOSED FOR DEMONSTRATION PLANT DESIGN

Sample	Montana Lignite	Illinois No. 6 Washed	Illinois No. 6 Run-of-Mine
Proximate Analysis, wt%			
Moisture	15.50	12.00	12.00
Volatile Matter	35.47	36.32	32.90
Fixed Carbon	38.40	42.15	38.21
Ash	10.63	9.53	16.89
Ultimate Analysis, dry, wt%			
Carbon	61.27	69.47	62.70
Hydrogen	4.20	5.25	4.67
Oxygen	20.10	9.60	7.85
Nitrogen	0.97	1.03	1.18
Sulfur	0.88	3.80	4.25
Chlorine	Not specified	0.02	0.16
Ash	12.58	10.83	19.19

based upon coals that have been used in pilot plant operations, but have slight variations in moisture or sulfur content to make them more representative of the overall basin that the coal represents.

IGT has also prepared preliminary heat and material balances for both lignite and subbituminous coals. The preliminary lignite balance was prepared so that Procon could initiate its efforts and define a base-case, primary system design. The basic heat and material balances for the gasifier, which were supplied in this preliminary design, are presented in Table 8 and Figures 28 and 29.

Optimizing the process design of the lignite gasifier was the next step. Basic costs of coal, steam, oxygen, gasifier construction, and CO_2 removal were estimated from C. F. Braun & Co.'s report on factored cost estimates for Western coal. Alternative gasification design schemes were assessed to determine minimum annualized costs. Other constraints in this effort are maximum plate thicknesses, maximum and minimum gasifier bed L/D ratios, and maximum temperatures in the gasifiers. These designs are being prepared at pressure levels of 800, 1000, and 1200 psig so that Procon can do a process trade-off study on the effects of pressure in the system.

A preliminary heat and material balance was also prepared for subbituminous coal so that Procon could assess the differences in the processing schemes (for a generic "Western" coal) for the two coals. These balances are presented in Table 9 and Figures 30 and 31.

Task 9. Support Studies

Plant Effluent Processing

The light-oil stripper was cleaned after Test 63. The light-oil and solids recovery units both worked satisfactorily for Test 64. No unusual problems were evident during an inspection of the effluent cleanup section following Test 64. Both the light-oil and the solids recovery units were operated for Test 65.

Table 8. CALCULATED GASIFIER MATERIAL AND HEAT BALANCE FOR MONTANA LIGNITE COAL

Stream No.	1			2		
Description	Lignite Feed			Ash Residue		
Temperature, ^{o}F	240			1700		
Components	lb/hr	wt%	mol/hr	lb/hr	wt%	mol/hr
C	612.7	61.27	51.02	61.2	31.97	5.09
H_2	42.0	4.20	20.83	1.6	0.84	0.80
O_2	201.0	20.10	12.56	--	--	--
N_2	9.7	0.97	0.35	1.9	0.99	0.07
S	8.8	0.88	0.27	0.9	0.47	0.03
Ash	125.8	12.58	--	125.8	65.73	--
Total	1000.0	100.00		191.4	100.00	
Moisture	183.4					
Slurry Oil	1222					

Stream No.	3	
Description	Raw Product Gas	
Temperature, ^{o}F	600	
Components	mol/hr	mol %
CO	10.75	9.95
CO_2	19.56	18.11
H_2	19.82	18.35
H_2O	46.92	43.42
CH_4	9.45	8.75
C_2H_6	0.77	0.71
NH_3	0.44	0.41
N_2	0.04	0.04
HCN	0.03	0.03
H_2S	0.24	0.22
COS	0.01	0.01
Total (Oil-Free Gas)	108.03	100.00
	mol/hr	wt%
C_6H_6	0.12	15.0
C_7H_8	0.56	85.0
Total (Product Oil)	0.68	100.0
Total (Oil-Free Gas + Product Oil)	108.71	
Slurry Oil, lb/hr	1222	

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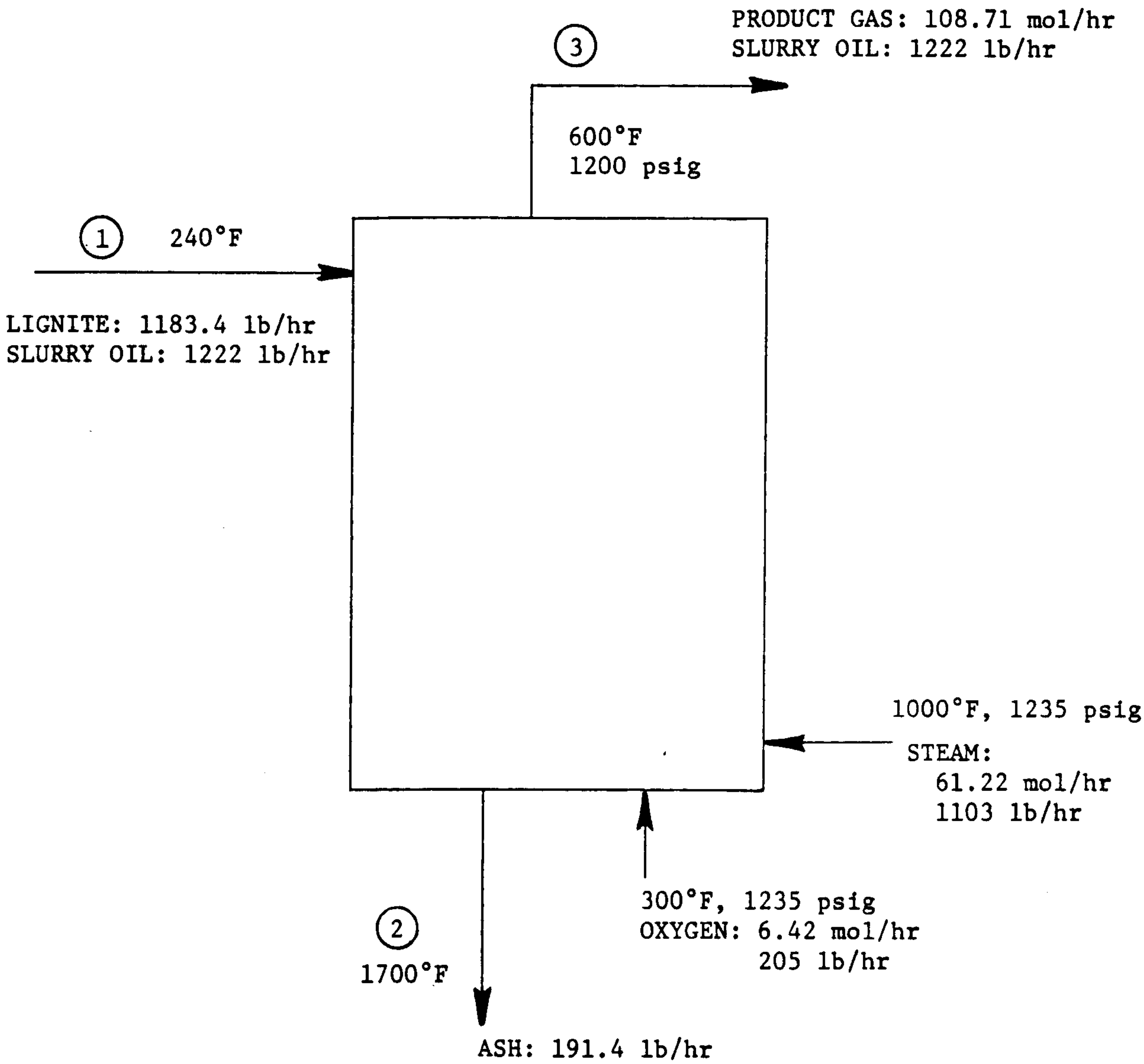
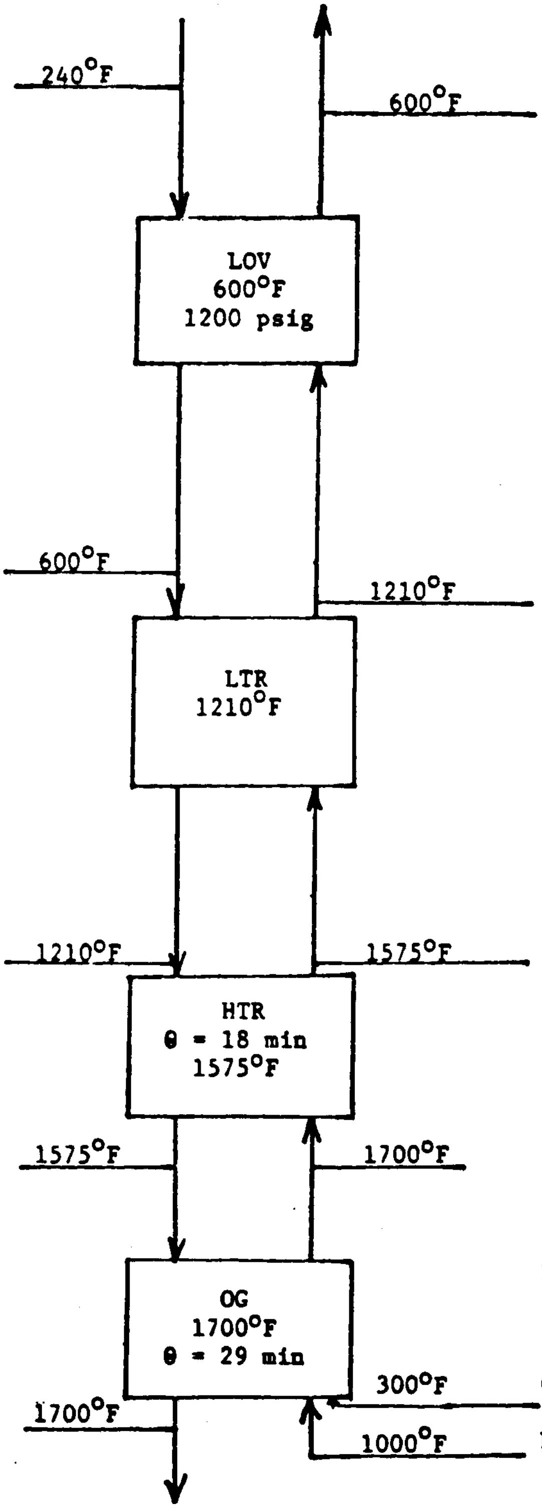


Figure 28. CALCULATED OVERALL GASIFIER MATERIAL BALANCE FOR MONTANA LIGNITE COAL

Basis: 1 hr



Solids streams (moles)

	Feed to LOV (240°F)	LOV to LTR (600°F)	LTR to HTR (1210°F)	HTR to OG (1575°F)	OG discharge (1700°F)
C	51.0160	51.0160	37.3384	28.5437	5.0940
H_2	20.8334	20.8334	2.5837	1.9751	0.7966
O_2	12.5625	12.5625	– –	– –	– –
N_2	0.3462	0.3462	0.0692	0.0692	0.0692
S	0.2745	0.2745	0.0885	0.0885	0.0275
Ash*	125.8	125.8	125.8	125.8	125.8
Moisture	10.1804				
Slurry* Oil	1222				

Gas streams (moles)

	LOV off-gas (600°F)	LTR to LOV (1210°F)	HTR to LTR (1575°F)	OG to HTR (1700°F)
CO	10.7505	10.7505	10.1759	8.5079
CO_2	19.5596	19.5596	15.6901	12.7736
H_2	19.8258	19.8258	17.6896	18.0004
H_2O	46.9181	36.7377	32.4987	39.9997
CH_4	9.4531	9.4531	6.3784	2.1682
C_2H_6	0.7652	0.7652		
NH_3	0.4439	0.4439		
N_2	0.0416	0.0416		
HCN	0.0269	0.0269		
H_2S	0.2371	0.2371	0.0610	0.0610
COS	0.0099	0.0099		
C_6H_6	0.1159	0.1159		
C_7H_8	0.5566	0.5566		
Slurry* Oil	1222			

Feeds to OG (moles)

O_2 (300°F)	6.4179
H_2O (1000°F)	61.2190

* These quantities in lbs.

Figure 29. CALCULATED GASIFIER MATERIAL BALANCE FOR MONTANA LIGNITE COAL SHOWING INTERSTAGE COMPOSITION

Table 9. CALCULATED GASIFIER MATERIAL AND HEAT BALANCE FOR MONTANA SUBBITUMINOUS COAL

Stream No.	1			2		
Description	Coal Feed			Ash Residue		
Temperature, °F	165			1800		
Components	lb/hr	wt %	mol/hr	lb/hr	wt %	mol/hr
C	680.0	68.00	56.62	66.4	40.81	5.53
H_2	44.5	4.45	22.07	1.7	1.04	0.84
O	166.0	16.60	10.38	--	--	--
N_2	8.1	0.81	0.29	1.6	0.98	0.06
S	9.3	0.93	0.29	0.9	0.55	0.03
Ash	92.1	9.21	--	92.1	56.62	--
Total	1000.0	100.00		162.7	100.00	
Moisture	136.4					
Slurry Oil	1857					

Stream No.	3	
Description	Raw Product Gas	
Temperature, °F	600	
Components	mol/hr	mol %
CO	9.29	8.50
CO_2	21.48	19.64
H_2	16.47	15.06
H_2O	46.97	42.95
CH_4	13.74	12.57
C_2H_6	0.71	0.65
NH_3	0.37	0.34
N_2	0.03	0.03
HCN	0.02	0.02
H_2S	0.25	0.23
COS	0.01	0.01
Total (Oil-Free Gas)	109.34	100.00
	mol/hr	wt %
C_6H_6	0.13	15.0
C_7H_8	0.62	85.0
Total (Product Oil)	0.75	100.0
Total (Oil-Free Gas + Product Oil)	110.09	
Slurry Oil, lb/hr	1857	

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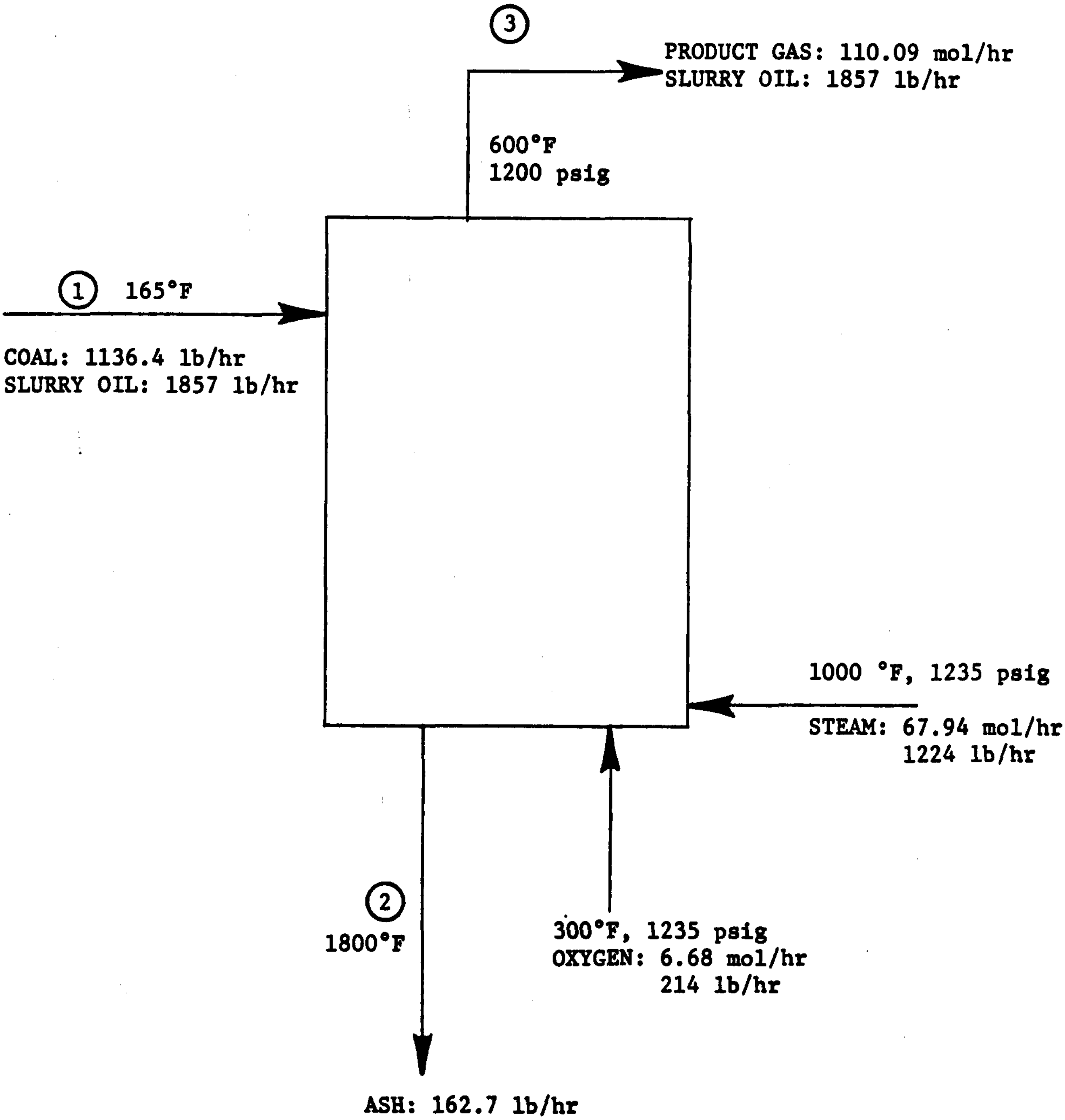
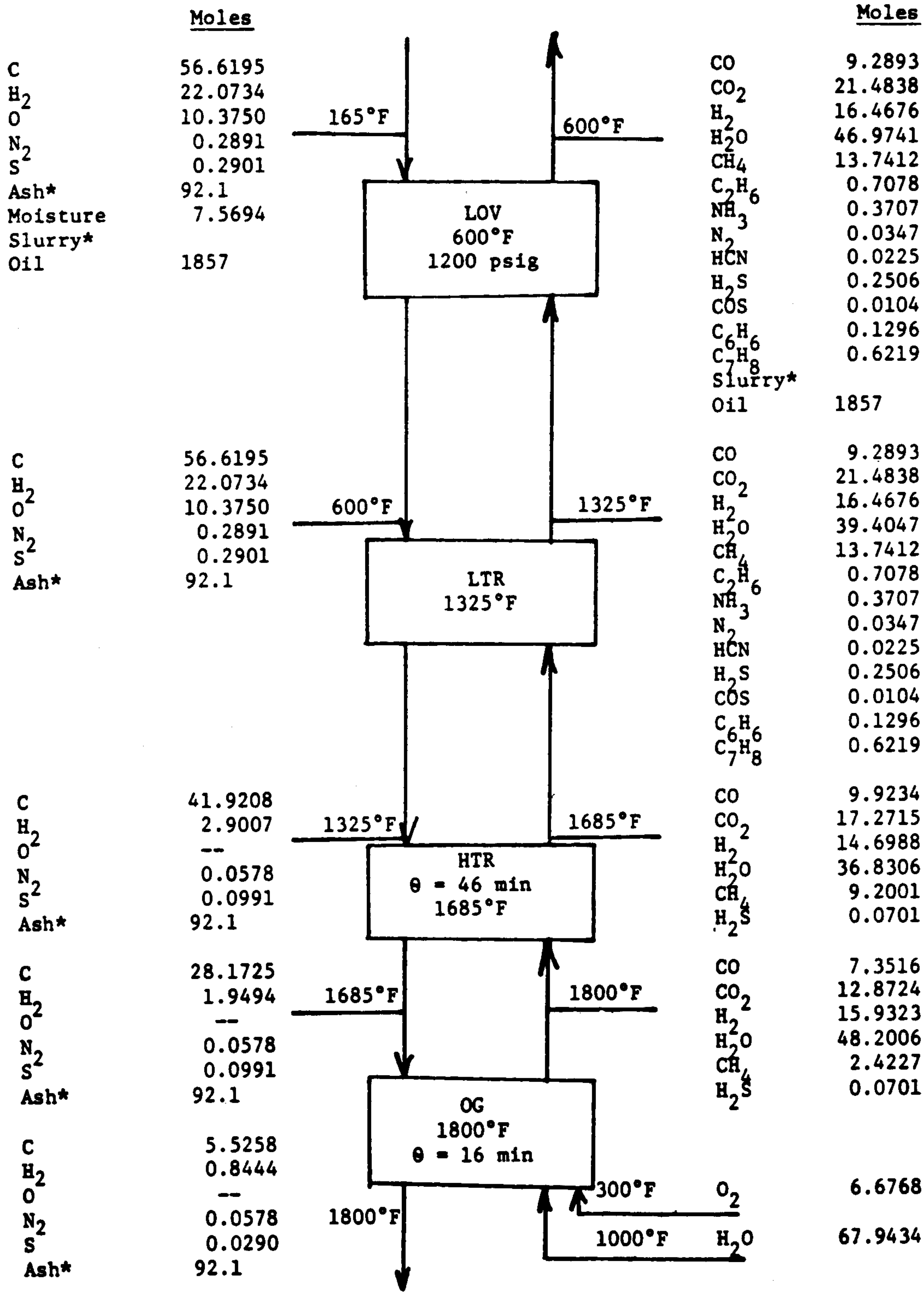


Figure 30. CALCULATED OVERALL GASIFIER MATERIAL BALANCE FOR MONTANA SUBBITUMINOUS COAL

Basis: 1 hr



* These quantities in lbs.

Figure 31. CALCULATED GASIFIER MATERIAL BALANCE FOR MONTANA SUBBITUMINOUS COAL SHOWING INTERSTAGE COMPOSITION

After Test 63, a new conveyor was installed to remove the filter cake from the Alar rotary vacuum filter. Piping modifications were made in the pressure letdown chokes. The new surge pot was not put into service this quarter because the packing on the mechanical stirrer leaked during Test 64.

All construction work and installations on the new high-capacity incinerator were completed. Some difficulties were originally encountered in aligning a section of the incinerator stack, and patchwork on the refractory slowed the installation. Incinerator shakedown was also finished this quarter. Preparation of start-up, maintenance, and shutdown manuals was begun.

Test Methanation Systems and Catalysts

The IGT fixed-bed methanation system was on standby this quarter.

Piping modifications for Chem Systems liquid-phase methanation (LPM) unit were completed early this quarter. X-rays of major welds on the 4-inch piping were all satisfactory. Purified product gas from the HYGAS reactor was fed to Chem Systems LPM pilot unit during Test 64, 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 was very important because it was the first time that this unit had 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.

Recent Methanation Catalyst Evaluation Studies

Evaluation studies of methanation catalyst LDI X-826 were continued this quarter. The first methanation catalyst, supplied by the LDI Catalyst Company (called LDI X-825 and identical to the CRG-A catalyst manufactured in the United Kingdom), was tested during November 1974. The results were presented in Interim Report No. 2 for ERDA Contract E(49-18)-1221. Its poor performance was explained when it was found that a bad batch of catalyst had been sent (ERDA Report No. FE-2434-125, February 1975).

The second LDI catalyst, LDI X-826, was received and first tested during January 1977 (Runs 476 through 487). This set of experiments showed that, under ideal conditions, LDI X-826 is as active and as durable as other high-activity catalysts. Detailed results of the tests, made in January 1977, can be found in the Project 9000 Third Quarter ERDA Report No. FE-2434-12, March 1977.

Evaluation studies of the methanation catalyst LDI X-826 were continued with the same batch of catalyst that was used in January 1977. The reactor had been sealed off in a hydrogen atmosphere after Run 487 at 52 psig. A feed gas, similar to that used in Run 480, was introduced to the catalyst bed after 2000 hours of inactivity to determine if the catalyst had been deactivated by aging. It was found that it had not. 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 32 and Table 10. Again, as expected, the activity of the catalyst was

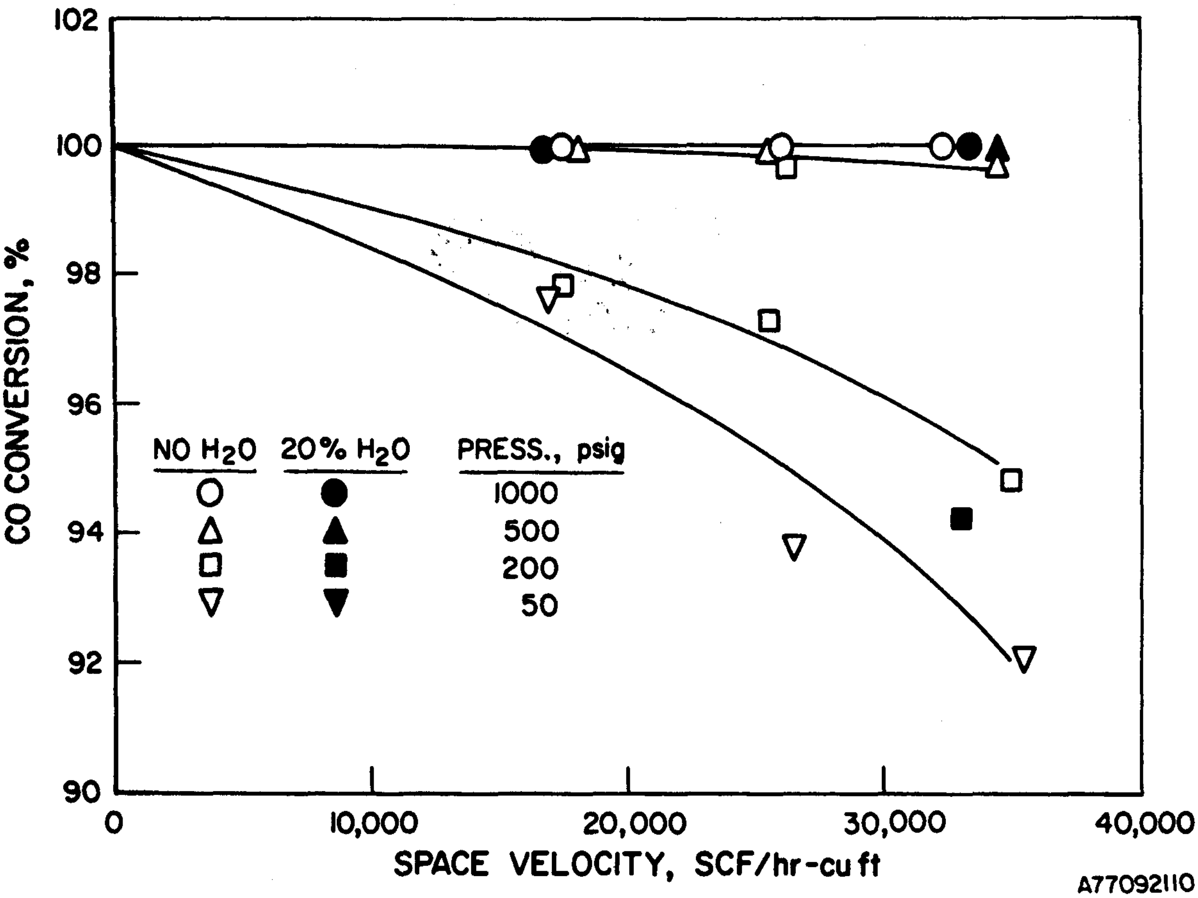


Figure 32. EFFECT OF SPACE VELOCITY AND STEAM ON THE CONVERSION OF CARBON MONOXIDE AT TEMPERATURES RANGING FROM 550° TO 680°F (LDI X-826 Catalyst, 1/8-Inch Cylinders)

Table 10. 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	0.002233	--	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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not affected by this amount of steam. These experiments, consisting of Runs 476 through 492, concluded the first evaluation step of IGT's standard program in which the activity of the catalyst was established within an ideal temperature range and with a feed composition simulating the gas effluent from the coal gasification reactor.

More tests were conducted on LDI X-826 to determine the upper temperature limit for use of this catalyst. Fresh catalyst was loaded into a continuous-stirred-tank-reactor (CSTR), which is an isothermal reactor. The rate of reaction was determined at 590° and 750°F; the results are presented in Table 11. These rates are typical of a high-activity nickel catalyst and are comparable to those obtained using Harshaw Ni-0104T, Union Carbide MC-100, and Girdler G-87p catalysts. When leaks from the CSTR were detected at 1000 psig and 885°F, this test was discontinued.

Tests were also conducted this quarter on Corning Glass Works' CELCOR® methanation catalyst, which was developed in 1975 by Corning when it began searching for nonautomotive applications for its CELCOR substrates. (IGT will not conduct tests on a catalyst unless the manufacturer first tests it under a specified set of minimal conditions.) To meet IGT's prerequisites, Corning tested its catalysts using a feed gas of 79% CH_4, 4% CO, 2% CO_2, and 15% H_2 at 1000 psi and 575°F with a liquid feed rate of about 0.05 cubic centimeters of benzene per minute. Corning prepared and screened eight different catalysts made of RuO or NiO, one of which was selected and submitted to IGT for its evaluation.

Corning's catalysts are supported by (i.e., placed in and around) a honeycombed CELCOR cylinder called a monolith. This arrangement offers many potential advantages, including a very low pressure drop and a strong support with high-temperature resistance. Also, with this type of support, the methanation reactor is less likely to become plugged by carbon that is deposited on the catalyst.

The specific catalyst submitted to IGT for evaluation was made up of Calsicat Ni-230T, Alcoa C-333 Al_2O_3, Conoco Dispal-M Al_2O_3, HNO_3, and H_2O.

Table 11. METHANATION CATALYSIS — CATALYST EVALUATION (CSTR Data)

Catalyst	LDI X-826*		Corning Monoliths†	
Run No.	493	494	495	496
Time, hr	4	28	4	28
Feed Gas Rate, SCF/hr	10.4896	10.5380	10.4832	10.1960
Feed Gas Composition, mol%				
Nitrogen	4.8	4.8	4.4	4.6
Helium	0.11	0.12	0.1	0.11
Carbon Monoxide	3.3	3.3	2.7	3.2
Carbon Dioxide	1.9	1.8	1.4	1.6
Hydrogen	8.7	8.1	18.7	12.7
Methane	81.01	81.77	72.62	77.73
Ethane	0.07	0.05	0.06	0.05
Propane	0.1	0.05	0.01	0.01
Butanes	0.01	0	0	0
Water	0	0	0	0
Total	100.00	100.00	100.00	100.00
Reactor Temperature, °F	591	750	575	752
Reactor Pressure, psig	1002	998	997	1002
Product Gas Rate, SCF/hr	9.6767	9.7187	9.4517	9.1859
Product Gas Composition, mol%				
Nitrogen	5.2	5.4	4.9	5.0
Helium	0.1	0.11	0.11	0.12
Carbon Monoxide	0.6	0.3	0.1	0.2
Carbon Dioxide	2.6	2.5	0.9	1.1
Hydrogen	1.1	1.4	9.2	2.7
Methane	90.02	89.82	84.21	89.73
Ethane	0.03	0.01	0.06	0.05
Propane	0.05	0	0.01	0
Butanes	0	0	0	0
Water	0.3	0.46	0.51	1.1
Total	100.00	100.00	100.00	100.00
Feed H_2/CO Ratio	2.64	2.45	6.93	3.97
Water Collected, lb-mol/hr	0.000075	0.000121	0.000129	0.000272
Rate of CO Conversion X 10^4, lb-mol/hr-g catalyst	1.88	2.08	0.66	0.75

* LDI Catalyst Co.'s LDI X-826 catalyst, 1/8-inch cylinders, 4.0445 grams.

†Corning Glass Works' Corning CELCOR® monoliths support methanation catalyst, 10.9016 grams.

The monoliths contained an average of 10 weight percent NiO. In Corning's tests, the catalyst was subjected to both high-temperature aging studies (600°C) and benzene-poisoning studies (3 mole percent). The results of both tests were satisfactory. Although each experimental run was no longer than 1 day, the total number of tests lasted more than 1 year. During this time, close communication was maintained between Corning and IGT.

Because of the nature of the support (the monolith) and the low concentration of catalytic material (approximately 10 weight percent), the effectiveness of the catalyst is largely dependent on the space velocity. This does not make the catalyst unattractive, however. The durability of the catalytic material and its performance at high temperatures are what is important.

To define the upper temperature limit at which this catalyst is effective, the CSTR was loaded with it and a test was begun. At 1000 psia and 900°F, leaks were detected in the CSTR and the test was discontinued. Valuable data were obtained for temperatures up to 900°F, and the rates of reaction at these lower temperatures are presented in Table 12.

Fresh catalyst was then loaded into a packed-bed reactor to determine the effects of pressure, space velocity, and steam on the conversion of CO. The results are presented in Table 12. The dependence of CO conversion on the space velocity and pressure is shown in Figure 33. This particular phase of testing showed that, on a volumetric basis, the catalyst is not as active as some of the other nickel catalysts. This was expected, because this catalyst has about 10 weight percent NiO, whereas the other high-activity catalysts contain as much as 75 weight percent NiO. It is not necessarily a disadvantage

After 500 hours of testing, the catalyst began to show deactivation due to aging, making it undesirable: It should show virtually no deactivation due merely to aging after this test period. As shown in Figure 34, the catalyst lost about 14% of its weight. The presence of 20 mole percent of steam definitely hindered conversion (Figure 35), but the exact effect of steam on conversion is not well-defined because the reason for the catalytic weight loss is unknown.

The effect of temperature on conversion is presented in Figure 36. This was not investigated in great detail, because the high deactivation rate had already been discovered.

Table 12, Part 1. METHANATION CATALYSIS — CATALYST EVALUATION FOR A PACKED-BED REACTOR (Corning Glass Works, Corning CELCOR Monolith Supported Methanation Catalyst, 9.6981 g)

Run No.	497		498		499	
Time, hr	3		27		51	
Basis for Analysis	Dry	Wet	Dry	Wet	Dry	Wet
Pressure, psig	1003	1003	1003	1003	503	503
Reactor Temperature, °F						
Inlet	595	595	520	520	600	600
Quarter Bed	--	--	--	--	--	--
Middle Bed	759	759	749	749	750	750
Outlet	625	625	745	745	660	660
Furnace Temperature, °F						
Top Zone	595	595	591	591	590	590
Bottom Zone	450	450	525	525	490	490
Flow Rate, lb-mol/hr						
Feed	0.05002	0.05002	0.09687	0.09687	0.05080	0.05080
H_2O	0	0	0	0	0	0
Feed Composition, mol%						
H_2	17.8	17.8	15.3	15.3	16.0	16.0
N_2	4.4	4.4	4.6	4.6	4.7	4.7
CH_4	73.11	73.11	75.03	75.03	74.34	74.34
C_2H_6	0.07	0.07	0.06	0.06	0.05	0.05
C_2H_4	0.01	0.01	0.01	0.01	0.01	0.01
C_3H_8	0	0	0	0	0	0
CO_2	1.5	1.5	1.7	1.7	1.7	1.7
CO	3.0	3.0	3.2	3.2	3.1	3.1
He	0.11	0.11	0.1	0.1	0.1	0.1
H_2O	0	0	0	0	0	0
Total	100.00	100.00	100.00	100.00	100.00	100.00
Flow Rate, lb-mol/hr						
Product	0.04335	0.04688	0.08516	0.09246	0.04463	0.04779
H_2O in Product	0	0.00353	0	0.00730	0	0.00316
Product Composition, mol%						
H_2	5.0	4.7	3.6	3.4	3.9	3.6
N_2	5.0	4.7	5.2	4.8	5.4	5.0
CH_4	89.37	82.52	89.54	82.39	89.37	83.54
C_2H_6	0	0	0.04	0.04	0.06	0.05
C_2H_4	0.01	0.01	0.01	0.01	0	0
C_3H_8	0	0	0	0	0	0
CO_2	0.43	0.40	1.2	1.1	0.9	0.84
CO	0.07	0.06	0.3	0.27	0.26	0.24
He	0.12	0.11	0.11	0.10	0.11	0.11
H_2O	0	7.5	0	7.89	0	6.62
Total	100.00	100.00	100.00	100.00	100.00	100.00
CO Consumed, lb-mol/hr	0.001473	--	0.002850	--	0.001460	--
CO_2 Changed, lb-mol/hr	−0.00056	--	−0.00063	--	−0.00046	--
H_2 Consumed, lb-mol/hr	0.006700	--	0.001168	--	0.006407	--
H_2O Produced, lb-mol/hr	0.003532	--	0.007298	--	0.003164	--
CH_4 Produced, lb-mol/hr	0.002114	--	0.003496	--	0.002167	--
C_2H_6 Consumed, lb-mol/hr	0.000035	--	0.000021	--	0	--
C_3H_8 Consumed, lb-mol/hr	--	--	--	--	--	--
Space Velocity, SCF/hr-cu ft	18,526	--	35,878	--	18,816	--

Table 12, Part 2. METHANATION CATALYSIS — CATALYST EVALUATION FOR A PACKED-BED REACTOR (Corning Glass Works, Corning CELCOR Monolith Supported Methanation Catalyst, 9.6981 g)

Run No.	500		501		502	
Time, hr	99		171		243	
Basis for Analysis	Dry	Wet	Dry	Wet	Dry	Wet
Pressure, psig	498	498	200	200	202	202
Reactor Temperature, °F						
Inlet	520	520	600	600	539	539
Quarter Bed	--	--	--	--	--	--
Middle Bed	730	730	765	765	723	723
Outlet	780	780	720	720	790	790
Furnace Temperature, °F						
Top Zone	592	592	590	590	595	595
Bottom Zone	620	620	565	565	640	640
Flow Rate, lb-mol/hr						
Feed	0.10183	0.10183	0.05103	0.05103	0.10076	0.10076
H_2O	0	0	0	0	0	0
Feed Composition, mol%						
H_2	13.1	13.1	15.4	15.4	15.1	15.1
N_2	4.6	4.6	4.6	4.6	4.7	4.7
CH_4	76.6	76.6	74.5	74.5	74.62	74.62
C_2H_6	0.09	0.09	0.07	0.07	0.06	0.06
C_2H_4	0.01	0.01	0	0	0.02	0.02
C_3H_8	0	0	0	0	0	0
CO_2	1.7	1.7	1.7	1.7	1.9	1.9
CO	3.3	3.3	3.2	3.2	3.5	3.5
He	0.6	0.6	0.53	0.53	0.1	0.1
H_2O	0	0	0	0	0	0
Total	100.00	100.00	100.00	100.00	100.00	100.00
Flow Rate, lb-mol/hr						
Product	0.08940	0.09482	0.04496	0.04758	0.08840	0.09248
H_2O in Product	0	0.00542	0	0.00262	0	0.00407
Product Composition, mol%						
H_2	3.9	3.7	4.7	4.4	4.6	4.4
N_2	5.2	4.9	5.2	4.9	5.3	5.1
CH_4	88.44	83.37	87.97	83.17	87.0	83.1
C_2H_6	0.02	0.02	0.01	0.01	0.01	0.01
C_2H_4	0.01	0.01	0	0	0.02	0.02
C_3H_8	0	0	0	0	0	0
CO_2	1.48	1.40	1.41	1.33	1.69	1.61
CO	0.69	0.65	0.05	0.05	1.27	1.22
He	0.26	0.25	0.66	0.63	0.11	0.11
H_2O	0	5.7	0	5.51	0	4.43
Total	100.00	100.00	100.00	100.00	100.00	100.00
CO Consumed, lb-mol/hr	0.002744	--	0.001334	--	0.002399	--
CO_2 Changed, lb-mol/hr	−0.00040	--	−0.00023	--	−0.00042	--
H_2 Consumed, lb-mol/hr	0.009830	--	0.005764	--	0.009431	--
H_2O Produced, lb-mol/hr	0.005426	--	0.002625	--	0.004071	--
CH_4 Produced, lb-mol/hr	0.003196	--	0.001556	--	0.002847	--
C_2H_6 Consumed, lb-mol/hr	0.000072	--	0.000031	--	0.000051	--
C_3H_8 Consumed, lb-mol/hr	--	--	--	--	--	--
Space Velocity, SCF/hr-cu ft	36,784	--	18,900	--	36,799	--

Table 12, Part 3. METHANATION CATALYSIS — CATALYST EVALUATION FOR A PACKED-BED REACTOR (Corning Glass Works, Corning CELCOR Monolith Supported Methanation Catalyst, 9.6981 g)

Run No.	503		504		505	
Time, hr	363		387		435	
Basis for Analysis	Dry	Wet	Dry	Wet	Dry	Wet
Pressure, psig	51	51	51	51	202	202
Reactor Temperature, °F						
Inlet	592	592	539	539	410	410
Quarter Bed	--	--	--	--	--	--
Middle Bed	758	758	695	695	505	505
Outlet	768	768	775	775	550	550
Furnace Temperature, °F						
Top Zone	595	595	595	595	482	482
Bottom Zone	640	640	665	665	550	550
Flow Rate, lb-mol/hr						
Feed	0.05077	0.05077	0.09965	0.09965	0.10107	0.10107
H_2O	0	0	0	0	0	0
Feed Composition, mol%						
H_2	14.3	14.3	16.6	16.6	14.8	14.8
N_2	4.8	4.8	4.4	4.4	4.3	4.3
CH_4	75.28	75.28	73.65	73.65	75.62	75.62
C_2H_6	0.07	0.07	0.07	0.07	0.07	0.07
C_2H_4	0	0	0	0	0.01	0.01
C_3H_8	0	0	0	0	0	0
CO_2	1.8	1.8	1.73	1.73	1.7	1.7
CO	3.3	3.3	3.25	3.25	3.2	3.2
He	0.45	0.45	0.3	0.3	0.3	0.3
H_2O	0	0	0	0	0	0
Total	100.00	100.00	100.00	100.00	100.00	100.00
Flow Rate, lb-mol/hr						
Product	0.04704	0.04870	0.09301	0.09538	0.09905	0.10015
H_2O in Product	0	0.00166	0	0.00237	0	0.00109
Product Composition, mol%						
H_2	7.4	7.2	10.7	10.4	13.1	13.0
N_2	5.2	5.0	4.7	4.6	4.4	4.4
CH_4	84.03	81.17	80.88	78.92	77.69	76.75
C_2H_6	0.01	0.01	0.02	0.02	0.07	0.07
C_2H_4	0	0	0	0	0	0
C_3H_8	0	0	0	0	0	0
CO_2	1.9	1.8	1.8	1.75	1.61	1.59
CO	1.0	0.95	1.6	1.56	2.82	2.79
He	0.46	0.45	0.3	0.3	0.31	0.30
H_2O	0	3.42	0	2.48	0	1.1
Total	100.00	100.00	100.00	100.03	100.00	100.00
CO Consumed, lb-mol/hr	0.001212	--	0.001754	--	0.000467	--
CO_2 Changed, lb-mol/hr	−0.00004	--	−0.00005	--	−0.00014	--
H_2 Consumed, lb-mol/hr	0.003753	--	0.006600	--	0.001961	--
H_2O Produced, lb-mol/hr	0.001665	--	0.002373	--	0.001097	--
CH_4 Produced, lb-mol/hr	0.001304	--	0.001838	--	0.000433	--
C_2H_6 Consumed, lb-mol/hr	0.000031	--	0.000005	--	0.000001	--
C_3H_8 Consumed, lb-mol/hr	--	--	--	--	--	--
Space Velocity, SCF/hr-cu ft	18,806	--	36,909	--	37,433	--

Table 12, Part 4. METHANATION CATALYSIS — CATALYST EVALUATION FOR A PACKED-BED REACTOR (Corning Glass Works, Corning CELCOR Monolith Supported Methanation Catalyst, 9.6981 g)

Run No.	506		507	
Time, hr	507		531	
Basis for Analysis	Dry	Wet	Dry	Wet
Pressure, psig	502	502	997	997
Reactor Temperature, °F				
Inlet	593	593	365	365
Quarter Bed	--	--	--	--
Middle Bed	749	749	592	592
Outlet	758	758	650	650
Furnace Temperature, °F				
Top Zone	590	590	495	495
Bottom Zone	617	617	590	590
Flow Rate, lb-mol/hr				
Feed	0.05211	0.05211	0.05148	0.06457
H_2O	0	0	0	0.01309
Feed Composition, mol%				
H_2	14.8	14.8	14.5	11.6
N_2	4.4	4.4	4.4	3.5
CH_4	75.76	75.76	76.37	60.83
C_2H_6	0.05	0.05	0.02	0.01
C_2H_4	0.01	0.01	0.04	0.03
C_3H_8	0	0	0	0
CO_2	1.7	1.7	1.5	1.2
CO	3.2	3.2	3.0	2.4
He	0.08	0.08	0.17	0.13
H_2O	0	0	0	20.3
Total	100.00	100.00	100.00	100.00
Flow Rate, lb-mol/hr				
Product	0.04668	0.04954	0.04940	0.06422
H_2O in Product	0	0.00286	0	0.01482
Product Composition, mol%				
H_2	5.7	5.4	10.9	8.4
N_2	4.9	4.6	4.6	3.5
CH_4	87.17	82.14	80.98	62.28
C_2H_6	0.01	0.01	0.02	0.02
C_2H_4	0.01	0.01	0.03	0.02
C_3H_8	0	0	0	0
CO_2	1.3	1.22	1.65	1.27
CO	0.9	0.85	1.65	1.27
He	0.01	0.01	0.17	0.14
H_2O	0	5.76	0	23.1
Total	100.00	100.00	100.00	100.00
CO Consumed, lb-mol/hr	0.001247	--	0.000728	--
CO_2 Changed, lb-mol/hr	−0.00028	--	0.000043	--
H_2 Consumed, lb-mol/hr	0.005037	--	0.002070	--
H_2O Produced, lb-mol/hr	0.002856	--	0.01482	--
CH_4 Produced, lb-mol/hr	0.001211	--	0.000689	--
C_2H_6 Consumed, lb-mol/hr	0.000021	--	0.000090	--
C_3H_8 Consumed, lb-mol/hr	--	--	--	--
Space Velocity, SCF/hr-cu ft	19,246	--	19,068	--

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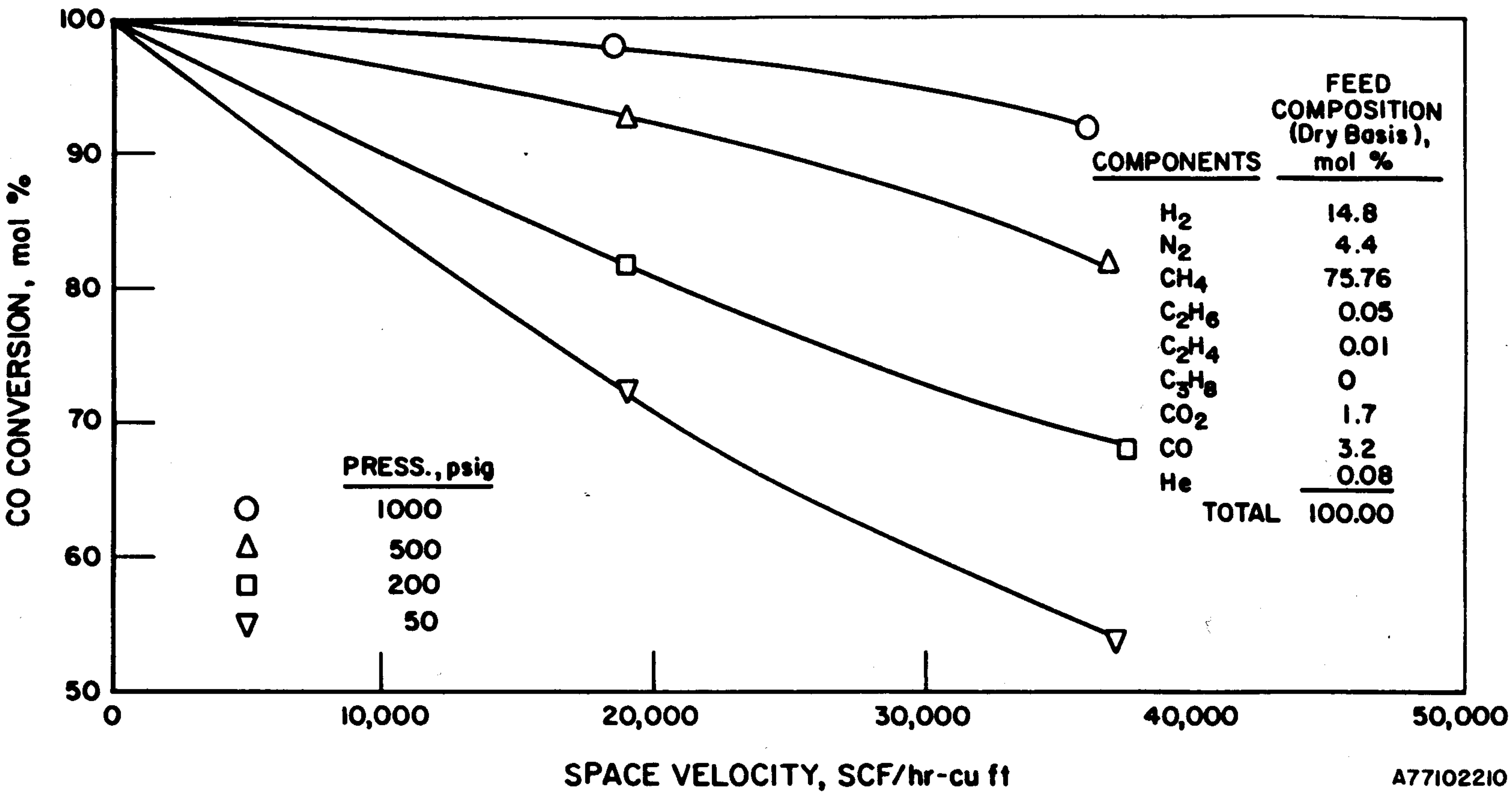


Figure 33. EFFECT OF SPACE VELOCITY ON THE CONVERSION OF CARBON MONOXIDE (Corning CELCOR Monolith Supported Methanation Catalyst, 9.6981 g)

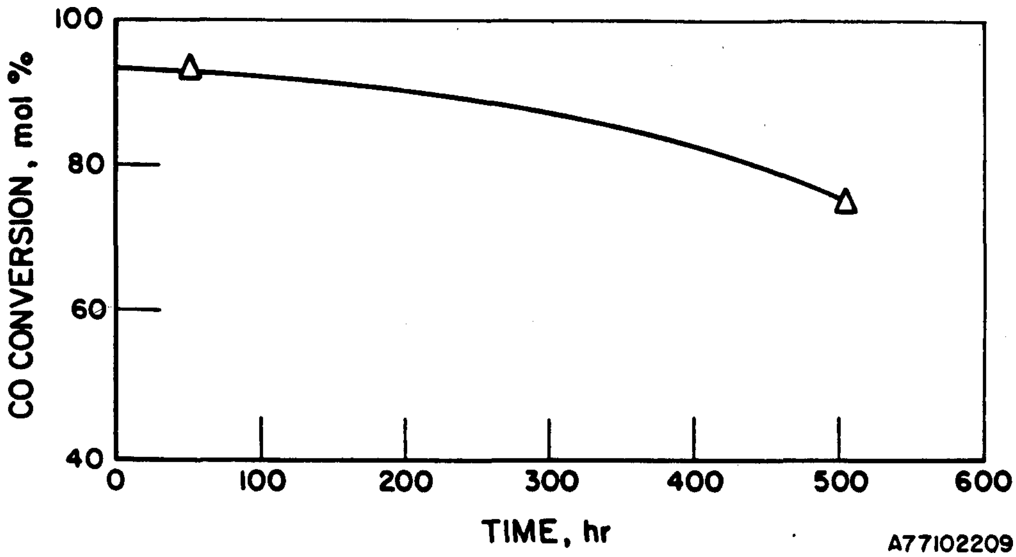


Figure 34. DEACTIVATION DUE TO AGING AT 750°F AND 500 psig (Corning CELCOR Monolith Supported Methanation Catalyst, 9.6981 g)

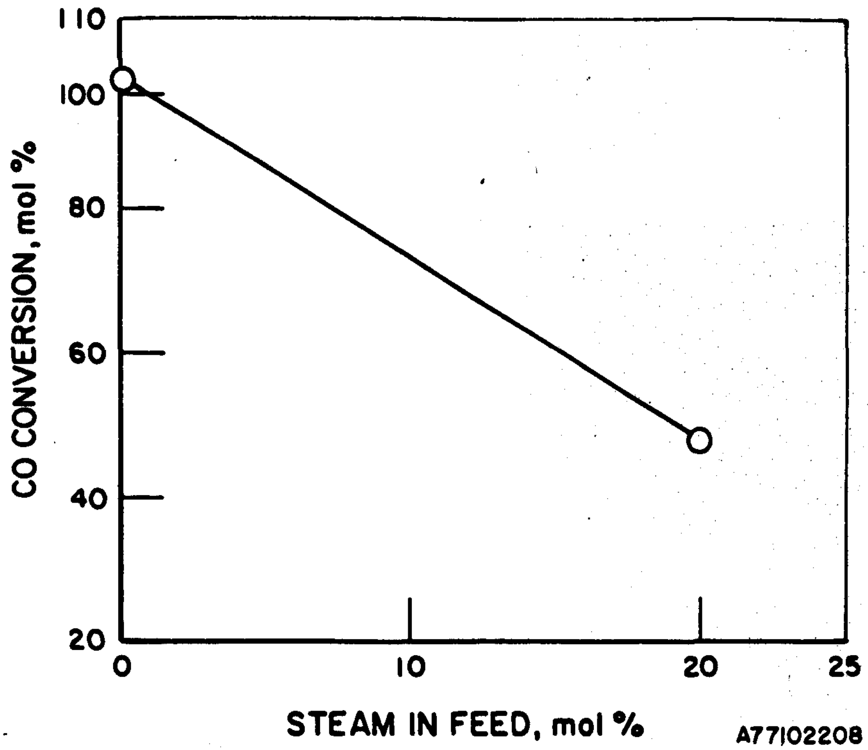


Figure 35. EFFECT OF STEAM ON THE CONVERSION OF CARBON MONOXIDE AT 750°F, 1000 psig, AND 20,000 SCF/hr-cu ft (Corning CELCOR Monolith Supported Methanation Catalyst, 9.6981 g)

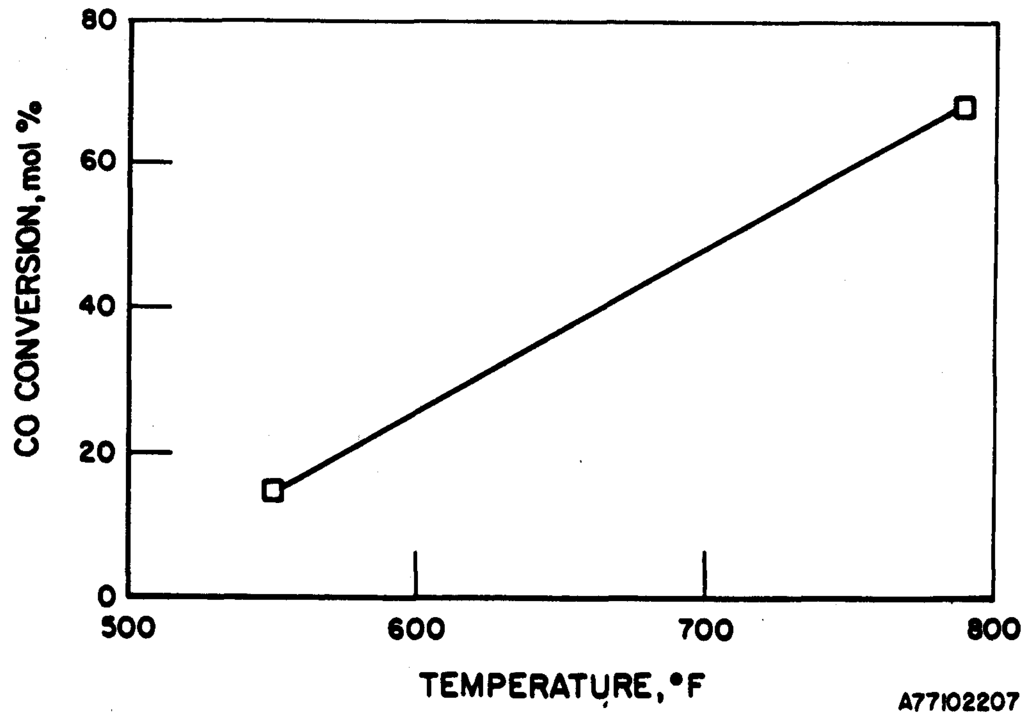


Figure 36. EFFECT OF TEMPERATURE ON THE CONVERSION OF CARBON MONOXIDE AT 200 psig AND 40,000 SCF/hr-cu ft (Corning CELCOR Monolith Supported Methanation Catalyst, 9.6981 g)

Because of the high deactivation rate, evaluation of this catalyst was discontinued. A better technique to put the catalytic material on the monolith is needed to make this type of catalyst more durable, and thus more attractive. Corning is currently analyzing the catalyst's physical properties.

Investigation of the Hot-Oil Quench System

Initial design of the hot-oil quench system was begun this quarter. A study of equilibrium between light oil, gas, and water was conducted to determine what condensable compositions could be expected.

Materials Testing

X-ray and ultrasonic testing of all piping in erosive service was completed during the plant turnaround period. No major wear areas were found. An inspection of the high-pressure reactor cyclone revealed additional wear, and a removable sleeve was installed in the solids discharge pipe to prevent additional wear on the base metal.

MPC corrosion and erosion test coupons were exposed during Tests 64 and 65. Nondestructive testing was performed on slurry lines and high-pressure lines in the HYGAS pilot plant prior to Test 64.

Engineering Services

Cold-Flow Model

A cold-flow model simulating the second-stage gasifier and the steam-oxygen gasifier was fabricated to investigate the conditions by which the steam-oxygen gasifier fluidized bed is established and the sensitivity of the system to normal variations in operating conditions. Figure 37 is a schematic of the test unit. To make the model, an existing low-pressure solids recirculation unit (used for nonmechanical valve testing) was modified to include two Plexiglas fluidized beds as shown in Figures 38 and 39.

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 high-temperature reactor). 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 40). The solids from the lower bed were then transferred into the lift line, which transported the solids back into the solids receiver.

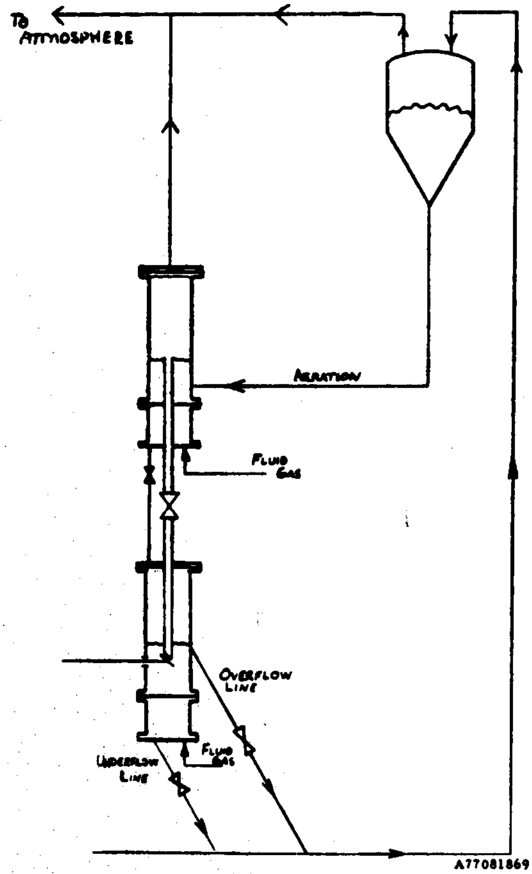
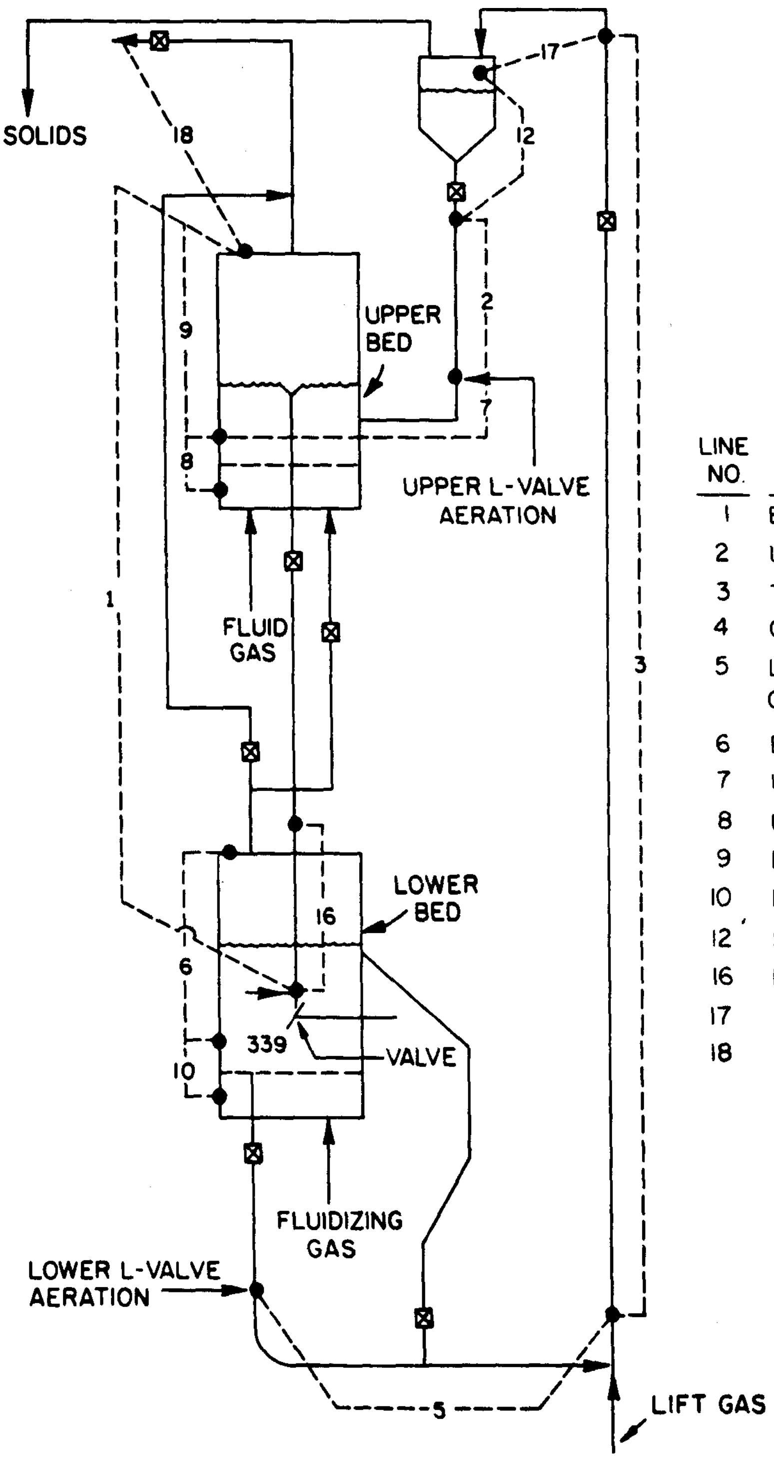


Figure 37. SCHEMATIC OF COLD-FLOW MODEL FOR LOWER REACTOR SOLIDS-FLOW STUDY

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 simulated line 339 using two different techniques. With the first technique, flapper valve 339 was first closed, and gas flow was started through the lower bed. Solids were then charged to the upper high-temperature reactor bed through the L-valve, and a bed level higher



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

Figure 38. INSTRUMENTATION SCHEMATIC FOR THE COLD-MODEL UNIT

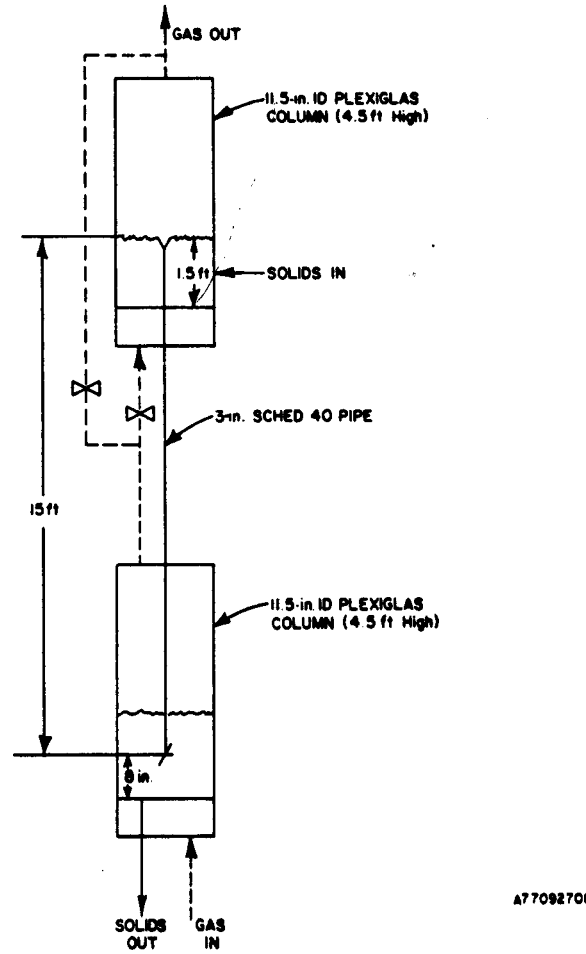


Figure 39. COLD-MODEL FLOWSHEET USED TO SIMULATE THE OPERATION OF LINE 339

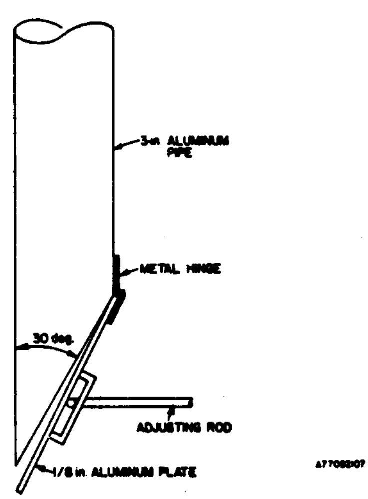


Figure 40. CROSS SECTION OF SIMULATED LINE 339 FLAPPER VALVE

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 size of 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 rate of aeration gas fed to line 339 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 through 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. A gas velocity of 6.6 ft/s in line 339 was found to be the 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 at 6.8 ft/s. (See the section on calculations.) This agrees excellently with the experimental results.

Apparently, excessive gas flow through the "closed" flapper valve and up line 339 prevents solids from passing down line 339 while the steam-oxygen bed is being filled. 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, 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 lower 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 high-temperature reactor, 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 the steam-oxygen bed was lost because of a faulty instrument reading. Attempts to restart the solids flow failed, i.e., line 339 could not be filled with solids. This failure was probably due to the fact that there was a full bed in the high-temperature reactor (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 have been done to solve the gas bypassing problem, but each solution had advantages and disadvantages. If a positive shutoff valve (such as a ball valve) was 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 have been 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 had some means of closing off its end for initial start-up. It would also be difficult to install.

Another approach would have been to convert overflow line 399 to an underflow line. This would have had 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 high-temperature reactor 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 —

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

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

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

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

Process Development Unit Operation

A 6-inch process development unit used to simulate the HYGAS pilot plant steam-oxygen gasification zone was operated with Peabody No. 10 mine char at residence times equivalent to those in the HYGAS reactor. The objective of these tests was to determine char conversion as a function of residence time. Results will be presented in a later report.

Computer Modeling

IGT has developed a generalized kinetic and mathematical model for use in a computer simulation to describe the expected performance of the HYGAS gasifier as a function of a wide range of variables. This model is used for the initial design of the HYGAS gasifier for large-scale plants. The philosophy of this approach is that a generalized model should be developed so that each specific coal need not be tested in the pilot plant before commercialization. However, for the first large-scale plant the design basis has to be verified by pilot plant tests on the coal of interest.

The model is based upon several hundred laboratory determinations, bench-scale tests, and pilot plant runs. It is continually being updated by results from the HYGAS pilot plant and other on-going studies at IGT. The basic kinetic correlations for the gasification rates of different coals have been developed from laboratory thermobalance data. Experimental results from integral gasification process development unit studies were used to characterize gas-solids contacting configurations for fluidized-bed gasifiers. This information, along with the constraints of the operating conditions and the requirements of energy and material balances are utilized to simulate the behavior of the three subelements of the HYGAS gasifier: 1) the low-temperature reactor, 2) the high-temperature reactor, and 3) the steam-oxygen gasifier. The computer model then predicts the operating characteristics, the utility requirements, and the size of the gasifier.

Representatives of ERDA-MFPM, ERDA's Coal Conversion Utilization Division, Darcon, and IGT met on September 14 to discuss, among other things, computer plots generated by this model.

Comparison of Pilot Plant and Process Development Unit Data With That Predicted by the Model

The computer model has been tested by comparing the data from steady-state periods of HYGAS pilot plant operations with that predicted by the model under several sets of operating conditions and different types of coal feeds. Data from a 6-inch-diameter process development unit steam-oxygen reactor have also been used to test the model. The comparison covers a wide range of operating variables and illustrates the versatility of the model as a design tool.

Pilot Plant Comparison

The pilot plant data were obtained from several steady-state periods of the plant operations covering all three types of coal feeds — lignite subbituminous, and bituminous — and several sets of operating conditions. The selected runs are for fully integrated gasification operations (no source of external heat or hydrogen) and of long enough duration to provide good heat and material balances. A wide carbon conversion range from 40% to 90% is covered in the selected steady-state periods.

To use the model for simulating a steady-state period of a pilot plant run, the coal feed, the steam feed, the solids residence time, and the temperature in various zones of the gasifier are set similar to those experienced in the run. Based on this information, the model predicts the percentage of carbon conversion, the off-gas composition, and the oxygen consumption. The comparison of the actual and predicted carbon conversion is shown in Table 13 and Figure 41. The gas yields comparison is shown in Table 14 and in Figures 42 and 43. The model predicts the pilot plant results with good accuracy over a wide range of operations. A comparison of the pilot plant data to date shows that use of the model will generally result in a conservative gasifier design.

In some instances, the pilot plant operating conditions were such that the available coal-characterizing factors used in the model were not applicable. In these cases, the coal parameters had to be estimated and the model updated to simulate pilot plant operating conditions. Moreover, in the gas yield data, the methane and ethane yields are considered together because the present model does not account for hydrogenation of the ethane that has evolved during devolatilization. A project presently under way at IGT is a detailed study of the overall kinetics of fast devolatilization reactions. As results become available, they will be incorporated in the model.

Process Development Unit Data Comparison

The process development unit data used in evaluating the model are from a 6-inch-diameter steam-oxygen reactor used in the development of the HYGAS Process.[1] The reactor used a fluidized bed to gasify, with steam-oxygen

[1] Institute of Gas Technology, "HYGAS: 1964 to 1972. Pipeline Gas From Coal — Hydrogenation (IGT Hydrogasification Process)" IGT Res. Rep. No. 22, OCR Contract No. 14-01-0001-381. Chicago, 1975.

Table 13. COMPARISON OF ACTUAL VS. PREDICTED* CARBON CONVERSION — PILOT PLANT DATA (IGT HYGAS Steam-Oxygen Pilot Plant)

Run No.	Period, hours	Carbon Conversion, % Actual†	Predicted
Lignite			
34	19	83.4	83.9
37	32	83.5	84.10
37	30	83.2	86.0
37	39	79.8	81.7
37	40	88.4	90.6
Bituminous Coal			
43	4	62.9	54.1
46	32	62.4	59.8
48	8	53.5	53.2
52	88	45.4	40.9
54	32	40.0	37.6
54	79	47.3	46.6
54	24	45.5	45.0
61	4	90.1	88.1
Subbituminous Coal			
55	22	63.8	62.7
55	16	66.4	63.8
55	13	65.3	64.5
58	19	65.6	63.8
58	18	69.5	75.1
58	24	67.0	68.6
58	32	61.2	58.7
58	15	66.8	70.2

* The pilot plant operation is simulated using the HYGAS computer model. Operating temperatures, feed steam and solids residence times similar to the pilot plant run are used in the model to predict the carbon conversions.

† The actual values are based on an ash balance.

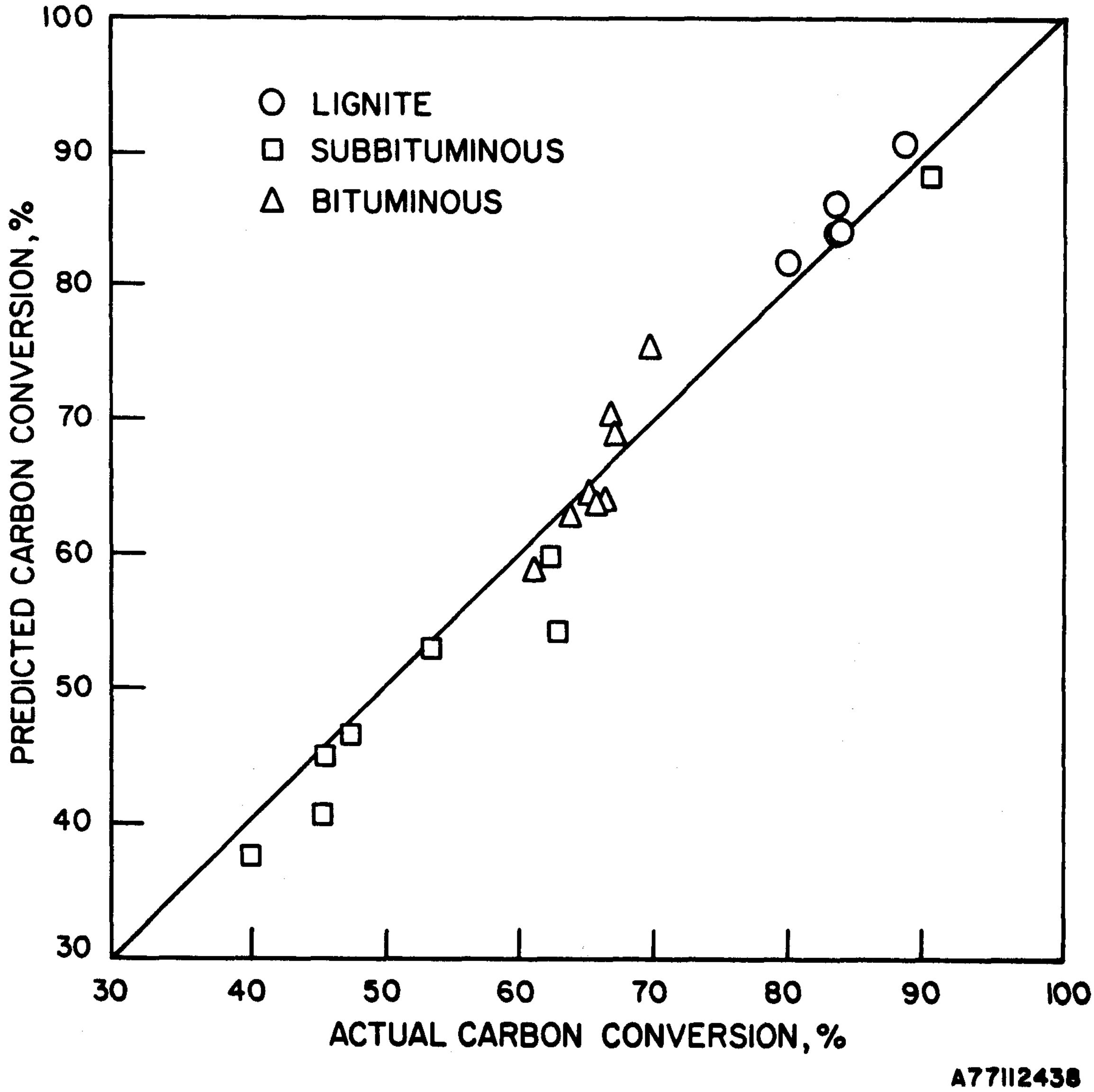


Figure 41. COMPARISON OF ACTUAL VS. PREDICTED CARBON CONVERSION — PILOT PLANT DATA (IGT HYGAS Steam-Oxygen Pilot Plant)

Table 14. COMPARISON OF ACTUAL VS. PREDICTED GAS YIELDS — PILOT PLANT DATA (IGT HYGAS Steam-Oxygen Pilot Plant)

Run No.	Period, hours	Moles of Carbon in Methane + Ethane Formed/ Mole of Carbon in Feed		Moles of Carbon Oxides Formed/ Mole of Carbon in Feed	
		Actual	Predicted	Actual	Predicted
Lignite					
34	19	0.232	0.183	0.543	0.578
37	32	0.160	0.161	0.465	0.468
37	30	0.159	0.162	0.440	0.429
37	39	0.171	0.163	0.449	0.461
37	40	0.150	0.165	0.486	0.496
Bituminous Coal					
43	4	0.151	0.153	0.276	0.287
46	32	0.132	0.131	0.246	0.251
48	8	0.113	0.107	0.185	0.196
52	88	0.146	0.128	0.242	0.260
54	32	0.103	0.103	0.149	0.164
54	79	0.124	0.123	0.196	0.218
54	24	0.123	0.123	0.193	0.206
61	4	0.235	0.182	0.504	0.608
Subbituminous Coal					
55	22	0.197	0.194	0.355	0.358
55	16	0.222	0.215	0.392	0.407
55	13	0.182	0.189	0.291	0.305
58	19	0.194	0.190	0.322	0.326
58	18	0.197	0.189	0.372	0.379
58	24	0.212	0.198	0.359	0.364
58	32	0.215	0.194	0.312	0.358
58	15	0.195	0.193	0.375	0.399

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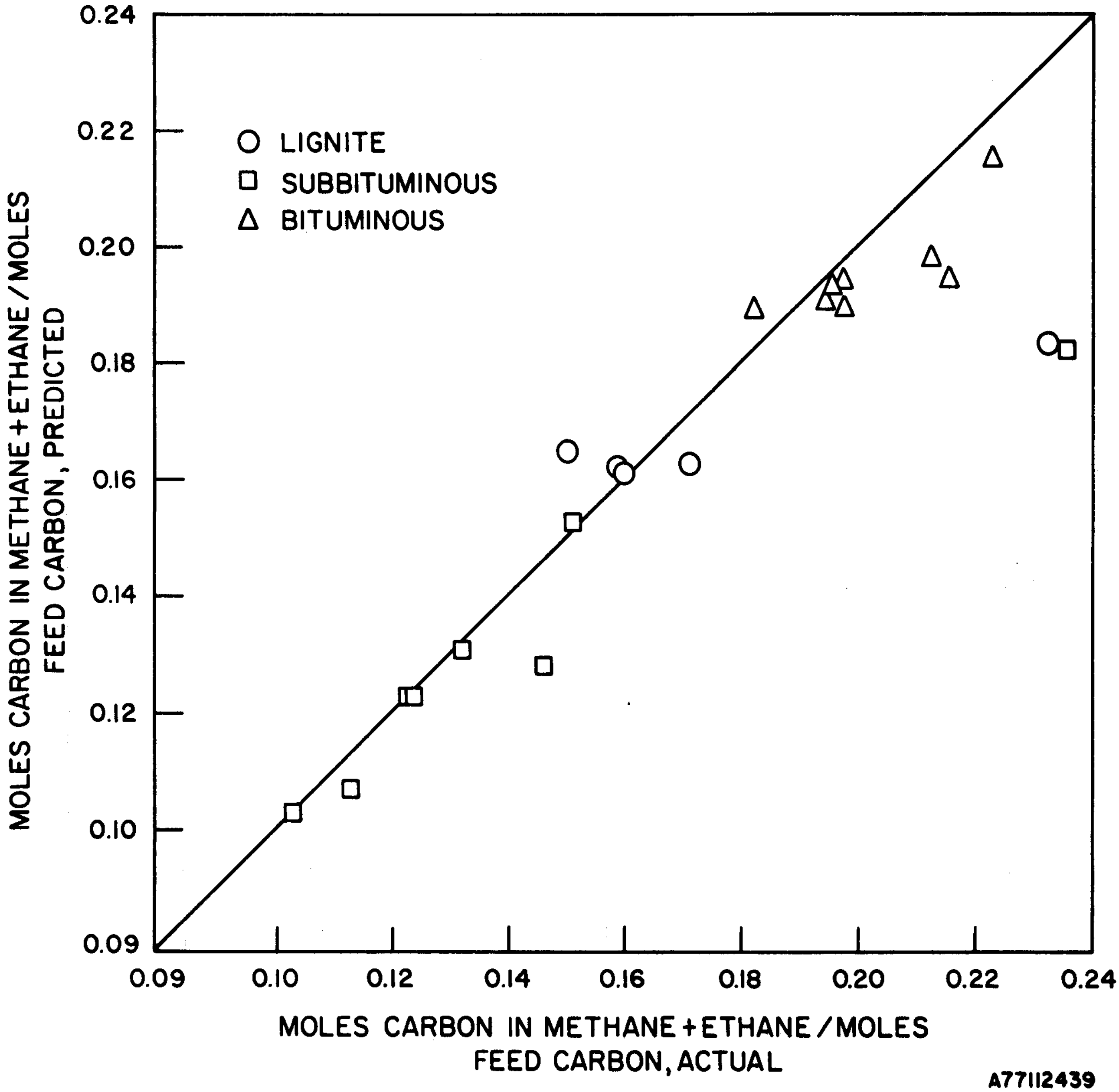

Figure 42. COMPARISON OF ACTUAL VS. PREDICTED METHANE PLUS ETHANE YIELDS — PILOT PLANT DATA

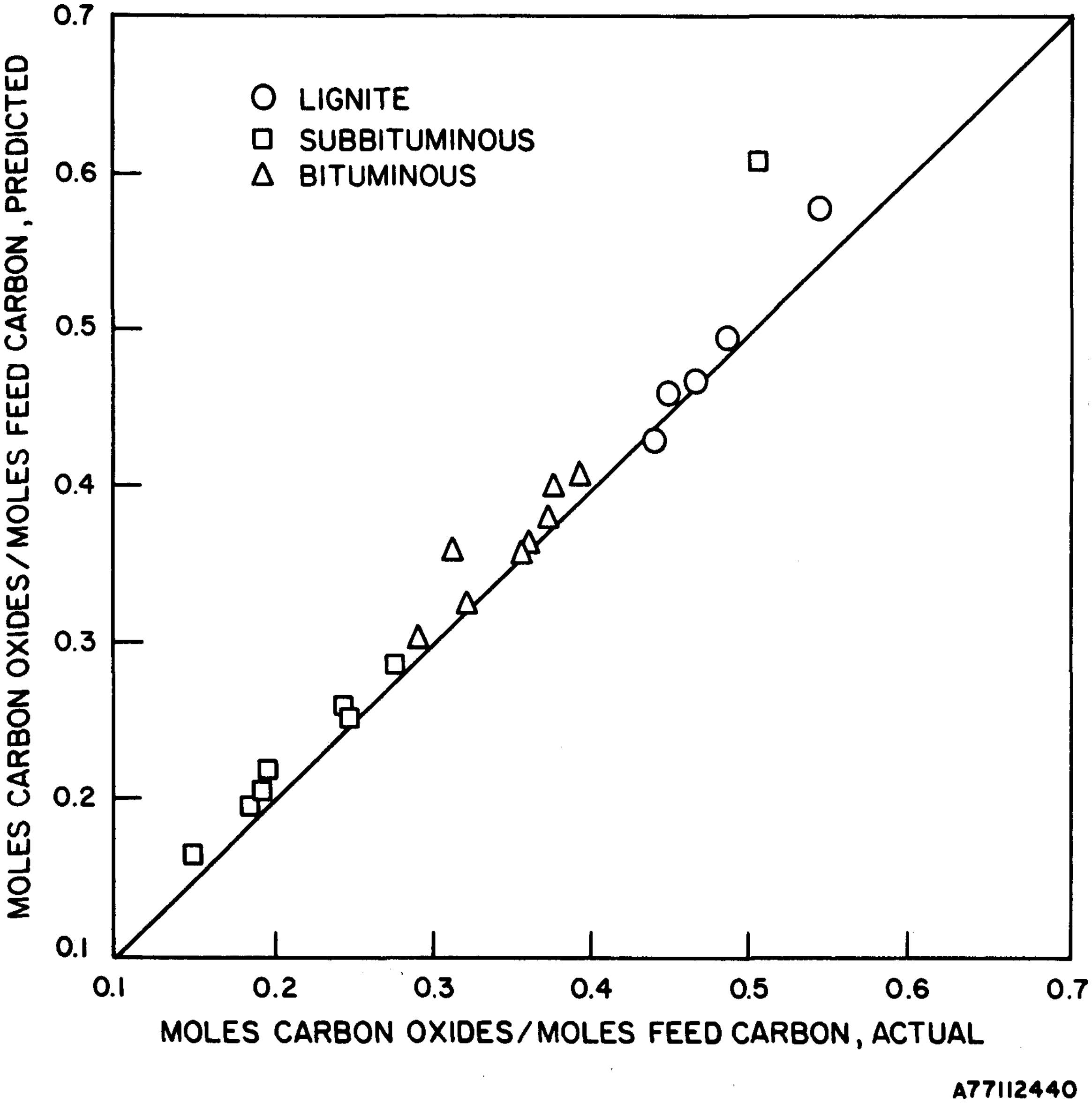

Figure 43. COMPARISON OF ACTUAL VS. PREDICTED CARBON OXIDES ($CO + CO_2$) YIELDS — PILOT PLANT DATA

mixtures, a char derived from an Illinois coal by the COED Process at a rate of about 60 to 80 pounds. The operating temperature ranged from 1700° to 1900°F, and the carbon conversion achieved ranged from 40% to 92%. Data from about 18 runs are used in comparing the carbon conversions and the yields of methane and carbon oxides ($CO + CO_2$) with those predicted by the model. The carbon conversion comparison is shown in Table 15 and Figure 44; the gas yields are compared in Table 16 and Figures 45 and 46. This comparison also shows the applicability of the model for predicting carbon conversions and gas yields over a wide range of operating variables. The range of process development unit gas yields data differs from the range of pilot plant data because the former corresponds to only the steam-oxygen section of the pilot plant, where the carbon oxide formation is higher and the methane formation much lower.

Table 15. COMPARISON OF ACTUAL VS. PREDICTED CARBON CONVERSION — PDU DATA*

Run No.	Carbon Conversion, %	
	Actual †	Predicted
21	76.9	77.0
23	61.1	60.5
24	47.1	46.8
26	62.5	64.9
27	46.4	48.6
29	89.2	87.1
30	64.5	65.9
31	63.7	64.7
34	39.6	41.4
39	58.3	62.3
44	61.8	65.0
45	53.3	54.5
49	79.0	78.3
51	62.3	62.3
55	90.9	96.0
57	92.0	96.5
58	92.0	96.0
60	85.8	88.5

*Source: 6-inch-diameter process development unit (PDU) steam-oxygen gasification tests.

Feed: Project COED char from Illinois No. 6 coal.

† The actual balance based on carbon in gas.

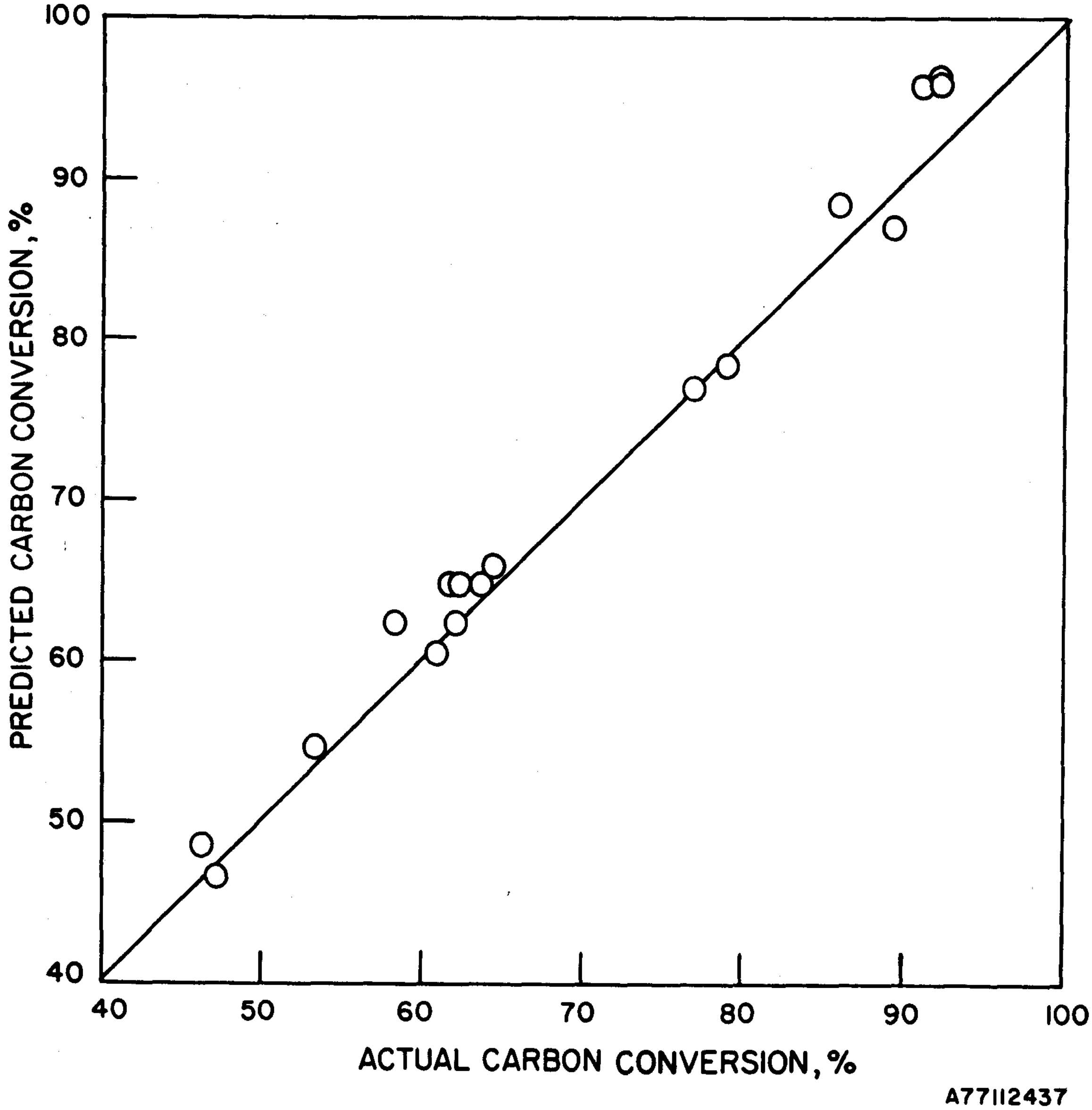

Figure 44. COMPARISON OF ACTUAL VS. PREDICTED CARBON CONVERSION — PDU DATA (6-Inch-Diameter PDU Steam-Oxygen Gasification Tests; Feed Was Project COED Char From Illinois No. 6 Coal)

Table 16. COMPARISON OF ACTUAL VS. PREDICTED GAS YIELDS — PDU DATA

Run No.	Moles of Methane Formed/ Mole of Carbon in Feed		Moles of Carbon Oxides Formed/ Mole of Carbon in Feed	
	Actual	Predicted *	Actual	Predicted
21	0.042	0.037	0.726	0.734
23	0.040	0.040	0.570	0.565
24	0.029	0.032	0.442	0.436
26	0.033	0.038	0.591	0.610
27	0.029	0.034	0.434	0.452
29	0.053	0.054	0.840	0.817
30	0.045	0.040	0.599	0.618
31	0.041	0.043	0.597	0.631
34	0.029	0.032	0.367	0.381
39	0.049	0.051	0.534	0.572
44	0.044	0.048	0.575	0.602
45	0.044	0.049	0.490	0.496
49	0.054	0.053	0.736	0.730
51	0.044	0.050	0.580	0.573
55	0.082	0.086	0.827	0.875
57	0.090	0.087	0.830	0.879
58	0.085	0.084	0.835	0.876
60	0.085	0.082	0.773	0.803

* The predicted methane has been adjusted to account for residual volatile matter present in COED char.

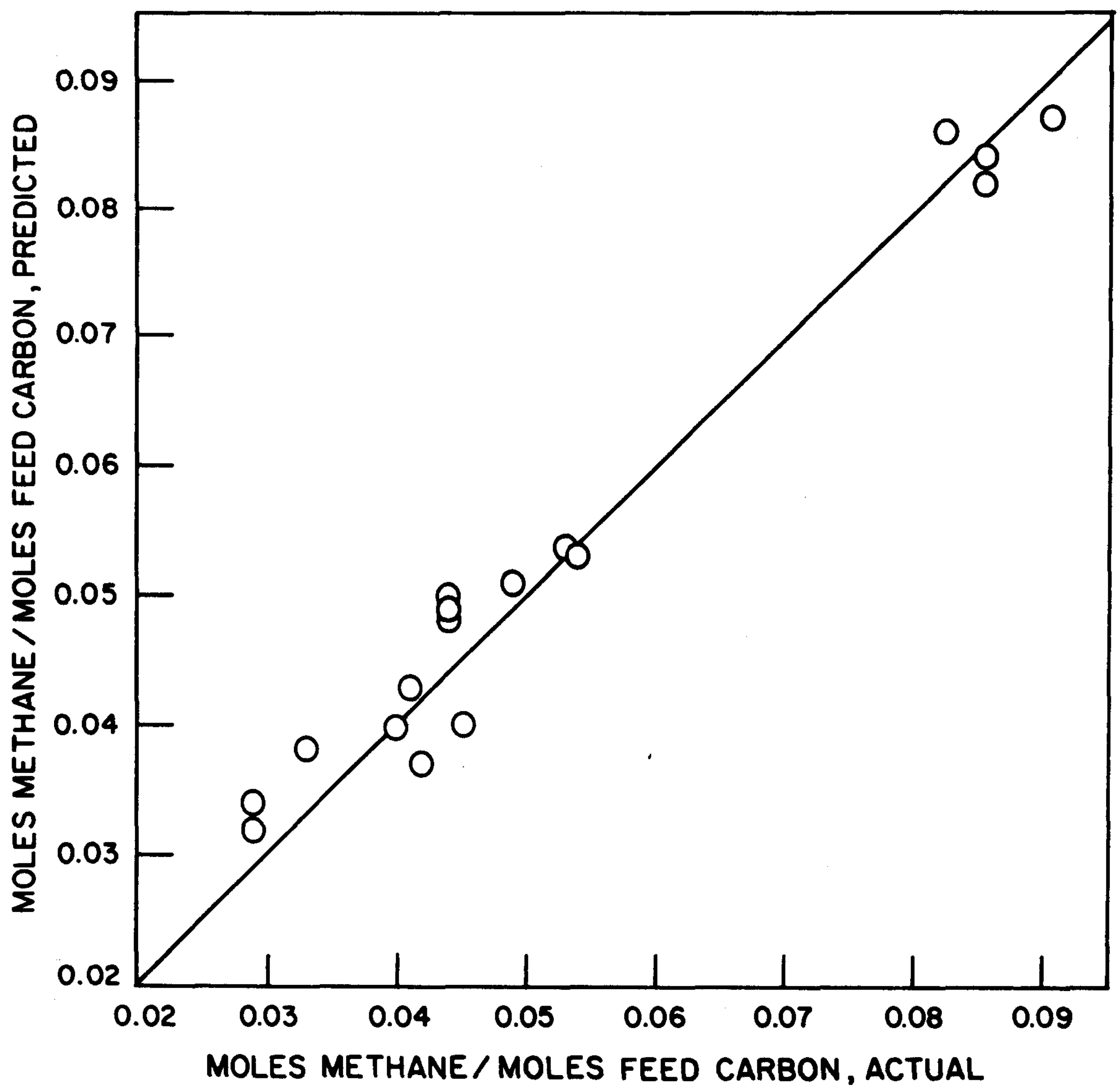

Figure 45. COMPARISON OF ACTUAL VS. PREDICTED METHANE YIELDS — PDU DATA

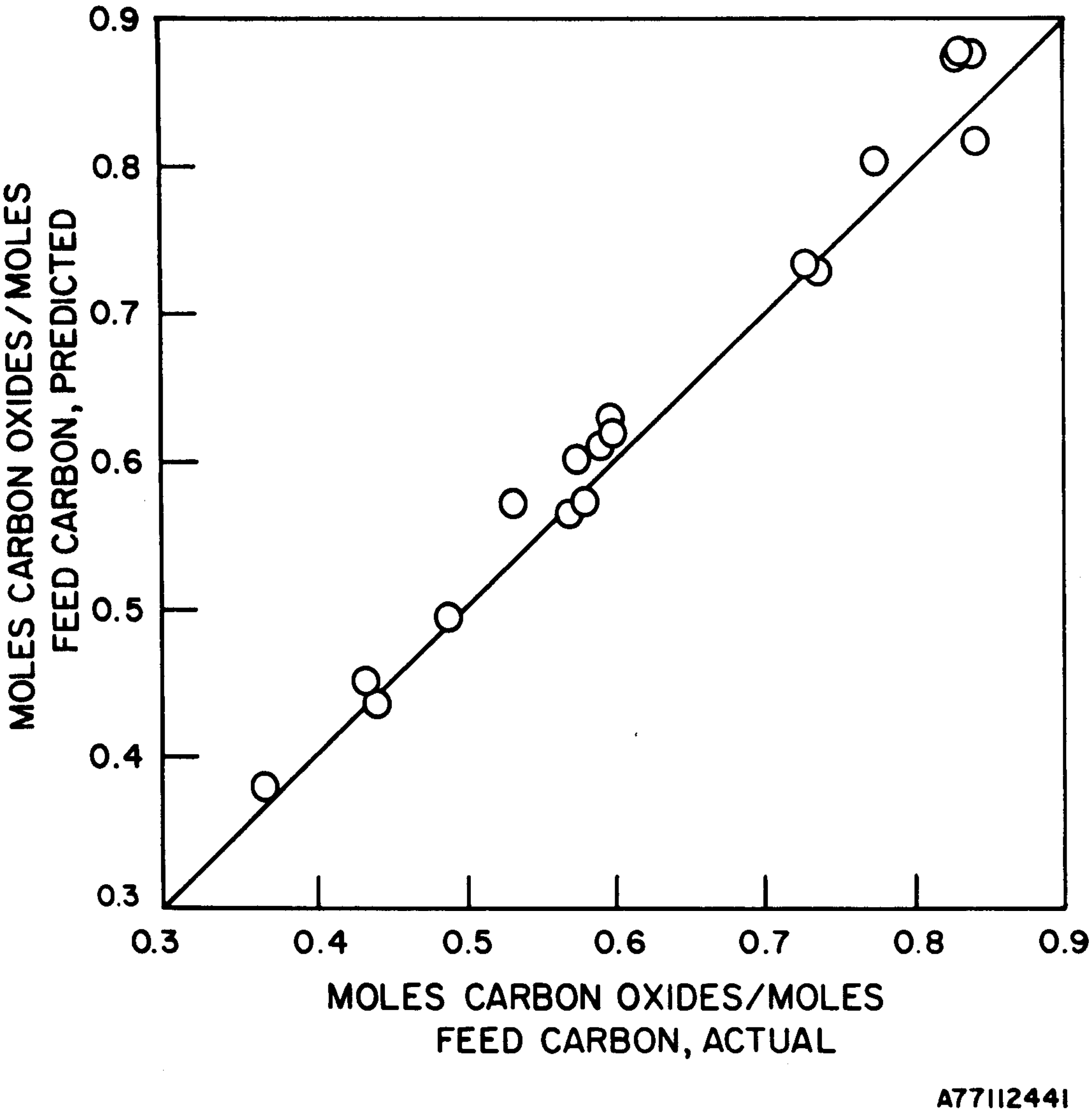

Figure 46. COMPARISON OF ACTUAL VS. PREDICTED CARBON OXIDES ($CO + CO_2$) YIELDS — PDU DATA

CONCLUSIONS

Two tests, Tests 64 and 65, were conducted during this reporting period. Test 64 was very successful with lengthy, smooth, steady-state operation of the reactor at char conversions ranging from about 80% for the major part of the test period to over 90%. Significant information was obtained by operating the reactor at these high char conversions. At the end of this quarter, the HYGAS plant was readied for Test 66.

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