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PIPELINE GAS FROM COAL-HYDROGENATION (IGT HYDROGASIFICATION PROCESS)

Project 9000 Quarterly Report No. 7, January 1—March 31, 1978

August 1978
Date Published

Work Performed Under Contract No. EX-76-C-01-2434

Institute of Gas Technology
IIT Center
Chicago, Illinois

U. S. DEPARTMENT OF ENERGY

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Prepared by
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IIT Center, 3424 S. State Street
Chicago, Illinois 60616

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SUMMARY

This quarterly report covers work performed between January 1 and March 31, 1978. The objective of this work was to conduct the necessary pilot plant operations and related support studies to acquire data for a commercial/demonstration plant based on the HYGAS® Process. Specifically, tests were conducted to obtain data on the operating conditions necessary for high char conversion at high throughputs, using Illinois No. 6 bituminous coal. Three tests, numbers 68, 69, and 70, were conducted. In addition, three major modifications were made to the HYGAS reactor to improve the operation of the steam-oxygen gasifier and to decrease fines loss from the reactor. The results of Test 67, conducted during November 1977, are also presented here.

Forty-five tons of char were fed to the HYGAS reactor during Test 68 conducted in late December and early January. A leak in Manway 0 forced early termination of this test. One-hundred-eighteen tons of char were fed to the reactor during Test 69 before a lack of high-pressure nitrogen forced termination of the test. The supply shortage was the direct result of a severe winter storm in Chicago which tied-up motor transport; therefore, the necessary delivery of liquid nitrogen could not be made.

The reactor operated very well during Test 70, and 279 tons of pretreated char were fed to the reactor during this test. A 3 ton/hr feed rate was achieved for 46 hours. The failure of a quench-water circulation pump forced the test to be terminated.

After Test 70, the Institute of Gas Technology (IGT) made three modifications to the plant. A new, six-nozzle, steam-oxygen sparger design was installed, the 339 valve was relocated, and the installation of double-screening equipment in the coal feed area was completed. Other plant turnaround activities were also conducted during the time required for the above modifications.

HYGAS personnel continued to supply Procon, Inc., with data to aid them in their design of a commercial/distribution HYGAS plant. IGT also made suggestions to Procon concerning the commercial plant reactor design.

In order to further achieve the objective of acquiring the data for a commercial/demonstration HYGAS plant, construction work continued on a cold flow model of the low-temperature transport stage of the commercial gasification reactor. Tests were conducted in a small plastic model to evaluate solids

feeding devices for the low-temperature reactor section of this gasifier. Four solids feeding devices were tested.: two lift-pot configurations, an L-valve, and a reverse-seal leg. Details of these tests are given in this report.

A preliminary engineering study was made of a hot-liquid quench system to be added to the HYGAS pilot plant. In addition, routine support studies were conducted during the quarter.

INTRODUCTION

This report covers work conducted between January 1 and March 31, 1978, under the Department of Energy (DOE) Contract No. EF-77-C-01-2434 for the purpose of performing the necessary pilot plant operations and related support studies to acquire data for a commercial/demonstration plant based on the HYGAS Process.

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

The extension of this contract began July 1, 1977. Its objective involves the completion of tasks 7 through 9, which are detailed in the body of this report.

ACHIEVEMENTS

Task 7. Pilot Plant Experimental Operation

Test 66

The results of Test 66, conducted in October 1977, are discussed in detail in the Project 9000 Quarterly Report No. 6 (DOE Report No. FE-2434-25). HYGAS reactor data for this test are presented in Table 1 and Figures 1 through 11.

Test 67

Test 67 was performed during November 1977, and the post-run inspection was conducted during December 1977. Over 14 days of self-sustained operation were achieved, including 6 hours of operation at a 3 ton/hr feed rate and at char conversion rates exceeding 85%. The test was terminated because of problems in the transfer of char from the second-stage gasifier to the steam-oxygen gasifier. The results of Test 67 are also discussed in detail in the Project 9000 Quarterly Report No. 6 (DOE Report No. FE-2434-25).

Analyses of the data from Test 67 were completed during this quarter. The results are shown in Figures 12 through 22 and Tables 2 through 4.

Test 68

The initial light-off for Test 68 occurred on December 22, 1977. A hot pressure test on December 24 revealed leaks in the reactor. These were fixed, and the reactor was relit on December 25. Char feed to the reactor began on December 31 at 1345 hours, and the reactor operation became self-sustaining at 0445 hours on January 1.

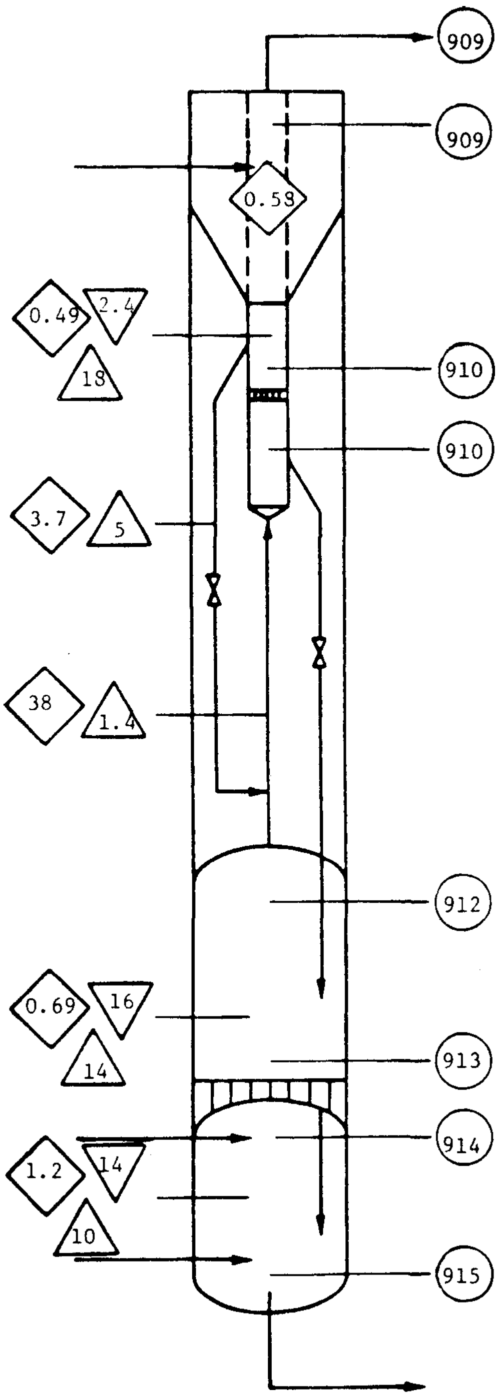
Test 68 was terminated at 1330 hours on January 3 due to a leak on Manway O of the reactor. During Test 68, problems were experienced in obtaining a proper mixing of the pretreater char-oil slurry in the slurry mix tank. High-density slurry plugged the low-pressure slurry circulation loop, and temporarily interrupted char feed to the reactor. During this test, 45 tons of char were fed to the reactor over a 20-hour period.

Pretreater operation during Test 68 started at 1400 hours on December 30. Eighty-five tons of coal were processed through the pretreater. Post-run inspection of the pretreater indicated that it and the char cooler were in good condition. Some tar-like material was found in the venturi scrubber, and the quench tower bottom liquid line contained some solids and tar.

Table 1. MATERIAL BALANCE SUMMARY FOR HYGAS GASIFIER FOR TEST 66 FOR STEADY PERIOD FROM 10/15/77 (0700 Hours) TO 10/16/77 (1000 Hours)

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

INPUT		C	H	O	N	S	ASH	TOTAL
Coal Feed	Wt % (Dry)	67.03	3.43	8.37	1.42	4.74	15.01	100
	Coal (Dry)	2806	144	350	60	198	628	4186
	Moisture		9	73				82
Sparger	Oxygen			1012				1012
	Steam		786	6234				7020
Burner	Oxygen			0				0
	Steam		0	0				0
	Hydrogen		0					0
Stripping Ring	Steam		263	2083				2346
Nitrogen From Purges					463			463
Pump Seal Flush			74	593				667
Water to Cyclone Pot			496	3938				4434
Light Oil In		10,280	970					11,250
TOTAL INPUT		13,086	2742	14,283	523	198	628	31,460
OUTPUT								
Reactor Overhead	Wt % (Dry)	72.80	2.69	5.07	1.34	2.83	15.36	100
	Dust (Dry)	738	27	51	13	29	156	1014
Spent Char	Wt % (Dry)	47.24	0.78	0.23	0.28	0.66	50.88	100
	Char (Dry)	462	7	2	3	6	497	977
Product Gas After Quench	Total (Dry)	1422	280	2385	468	20		4575
	Components H_2		140					140
	CO_2	782		2087				2869
	C_2H_5	0	0					0
	H_2S		1			20		21
	N_2				468			468
	CH_4	416	139					555
	CO	224		298				522
Water Out + Dissolved Materials		26	1428	11,313	19	40		12,826
Toluene Storage Tank Vent Gases		195	15	416	20	20		666
Stripper Vent Gas		64	7	116	27	1		215
Light-Oil Out		9978	941					10,919
Estimated Oil Losses		--	--					--
TOTAL OUTPUT		12,885	2705	14,283	550	116	653	31,192
Net (Output - Input)		−201	−37	0	27	−82	25	−262
% Balance (Output/Input)		98	99	100	105	59	104	99



○ 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) = 0.49
lb STEAM/lb CARBON (net) = 4.5
lb OXYGEN/lb COAL FED = 0.24
lb STEAM/lb COAL FED = 2.2
lb COAL FED/1000 SCF REACTOR PRODUCT GAS = 85

BY ASH BALANCE

MAF† COAL GASIFIED, % = 83
CARBON GASIFIED, % = 79

METHANE YIELD, SCF/lb COAL FED = 3.4

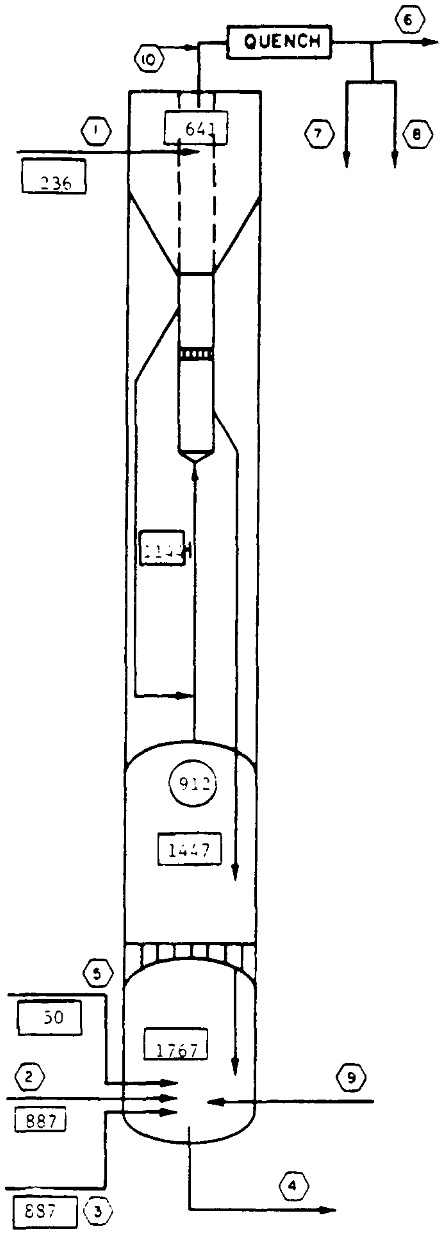
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 5.5

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 12
SOG = 14

†MOISTURE ASH FREE

Figure 1. HYGAS REACTOR ENGINEERING DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/15/77 (0700 Hours) TO 10/16/77 (1000 Hours)



○ STREAM No.
○ PRESSURE, psig
□ TEMPERATURE, °F

MATERIAL BALANCE, %

C	98
H	99
O	100
ASH	104

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	34.01	31.43
CO_2	31.67	35.26
C_2H_6	0	0.09
N_2	8.12	7.96
H_2S	0.30	0.54
CH_4	16.85	16.34
CO	9.05	8.33
mol/hr (dry)	206	231

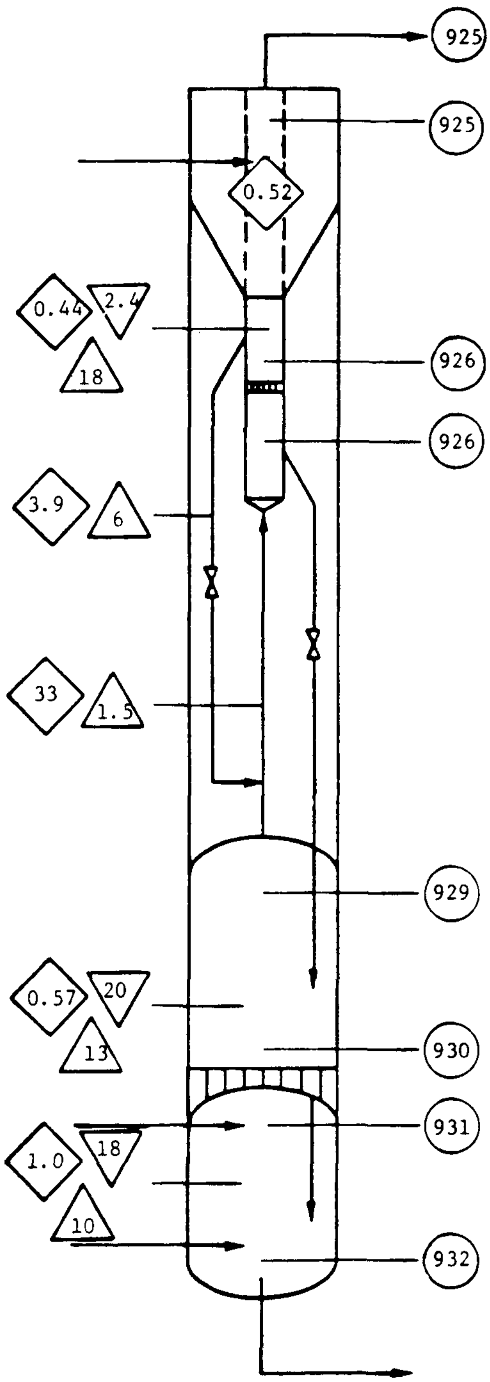
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	67.03	72.80	47.24
H	3.43	2.69	0.78
N	1.42	1.34	0.28
S	4.74	2.83	0.66
ASH	15.01	15.36	50.88
O	8.37	5.07	0.23
lb/hr (dry)	4186	1014	977
MOISTURE, wt %	1.9		
SLURRY CONCN, wt %	27		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	7643	--	--
LIGHT OIL	--	10,919	11,250

GAS FEED STREAMS			
2	3	5	9
mol/hr			
390	130	32	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 2. HYGAS REACTOR DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/15/77 (0700 Hours) TO 10/16/77 (1000 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) = *
lb STEAM/lb CARBON (net) = *
lb OXYGEN/lb COAL FED = 0.20
lb STEAM/lb COAL FED = 1.93
lb COAL FED/1000 SCF REACTOR PRODUCT GAS = 115

BY ASH BALANCE

MAF† COAL GASIFIED, % = 62
CARBON GASIFIED, % = 54

METHANE YIELD, SCF/lb COAL FED = 3.1

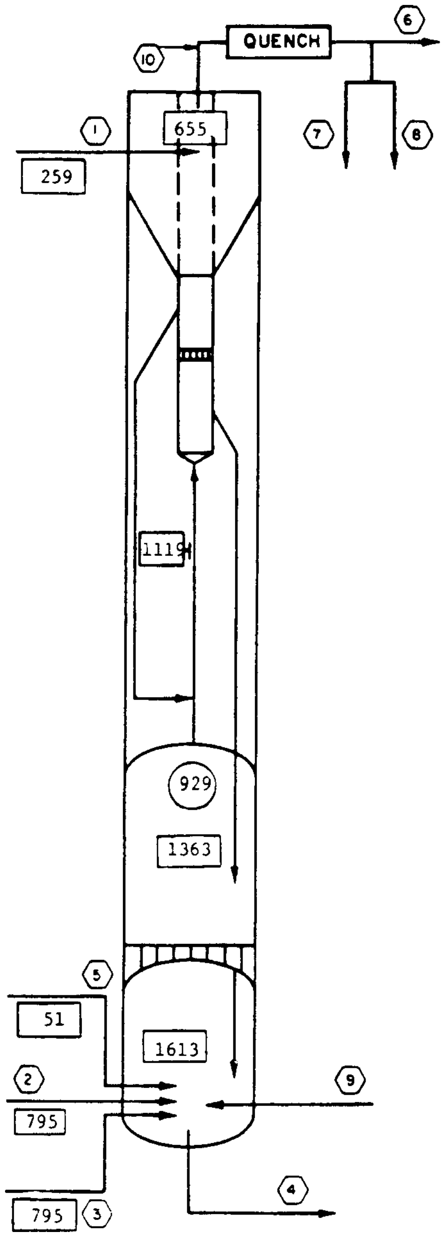
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 4.6

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 15
SOG = 22

†MOISTURE ASH FREE

Figure 3. HYGAS REACTOR ENGINEERING DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/6/77 (0600 Hours) TO 10/6/77 (1100 Hours)



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

MATERIAL BALANCE, %	
C	
H	
O	
ASH	

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	30.77	28.34
CO_2	30.65	34.62
C_2H_6	0.43	0.48
N_2	9.45	9.18
H_2S	1.38	1.44
CH_4	20.40	19.56
CO	6.92	6.38
mol/hr (dry)	161	181

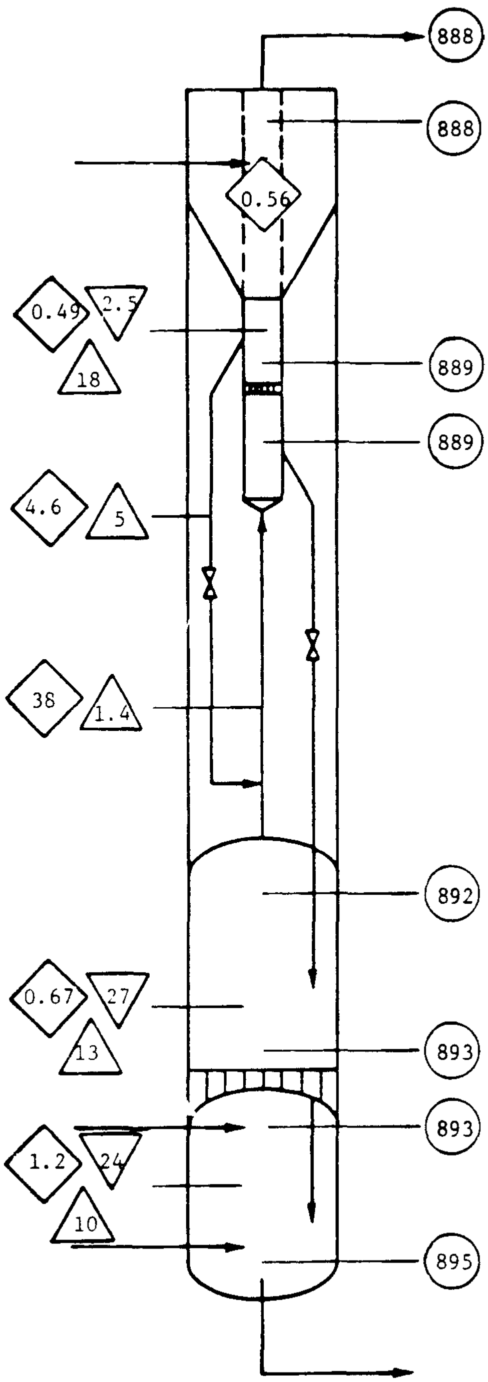
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	68.85	77.30	67.60
H	3.45	2.80	1.11
N	1.44	1.25	0.51
S	4.25	2.91	1.13
ASH	13.34	11.08	28.71
O	8.67	4.66	0.94
lb/hr (dry)	4327		
MOISTURE, wt %	2.5		
SLURRY CONCN, wt %	34		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	7418	--	--
LIGHT OIL	--	9054	8772

GAS FEED STREAMS			
2	3	5	9
mol/hr			
378	87	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 4. HYGAS REACTOR DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/6/77 (0600 Hours) TO 10/6/77 (1100 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) = *
lb STEAM / lb CARBON (net) = *
lb OXYGEN / lb COAL FED = 0.22
lb STEAM / lb COAL FED = 2.29
lb COAL FED / 1000 SCF REACTOR PRODUCT GAS = 128

BY ASH BALANCE

MAF† COAL GASIFIED, % = 68
CARBON GASIFIED, % = 62

METHANE YIELD, SCF/lb COAL FED = 2.5

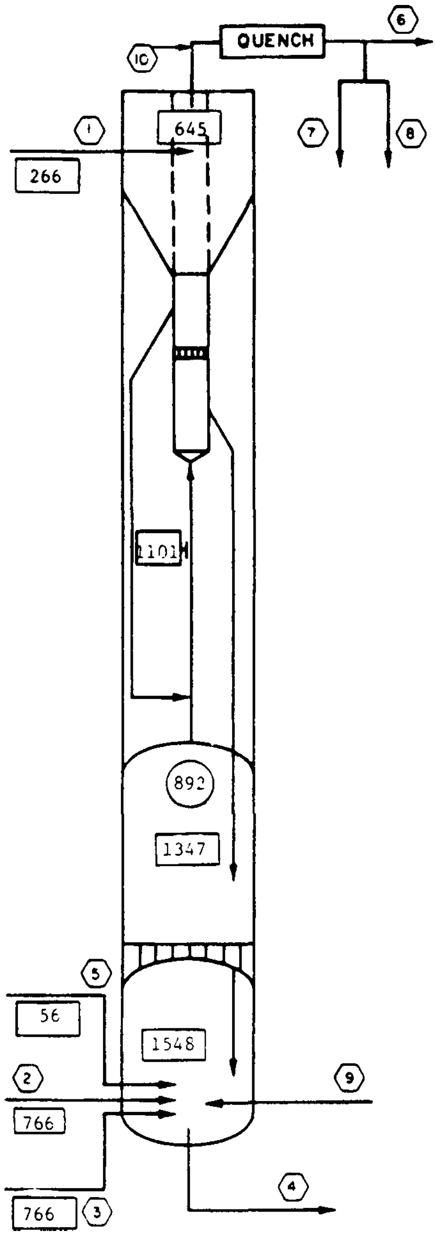
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 3.9

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 14
SOG = 20

†MOISTURE ASH FREE

Figure 5. HYGAS REACTOR ENGINEERING DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/6/77 (1600 Hours) TO 10/7/77 (0100 Hours)



○ STREAM No.
○ PRESSURE, psig
□ TEMPERATURE, °F

MATERIAL BALANCE, %

C
H
O
ASH

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	32.68	29.95
CO_2	31.84	35.80
C_2H_6	0.37	0.43
N_2	10.36	9.96
H_2S	1.21	1.30
CH_4	17.25	16.75
CO	6.29	5.81
mol/hr (dry)	142	160

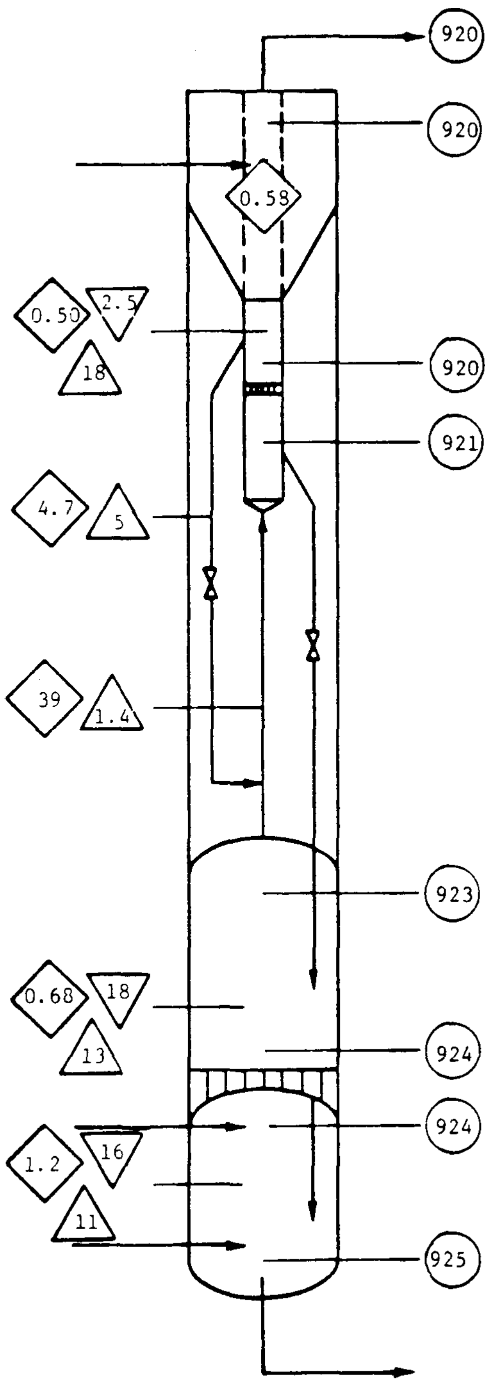
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	69.30	76.20	64.50
H	3.50	2.87	1.26
N	1.34	1.17	0.55
S	4.20	2.87	1.26
ASH	12.71	12.11	31.01
O	8.95	4.78	1.42
lb/hr (dry)	4123		
MOISTURE, wt %	2.6		
SLURRY CONCN, wt %	30		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	8723	--	--
LIGHT OIL	--	7905	9914

GAS FEED STREAMS			
2	3	5	9
mol/hr			
393	132	28	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 6. HYGAS REACTOR DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/6/77 (1600 Hours) TO 10/7/77 (0100 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) = *
lb STEAM / lb CARBON (net) = *
lb OXYGEN / lb COAL FED = 0.24
lb STEAM / lb COAL FED = 2.27
lb COAL FED / 1000 SCF REACTOR PRODUCT GAS = 90

BY ASH BALANCE

MAF† COAL GASIFIED, % = 76
CARBON GASIFIED, % = 71

METHANE YIELD, SCF/lb COAL FED = 3.4

EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 5.3

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 12
SOG = 15

†MOISTURE ASH FREE.

Figure 7. HYGAS REACTOR ENGINEERING DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/12/77 (1400 Hours) TO 10/12/77 (2400 Hours)

MATERIAL BALANCE, %

C

H

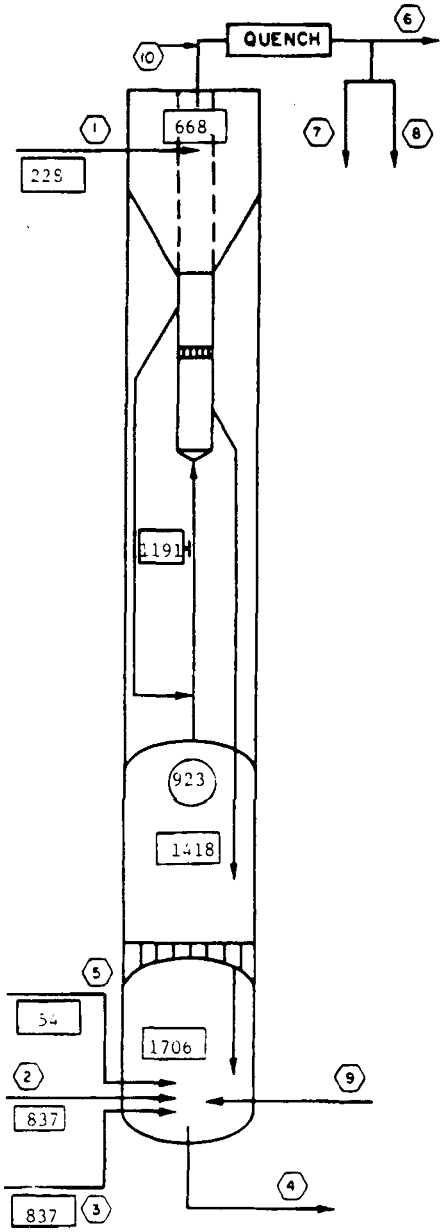
O

ASH

○ STREAM No.

○ PRESSURE, psig

□ TEMPERATURE, °F



GAS STREAMS

	6	10
mol % (dry)		
H_2	33.59	29.94
CO_2	32.72	37.94
C_2H_6	0	0.13
N_2	7.72	7.32
H_2S	0.85	1.04
CH_4	16.99	16.35
CO	8.13	7.29
mol/hr (dry)	195	227

SOLIDS STREAMS

	1	10	4
wt % (dry)			
C	69.30	73.65	58.55
H	3.47	2.78	0.95
N	1.36	1.08	0.51
S	4.35	2.91	0.83
ASH	13.16	14.74	38.76
O	8.36	4.84	0.40
lb/hr (dry)	4178		
MOISTURE, wt %	2.0		
SLURRY CONCN, wt %	29		

LIQUID STREAMS

	7	8	1
	lb/hr		
H_2O	7705	--	--
LIGHT OIL	--	10,049	10,292

GAS FEED STREAMS

2	3	5	9
mol/hr			
395	133	32	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

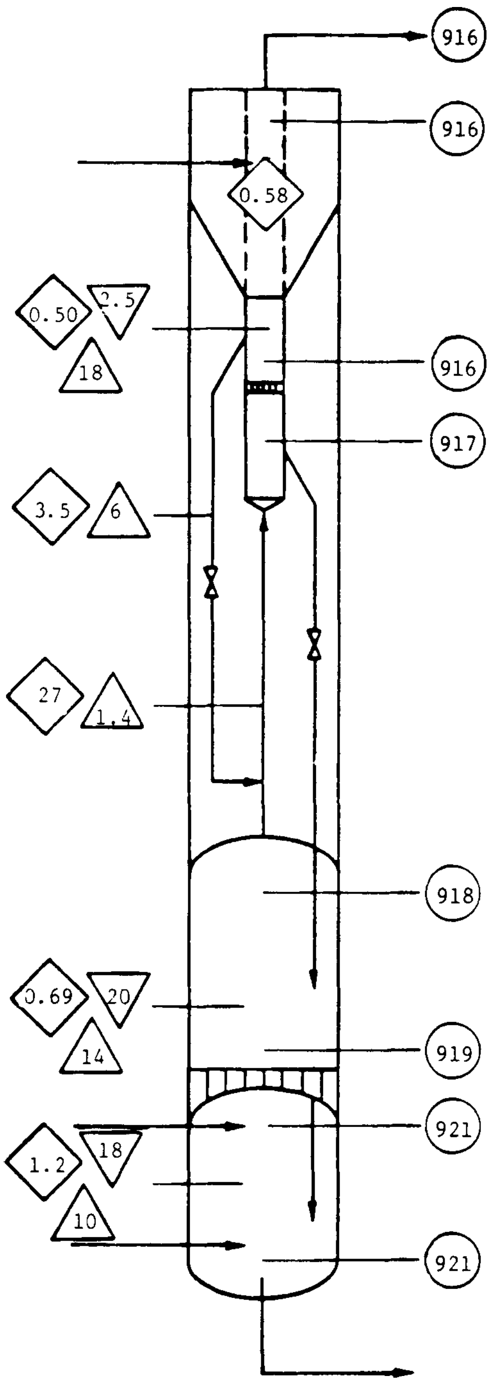
(7) WATER MADE

(8) LIGHT OIL TO RECYCLE

(9) GASES FROM EXTERNAL HEATER

(10) REACTOR OVERHEAD

Figure 8. HYGAS REACTOR DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/12/77 (1400 Hours) TO 10/12/77 (2400 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) = *
lb STEAM / lb CARBON (net) = *
lb OXYGEN / lb COAL FED = 0.25
lb STEAM / lb COAL FED = 2.28
lb COAL FED / 1000 SCF REACTOR PRODUCT GAS = 87

BY ASH BALANCE

MAF† COAL GASIFIED, % = 90
CARBON GASIFIED, % = 88

METHANE YIELD, SCF/lb COAL FED = 3.3

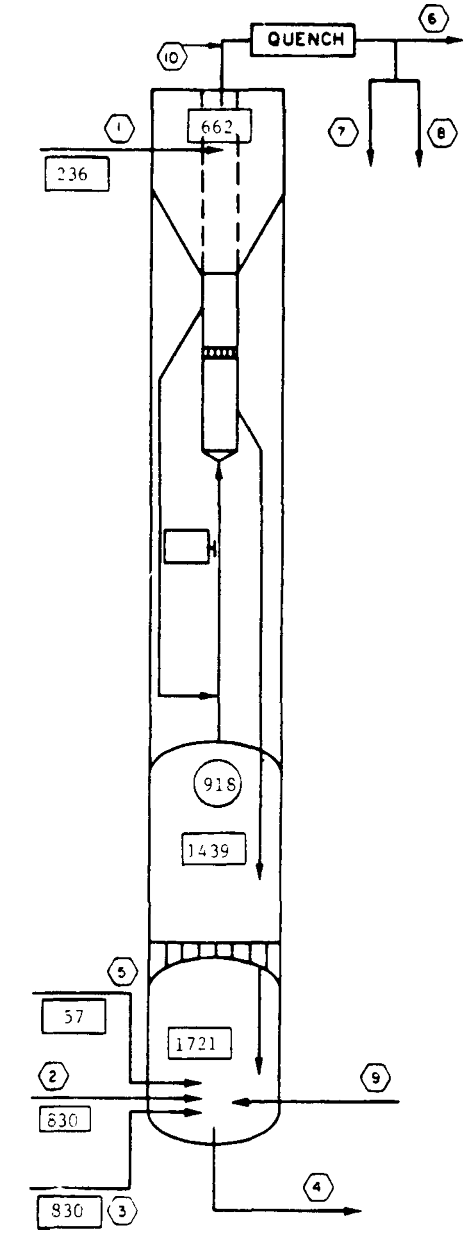
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 5.4

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 12
SOG = 16

†MOISTURE ASH FREE

Figure 9. HYGAS REACTOR ENGINEERING DATA FOR TEST 66 FOR STEADY PERIOD FROM 10/14/77 (0300 Hours) TO 10/14/77 (1400 Hours)



MATERIAL BALANCE, %	
C	
H	
O	
ASH	

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

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	33.36	30.37
CO_2	32.68	37.22
C_2H_6	0	0.10
N_2	7.86	8.02
H_2S	0.79	0.97
CH_4	16.66	15.73
CO	8.65	7.60
mol/hr (dry)	200	234

	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	68.13	72.83	35.30
H	3.47	2.75	0.83
N	1.37	1.11	0.35
S	4.42	2.93	1.21
ASH	14.27	15.54	62.21
O	8.34	4.84	0.10
lb/hr (dry)	4181		
MOISTURE, wt %	2.1		
SLURRY CONCN, wt %	29		

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	7697	--	--
LIGHT OIL	--	9861	10,318

GAS FEED STREAMS			
2	3	5	9
mol/hr			
396	133	32	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 66 FOR STEADY PERIOD FROM 10/14/77 (0300 Hours) TO 10/14/77 (1400 Hours)

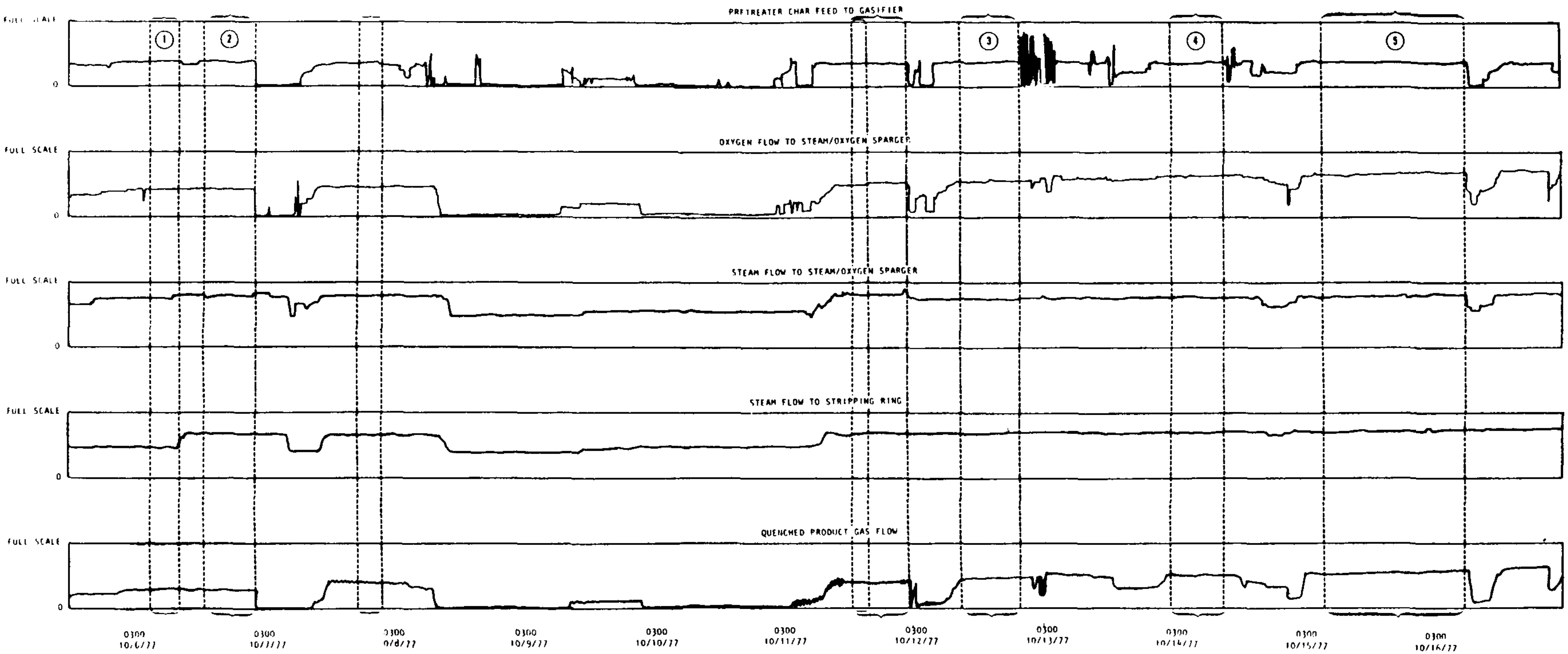


Figure 11. OVERVIEW OF REACTOR OPERATION FOR TEST 66

(Ed. note: Areas bounded by vertical dash-lines reflect periods of steady operation).

Table 2. MATERIAL BALANCE SUMMARY FOR HYGAS PRETREATER SECTION FOR TEST 67 FROM 11/17/77 (2100 Hours) TO 11/18/77 (1000 Hours)

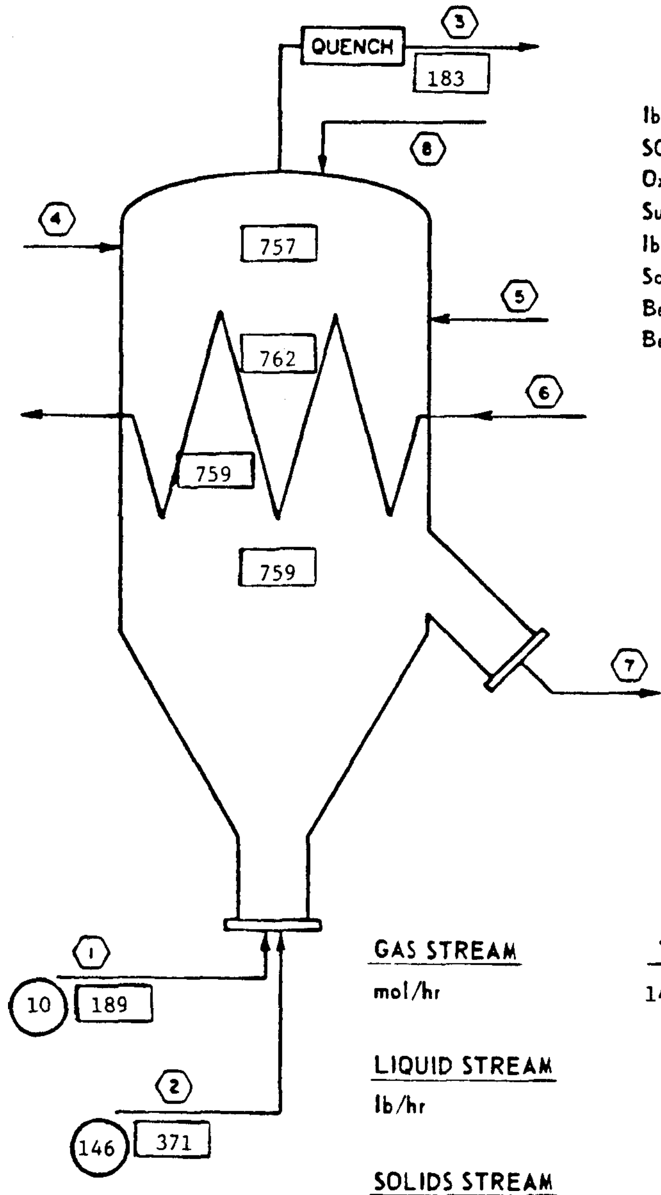
Basis = 1 hr. All units in lbs unless otherwise noted.

INPUT		C	H	O	N	S	Ar	Ash	Other	Total
Coal Feed	Wt % (Dry)	69.00	4.90	9.33	1.22	4.60		10.95		100
	Coal (Dry)	3623	257	490	64	242		575		5251
	Moisture		18	144						162
Streams to Pretreater	Air			992	3231		55			4278
	Steam		182	1444						1626
Nitrogen from purges					79					79
Air from purges				9	29					38
H_2O to venturi scrubber			1983	15736						17719
H_2O to quench tower			476	3776						4252
N_2 to char cooler					418					418
Cooling water to char cooler			108	854						962
TOTAL INPUT		3623	3024	23445	3821	242	55	575		34785
OUTPUT										
Pretreater Char	Wt.% (Dry)	69.50	3.47	8.54	1.40	4.09		13.00		100
	Char (Dry)	2888	144	355	58	170		540		4155
	Moisture		4	34						38
Slurry Waste from Quench	Wt.% (Dry)	65.35	2.80	11.89	1.20	3.84		14.92		100
	Solids (Dry)	152	7	28	3	9		35		234
	Tars & Oils	159	15	13	1	5				193
	H_2O & Dis. materials *	26	2473	19621	2	72				22194
Quench Tower Off-Gas	Total	236	364	3394	3979		55			8028
	Components: H_2		0							0
	CO_2	161		429						590
	C_2H_6	10	2							12
	N_2				3979					3979
	CH_4	16	5							21
	CO	49		65						114
	O_2			67						67
	Ar						55			55
	H_2O		357	2833						3190
TOTAL OUTPUT		3461	3007	23445	4043	256	55	575		34842
Net (Output – Input)		–162	–17	0	222	14	0	0		57
% Balance (Output/Input)		96	99	100	106	106	100	100		100

* Oxygen balance

STREAM No.
PRESSURE, psig
TEMPERATURE, °F

lb Char/lb Coal = 0.82
SCF O_2/lb Coal = 2.24
Oxygen Breakthrough, % fed = 7
Superficial Velocity, ft/s = 0.99
lb O_2 Reacted/lb Coal = 0.18
Solids Residence Time, hr = 1.84
Bed Density, lb/cu ft = 17
Bed Height, ft = 13



GAS STREAM	3
COMPONENTS, mol %	
H_2	0.05
CO_2	3.92
C_2H_6	0.12
O_2	1.12
N_2	41.49
Ar	--
CH_4	0.39
CO	1.19
H_2O	51.72
mol/hr	342
mol wt	23

GAS STREAM	1	2	8
mol/hr	148	90	68

LIQUID STREAM	5	6
lb/hr	0	1229

SOLIDS STREAM	4	7
lb/hr	5251	4155

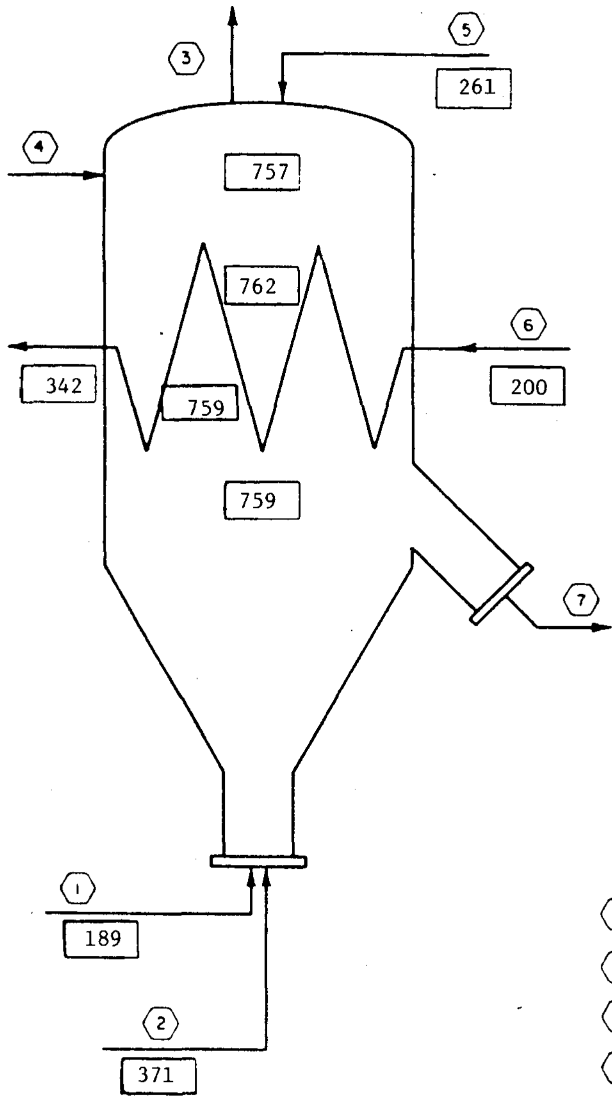
1 AIR TO PRETREATER
2 STEAM TO PRETREATER
3 PRETREATER OFF-GAS AFTER QUENCH
4 RAW COAL TO PRETREATER
5 DIRECT WATER QUENCH
6 COOLING WATER TO COOLING COIL IN PRETREATER
7 PRETREATER CHAR TO CHAR COOLER
8 GAS FROM CHAR COOLER

Figure 12. PRETREATMENT DATA FOR TEST 67 FOR STEADY PERIOD FROM 11/17/77 (2100 Hours) TO 11/18/77 (1000 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)	616,792
Heat of Combustion (Stream 4)	65,170,571
Steam Enthalpy (Streams 2 & 5)	2,880,000
Total	68,667,364
OUTPUT	
Sensible Heat (Streams 3 & 7)	3,657,936
Heat of Combustion (Streams 3 & 7)	55,838,607
Steam Enthalpy (Streams 3 & 6)	4,731,731
Total	64,228,274
% Balance	

(1) Air to Pretreater

(2) Steam to Pretreater

(3) Pretreater Overhead

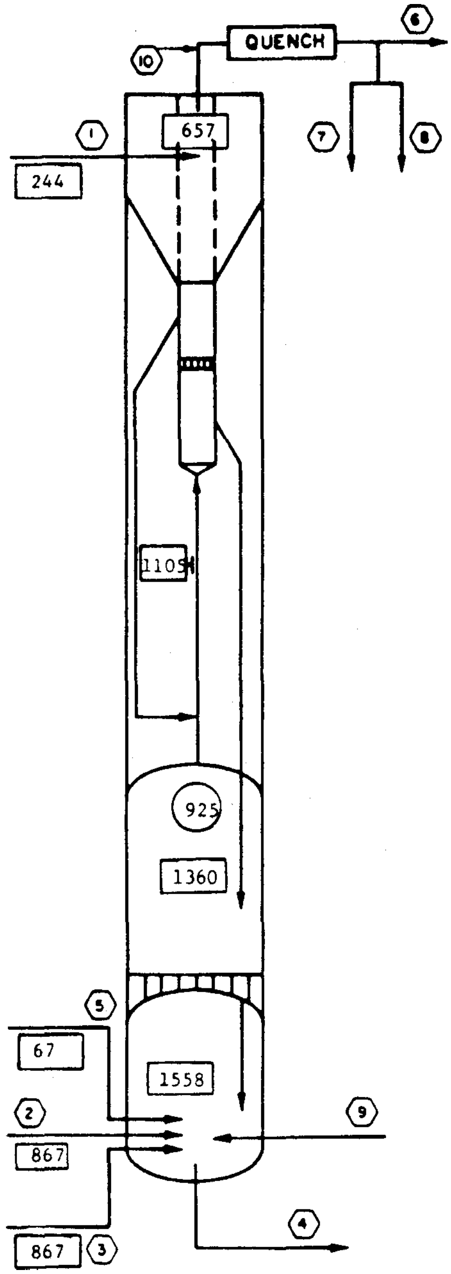
(4) Raw Coal to Pretreater

(5) Gas From Char Cooler

(6) Cooling Water to Cooling Coil in Pretreater

(7) Pretreated Char to Char Cooler

Figure 13. PRETREATER HEAT BALANCE DATA FOR TEST 67 FOR STEADY PERIOD FROM 11/17/77 (2100 Hours) TO 11/18/77 (1000 Hours)



MATERIAL BALANCE, %

C
H
O
ASH

○ STREAM No.
○ PRESSURE, psig
□ TEMPERATURE, °F

GAS STREAMS	6	10
mol % (dry)		
H_2	28.20	25.92
CO_2	33.60	37.46
C_2H_6	0.28	0.34
N_2	13.36	12.71
H_2S	0.82	1.06
CH_4	17.56	16.89
CO	6.18	5.60
mol/hr (dry)	150	169

SOLIDS STREAMS	1	10	4
wt % (dry)			
C	69.57	76.53	66.37
H	3.50	2.80	1.11
N	1.46	1.26	0.49
S	4.11	2.99	1.10
ASH	12.58	11.55	29.48
O	8.78	4.87	1.45
lb/hr (dry)	3986	*	*

FEED MOISTURE, wt % 1.4

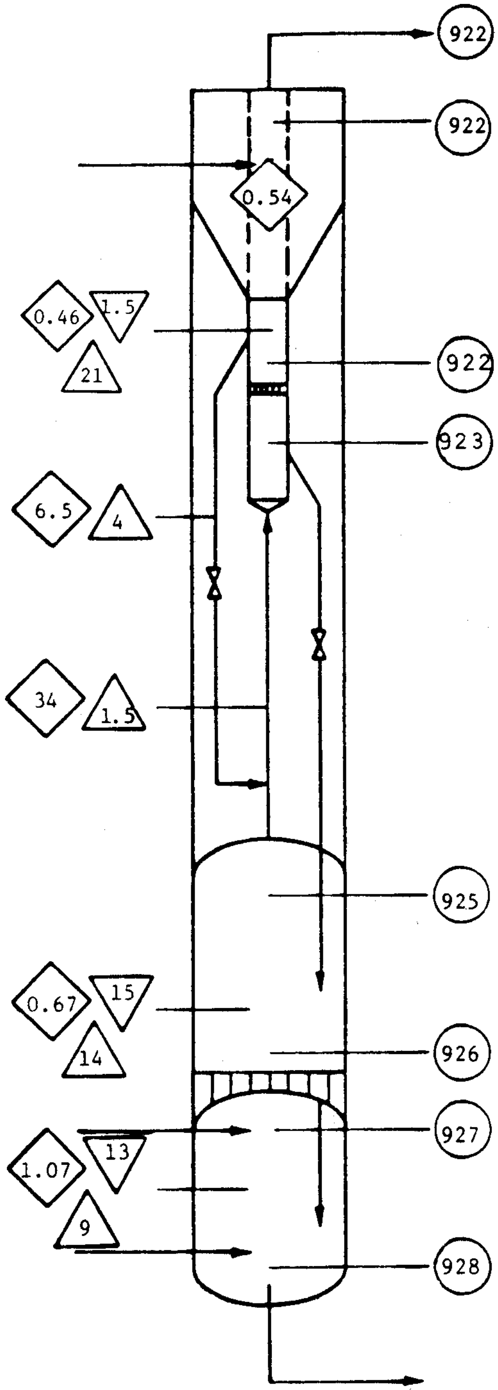
FEED SLURRY CONCN, wt % 30

LIQUID STREAMS	7	8	1
	lb/hr		
H_2O	8003	--	--
LIGHT OIL	--	9612	9221

GAS FEED STREAMS			
2	3	5	9
mol/hr			
385	122	25	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 67 FOR STEADY PERIOD FROM 11/8/77 (1500 Hours) TO 11/8/77 (2300 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) = *
lb STEAM/lb CARBON (net) = *
lb OXYGEN/lb COAL FED = 0.20
lb STEAM/lb COAL FED = 2.3
lb COAL FED/1000 SCF REACTOR PRODUCT GAS = 127

BY ASH BALANCE

MAF† COAL GASIFIED, % = 66
CARBON GASIFIED, % = 59

METHANE YIELD, SCF/lb COAL FED = 2.7

EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 4.1

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 12
SOG = 17

†MOISTURE ASH FREE

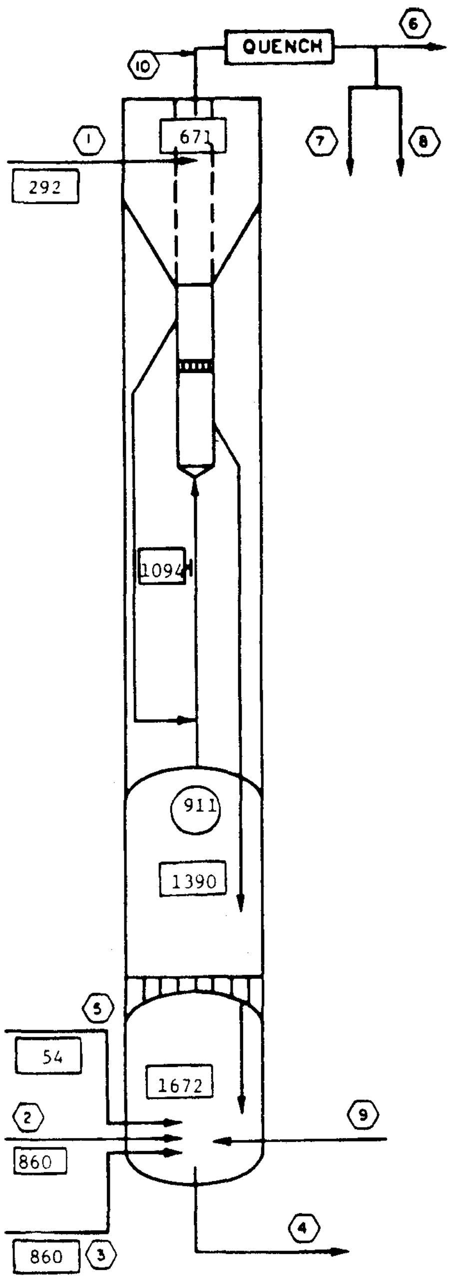
Figure 15. HYGAS REACTOR ENGINEERING DATA FOR TEST 67 FOR STEADY PERIOD FROM 11/8/77 (1500 Hours) TO 11/8/77 (2300 Hours)

Table 3. MATERIAL BALANCE SUMMARY FOR HYGAS GASIFIER FOR TEST 67 FROM 11/14/77 (2300 Hours) TO 11/15/77 (0700 Hours)

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

INPUT		C	H	O	N	S	ASH	TOTAL
Coal Feed	Wt % (Dry)	69.57	3.50	8.78	1.46	4.11	12.58	100
	Coal (Dry)	2950	149	372	62	174	534	4241
	Moisture		7	53				60
Sparger	Oxygen			859				859
	Steam		768	6099				6867
Burner	Oxygen			0				0
	Steam		0	0				0
	Hydrogen		0					0
Stripping Ring	Steam		247	1958				2205
Nitrogen From Purges					607			607
Pump Seal Flush			74	593				667
Water to Cyclone Pot			533	4232				4765
Light Oil In		8657	763					9420
TOTAL INPUT		11607	2541	14166	669	174	534	29691
OUTPUT								
Reactor Overhead	Wt % (Dry)	75.10	2.80	5.39	1.22	2.97	12.52	100
	Dust (Dry)	635	24	46	10	25	106	846
Spent Char	Wt % (Dry)	61.81	1.03	0.72	0.45	0.94	34.99	100
	Char (Dry)	756	13	9	6	11	428	1223
Product Gas After Quench	Total (Dry)	1208	234	2023	515	52		4032
	Components H_2		106					106
	CO_2	686		1829				2515
	C_2H_6	14	4					18
	H_2S		3			52		55
	N_2				515			515
	CH_4	363	121					484
	CO	145		194				339
Water Out + Dissolved Materials		12	1416	11664	1	29		13122
Toluene Storage Tank Vent Gases		170	12	357	17			556
Stripper Vent Gas		61	7	67	32			167
Light-Oil Out		8765	772					9537
Estimated Oil Losses		--	--					--
TOTAL OUTPUT		11607	2478	14166	581	117	534	29483
Net (Output - Input)		0	−63	0	−88	−57	0	−208
% Balance (Output/Input)		100	98	100	87	67	100	99

Ed. note: Due to operational problems, ash, carbon, and oxygen balances are forced.



MATERIAL BALANCE, %

C	100
H	98
O	100
ASH	100

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

GAS STREAMS

	6	10
mol % (dry)		
H_2	30.52	28.28
CO_2	33.05	36.37
C_2H_6	0.35	0.41
N_2	10.65	10.44
H_2S	0.94	0.84
CH_4	17.49	17.16
CO	7.00	6.50
mol/hr (dry)	173	193

SOLIDS STREAMS

	1	10	4
wt % (dry)			
C	69.57	75.10	61.81
H	3.50	2.80	1.03
N	1.46	1.22	0.45
S	4.11	2.97	0.94
ASH	12.58	12.52	34.99
O	8.78	5.39	0.72
lb/hr (dry)	4241	846	1223

FEED MOISTURE, wt % 1.4

FEED SLURRY CONCN, wt % 31

LIQUID STREAMS

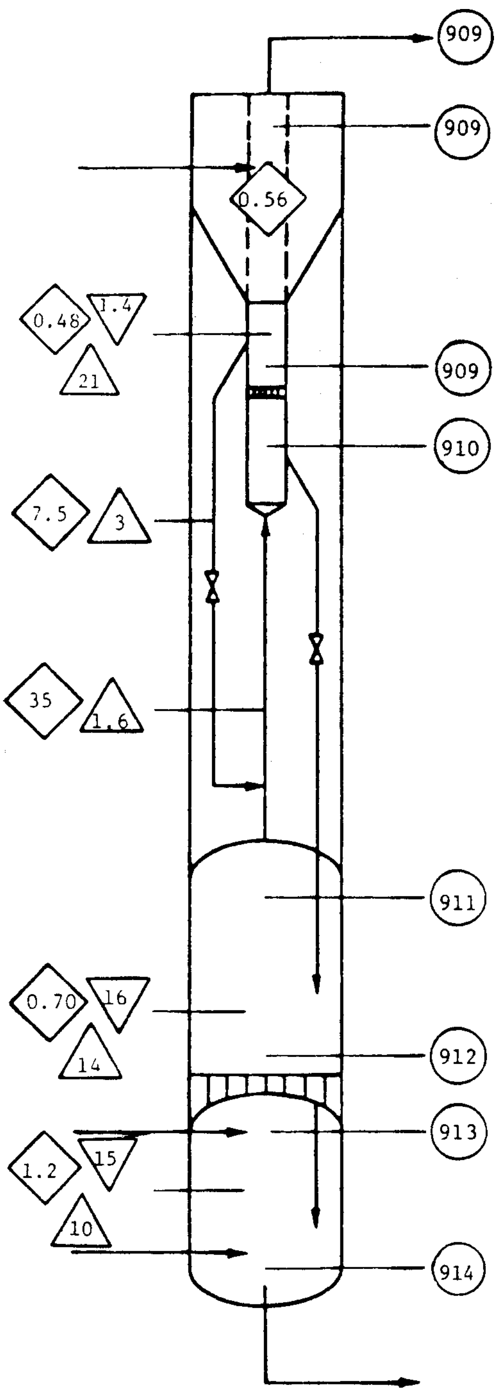
	7	8	1
	lb/hr		
H_2O	7690	--	--
LIGHT OIL	--	9537	9420

GAS FEED STREAMS

2	3	5	9
mol/hr			
382	123	27	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 67 FOR STEADY PERIOD FROM 11/14/77 (2300 Hours) TO 11/15/77 (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) = 0.37
lb STEAM / lb CARBON (net) = 3.9
lb OXYGEN / lb COAL FED = 0.20
lb STEAM / lb COAL FED = 2.1
lb COAL FED / 1000 SCF REACTOR PRODUCT GAS = 110

BY ASH BALANCE

MAF† COAL GASIFIED, % = 73
CARBON GASIFIED, % = 68

METHANE YIELD, SCF/lb COAL FED = 3.0

EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 4.6

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 11
SOG = 17

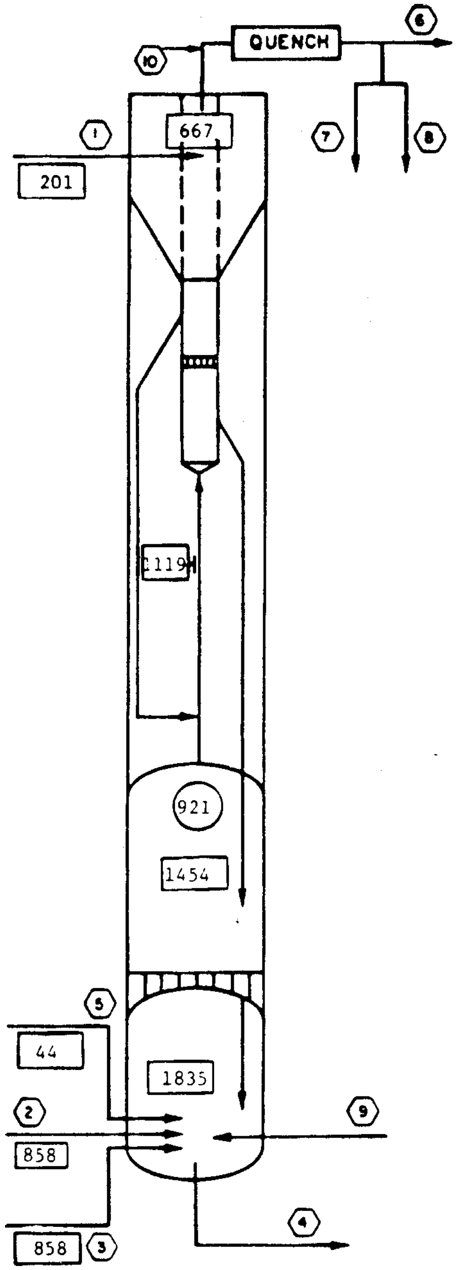
†MOISTURE ASH FREE.

Figure 17. HYGAS REACTOR ENGINEERING DATA FOR TEST 67 FOR STEADY PERIOD FROM 11/14/77 (2300 Hours) TO 11/15/77 (0700 Hours)

Table 4. MATERIAL BALANCE SUMMARY FOR HYGAS GASIFIER FOR TEST 67
11/17/77 (1400 Hours) TO 11/18/77 (0700 Hours)

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

INPUT			C	H	O	N	S	ASH	TOTAL
Coal Feed	Wt % (Dry)		69.57	3.50	8.78	1.46	4.11	12.58	100
	Coal (Dry)		3488	176	440	73	206	631	5014
	Moisture			8	63				71
Sparger	Oxygen				1076				1076
	Steam			790	6270				7060
Burner	Oxygen				0				0
	Steam			0	0				0
	Hydrogen			0					0
Stripping Ring	Steam			242	1923				2165
Nitrogen From Purges						641			641
Pump Seal Flush				74	593				667
Water to Cyclone Pot				408	3238				3646
Light Oil In			8922	848	-				9770
TOTAL INPUT			12410	2546	13603	714	206	631	30110
OUTPUT									
Reactor Overhead	Wt % (Dry)		74.86	2.86	4.74	1.26	2.82	13.46	100
	Dust (Dry)		827	32	52	14	31	149	1105
Spent Char	Wt % (Dry)		54.40	0.85	0.25	0.37	0.77	43.36	100
	Char (Dry)		621	10	3	4	9	495	1142
Product Gas After Quench	Total (Dry)		1699	336	2730	556	93		5414
	Components	H_2		153					153
		CO_2	885		2359				3244
		C_2H_6	22	6					28
		H_2S		6			93		99
		N_2				556			556
		CH_4	513	171					684
		CO	279		371				650
Water Out + Dissolved Materials			13	1323	10468	17	28		11849
Toluene Storage Tank Vent Gases			152	11	320	15			498
Stripper Vent Gas			52	6	57	27			142
Light-Oil Out			9046	860					9906
Estimated Oil Losses			--	--					--
TOTAL OUTPUT			12410	2578	13630	633	161	644	31033
Net (Output - Input)			0	32	27	−81	−45	13	923
% Balance (Output/Input)			100	101	100	89	78	102	103



○ STREAM No.
⬡ PRESSURE, psig
□ TEMPERATURE, °F

MATERIAL BALANCE, %

C	100
H	101
O	100
ASH	102

GAS STREAMS

	6	10
mol % (dry)		
H_2	31.90	30.31
CO_2	30.73	33.11
C_2H_6	0.39	0.43
N_2	8.28	8.29
H_2S	1.21	1.13
CH_4	17.81	17.57
CO	9.68	9.17
mol/hr (dry)	240	257

SOLIDS STREAMS

	1	10	4
wt % (dry)			
C	69.57	74.86	54.40
H	3.50	2.86	0.85
N	1.46	1.26	0.37
S	4.11	2.82	0.77
ASH	12.58	13.46	43.36
O	8.78	4.74	0.25
lb/hr (dry)	5014	1105	1142

FEED MOISTURE, wt % 1.4

FEED SLURRY CONCN, wt % 34

LIQUID STREAMS

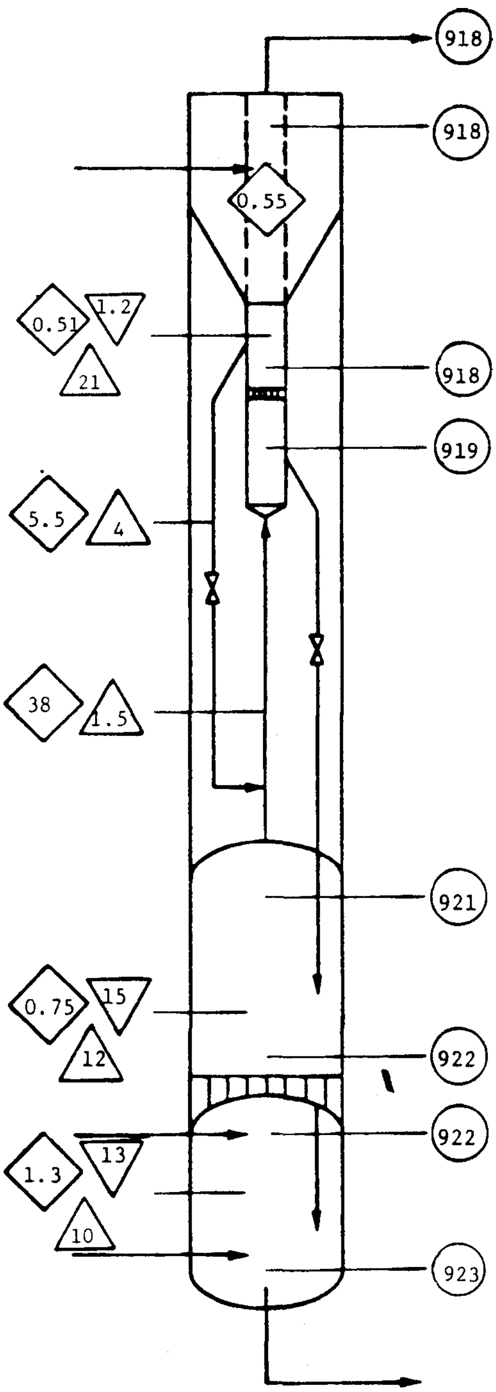
	7	8	1
	lb/hr		
H_2O	7536	--	--
LIGHT OIL	--	9906	9770

GAS FEED STREAMS

2	3	5	9
mol/hr			
392	120	34	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 67 FOR STEADY PERIOD FROM 11/17/77 (1400 Hours) TO 11/18/77 (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) = 0.40
lb STEAM/ lb CARBON (net) = 3.47
lb OXYGEN / lb COAL FED = 0.21
lb STEAM / lb COAL FED = 1.84
lb COAL FED / 1000 SCF REACTOR PRODUCT GAS = 89

BY ASH BALANCE

MAF† COAL GASIFIED, % = 81
CARBON GASIFIED, % = 77

METHANE YIELD, SCF/lb COAL FED = 3.42

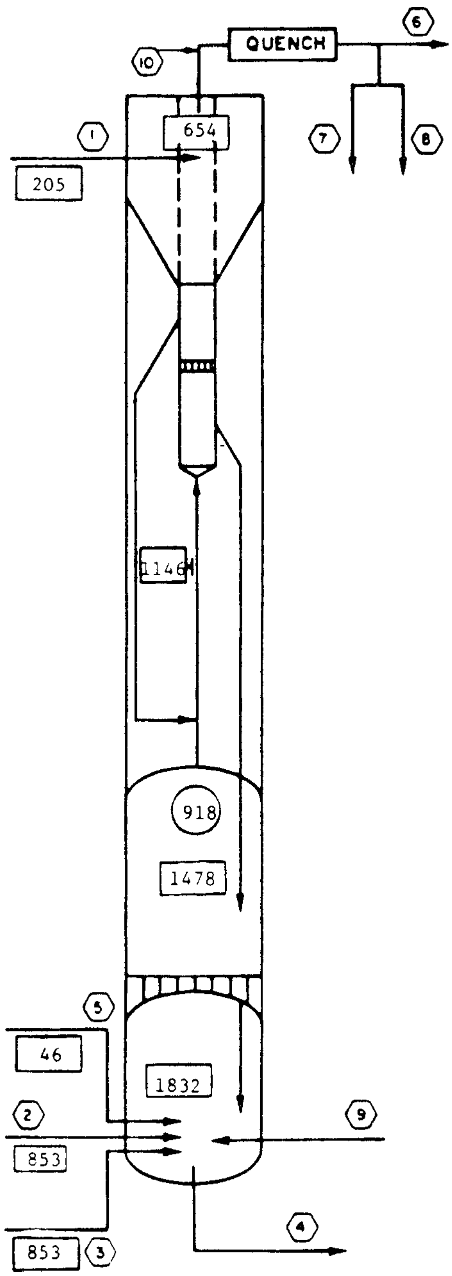
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 5.48

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 12
SOG = 14

†MOISTURE ASH FREE

Figure 19. HYGAS REACTOR ENGINEERING DATA FOR TEST 67 FOR STEADY PERIOD FROM 11/17/77 (1400 Hours) TO 11/18/77 (0700 Hours)



MATERIAL BALANCE, %
C
H
O
ASH

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

	GAS STREAMS	
	6	10
mol % (dry)		
H_2	30.42	27.68
CO_2	30.66	34.90
C_2H_6	0.44	0.50
N_2	7.68	7.63
H_2S	1.15	1.01
CH_4	18.48	17.93
CO	11.17	10.36
mol/hr (dry)	270	307

	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	69.57	75.00	59.85
H	3.50	2.86	0.90
N	1.46	1.27	0.38
S	4.11	3.04	0.82
ASH	12.58	12.20	38.12
O	8.78	5.63	0.07
lb/hr (dry)	5889	*	*

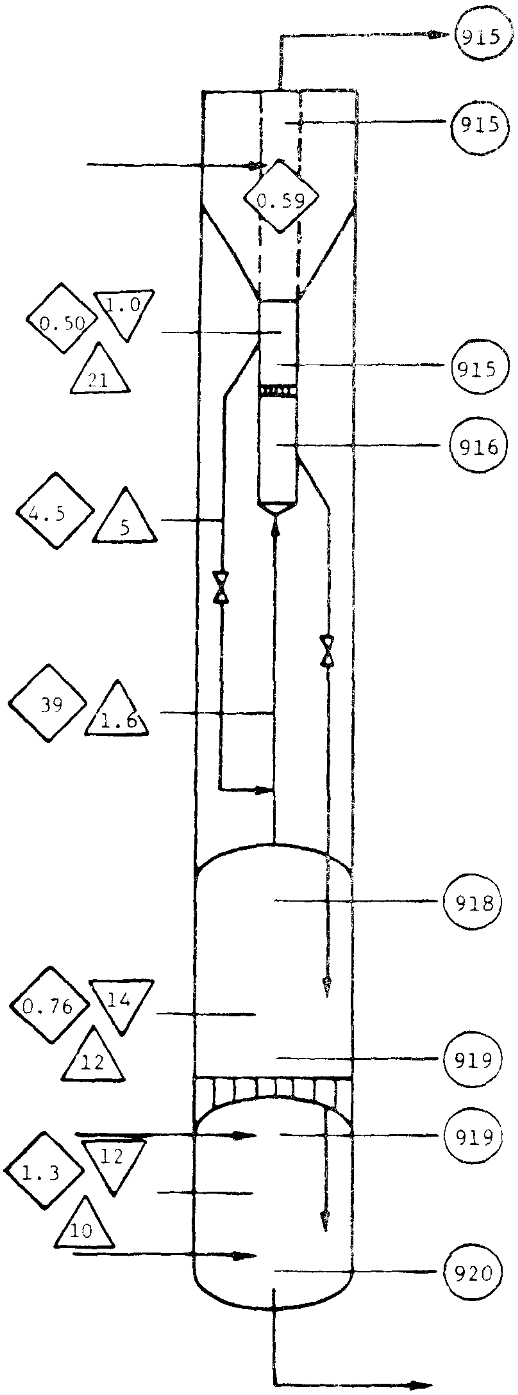
FEED MOISTURE, wt % 1.4
FEED SLURRY CONCN, wt % 35

	LIQUID STREAMS		
	7	8	1
	lb/hr		
H_2O	6560	--	--
LIGHT OIL	--	10,557	11,226

GAS FEED STREAMS			
2	3	5	9
mol/hr			
392	120	35	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 20. HYGAS REACTOR DATA FOR TEST 67 FOR STEADY PERIOD FROM 11/19/77 (0300 Hours) TO 11/19/77 (0900 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) = *
lb STEAM/lb CARBON (net) = *
lb OXYGEN/lb COAL FED = 0.19
lb STEAM/lb COAL FED = 1.57
lb COAL FED/1000 SCF REACTOR PRODUCT GAS = 89

BY ASH BALANCE

MAF† COAL GASIFIED, % = 77
CARBON GASIFIED, % = 72

METHANE YIELD, SCF/lb COAL FED = 3.5

EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 5.6

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 12
SOG = 14

†MOISTURE ASH FREE

Figure 21. HYGAS REACTOR ENGINEERING DATA FOR TEST 67 FOR STEADY PERIOD FROM 11/19/77 (0300 Hours) TO 11/19/77 (0900 Hours)

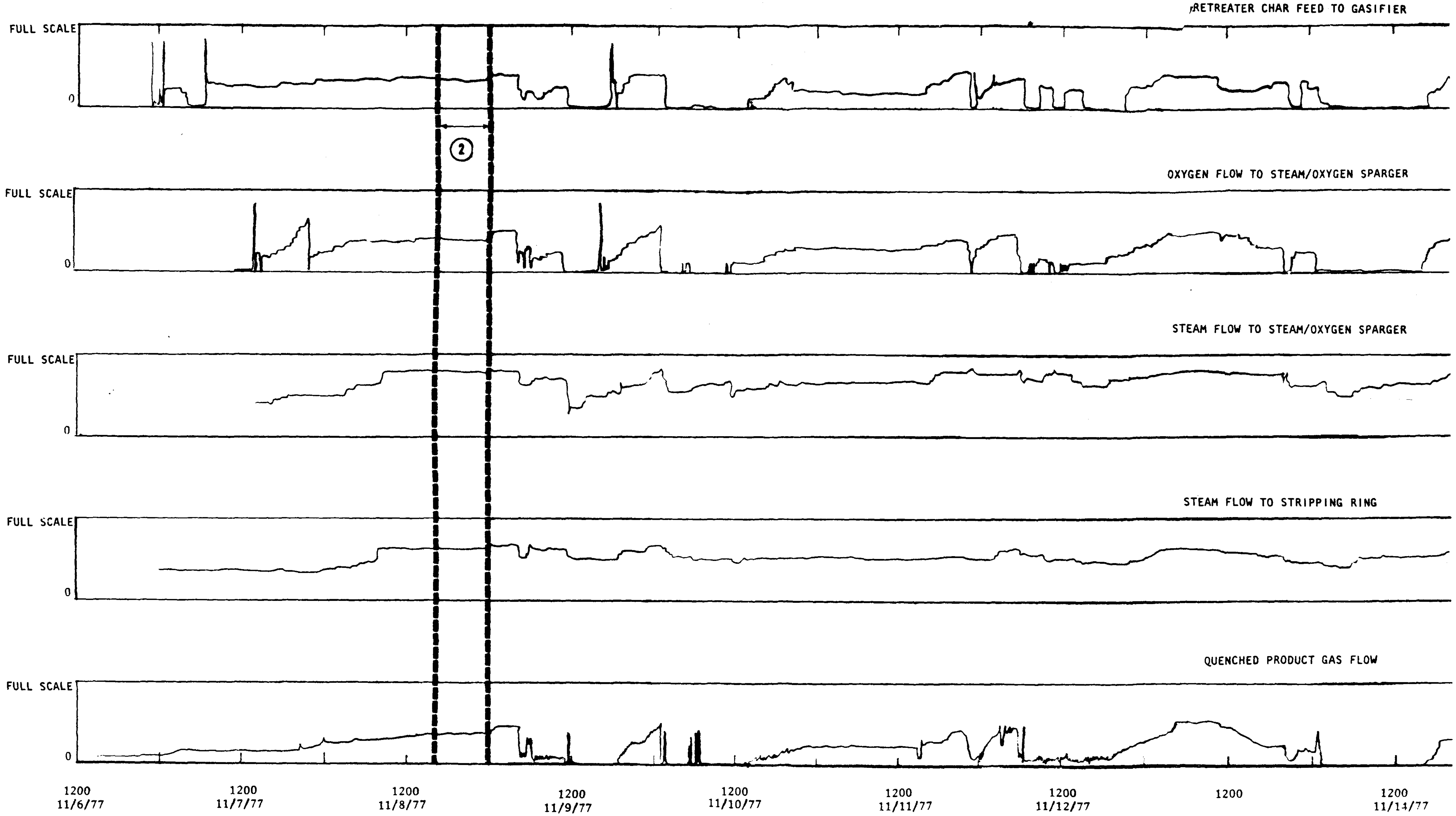
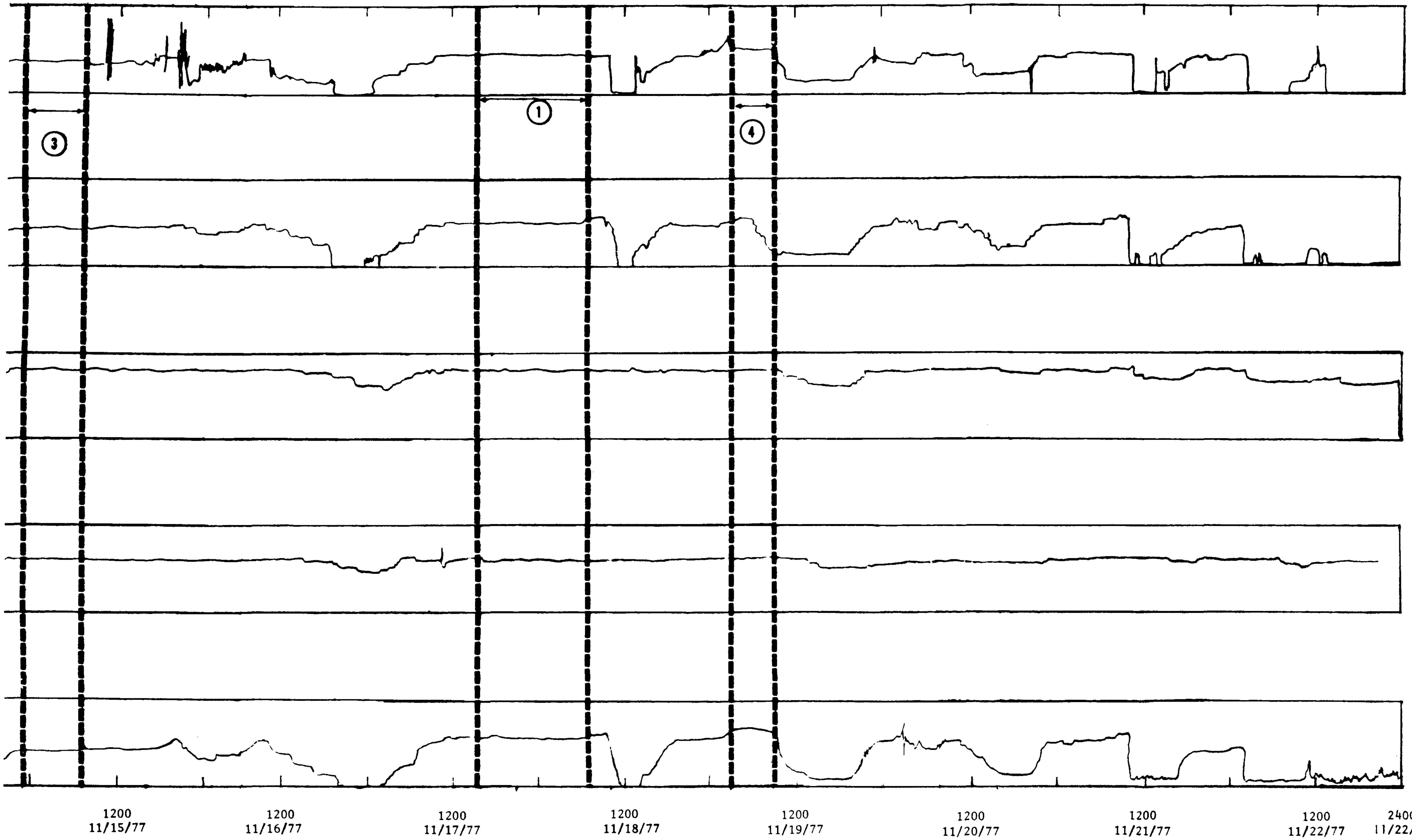


Figure 22. OVERALL REACTOR

(Ed. note: Areas bounded by vertical dash-lines reflect periods of steady operation.)



CONDITIONS FOR TEST 67

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When Manway 0 in the reactor was inspected after the test, it was found that a groove had been cut across the sealing surface on the east side of the vessel. The slurry dryer section was in good condition. Solids transfer Lines 321 and 322 were partially plugged with solids, but were readily cleared by blasting with nitrogen. The lift-line reactor and the high-temperature reactor were clean. A small, soft clinker was found lying on top of the steam-oxygen sparger in the steam-oxygen gasifier. It was believed that this 6-inch x 3-inch x 12-inch clinker fell from an area above the 339 valve, and had been formed during an earlier test. Line 339 was clear. The rest of the plant was also clean.

Due to the short duration of Test 68, no detailed analysis of operating results will be made.

Test 69

In preparation for Test 69, the coal mill speed was increased from 67 to slightly over 100 rpm to increase its crushing capacity. The pretreater section was cleaned and readied for Test 69. The highly concentrated slurry was removed from the slurry mix tank. Argonne National Laboratory installed two test meters in the low-pressure, slurry-circulation loop in preparation for Test 69. The reactor was cleaned, and the groove cut into the sealing surface of Manway 0 was repaired by Gray-Serv technicians. The reactor was reassembled and prepared for Test 69. The quench section was cleaned and readied for service. The purification section and the IGT fixed-bed catalyst methanation section were readied for operation in Test 69. At that time, the liquid-phase methanation pilot unit was still being modified. All utilities and the plant-effluent cleanup section were readied for Test 69.

Light-off for Test 69 occurred at 1130 hours on January 16; several previous attempts to light-off had been interrupted by electrical problems and instrument freezing. Char feed to the reactor was begun at 1700 hours on January 18. Test 69 was terminated at 2100 hours on January 26 due to a lack of high-pressure nitrogen for balancing the pressure in the HYGAS reactor and for instrument purges. The supply shortage was a direct result of a severe winter storm in Chicago that tied-up motor transport for the entire day of January 26. Prior to the forced termination of the test, conditions in the reactor had been stabilized at slightly over 2 tons of char feed per hour. More than 118 tons of char were fed to the reactor during Test 69.

In Test 69, the pretreater began operating at 2130 hours on January 18, and satisfactorily provided nonagglomerating char for the reactor feed. The pretreater feed system was interrupted several times by problems in operating the ball-valve in the lockhopper feeding system. The slurry preparation section operated satisfactorily for Test 69. Argonne National Laboratories' personnel operated their test slurry flowmeters during this test. The quench section operated well; however, the purification and methanation sections were not put on-stream. The effluent cleanup section was in service for Test 69. The utilities operated satisfactorily during the test. The hydrogen plant operated well, supplying reactor cooldown gases at the end of the test.

Post-run inspection for Test 69 was carried out in early February. The coal preparation section was found to be in satisfactory condition after the test, as were the pretreater reactor and char cooler. The gas and solids transfer lines in the pretreater section were all found to be clear. Small amounts of tar build-up were found in the venturi scrubber, and some tar-like coal material was found in the bottom of the pretreater quench tower. A crack was discovered on the bottom of one of the pretreater reactor cooling coils. This probably happened after the test when the vessel was opened and water, which had not completely drained from the coils, froze. Also, a leak was found on the top flange of another cooling coil.

The slurry preparation section was found to be in good condition. The HYGAS reactor was inspected after the test and found to be in satisfactory condition. The slurry dryer bed, the spouting bed, the lift-line, the 332 line, the second-stage reactor, the 339 line, and the steam-oxygen reactor were all found to be clean. A solid plug was found at the bottom of the 321 line at the L-valve, and a crack was found in the nitrogen-purge inlet-flange to the valve, which was subsequently repaired. Gray-Serv technicians remachined the sealing surface on Manway 0 which had been scratched. The reactor quench section and the effluent cleanup section were found to be in satisfactory condition.

Test 70

Following post-run inspection for Test 69, the plant was readied for Test 70. The coal preparation and pretreater sections were prepared. The reactor was buttoned up, and the quench section and effluent cleanup sections were cleaned and readied.

Light-off for Test 70 occurred at 2300 hours on February 11, 1978. Several problems upstream of the reactor delayed the start of char feed. In the pretreater section, a high-pressure drop across the fluidizing grid occurred. The pretreater was shut down, and all grid nozzles were checked; 6 were found to be partially plugged. Upon reassembly, the pressure drop across the grid was lowered to acceptable levels. Problems also developed in the lockhopper feed system as a result of a faulty level-control mechanism and erratic char cooler operation. These problems were solved.

Coal feed to the pretreater was started at 0100 hours on February 16, 1978. Pretreater char feed to the reactor was started at 0800 hours on February 17. The reactor became self-sustaining at 2330 hours on February 18 when oxygen was taken out of the start-up burner.

Char feed to the reactor was low and intermittent due to operating problems in the pretreater section until the afternoon of February 21. A 3 ton/hr char feed rate to the reactor was reached at 1400 hours on February 22. After this, char feed to the reactor was continuous except for a short interruption at 0500 hours on February 24 when the solids flow from the high-temperature reactor to the steam-oxygen gasifier was temporarily lost. Test 70 was terminated when the reactor product-gas quench-water circulation pump failed at 1830 hours on February 24.

The reactor operated very well during Test 70, with a total of 279 tons of pretreated char being fed to the reactor. A 3 ton/hr feed rate was achieved for 46 hours, 39 of which were continuous. Solids flow through the reactor was smooth. For Test 70, 321 tons of coal were processed through the pretreater. The purification section operated satisfactorily during Test 70, with 25 hours of operation being logged for the Liquid Phase Methanation unit while processing HYGAS reactor-purified product gas. The quench section operated smoothly until the termination of the test when the quench water circulation pump failed. The effluent cleanup section operated well for Test 70, as did the utilities.

Post-run inspection after Test 70 showed that the coal mill section was in satisfactory condition. The pretreater was found to be clean except for a few pieces of caked coal ranging from 2 to 3-inches in diameter and one of 9-inches. All of the pretreater grid nozzles were intact; however, some hairline cracks were discovered on the pretreater grid itself. Pressure testing revealed a crack in one of the internal cooling coils in the pretreater. The char cooler was found to be clean, as were all the gas and solids transfer lines in that section. In addition, the venturi scrubber and the pretreater quench tower were both in good condition after Test 70.

The slurry preparation section was found to be in satisfactory condition. However, an unusual amount of wear was discovered on the discharge valve seat of both high-pressure slurry pumps after 180 hours of operation each. The extent of wear was similar to that observed after 700 hours of operation with lignite or 400 hours with subbituminous coal. The difference between this case and the earlier ones is that during Test 70, both pumps were operated concurrently at a slower pump speed to achieve char feed rates of over 3 tons/hr.

The HYGAS reactor was inspected after Test 70. The slurry dryer bed, spouting bed, and the second stage reactor were all cleaned. All the solids transfer lines (321, 322, and 399) were found to be plugged with char. This was the result of the sudden termination of Test 70 which did not allow for the emptying of the lines prior to termination. A soft red clinker formation was found in the steam-oxygen gasifier above the gas sparger in the form of a cylinder with an inside diameter of 1 foot. The clinker extended from along the southwest wall up to about 18 inches above the 339 valve between the thermowell and the wall. There were some very minor clinker formations extending about two inches below the cones of the steam-oxygen sparger. The thermocouples and the steam-oxygen gasifier sparger were all found to be in good condition after the test. The high-pressure cyclone and the dipleg were found to be clean.

During the emergency shutdown of Test 70, the steam-oxygen filter assembly exhibited a dull red glow following the stoppage of oxygen feed to the steam-oxygen gasifier. This phenomenon is believed to have been due to the continued flow of superheated steam through the assembly. The filter unit was disassembled for complete testing following Test 70. The filter cartridge was found to be intact, but did show some discoloration similar to that resulting from high heat exposure. Brinnell hardness tests were performed on the filter shell and two downstream locations on the line. All were found to be well within the specifications for the material. There was no evidence of combustible gas backed into the filter element. Therefore, the dull red glow observed on the filter assembly was concluded to be a normal phenomenon related to the temperature of the superheated steam used.

Inspection of the failed quench water pump revealed a clean shear of the pump shaft on the motor side of the pump seals. It was speculated that the pump shaft had bent because of water freezing, and that this eventually caused a fatigue failure of the shaft due to excessive vibration. The quench separator and the prequench tower were found to have large accumulations of solids. The purification section was found to be in satisfactory condition, and the light oil recovery unit was also in good condition after the test.

Plant Turnaround

After Test 70, IGT prepared to make three modifications in the plant to improve the operation of the steam-oxygen gasifier and fines elutriation from the reactor slurry drier.

The first of these modifications was the incorporation of a new, six-nozzle, steam-oxygen sparger design (Figure 23). Another modification was the relocation of the 339 valve to a position approximately 9 feet above the steam-oxygen sparger. Figure 24 shows the new 339 valve location and its mechanical configuration for valve actuation. The installation of double-screening feed equipment upstream of the pretreater was the third major modification. Figure 25 shows the schematic of the new double-screening equipment installation.

Other plant turnaround activities conducted while these three modifications were being made included the inspection of all orifice plates and safety relief valves. Replacements, wherever necessary, were made. Also, the HYGAS high-pressure cyclone was sent to Argonne National Laboratories for inspection and non-destructive testing of all effluent high-pressure slurry lines. Other plant activities by sections are given below.

In the coal preparation section, the coal mill hot-flue gas line was replaced due to normal wear and tear incurred since its original installation. In the pretreater section, a crack found in the pretreater internal cooling bundle was repaired. The repairs on the pretreater gas distributor grid were completed, and new pretreater grid nozzles were installed to provide easy accessibility for cleaning. Electrical wiring and controls on the lockhopper feed system to the pretreater were checked out, and the pretreater section was cleaned.

In the slurry preparation section, the Wilson-Snyder high-pressure slurry pump valve seats were replaced, and the slurry mix tank was emptied of the residual char slurry. The reactor was cleaned. In the quench section, all

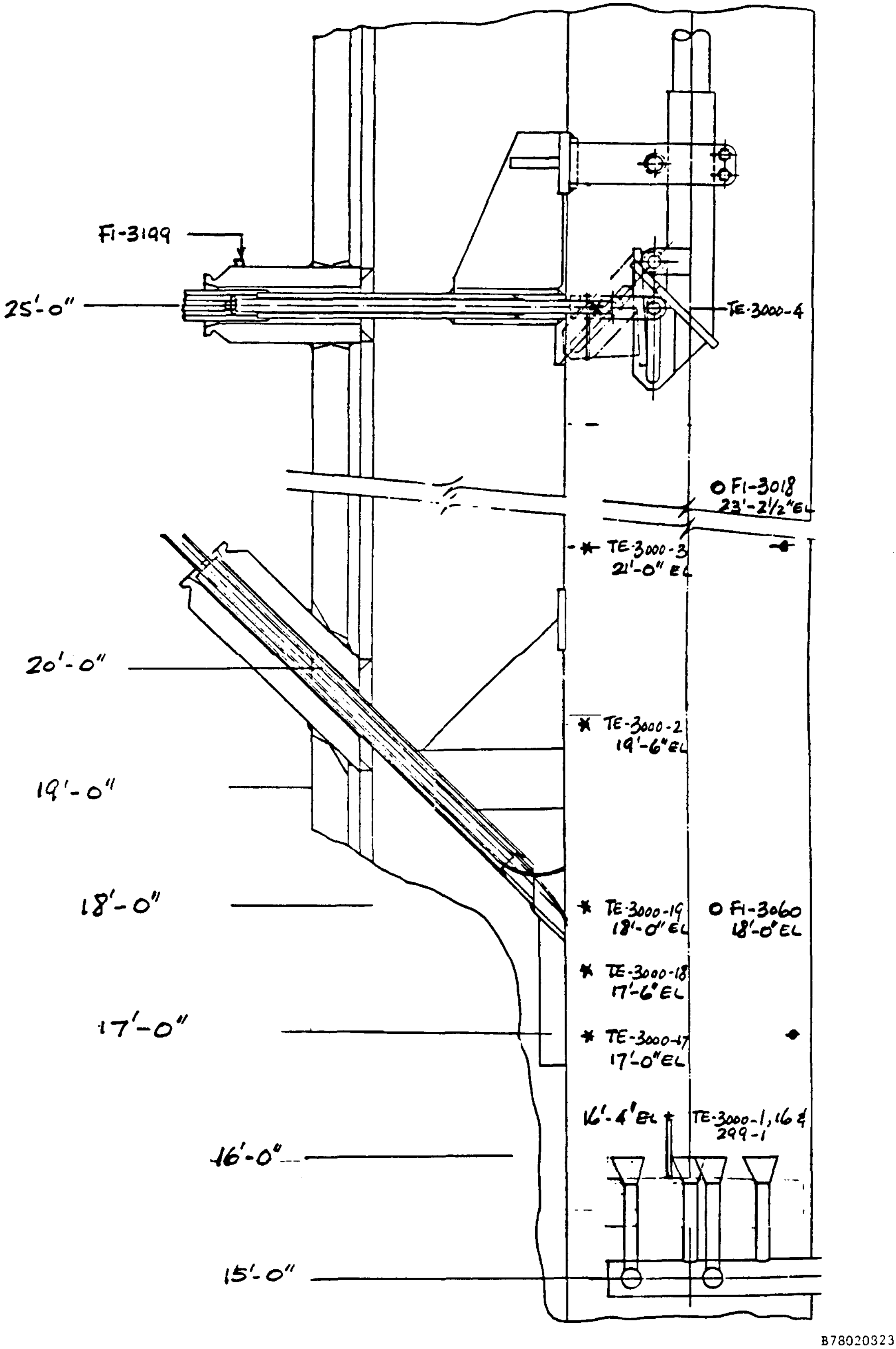


Figure 24. LOCATION OF 339 VALVE AND MECHANICAL ACTUATOR

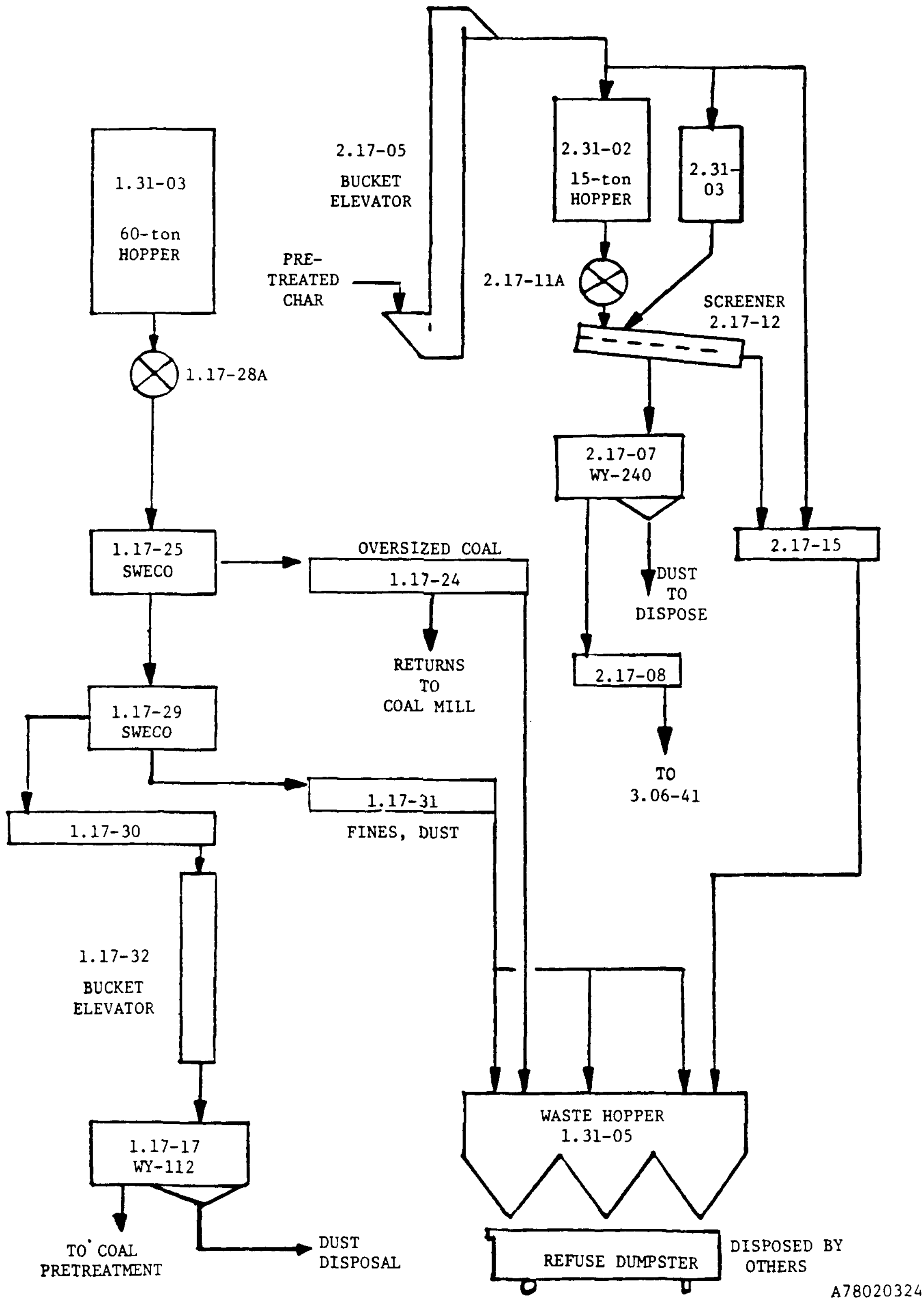


Figure 25. DOUBLE-SCREENING AND SOLIDS WASTE DISPOSAL SYSTEM

vessels and their transfer lines were cleaned, and the quench-water circulation pump was fixed. Repair of the liquid-phase methanation pilot unit was completed. Routine maintenance was performed on the plant utilities. Repair of a level control valve and of the fan on the low-pressure boiler was also completed.

Meetings and Debriefings

A debriefing session for Test 68 was held on January 18, and a debriefing for Tests 69 and 70 was held on March 28. Representatives of DOE, Scientific Design, C.F. Braun, Procon, Inc., and IGT attended both meetings.

A meeting was held at the U.S. Department of Energy (DOE) headquarters on January 10 to lay the ground rules for the transfer of environmental information from IGT to Procon and DOE personnel. On January 18, a meeting was held to review IGT's data on the operating requirements for sinter-free operation in the steam-oxygen gasifier. At the same meeting, an initial review of Procon's commercial HYGAS reactor design was presented. IGT personnel then attended a Procon monthly review meeting in the Edgewood area of the Aberdeen Proving Grounds on February 2. A Scientific Design Corporation debriefing meeting with representatives of DOE, Oak Ridge National Laboratories, and IGT was held in Washington, D.C., on February 7. A review meeting was held on February 8 at Procon on the HYGAS reactor design. Representatives of IGT, Procon, and Darcon attended. IGT personnel also attended a meeting in Washington, D.C., on March 17 on the subject of IGT pilot plant modifications. Representatives of DOE, Gas Research Institute (GRI), Scientific Design, Oak Ridge National Laboratories, and C.F. Braun attended.

The American Gas Association (AGA) project advisors held their quarterly review of the HYGAS Process at IGT on February 14. The previous three months of pilot plant operation were discussed.

Task 8. Demonstration Plant Support

The objective of the work done under this task is to provide engineering assistance to DOE and Procon in their design of a commercial/demonstration plant based on the HYGAS Process. One of the major activities under this task has been the transmittal of data to Procon. During this quarter, the following items were delivered to Procon:

a. Start-up and shutdown procedure for the HYGAS plant

b. The section of the "Coal Conversion Systems Technical Data Book" for the physical properties (such as density, specific heat, etc.) for both pretreated char and spent reactor char

c. Environmental data collected by IGT for the HYGAS pilot plant for the HYGAS Environmental Assessment Program

d. Coal crushing test data acquired by IGT at the T.J. Gundlach Machine Company of Belleville, Illinois, on high-ash, high-sulfur, run-of-mine Illinois No. 6 coal

e. Preliminary heat and material balances for the water-slurry feed case when operating the 1200 psig gasifier on the run-of-mine Illinois No. 6 bituminous coal (These energy and material balances are shown in Figure 26. They were determined by selecting a pumpable slurry concentration for the pretreated coal, determining the heat required to evaporate that water in the slurry, defining the heat available in the slurry dryer bed, and supplying additional heat for partial vaporization externally. Even though the external heat load is quite high, the overall system may be advantageous because the hot, lightly washed, raw off-gas can proceed directly to shift without cooling for oil removal.)

f. The effect of changes from the design operating conditions in the gasifier (The operation of the proposed gasifier on run-of-mine Illinois No. 6 coal was evaluated at conditions which varied from the design conditions over the anticipated range. These data are presented in Table 5.)

g. Heat and material balances for the pretreater section for Illinois No. 6 coal using both 6% and 12% moisture cases (These data are presented in Figures 27 and 28 and Tables 6 through 11.)

h. Heat and material balances for the pretreater and gasifier using washed coal as a feed to the system (These balances were the result of a brief cost minimization study and are comparable to the balances done on run-of-mine coal. These balances are shown in Tables 12 through 17 and Figures 29 and 30.)

i. The following documents on the IGT data base for the pretreatment section:

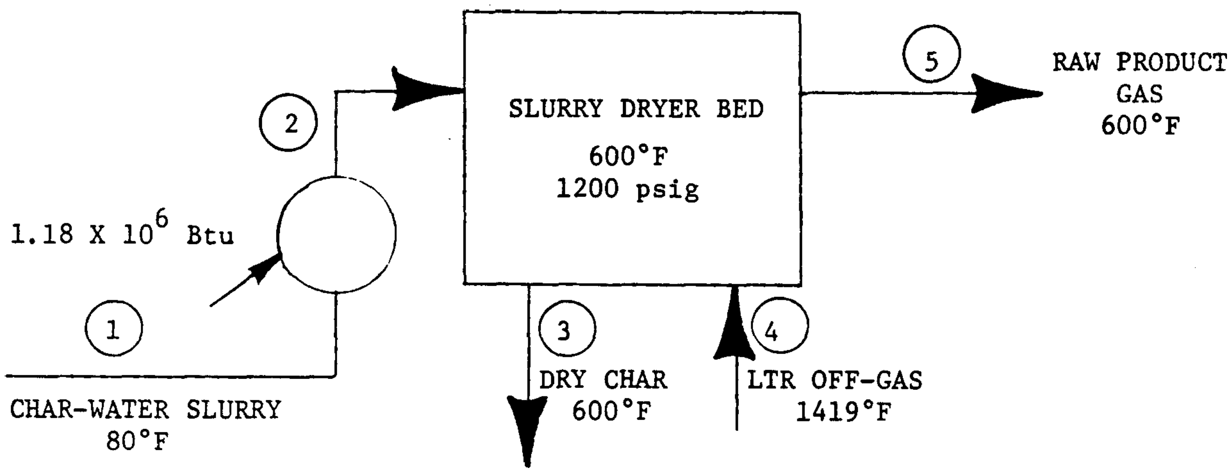
1) IGT Research Bulletin No. 39, entitled "The Production of Pipeline Gas by Hydrogasification of Coal," covering work continued through 1964

2) The copies of test results of 10-inch PDU work on pretreatment for Eastern Coals, published in Part VII of Vol. 3 of the IGT report of work conducted under OCR Contract No. 14-01-0001-381, entitled "HYGAS 1964 to 1972 — Pipeline Gas from Coal — Hydrogenation (IGT Hydrogasification Process)"

3) Copies of pretreatment data sheets taken from IGT monthly work reports on the HYGAS Process on an 8-foot-diameter pretreater which is presently under operation at the pilot plant

4) Excerpts from the final report being prepared for the EPA on a program for coal desulfurization.

BASIS: 1000 lb Pretreated Char



Stream No.	1	2	3	4	5
Description	Char-water slurry	Preheated slurry	Dry char	LTR off-gas	Product gas
Temperature, °F	80	569	600	1419	600
Pressure, psig	Ambient	1200	1200	1200	1200
Composition					
Char Streams, lb					
Char	1000	1000	1000	--	--
H_2O (l)	1500	860	--	--	--
H_2O (v)	--	640	--	--	--
Gas Streams, mol					
CO				12.6020	12.6020
CO_2				16.8722	16.8722
H_2				17.7542	17.7542
H_2O				24.5919	107.8558
CH_4				12.9491	12.9491
C_2H_6				0.2624	0.2624
NH_3				0.5247	0.5247
N_2				0.0505	0.0505
HCN				0.0481	0.0481
H_2S				0.9217	0.9217
COS				0.0384	0.0384
C_6H_6				0.1050	0.1050
C_7H_8				0.5098	0.5098

Figure 26. PRELIMINARY HEAT AND MATERIAL BALANCE FOR WATER SLURRY CASE FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL (Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies.)

Table 5. EFFECTS OF CHANGES IN THE GASIFIER DESIGN OPERATING CONDITIONS

Change	ΔT OG	ΔT HTR	Δ Coal	Conver	ΔH_2O or Steam	ΔO_2	Prodn, 10^9 Btu/day	Raw Gas ΔH_2O	Raw Gas Δ CO	Raw Gas ΔH_2	Raw Gas ΔCH_4	Processing ΔH_2O Condn	Processing Δ CO Shiftd	Processing Δ CO Meth	Processing ΔCO_2 Rmvd
	°F	°F	%	%	%	%		%	%	%	%	%	%	%	%
None	0	0	0	90.0	0	0	250.0	0	0	0	0	0	0	0	
Base Case	1850*	1725*	655.67†		68,821††	9258††		32,248††	16,527††	23,282††	17,583††	25,200††	7048††	9478††	29,173††
Increased Coal Reactivity, rapid	0	+4.5	0	90.9	0	-1.5	253.4	-1.09	-2.73	-4.92	+4.40	-1.11	-1.02	-4.01	+0.649
Decreased Coal Reactivity															
Rapid	0	-3.0	0	89.3	0	+0.7	247.4	+0.83	+1.69	+3.51	-3.20	+0.99	+0.27	+2.76	-0.643
Low	0	+0.5	0	89.0	0	-1.0	247.3	+1.36	-2.57	-0.79	-0.83	+2.85	-3.97	-1.53	-1.31
Coal Feed															
Decreased	0	-1.7	-5.0	91.6	0	-2.6	241.3	+1.97	-4.52	-1.20	-3.96	+4.51	-7.14	-2.58	-2.84
Decreased	0	-3.8	-10.0	93.2	0	-5.2	232.3	+4.20	-9.32	-2.55	-8.08	+9.47	-14.64	-5.36	-5.84
Increased	0	+1.8	+5.0	88.5	0	+2.5	258.4	-1.83	+4.39	+1.11	+3.88	-4.28	+6.97	+2.47	+2.71
Increased	0	+4.0	+10.0	87.1	0	+4.7	266.8	-3.63	+8.77	+2.16	+7.78	-8.55	+13.97	+4.90	+5.35
Steam															
Increased	0	-1.8	0	91.5	+5.0	+2.4	253.9	+6.94	+0.49	+3.79	+1.07	+9.47	-2.10	+2.42	+2.18
Decreased	0	+1.7	0	88.4	-5.0	+2.6	246.0	-6.84	-0.61	-3.91	-1.09	-9.31	+1.99	-2.54	-2.28
OG Temperature															
Decreased	-10	-6.3	0	88.5	0	-2.6	246.4	+1.87	-4.53	-1.97	-0.57	+4.23	-6.54	-3.03	-2.17
Decreased	-20	-12.9	0	87.0	0	-5.3	242.7	+3.79	-8.97	-3.96	-1.16	+8.47	-12.91	-6.04	-4.35
Decreased	-30	-19.1	0	85.5	0	-7.9	238.8	+5.82	-13.37	-5.92	-1.89	+12.83	-19.22	-9.01	-6.58

* Base case = °F.

† Base case = tons/hr.

†† Base case = mol/hr.

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Table 6. PRELIMINARY MATERIAL AND HEAT BALANCE FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL*
(Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies.)

Run of Mine Coal Analyses

Proximate Analysis, wt %	
Volatile Matter	32.90
Moisture	12.00
Fixed Carbon	38.21
Ash	16.89
Total	100.00
Ultimate Analysis, wt %	
Carbon	62.70
Hydrogen	4.67
Oxygen	7.85
Nitrogen	1.18
Sulfur	4.25
Chloride	0.16
Ash	19.19
Total	100.00

* Coal contains 12% moisture

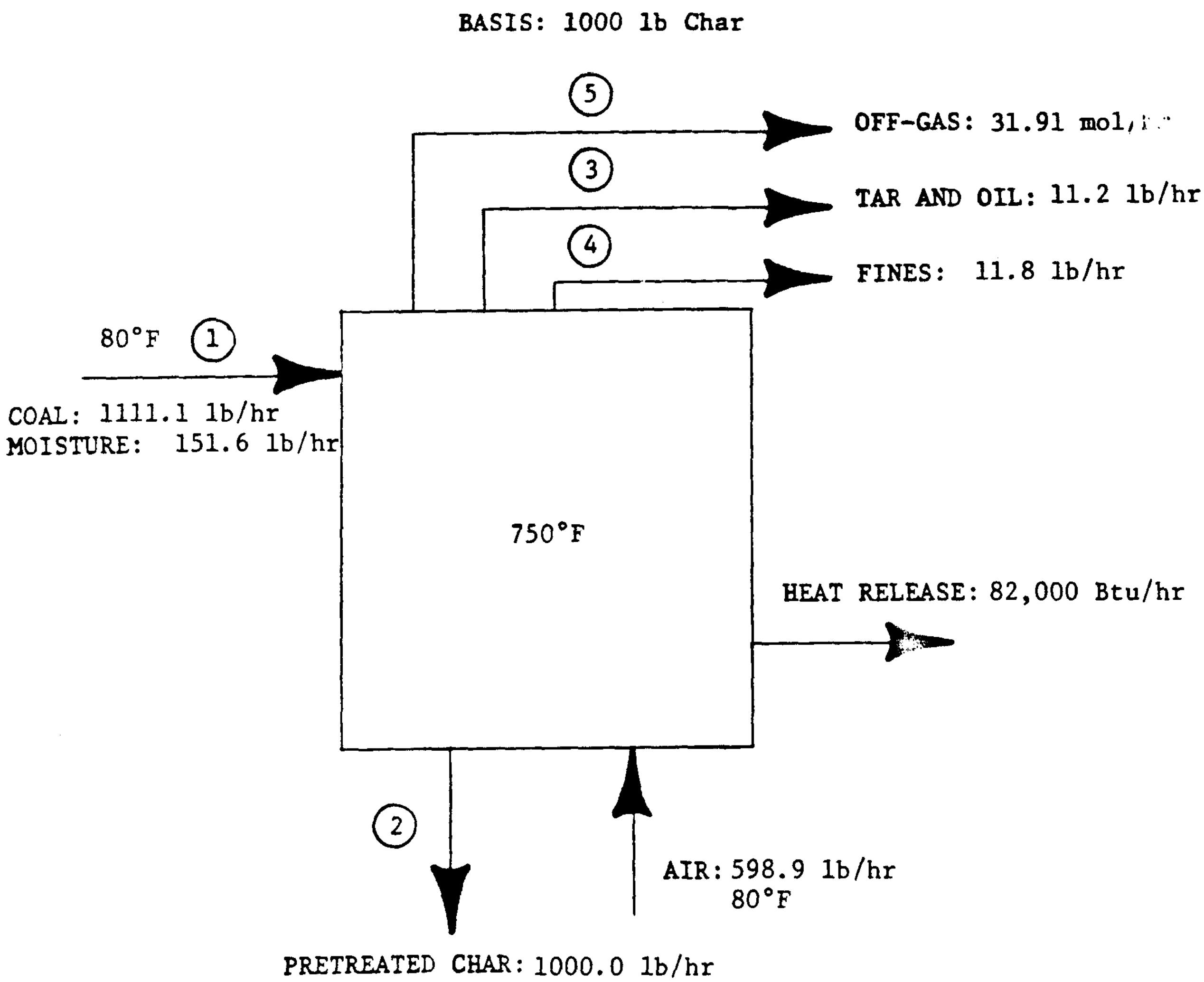

Figure 27. PRELIMINARY MATERIAL AND HEAT BALANCE FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL*
(Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies.)

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* Coal contains 12% moisture.

Table 7. PRELIMINARY HEAT AND MATERIAL BALANCE FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL*
(Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies.)

Stream No.	1		2		3		4	
Description	Coal Feed		Pretreated Char		Tar and Oil in Off-Gas		Fines in Off-Gas	
Temperature, °F	80		750					
Components	lb/hr	wt %	lb/hr	wt %	lb/hr	wt %	lb/hr	wt %
Carbon	696.7	62.70	630.3	63.03	--	80.5	--	51.5
Hydrogen	51.9	4.67	43.4	4.34	--	8.0	--	2.7
Oxygen	87.2	7.85	70.0	7.00	--	7.5	--	9.1
Nitrogen	13.1	1.18	11.8	1.18	--	0.5	--	1.2
Sulfur	47.2	4.25	34.2	3.42	--	3.5	--	4.1
Chloride	1.8	0.16	0.9	0.09	--	--	--	--
Ash	213.2	19.19	209.4	20.94	--	--	--	31.4
Total	1111.1	100.00	1000.0	100.00	11.2	100.0	11.8	100.0
Moisture	151.6							

* Coal contains 12% moisture.

Table 8. PRELIMINARY HEAT AND MATERIAL BALANCE FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL IN STREAM 5*
(Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies)

Stream No.	5	
Description	Pretreater Off-Gas	
Temperature, °F	750	
Components	mol/hr	mol %
CO	0.598	1.87
CO_2	2.820	8.84
H_2	0.471	1.48
H_2O	10.088	31.62
SO_2	0.378	1.18
CH_4	0.402	1.26
C_2H_6	0.100	0.31
C_3H_8	0.081	0.25
HCl	0.125	0.08
N_2	16.438	51.53
O_2	0.505	1.58
Total	31.906	100.00

* Coal contains 12% moisture.

Table 9. PRELIMINARY BALANCE OF HEAT AND MATERIAL FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL*
(Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies.)

Proximate Analysis, wt %	
Volatile Matter	35.14
Moisture	6.00
Fixed Carbon	40.82
Ash	18.04
Total	100.00
Ultimate Analysis, wt %	
Carbon	62.70
Hydrogen	4.67
Oxygen	7.85
Nitrogen	1.18
Sulfur	4.25
Chloride	0.16
Ash	19.19
Total	100.00

* Coal contains 6% moisture.

Pretreater Balance

BASIS: 1000 lb Char

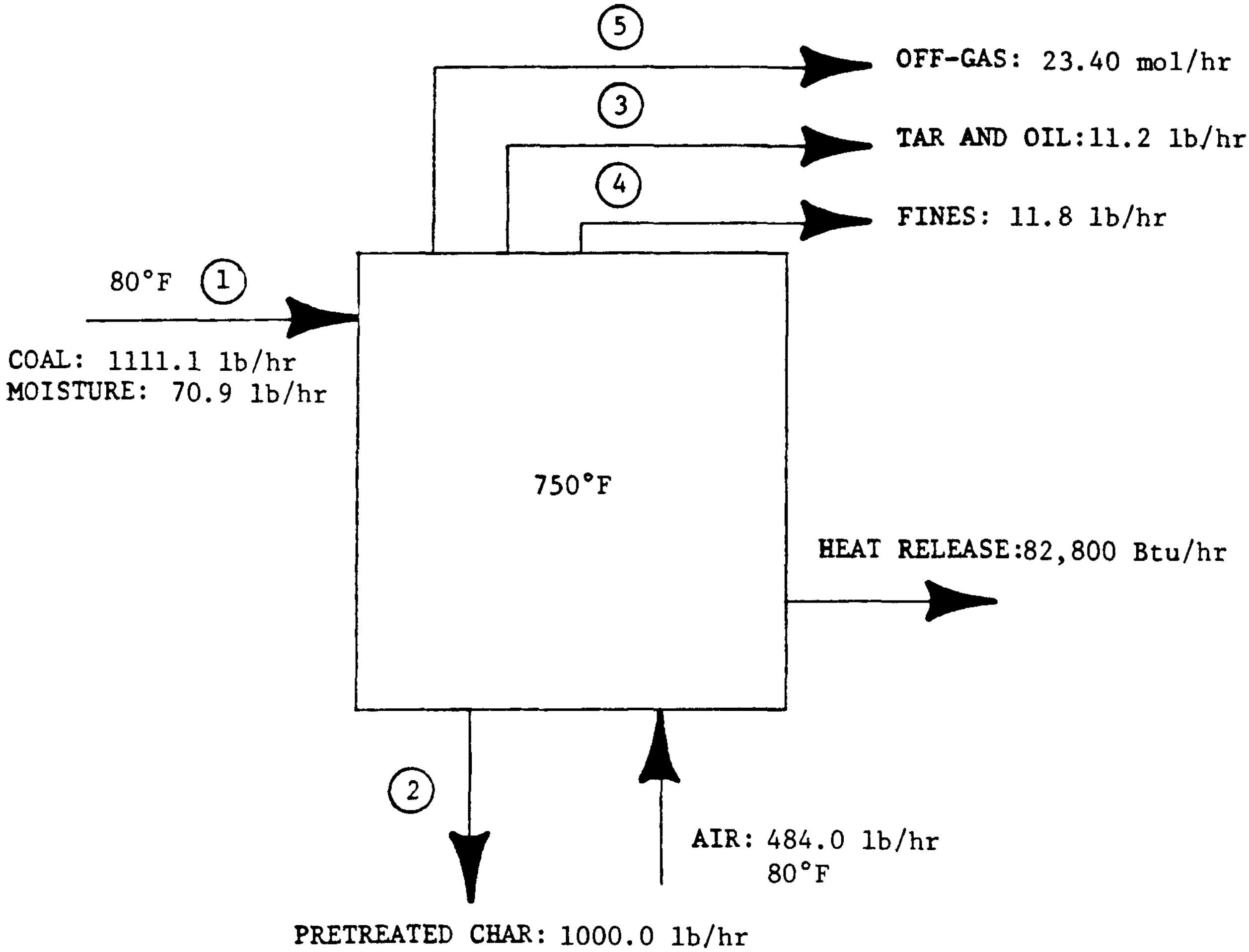


Figure 28. PRELIMINARY HEAT AND MATERIAL BALANCE FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL*
(Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies.)

---

* Coal contains 6% moisture.

Table 10. PRELIMINARY BALANCE OF MATERIAL AND HEAT FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL*
(Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies.)

Stream No.	1		2		3		4	
Description	Coal Feed		Pretreated Char		Tar and Oil in Off-Gas		Fines in Off-Gas	
Temperature, °F	80		750					
Components	lb/hr	wt %	lb/hr	wt %	lb/hr	wt %	lb/hr	wt%
Carbon	696.7	62.70	630.3	63.03	--	80.5	--	51.5
Hydrogen	51.9	4.67	43.4	4.34	--	8.0	--	2.7
Oxygen	87.2	7.85	70.0	7.00	--	7.5	--	9.1
Nitrogen	13.1	1.18	11.8	1.18	--	0.5	--	1.2
Sulfur	47.2	4.25	34.2	3.42	--	3.5	--	4.1
Chloride	1.8	0.16	0.9	0.09	--	--	--	--
Ash	213.2	19.19	209.4	20.94	--	--	--	31.4
Total	1111.1	100.00	1000.0	100.00	11.2	100.0	11.8	100.0
Moisture	70.9							

* Coal contains 6% moisture.

Table 11. PRELIMINARY PRETREATER OFF-GAS MATERIAL AND HEAT BALANCE FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL*
(Note: This Balance is Preliminary and Must Be Confirmed by Additional Studies.)

Stream No.	5	
Description	Pretreater Off-Gas	
Temperature, °F	750	
Components	mol/hr	mol %
CO	0.598	2.56
CO_2	2.428	10.37
H_2	0.471	2.01
H_2O	4.920	21.02
SO_2	0.378	1.62
CH_4	0.627	2.68
C_2H_6	0.152	0.65
C_3H_8	0.102	0.44
HCl	0.025	0.11
N_2	13.294	56.80
O_2	0.408	1.74
Total	23.403	100.00

* Coal contains 6% moisture.

Table 12. PRETREATER AND GASIFIER MATERIAL AND HEAT BALANCE FOR ILLINOIS NO. 6 BITUMINOUS WASHED COAL

	wt %
Proximate Analysis	
Volatile Matter	36.32
Moisture	12.00
Fixed Carbon	42.15
Ash	9.53
Total	100.00
Ultimate Analysis	
Carbon	69.47
Hydrogen	5.25
Oxygen	9.60
Nitrogen	1.03
Sulfur	3.80
Chloride	0.02
Ash	10.83
Total	100.00

Table 13. PRETREATER AND GASIFIER HEAT AND MATERIAL BALANCE FOR ILLINOIS NO. 6 BITUMINOUS WASHED COAL

Stream No.	1		2		3		4	
Description	Coal Feed		Pretreated Char		Tar and Oil in Off-Gas		Fines in Off-Gas	
Temperature, °F	80		750					
Components	lb/hr	wt %	lb/hr	wt %	lb/hr	wt %	lb/hr	wt %
Carbon	771.9	69.47	714.5	71.45	9.0	80.4	6.1	51.5
Hydrogen	58.3	5.25	49.0	4.90	0.9	8.0	0.3	2.7
Oxygen	106.7	9.60	78.9	7.89	0.8	7.1	1.1	9.0
Nitrogen	11.5	1.03	10.3	1.03	0.1	0.9	0.1	1.2
Sulfur	42.2	3.80	30.6	3.06	0.4	3.6	0.5	4.1
Chloride	0.2	0.02	0.1	0.01	--	--	--	--
Ash	120.3	10.83	116.6	11.66	--	--	3.7	31.5
Total	1111.1	100.00	1000.0	100.00	11.2	100.0	11.8	100.0
Moisture	151.5							

Table 14. PRETREATER AND GASIFIER HEAT AND MATERIAL BALANCE FOR ILLINOIS NO. 6 BITUMINOUS WASHED COAL IN STREAM 5

Stream No.	5	
Description	Pretreater Off-Gas	
Temperature, °F	750	
Components		
	mol/hr	mol %
CO	0.49	1.64
CO_2	2.32	7.76
H_2	0.21	0.70
H_2O	11.04	36.95
SO_2	0.34	1.14
CH_4	0.34	1.14
C_2H_6	0.08	0.27
C_3H_8	0.07	0.23
HCl	0.00	0.00
N_2	14.54	48.66
O_2	0.45	1.51
Total	29.88	100.00

Table 15. PRETREATER OFF-GAS AND GASIFIER MATERIAL AND HEAT BALANCE FOR ILLINOIS NO. 6 BITUMINOUS WASHED COAL

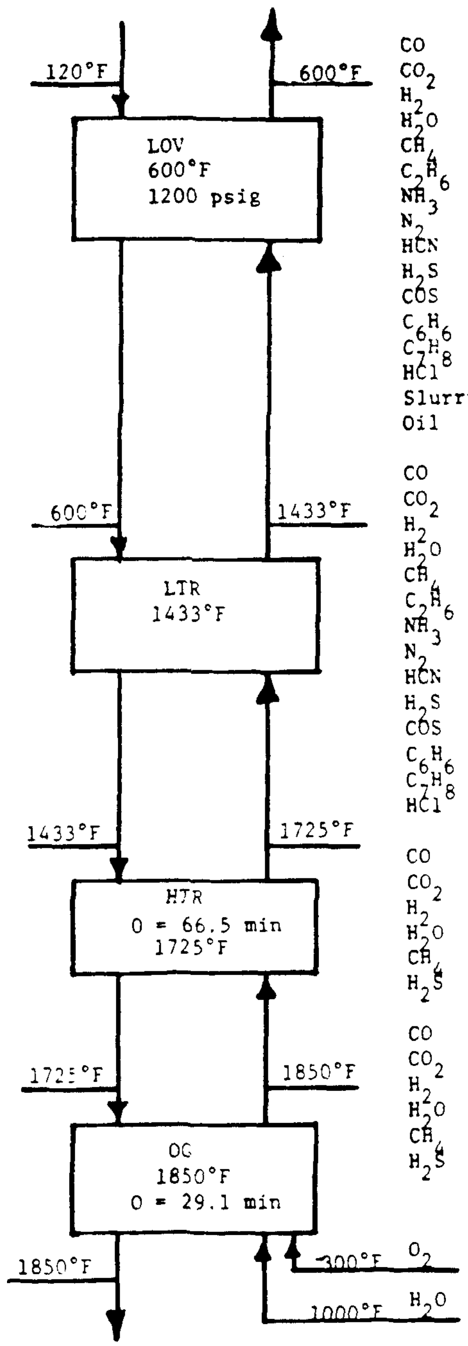
Stream No.	1			2		
Description	Coal Feed			Ash Residue		
Temperature °F	120			1850		
Components						
	lb/hr	wt %	mol/hr	lb/hr	wt %	mol/hr
Carbon	714.5	71.45	59.49	71.3	36.79	5.94
Hydrogen	49.0	4.90	24.31	1.9	0.98	0.95
Oxygen	78.9	7.89	4.93	--	--	--
Nitrogen	10.3	1.03	0.37	0.9	0.46	0.03
Sulfur	30.6	3.06	0.95	3.1	1.60	0.10
Chloride	0.1	0.01	0.003	--	--	--
Ash	116.6	11.66	--	116.6	60.17	--
Total	1000.0	100.00		193.8	100.00	
Slurry Oil	2249.9					

Table 16. PRETREATER AND GASIFIER MATERIAL AND HEAT BALANCE FOR ILLINOIS NO. 6 BITUMINOUS WASHED COAL IN STREAM 3

Stream No.	3	
Description	Raw Product Gas	
Temperature, °F	600	
Components		
	mol/hr	mol %
CO	14.90	15.69
CO_2	18.42	19.39
H_2	19.82	20.87
H_2O	25.28	26.62
CH_4	14.79	15.57
C_2H_6	0.30	0.32
NH_3	0.52	0.55
N_2	0.05	0.05
HCN	0.05	0.05
H_2S	0.82	0.86
COS	0.03	0.03
HCl	--	--
Total (Oil-Free Gas)	94.98	100.00
	mol/hr	wt %
C_6H_6	0.12	15.0
C_7H_8	0.58	85.0
Total (Product Oil)	0.70	100.0
Total (Oil-Free Gas + Product Oil)	95.68	
Slurry Oil, lb/hr	2249.9	

Table 17. GASIFIER AND PRETREATER HEAT AND MATERIAL BALANCE FOR ILLINOIS NO. 6 BITUMINOUS WASHED COAL

Basis: 1000 lbs Char



Feed to LOV (120°F)

	Moles
C	59.4921
H_2	24.3056
O_2	4.9312
N_2	0.3676
S_2	0.9545
Cl	0.0030
Ash*	116.6000
Moisture	0.0
Slurry*	
Oil	2249.9

LOV gas exit (600°F)

	Moles
CO	14.9047
CO_2	18.4149
H_2	19.8161
H_2O	25.2775
CH_4	14.7885
C_2H_6	0.2974
NH_3	0.5247
N_2	0.0505
HCN	0.0481
H_2S	0.8246
COS	0.0344
C_6H_6	0.1203
C_7H_8	0.5779
HCl	0.0030
Slurry*	
Oil	2249.9

LOV to LTR (600°F)

	Moles
C	59.4921
H_2	24.3056
O_2	4.9312
N_2	0.3676
S_2	0.9545
Cl	0.0030
Ash*	116.6000

LTR gas to LOV (1433°F)

	Moles
CO	14.9047
CO_2	18.4149
H_2	19.8161
H_2O	25.2775
CH_4	14.7885
C_2H_6	0.2974
NH_3	0.5247
N_2	0.0505
HCN	0.0481
H_2S	0.8246
COS	0.0344
C_6H_6	0.1203
C_7H_8	0.5779
HCl	0.0030

LTR to HTR (1433°F)

	Moles
C	46.5707
H_2	3.2225
O_2	---
N_2	0.0307
S_2	0.1874
Ash*	116.6000

HTR gas to LTR (1725°F)

	Moles
CO	16.1338
CO_2	15.6900
H_2	16.4814
H_2O	24.6013
CH_4	8.8073
H_2S	0.0920

HTR to OG (1725°F)

	Moles
C	34.8835
H_2	2.4138
O_2	---
N_2	0.0307
S_2	0.1874
Ash*	116.6000

OG gas to HTR (1850°F)

	Moles
CO	13.1632
CO_2	12.7829
H_2	18.5069
H_2O	33.3861
CH_4	2.9978
H_2S	0.0920

OG solids exit (1850°F)

	Moles
C	5.9395
H_2	0.9506
O_2	---
N_2	0.0307
S_2	0.0955
Ash*	116.6000

OG inputs

	Moles
O_2 (300°F)	7.7988
H_2O (1000°F)	56.5175

* These quantities in lbs.

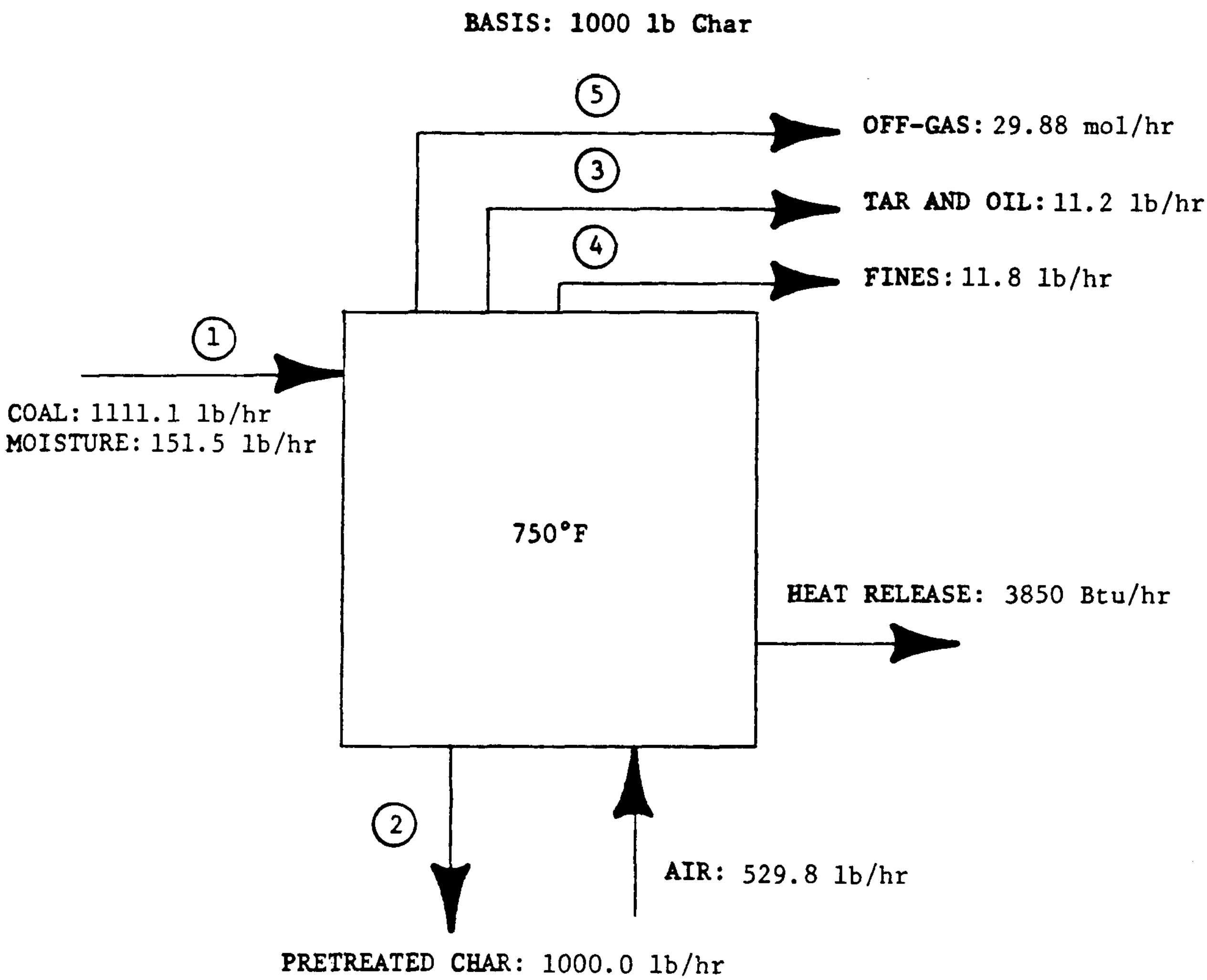


Figure 29. PRETREATER AND GASIFIER MATERIAL AND HEAT BALANCE FOR ILLINOIS NO. 6 BITUMINOUS WASHED COAL

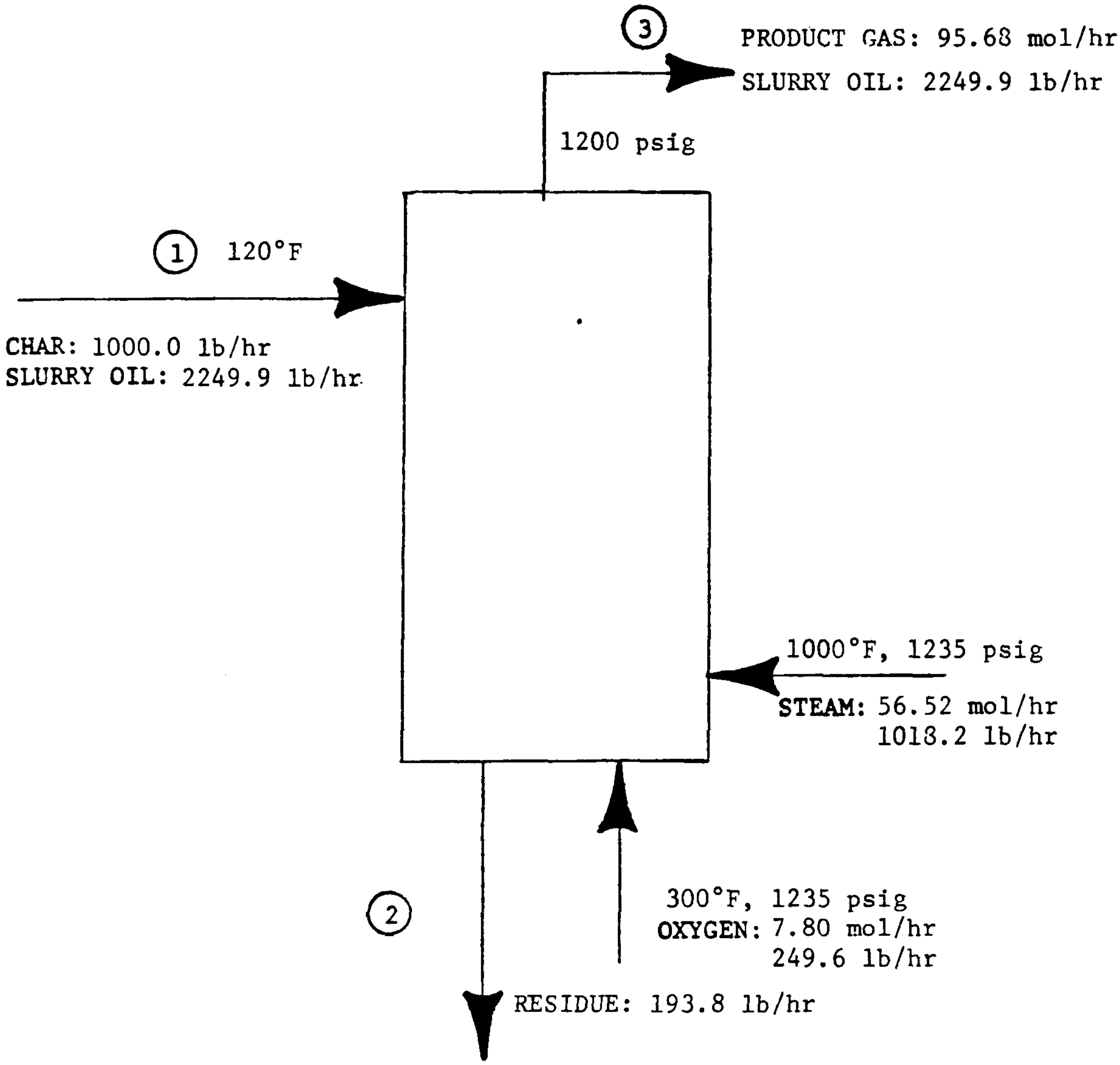


Figure 30. PRETREATER AND GASIFIER HEAT AND MATERIAL BALANCE FOR ILLINOIS NO. 6 BITUMINOUS WASHED COAL

Commercial Plant Reactor Design

A meeting was held at IGT on January 18, 1978, with DOE and Procon. At this meeting, R. Pfeiffer, Procon's fluidization consultant, discussed the proposed reactor design which consists of an external low temperature reactor (LTR). The following advantages were cited by Pfeiffer for the design: 1) easy maintenance because all LTR equipment would be located externally and 2) an external solids flow control valve which operates at 600°F.

Pfeiffer also discussed other advantages such as the use of a "waffle" refractory grid in the steam-oxygen gasifier, in the HTR and in the slurry drier bed. The only disadvantage noted was that this design would require an additional three, hot, steel penetrations, which could be expensive.

Pfeiffer also briefly described two other designs, one of which has an external LTR, but no cyclone. In this design, the plenum chamber between the HTR and the slurry drier bed would be extended to be used as the disengaging vessel. The other design utilized a completely internal LTR gasifier.

Some of the major points of the proposed reactor design discussed were: 1) reduction of the superficial gas velocity in the slurry drier to 0.4 feet to minimize fines carry-over, 2) provision for a steam-oxygen gas distributor inside the plenum below the steam-oxygen gasifier grid, and 3) possible removal of the slurry drier cyclone fines entirely from the reactor system. The last item of concern would depend upon the size distribution of the char feed to the reactor which, in turn, would depend on the crushing and screening operation in the coal preparation area.

Another follow-up meeting on the reactor design was held on February 9, 1978, at Procon to review the revised reactor drawing and to discuss other questions raised by IGT.

Following this meeting, IGT prepared a revised drawing for the proposed HYGAS reactor which incorporated IGT's proposed modifications. This drawing, shown in Figure 31, was developed after in-house discussions at IGT on the Procon reactor design labeled 1-A. The modifications made include the following:

a. Modification of the steam-oxygen gasifier bed solids removal system from an overflow pipe to an underflow pipe (Design 1A used an overflow pipe with a slip fit of the overflow pipe through the waffle grid.)

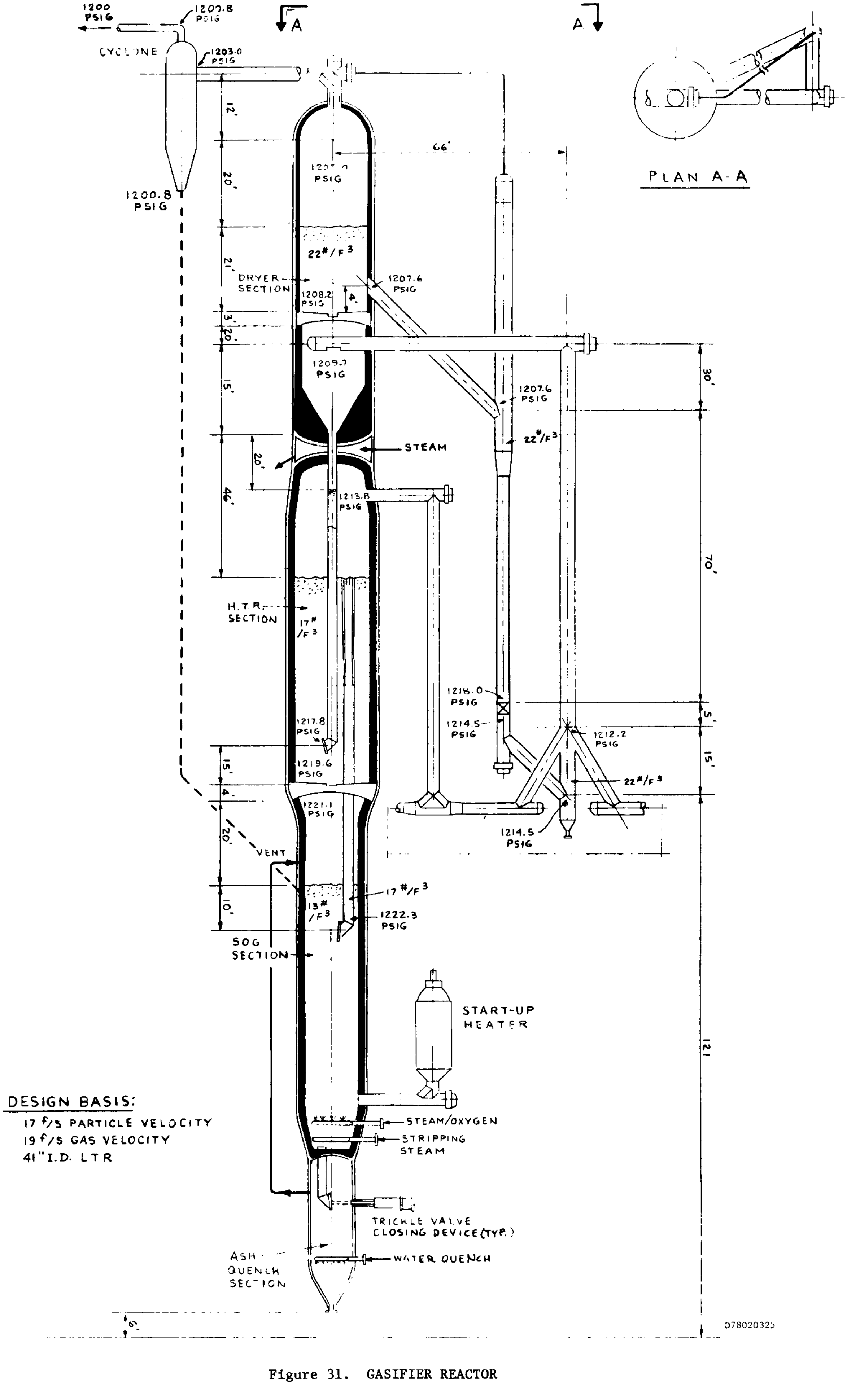


Figure 31. GASIFIER REACTOR

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b. Elimination of the LTR cyclone and direct return routing of the solids to a separation chamber within the main vessel (IGT's work on attrition in gas cyclones has indicated that this unit would cause significant attrition of the coal particles. The correlations for lignite predict a 19.6% size reduction with the design conditions for this cyclone. The elimination of attrition is important in minimizing fines production within the gasifier system and the loss of this coal from the process.)

c. Elimination of the hot head at the top of the HTR gasifier stage by utilization of two pressure heads and by the use of steam for interhead cooling (In cases of steam loss, a backup stream of water sprays could be used.)

d. Utilization of a steam-oxygen sparger system for gas distribution (IGT also recommended underflow discharge of solids with a valve on the underflow, and steam stripping of the gasification residue to recover heat and eliminate synthesis gas from voids of the discharged solids.)

e. Reversal of the relative locations of the valve and water quench sprays in the water quench section to eliminate the potential blockage of the valve, and the connection of the vent for this section to the top of the steam-oxygen gasifier.

Cold Flow Model

During this quarter, work continued on the construction of a cold flow model of the upper stage of the gasification reactor. This stage of the system is the only section of the proposed demonstration unit that is not a direct mechanical transfer of technology from the pilot plant reactor. The model is being constructed to determine gas-solids behavior, on a large scale and at elevated pressures, in systems similar to the proposed demonstration plant design.

The procurement status of the various elements of this model at the end of this quarter was as follows:

a. Compressor: On Order

b. Instrumentation: Partially received

c. Building Foundation: Completed

d. Building Structural Steel: Drawings have been approved; steel is being cut

e. Vessels: Under Construction

f. Cyclone: Under Construction

g. Pipeline Filters: On Order

h. Control Valves: Received.

A drawing showing the process equipment and instrumentation layout for this test unit is shown in Figure 32.

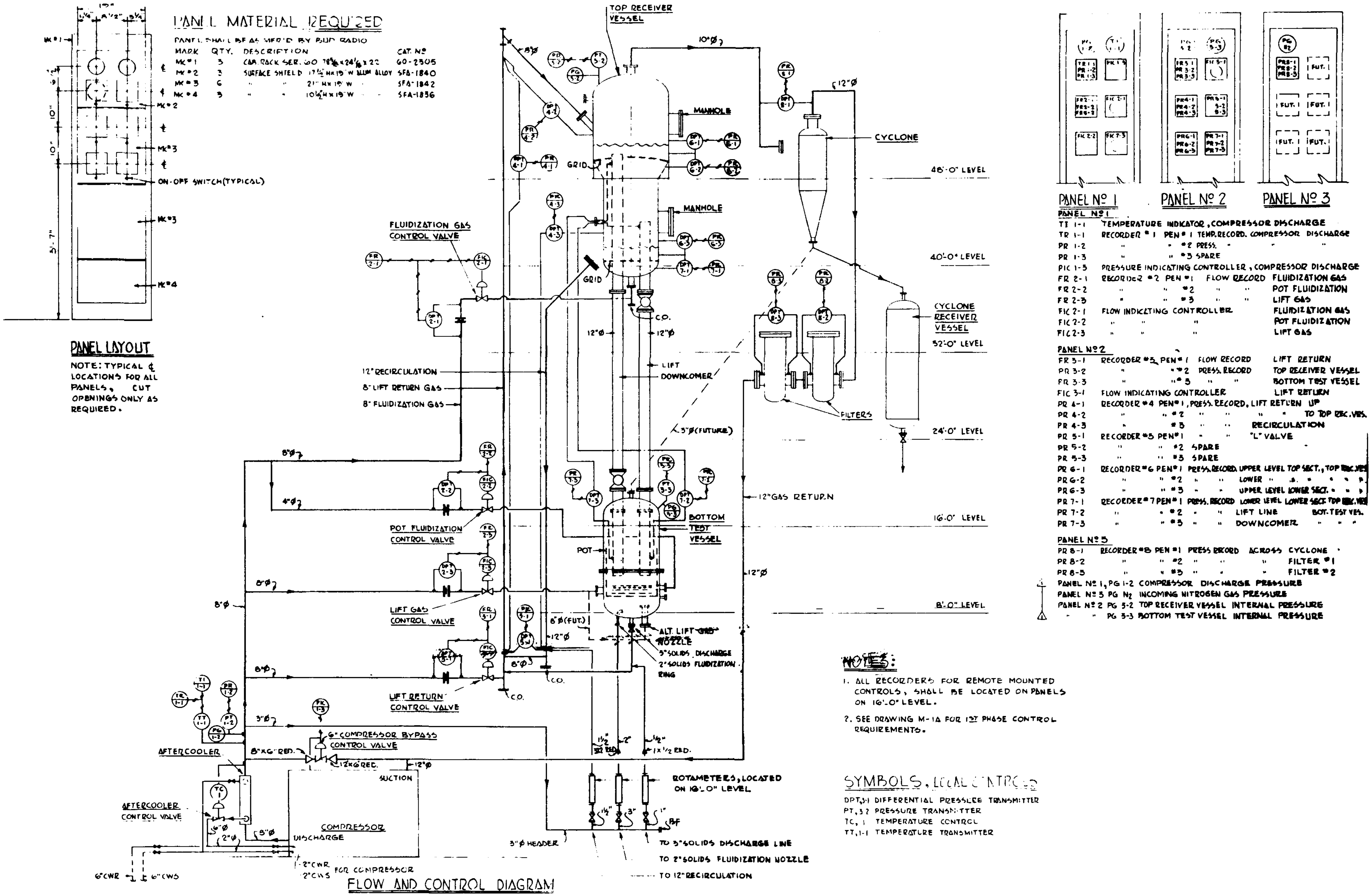

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Figure 32. FLUIDIZATION TEST BUILDING SECOND-PHASE FLOW AND CONTROL DIAGRAM

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Evaluation of LTR Feeder Devices

In support of the modeling effort discussed above, a small, ambient temperature, low-pressure plastic model was constructed to evaluate various solids feeding devices for the LTR section of the demonstration HYGAS plant. Six configurations were selected for testing: three lift-pot configurations, an L-valve, a reverse seal leg, and a reverse-seal pot configuration. During this quarter, tests were completed on one lift-pot device, and a second lift-pot device was tested. In addition, the L-valve and reverse-seal leg configurations were tested.

Lift-Pot LTR Feeder Devices

Lift-Pot I

A lift-pot feeder device was constructed and tested with sand in December 1977. Early in this quarter, tests were continued with this configuration using —20+200 mesh pretreated Illinois No. 6 bituminous coal. The purpose of these tests was to evaluate the operation of the device using a material similar to that which will be used in the LTR section of the HYGAS demonstration plant. Figure 33 shows the configuration of this device.

In a typical test, the 9.5-inch-diameter lift pot is first filled with coal and fluidized with air passing through a ring distributor. A 2-inch, Schedule 40, clear polyvinyl chloride (PVC) standpipe transfers the coal from a fluidized bed (not shown in Figure 33) to the lift pot. The flow rate of coal into the lift pot is controlled by a full-port ball valve in the downcomer. The coal flow rate is determined by timing the particles as they pass between two marks, 12 inches apart, on the downcomer. During normal operation, the coal is in packed-bed flow above the ball valve and in streaming-flow below it.

The fluidization gas for the lift pot passes up the lift line. Additional air is added to the 11.5-inch inside diameter Plexiglas column to ensure an adequate gas velocity up the lift line. This air sweeps the solids from the top of the lift-pot bed into the lift line, and carries them into the fluidized bed above.

In a typical run, the upper bed is first fluidized. The desired lift-pot fluidization velocity and the lift-line velocity are then set. Readings are taken at several different solids flow rates and the results are then analyzed.

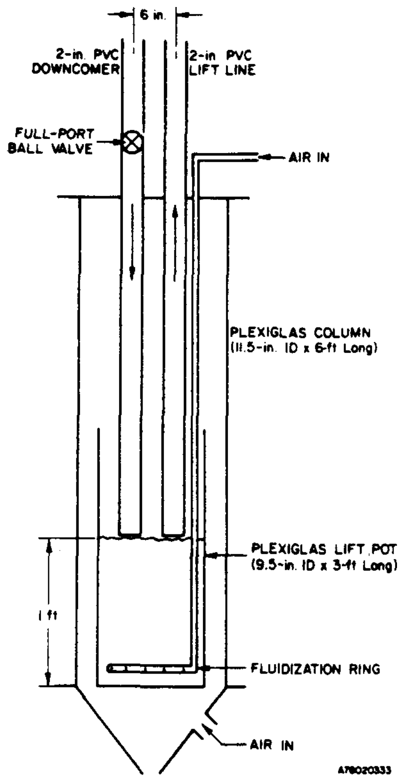
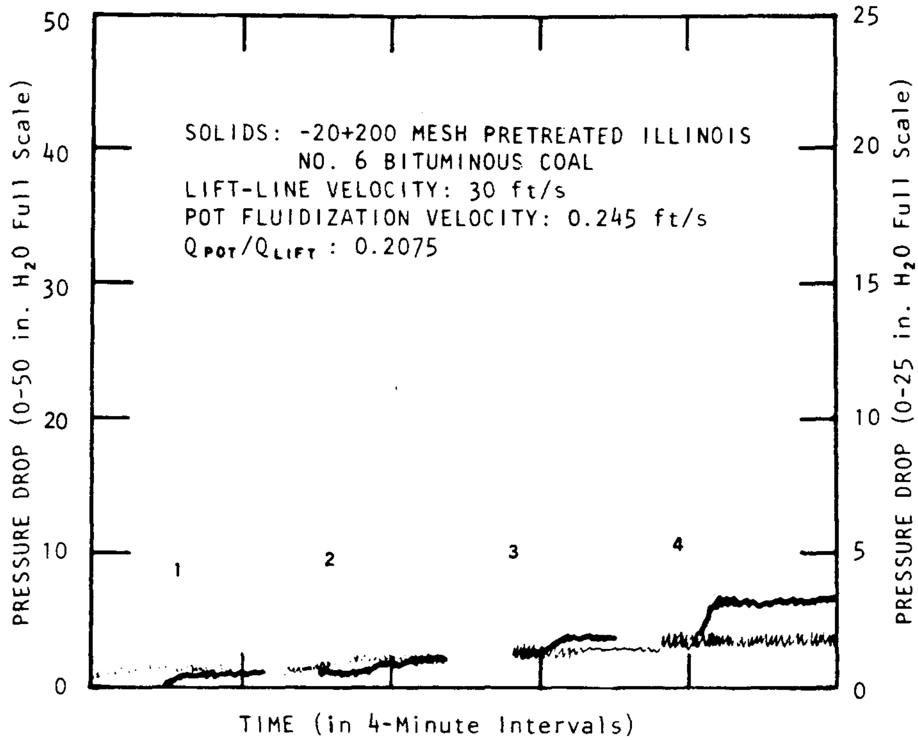


Figure 33. LIFT-POT TEST CONFIGURATION

It is important that the solids be injected into the lift line smoothly and controllably to prevent slugging and poor conversion in the LTR. The fluctuations in the recorder tracing for the lift-line pressure drop are used to analyze the smoothness of the lift-line's operation. Two lift-line pressure drops are monitored: 1) a lower acceleration section, where the solids are accelerated to their final velocity by the lift gas, and 2) an upper steady-state section where the solids have completed their acceleration.

The effects of lift-pot and lift-line velocities on the smoothness of lift-line operation were determined in the lift-pot test using coal. Lift-line velocities of 30, 35, and 40 ft/s and lift-pot velocities of 0.127, 0.182, 0.245, and 0.3 ft/s were tested.

The first test, Run HGD-2A, was made at a lift-pot velocity of 0.245 ft/s and at a lift-line velocity of 30 ft/s. The recorder tracings and the conditions used for this run are shown in Figure 34. As the ball valve in the



READING NO.	SOLIDS FLOW RATE, lb/hr	PRESSURE DROP	SCALE, in. H_2O
1	285	▬ ACROSS LOWER SECTION OF LIFT LINE	0-50
2	580		
3	1100	— ACROSS UPPER SECTION OF LIFT LINE	0-25
4	1685		

Figure 34. LIFT-LINE PRESSURE DROP FOR RUN HGD-2A

downcomer opened increasingly wider, the solids flow rate to the lift-pot bed increased, as did the solids flow into the lift line, thus causing a rise in the lift-line pressure-drop. Reading 4 was taken with the downcomer ball valve wide open. At this reading, the lift-line pressure drop fluctuated approximately ±0.1 inch of water from the average lift-line pressure drop reading. This fluctuation was approximately one-tenth of that observed in the tests with sand.

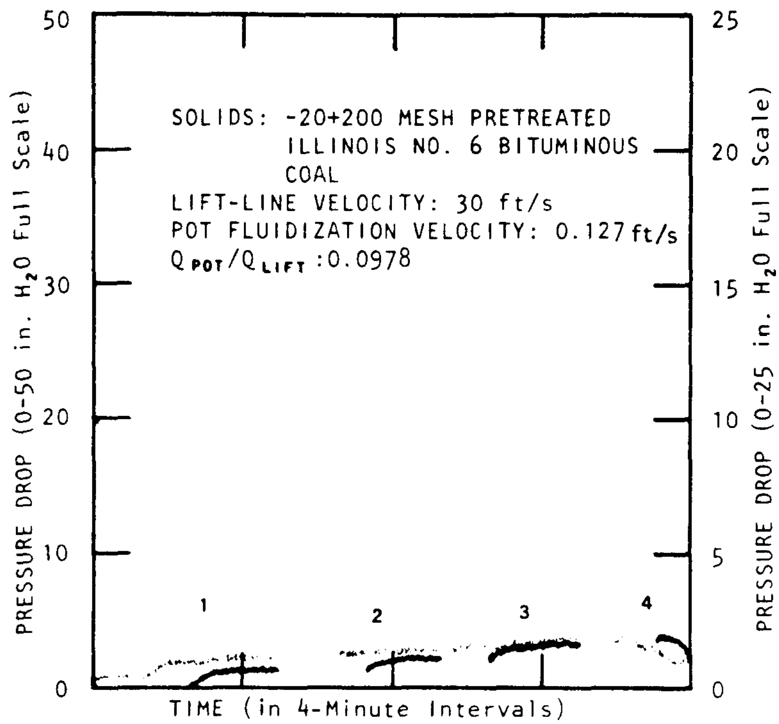
In Run HGD-2B, the lift-line velocity was kept at 30 ft/s, but the lift-pot velocity was reduced to 0.127 ft/s. The results obtained for this run are shown in Figure 35. In the first three readings, as the valve in the downcomer opened increasingly wider, the solids flow rate to the lift pot, and the lift-line pressure-drop, increased as expected. In Reading 4, however, the solids flow rate dropped. At this reading, the rate of solids flow to the lift-pot bed was greater than the rate at which they could be injected into the lift line. This occurred because the bed fluidization velocity was not high enough to transfer the solids to the lift line; consequently, the downcomer became packed below the ball valve, and the solids flow rate dropped to the value at which the bed could transfer solids to the lift line. The lift-line pressure drop fluctuations were about ±0.1 inch of water from the average pressure drop reading in this run.

In Run HGD-2C, a lift-line velocity of 30 ft/s was used once again, but this time with a lift-pot velocity of 0.3 ft/s. The results (Figure 36) were similar to those for Run HGD-2A; however, the amplitude of the lift-line fluctuations was somewhat higher than that in Run HGD-2A.

The lift-line velocity in Run HGD-2D (Figure 37) was increased to 35 ft/s, and the lift-pot velocity was set at 0.245 ft/s. At the maximum solids flow rate in this run, the entire downcomer was in streaming flow, and the lift-line fluctuations were twice as large as those in Run HGD-2A, which had the same lift-pot velocity, but a lower lift-line velocity.

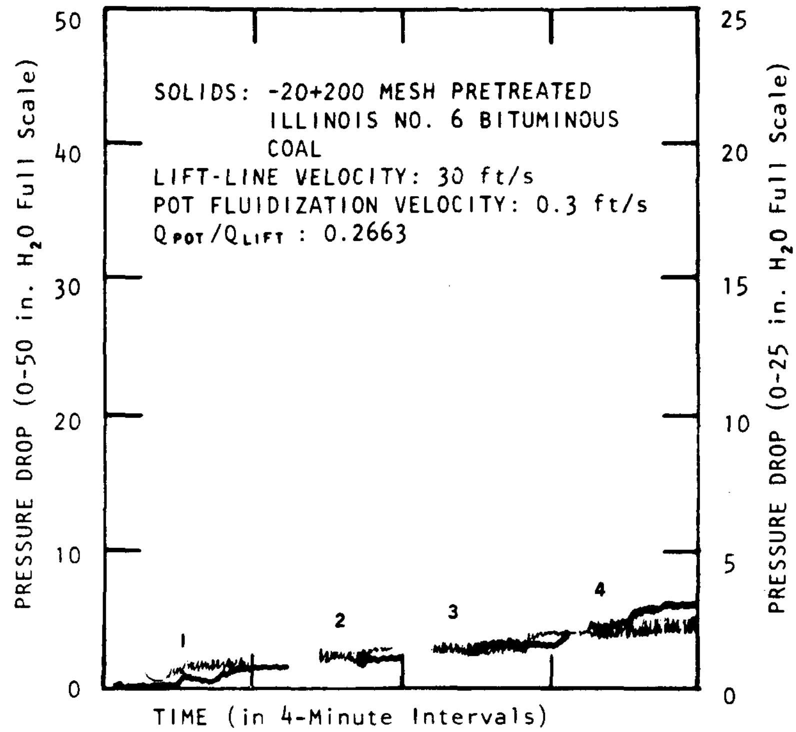
Run HGD-2E (Figure 38) was made at a lift-line velocity of 40 ft/s, while the lift-pot velocity was maintained at 0.245 ft/s. Lift-line fluctuations in this run were about the same as those in Run HGD-2D. When the ball-valve in the downcomer was fully opened, the entire downcomer became dilute, and the solids flow rate fell sharply.

3



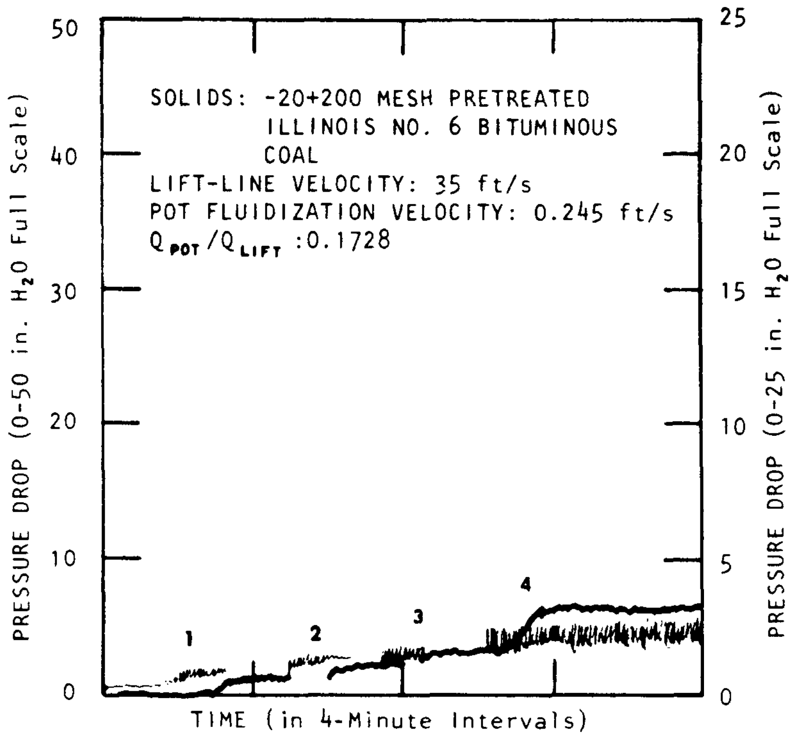
READING NO.	SOLIDS FLOW RATE, lb/hr		PRESSURE DROP	SCALE, in. H_2O
1	330	━━━	ACROSS LOWER SECTION	0-50
2	585		OF LIFT LINE	
3	1010	———	ACROSS UPPER SECTION	0-25
4	320		OF LIFT LINE	

Figure 35. LIFT-LINE PRESSURE DROP FOR RUN HGD-2B



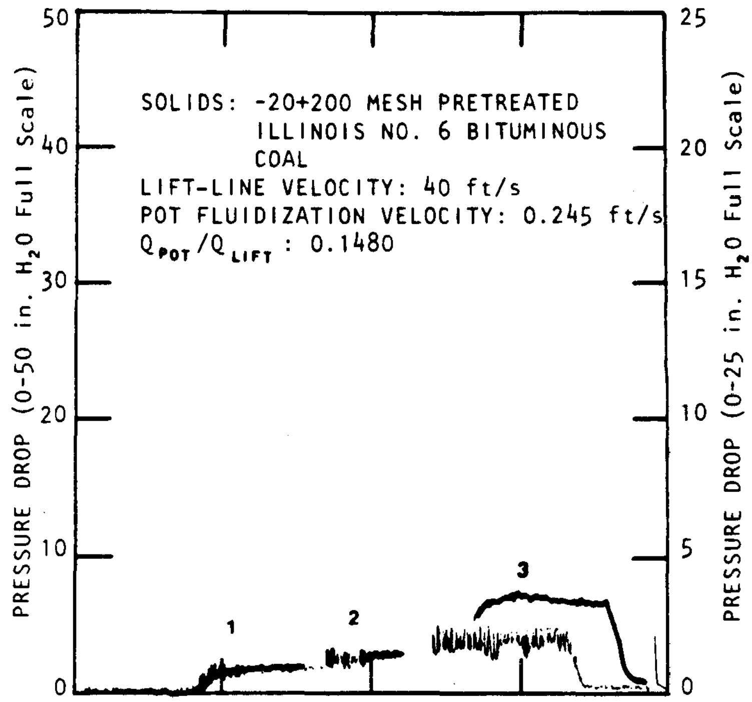
READING NO.	SOLIDS FLOW RATE, lb/hr		PRESSURE DROP	SCALE, in. H_2O
1	380	━━━	ACROSS LOWER SECTION OF LIFT LINE	0-50
2	575			
3	895	───	ACROSS UPPER SECTION OF LIFT LINE	0-25
4	1750			

Figure 36. LIFT-LINE PRESSURE DROP FOR RUN HGD-2C



READING NO.	SOLIDS FLOW RATE, lb/hr	PRESSURE DROP	SCALE, in. H_2O
1	275	——— ACROSS LOWER SECTION OF LIFT LINE	0-50
2	510		
3	805	——— ACROSS UPPER SECTION OF LIFT LINE	0-25
4	1530		

Figure 37. LIFT-LINE PRESSURE DROP FOR RUN HGD-2D



READING NO.	SOLIDS FLOW RATE, lb/hr		PRESSURE DROP	SCALE, in. H_2O
1	390	———	ACROSS LOWER SECTION OF LIFT LINE	0-50
2	645			
3	1570	———	ACROSS UPPER SECTION OF LIFT LINE	0-25

Figure 38. LIFT-LINE PRESSURE DROP FOR RUN HGD-2E

In Run HGD-2F (Figure 39), the lift-line velocity was set at 30 ft/s, and the lift-pot velocity was set at 0.182 ft/s. This combination of lift-line and lift-pot velocities gave results similar to those obtained in Run HGD-2A. Lift-line fluctuations were very small (±0.1 inch of water), and a stable downcomer flow pattern was maintained even at the wide-open ball valve position.

The results of the lift-pot tests with coal were somewhat different from those obtained with sand. With coal, the lowest lift-line velocity resulted in the smoothest lift-line operation; whereas, with sand, the highest lift-line velocity resulted in the smoothest lift-line operation. The latter results are probably due to the fact that the low lift-line velocities used in the sand tests were near the choking threshold, thus causing large pressure-drop fluctuations compared with the higher lift velocities.

Since coal is much lighter than sand, the lift velocities used with coal were not as close to choking. The results probably mean that there is an optimum lift-line velocity that will minimize the lift-line, pressure-drop fluctuations — one that is not too far from (nor too close to) choking.

With both materials, the lowest practical lift-pot fluidization velocity minimized lift-line pressure-drop fluctuations.

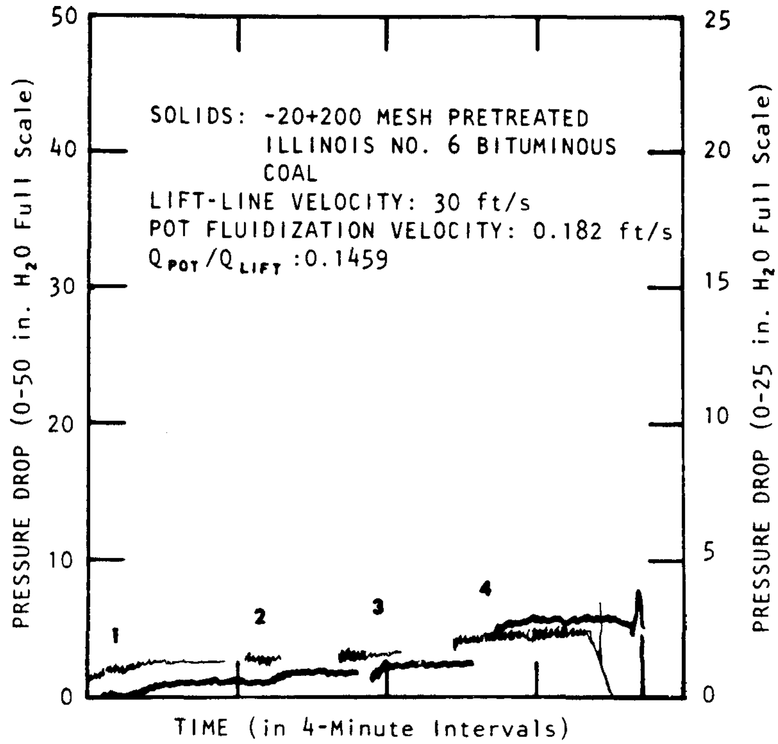
Lift-Pot II

Late in the quarter, a second lift-pot configuration (Lift-Pot II) was tested as a possible lift-line feeder configuration for the demonstration plant LTR. A drawing of this configuration is shown in Figure 40.

During operation, the solids passed down a 2-inch-diameter downcomer into the fluidized bed in the lift-pot. The solids flow rate was controlled by a ball valve in the lower section of the downcomer. The solids were in packed-bed flow above the ball valve, and in dilute-phase flow below it.

The solids were carried into the 2-inch-diameter lift line by the lift gas, which was routed into the lift line by a 3-inch-diamter lift-line feed tube (LLFT). The LLFT extended down several inches beneath the entrance to the lift line. As the solids passed upwards from the surface of the fluidized bed into the lift line, they formed a cone-shaped flow pattern.

The entire lift-pot configuration was constructed of Plexiglas and clear PVC pipe so that solids flow could be visually monitored. The dimensions of the Lift-Pot II configuration are shown in Figure 41.



READING NO.	SOLIDS FLOW RATE, lb/hr		PRESSURE DROP	SCALE, in. H_2O
1	325	—— (heavy)	ACROSS LOWER SECTION OF LIFT LINE	0-50
2	505			
3	690	—— (light)	ACROSS UPPER SECTION OF LIFT LINE	0-25
4	1720			

Figure 39. LIFT-LINE PRESSURE DROP FOR RUN HGD-2F

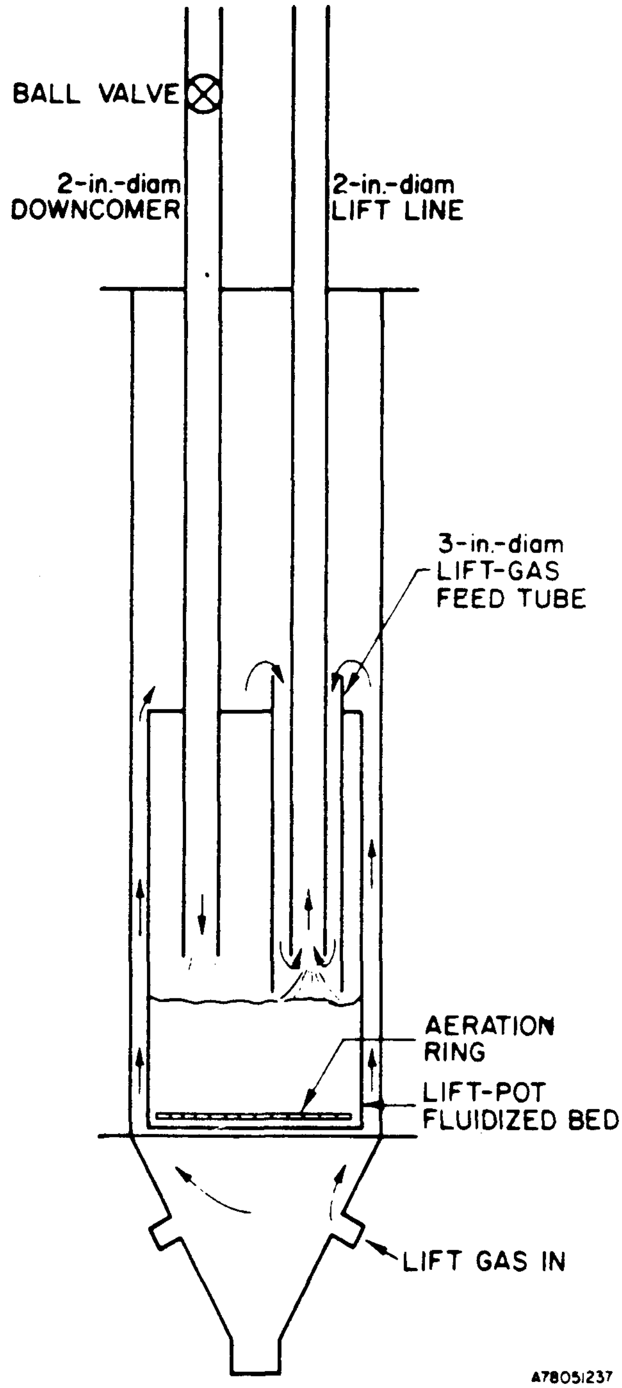


Figure 40. LIFT-POT II CONFIGURATION

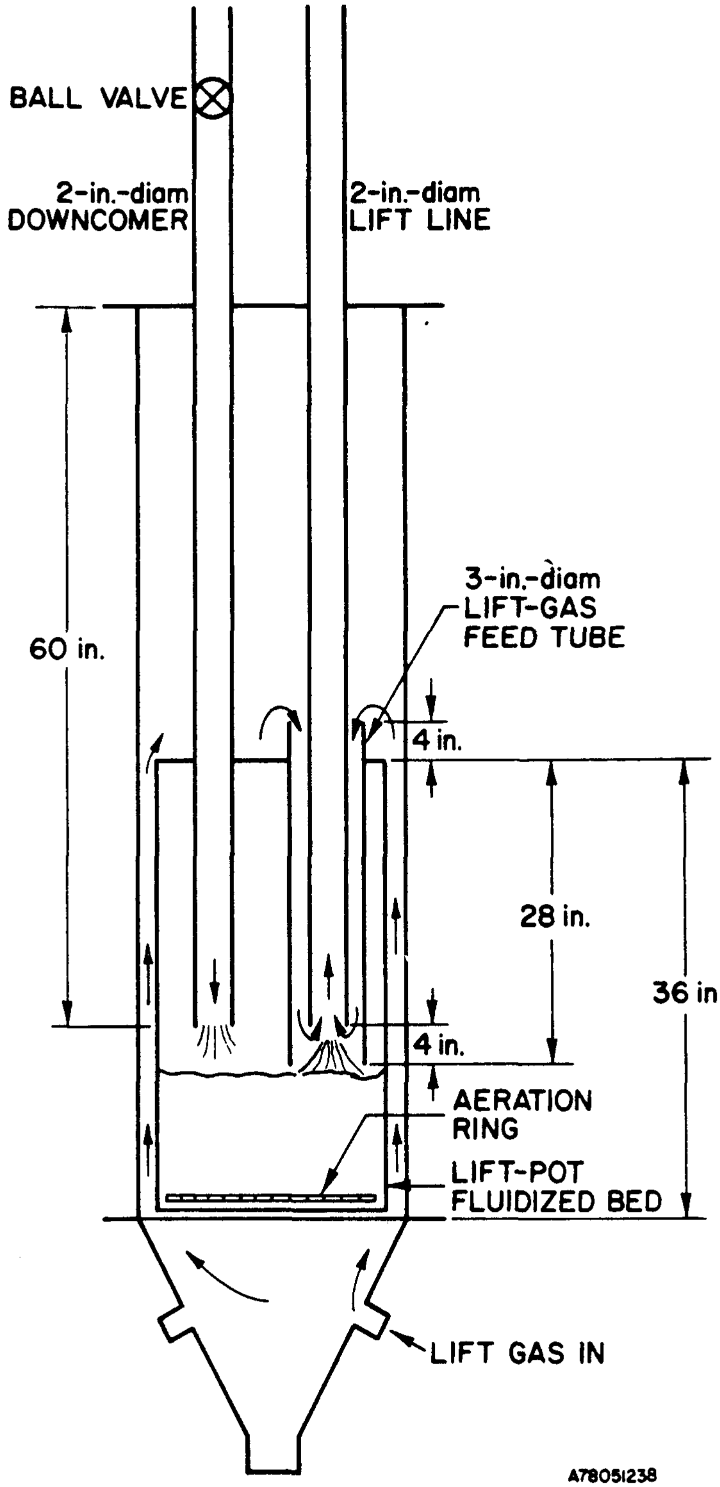


Figure 41. DIMENSIONS OF LIFT-POT II CONFIGURATION

The low-pressure test unit, as modified to test the Lift-Pot II configuration, is shown in Figure 42. In a typical run, the upper fluid bed is fluidized and the desired fluidization velocity set in the lift pot. The lift-line gas flow rate is then set. Readings are taken at several different solids flow rates and the results analyzed. Solids flow rates are determined by timing individual solid particles as they pass between two marks, 12-inches apart, in the clear PVC downcomer. The fluctuations in the recorder trace for the lean-phase lift-line pressure drop are used to analyze the smoothness of the lift-line's operation.

As with the other configuration tested, the Lift-Pot II configuration was operated using —20+80 mesh sand and —20+200 mesh pretreated Illinois No. 6 bituminous coal.

Six tests were conducted with the —20+80 mesh sand material. In Run HGD-7A (Figure 43), the lift-line velocity was set at 40 ft/s and the pot fluidization velocity at 0.4 ft/s. A maximum solids flow rate of 11,200 lb/hr was obtained with the controlling ball valve fully open. Fluctuations in the lift-line pressure were ±3-4 inches of water from the average pressure drop reading when the valve was fully open. Operation was relatively smooth and controllable.

In Runs HGD-7B and HGD-7C (Figures 44 and 45), the lift-pot fluidization velocity was maintained at 0.4 ft/s, and the lift-line velocity was set at 35 and 30 ft/s, respectively. The lift-line pressure-drop fluctuations in these runs were approximately the same as reported for Run HGD-7A. However, the maximum solids flow rate with the ball fully open for both runs was less than that obtained in Run HGD-7A. Thus, the solids flow rate up the lift line is a function of lift-line velocity.

Run HGD-7D was made with a lift-pot velocity of 0.5 ft/s and a lift-line velocity of 40 ft/s (Figure 46). Fluctuations in the lift-line pressure-drop trace were slightly better than those occurring in Run HGD-7A, while the maximum solids flow rate was approximately the same. The operation of the lift line was smooth and controllable.

Run HGD-7E (Figure 47) was made with a lift-line velocity of 35 ft/s and a lift-pot velocity of 0.3 ft/s. Very low solids flow rates were obtained, and the operation of the lift line was not as desirable as in the other runs. Considerable solids refluxing was observed in the lift line, and the lift pot was just barely fluidized.

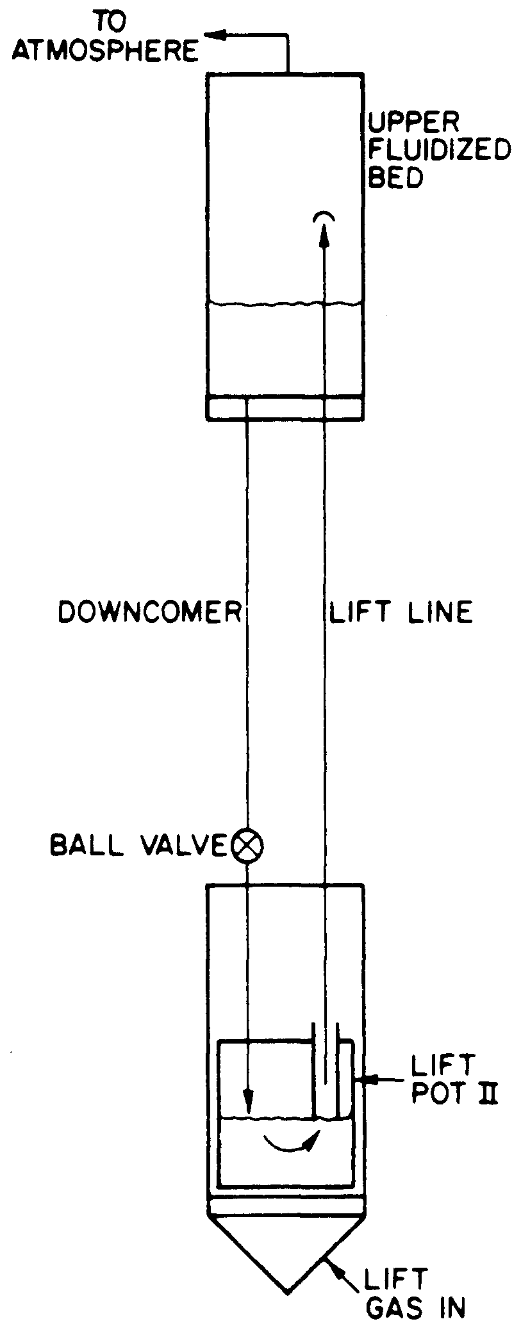
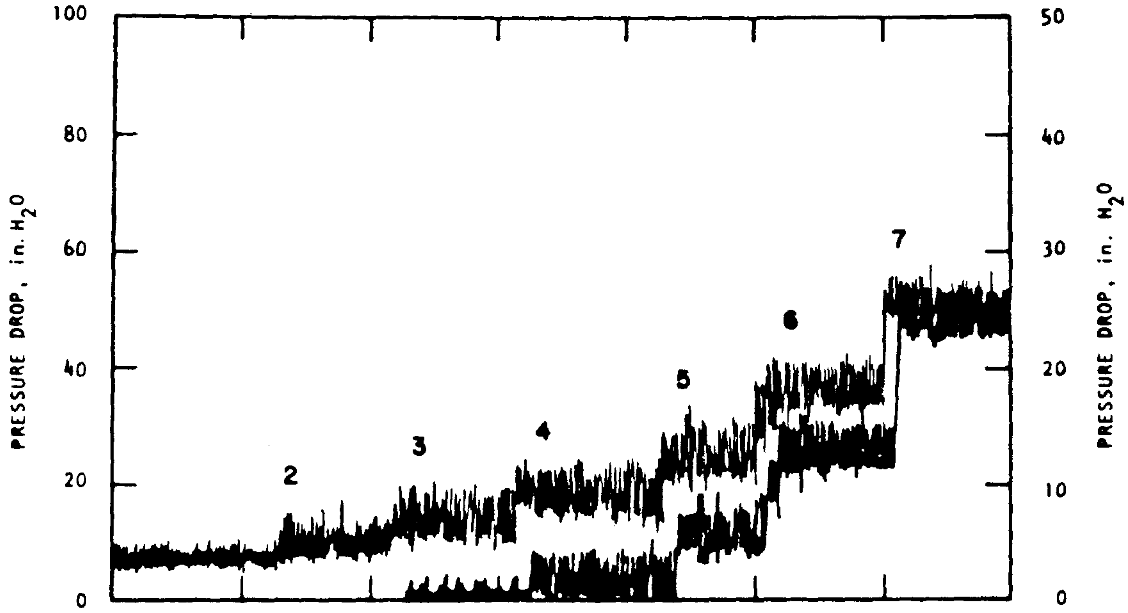


Figure 42. LIFT-POT II TEST LOOP



Lift-Line Velocity: 40 ft/s

Solids: Ottawa Sand,
-20 + 80 mesh

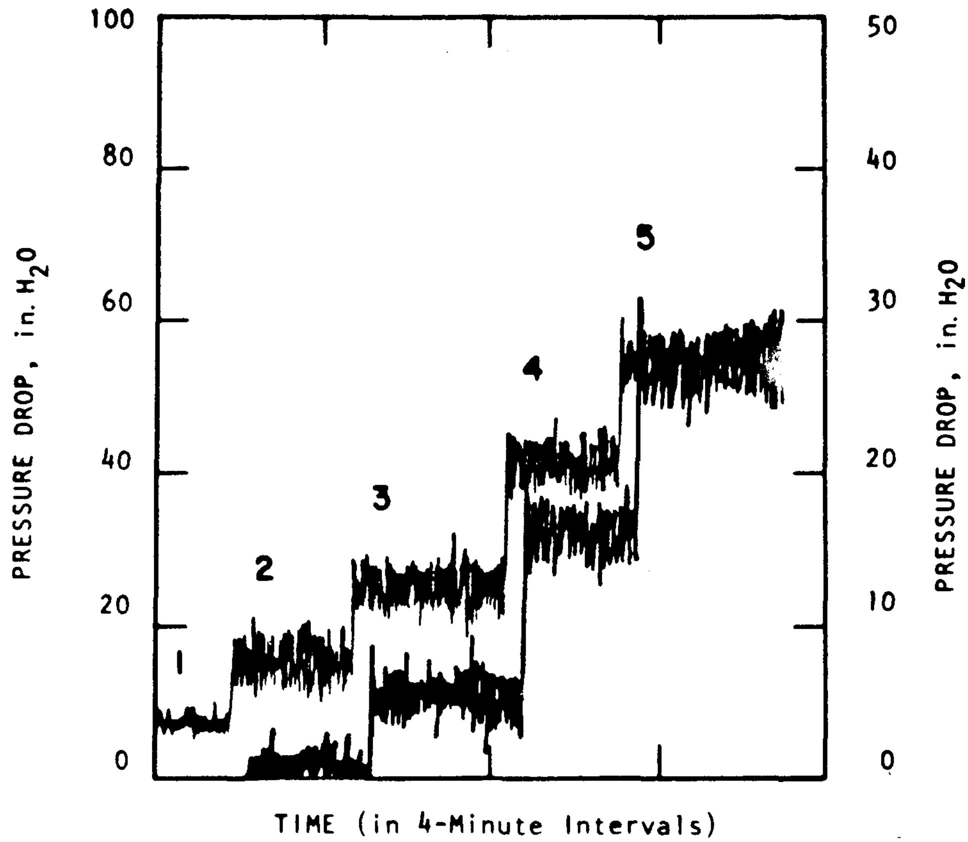
Lift-Pot Fluidization
Velocity: 0.4 ft/s

Reading Number	Solids Flow Rate, lb/hr
1	900
2	1,700
3	2,650
4	3,550
5	4,950
6	7,500
7*	11,200

Pressure Drop	Scale, in. H_2O
—— Across Lift Line	0 to 100
—— Across Downcomer	0 to 50

* Valve fully open

Figure 43. RESULTS AND CONDITIONS FOR RUN HGD-7A



Lift-Line Velocity: 35 ft/s

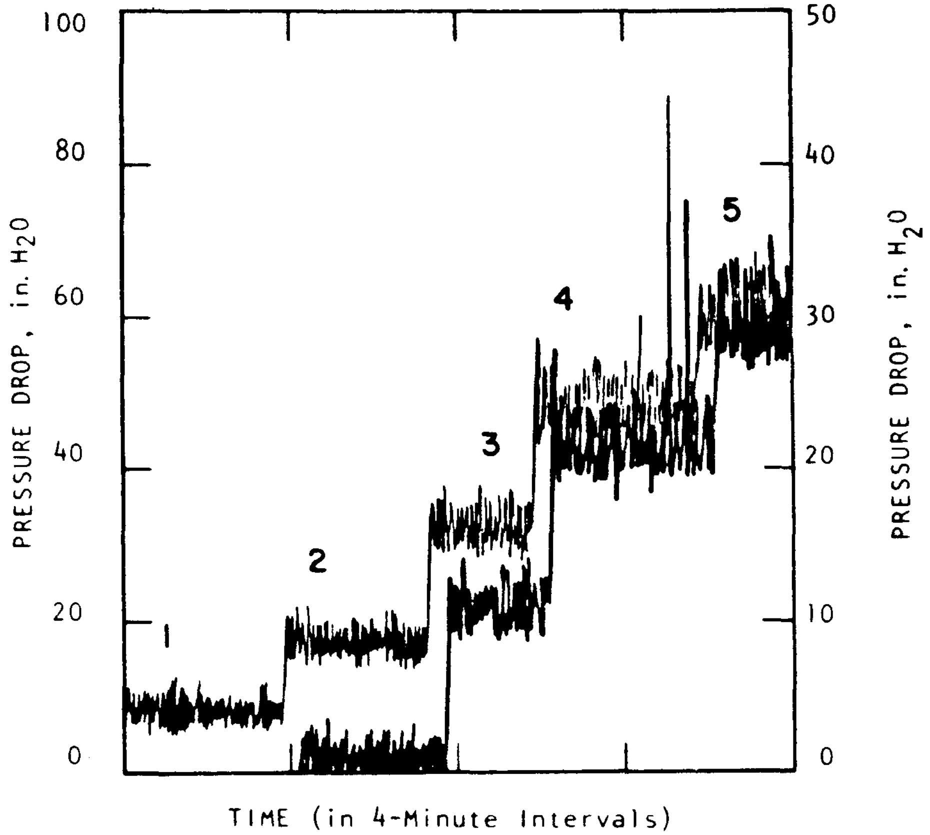
Solids: Ottawa Sand,
-20 + 80 mesh

Lift-Pot Fluidization
Velocity: 0.4 ft/s

Reading Number	Solids Flow Rate, lb/hr	Pressure Drop	Scale, in. H_2O
1	400	—— Across Lift Line	0 to 100
2	2,000		
3	4,000	—— Across Downcomer	0 to 50
4	7,200		
5*	9,800		

* Valve fully open

Figure 44. RESULTS AND CONDITIONS FOR RUN HGD-7B



Lift-Line Velocity: 30 ft/s

Solids: Ottawa Sand,
-20 + 80 mesh

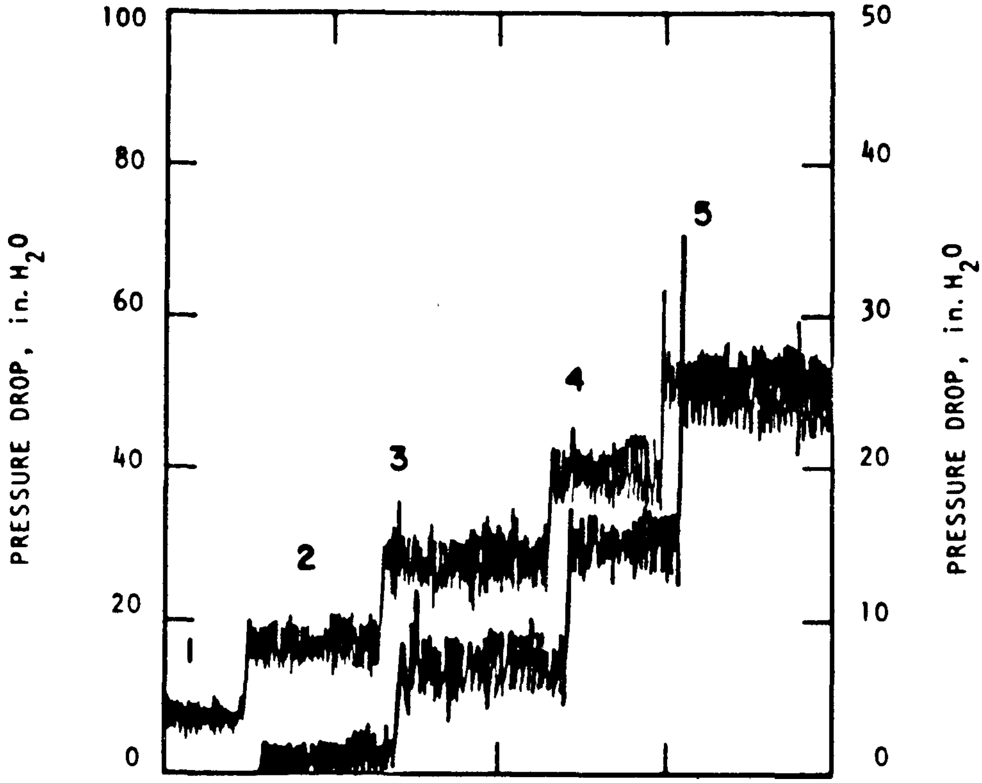
Lift-Pot Fluidization
Velocity: 0.4 ft/s

Reading Number	Solids Flow Rate, lb/hr
1	1,000
2	2,600
3	5,250
4	7,700
5*	9,350

Pressure Drop	Scale, in. H_2O
——Across Lift Line	0 to 100
——Across Downcomer	0 to 50

* Valve fully open

Figure 45. RESULTS AND CONDITIONS FOR RUN HGD-7C



Lift-Line Velocity: 40 ft/s

Solids: Ottawa Sand,
-20 + 80 mesh

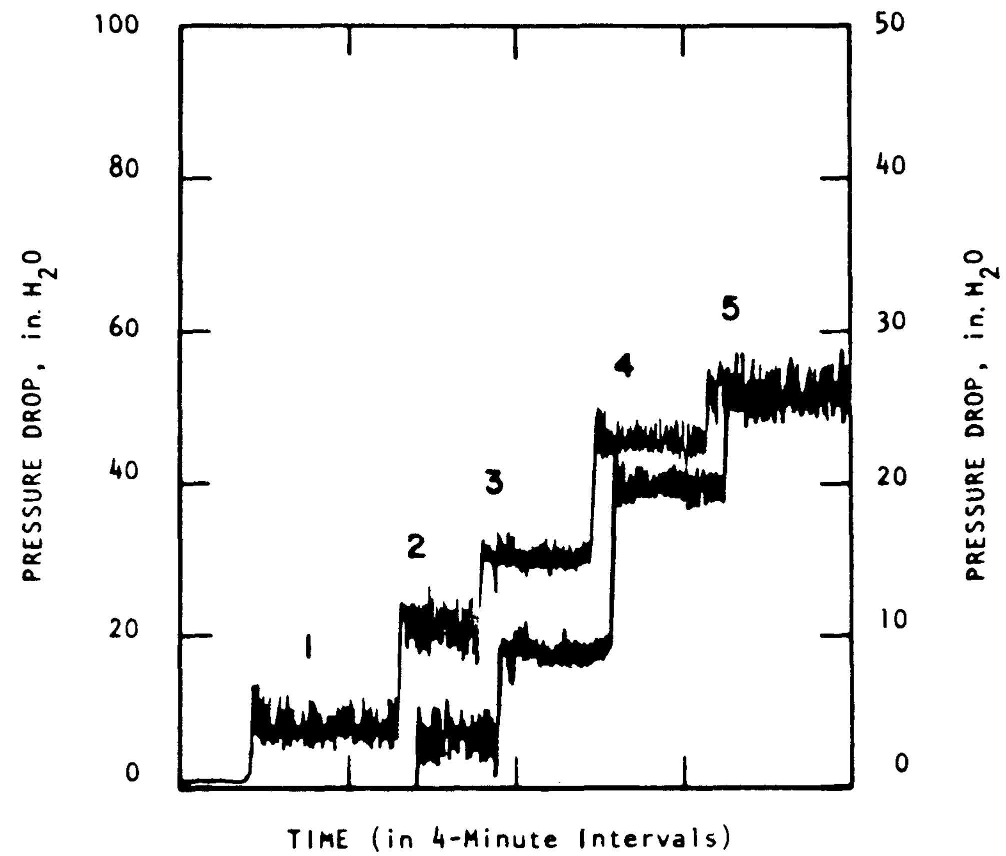
Lift-Pot Fluidization
Velocity: 0.5 ft/s

Reading Number	Solids Flow Rate, lb/hr
1	750
2	3,100
3	5,650
4	8,450
5*	11,000

	Pressure Drop	Scale, in. H_2O
——	Across Lift Line	0 to 100
——	Across Downcomer	0 to 50

* Valve fully open

Figure 46. RESULTS AND CONDITIONS FOR RUN HGD-7D



Lift-Line Velocity: 35 ft/s

Solids: Ottawa Sand,
-20 + 80 mesh

Lift-Pot Fluidization
Velocity: 0.3 ft/s

Reading Number	Solids Flow Rate, lb/hr
1	600
2	3,400
3	5,250
4	8,250
5*	9,400

Pressure Drop	Scale, in. H_2O
—— Across Lift Line	0 to 100
—— Across Downcomer	0 to 50

* Valve fully open

Figure 47. RESULTS AND CONDITIONS FOR RUN HGD-7E

In Run HGD-7F (Figure 48), the lift-pot velocity was set at 0.6 ft/s and the lift-line velocity at 35 ft/s. Pot fluidization was extremely good, but the lift-line velocity was low, and considerable refluxing in the lift line was observed.

The results of these tests have shown that this lift-pot feeder configuration can be made to work relatively well. It is necessary to have sufficient fluidization gas in the lift pot for good fluidization, and a sufficient lift-line velocity for efficient operation of the device because the solids flow rate to the lift line depends on these two velocities.

The Lift-Pot II configuration was also tested with −20+200 mesh pretreated Illinois No. 6 bituminous coal. Five tests were made.

In the first three tests, Runs HGD-8A, HGD-8B, and HGD-8C, the lift-line velocity was kept constant at 30 ft/s, while the pot velocity was 0.245, 0.3, and 0.182 ft/s, respectively. The results of these three runs are shown in Figures 49, 50, and 51. Operation was smooth and controllable for all three runs. The maximum solids flow rate obtained in all 3 runs was approximately 2400 lb/hr. Fluctuations of the lift-line pressure drop were about 1-3 inches of water from the average pressure-drop reading.

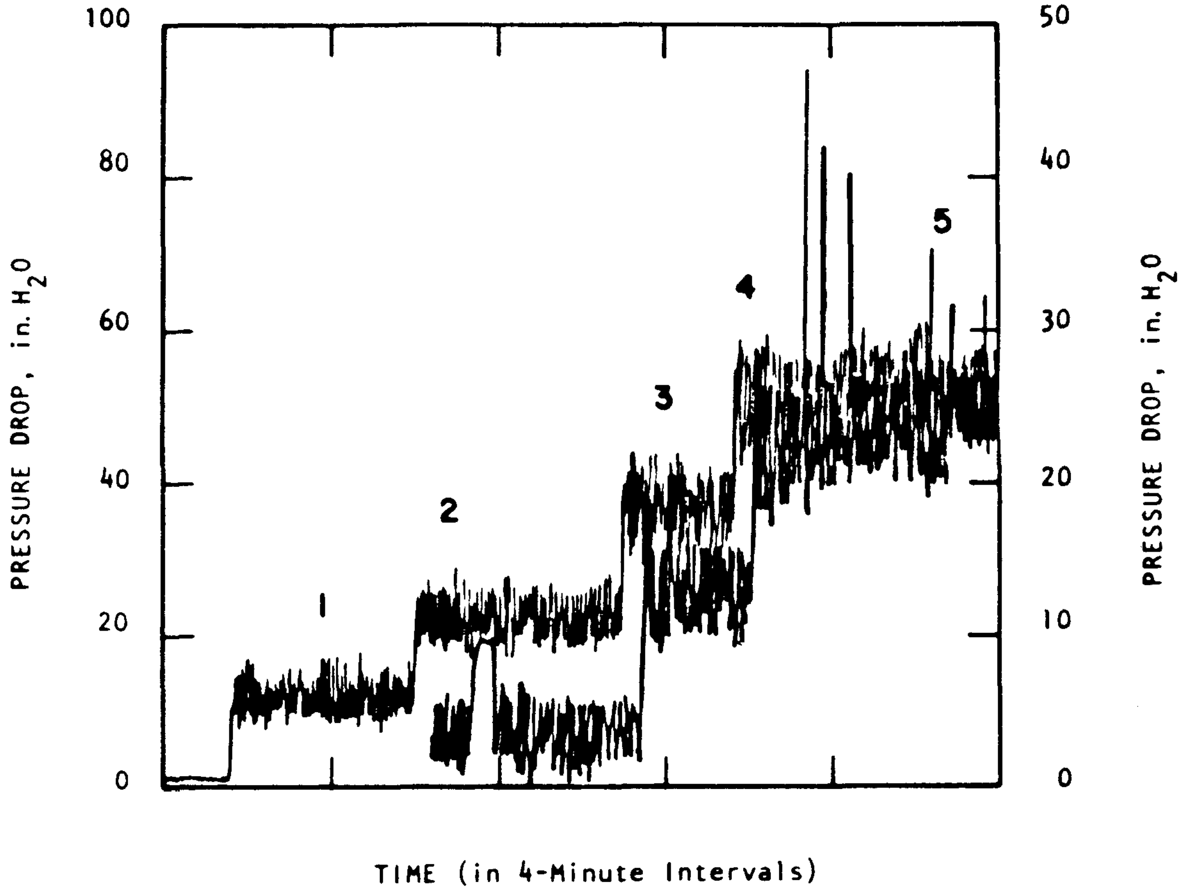
In the next two runs, HGD-8D and HGD-8E, the pot fluidization velocity was held constant at 0.245 ft/s and the lift-line velocity varied. Lift-line velocities were 35 and 40 ft/s, respectively. The results of these two runs are shown in Figures 52 and 53. Higher solids flow rates were obtained with two 30 ft/s lift-line runs than with the higher lift velocity runs.

From the results of these tests, it is evident that the effects of lift-line velocity and pot velocity on lift-pot operation is extremely important. In the tests using pretreated coal, the best results were obtained with a lift-line velocity of 30 ft/s. This is the velocity presently used in the LTR section of the HYGAS gasifier.

L-Valve LTR Feeder Device

A second lift-line feeder device, the L-valve, was also tested during the quarter. A sketch of this device is shown in Figure 54.

The L-valve was constructed of 2-inch-diameter PVC pipe so that the flow of solids through it could be observed. The flow was controlled using



Lift-Line Velocity: 35 ft/s

Solids: Ottawa Sand, -20 + 80 mesh

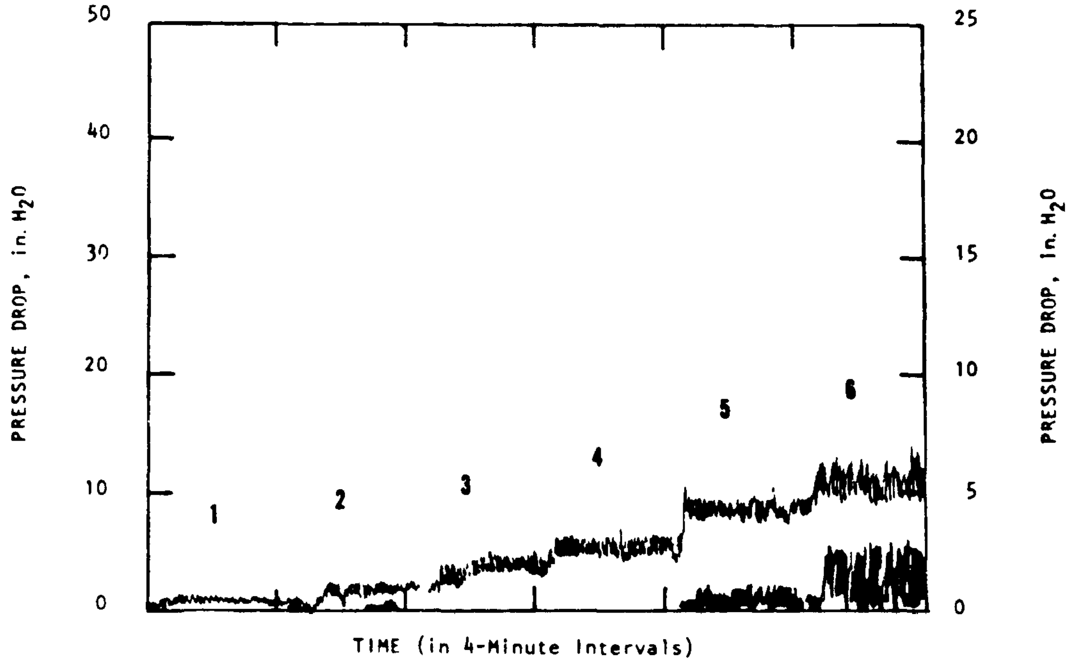
Lift-Pot Fluidization Velocity: 0.6 ft/s

Reading Number	Solids Flow Rate, lb/hr
1	1,400
2	3,600
3	6,600
4	9,350
5*	9,200

Pressure Drop	Scale, in. H_2O
—— Across Lift Line	0 to 100
—— Across Downcomer	0 to 50

* Valve fully open

Figure 48. RESULTS AND CONDITIONS FOR RUN HGD-7F



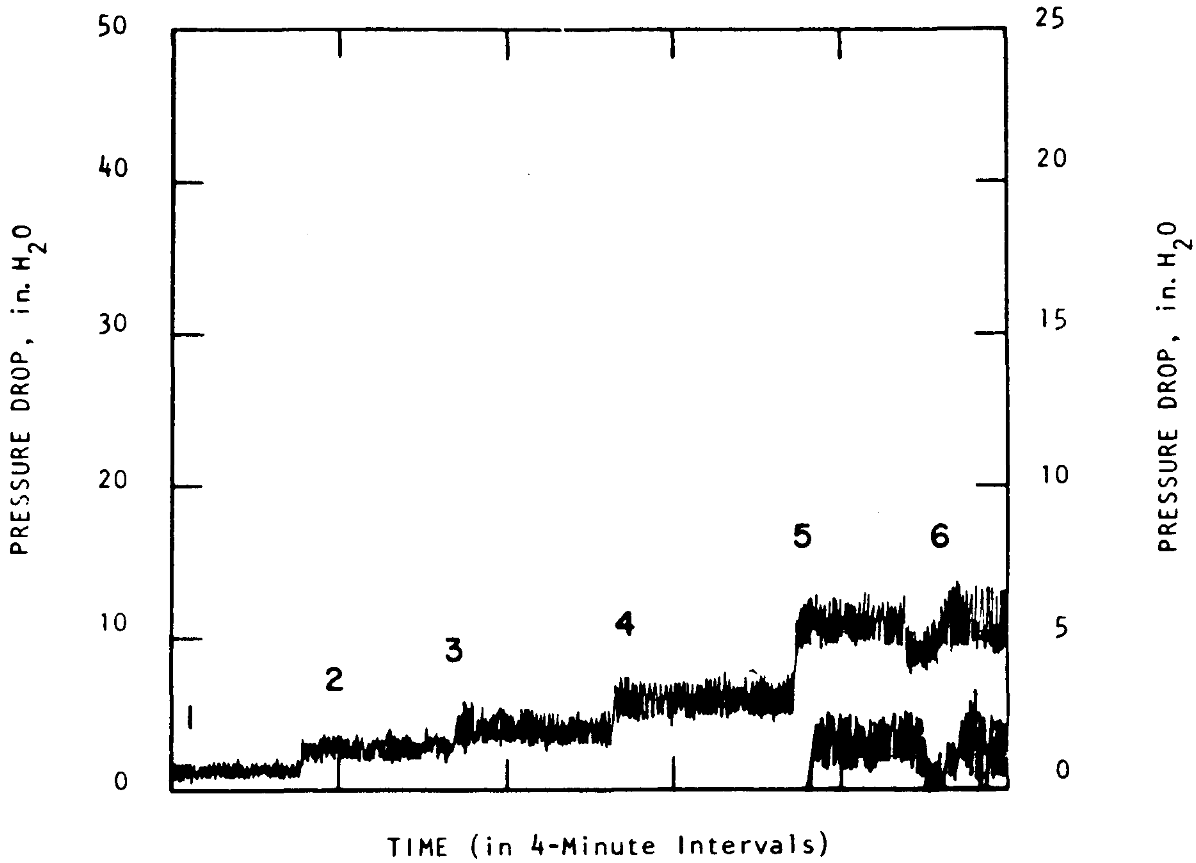
Lift-Line Velocity: 30 ft/s

Solids: Pretreated Illinois No. 6 Bituminou. Coal
-20 + 200 mesh

Lift-Pot Fluidization Velocity: 0.245 ft/s

Reading Number	Solids Flow Rate, lb/hr	Pressure Drop	Scale, in. H_2O
1	0		
2	230	—— Across Lift Line	0 to 50
3	700		
4	1,100	—— Across Downcomer	0 to 25
5	1,850		
6	2,430		

Figure 49. RESULTS AND CONDITIONS FOR RUN HGD-8A



Lift-Line Velocity: 30 ft/s

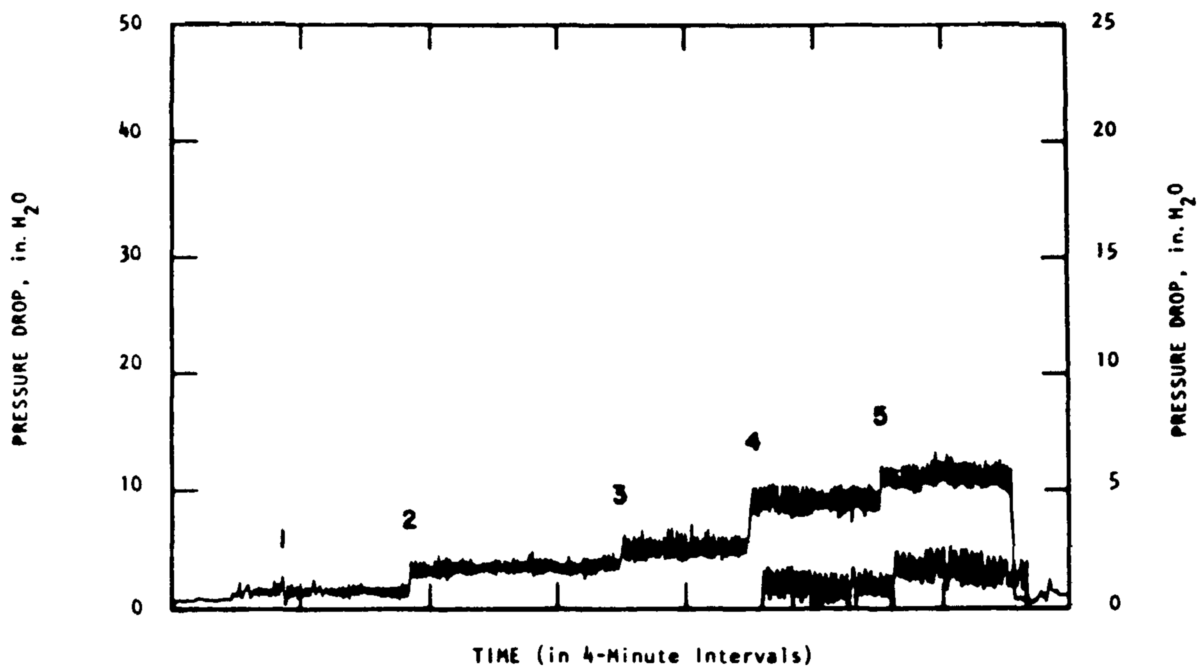
Solids: Pretreated Illinois No. 6 Bituminous Coal
-20 + 200 mesh

Lift-Pot Fluidization Velocity: 0.3 ft/s

Reading Number	Solids Flow Rate, lb/hr	Pressure Drop	Scale, in. H_2O
1	50		
2	520	Across Lift Line	0 to 50
3	700		
4	1,150	Across Downcomer	0 to 25
5	2,370		
6*	2,430		

* Valve fully open

Figure 50. RESULTS AND CONDITIONS FOR RUN HGD-8B



Lift-Line Velocity: 30 ft/s

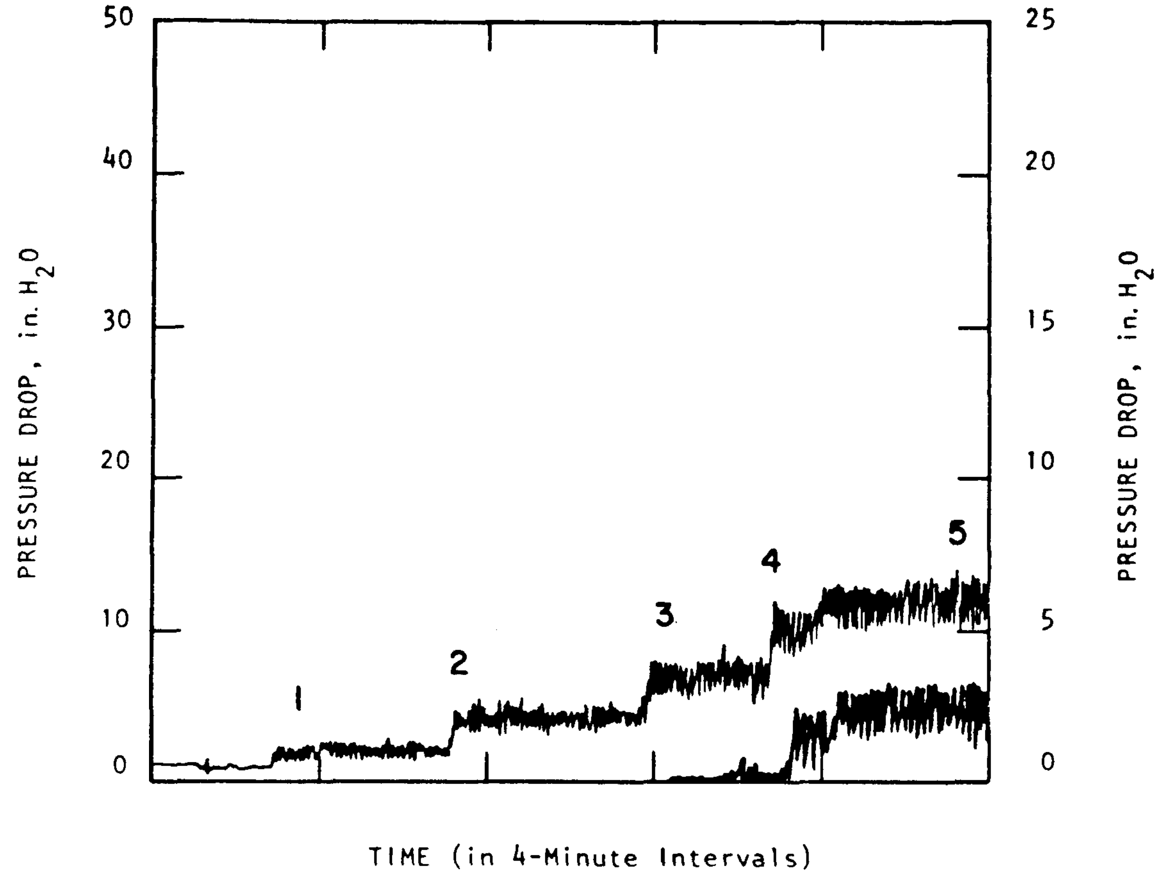
Solids: Pretreated Illinois No. 6 Bituminous Coal -20 + 200 mesh

Lift-Pot Fluidization Velocity: 0.182 ft/s

Reading Number	Solids Flow Rate, lb/hr	Pressure Drop	Scale, in. H_2O
1	110		
2	580	——Across Lift Line	0 to 50
3	920		
4	1,850	——Across Downcomer	0 to 25
5*	2,370		

* Valve fully open

Figure 51. RESULTS AND CONDITIONS FOR RUN HGD-8C



Lift-Line Velocity: 35 ft/s

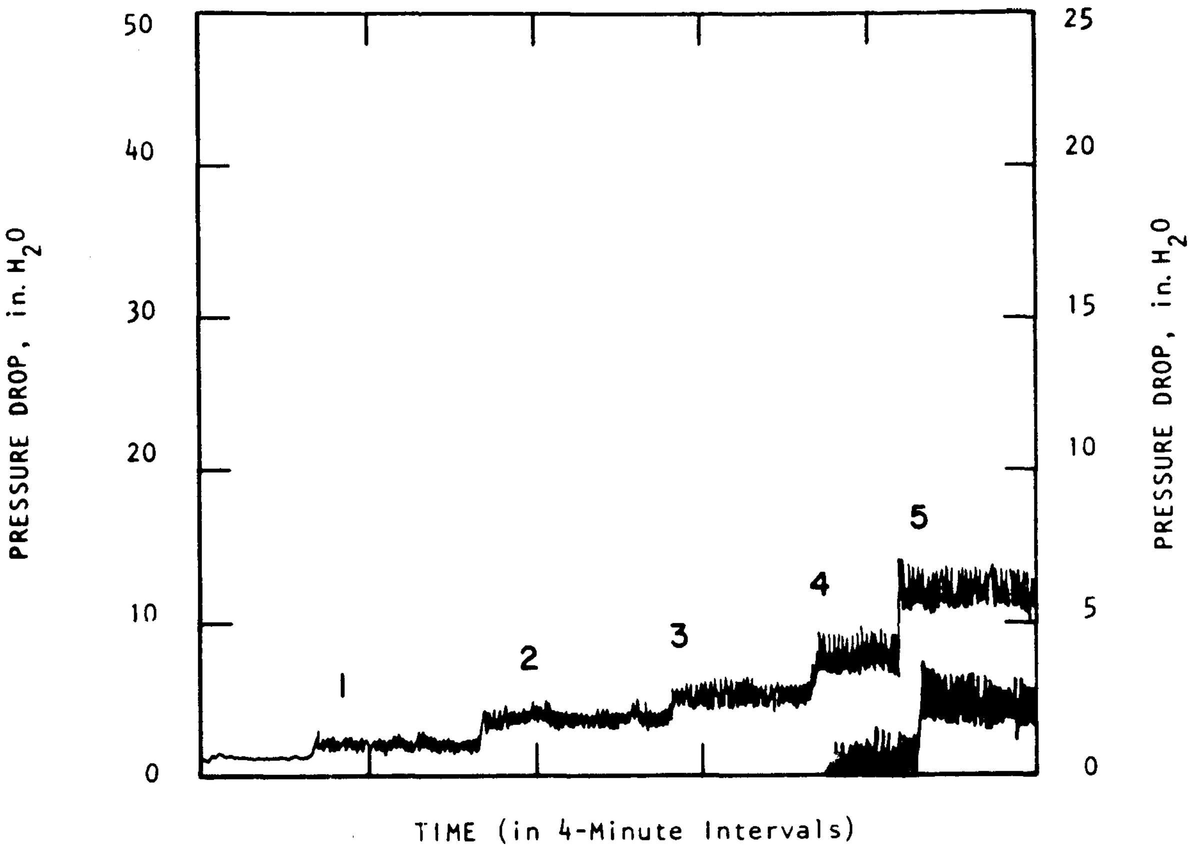
Solids: Pretreated Illinois No. 6 Bituminous Coal -20 + 200 mesh

Lift-Pot Fluidization Velocity: 0.245 ft/s

Reading Number	Solids Flow Rate, lb/hr	Pressure Drop	Scale, in. H_2O
1	190		
2	540	Across Lift Line	0 to 50
3	1,080		
4	1,610	Across Downcomer	0 to 25
5*	1,970		

* Valve fully open

Figure 52. RESULTS AND CONDITIONS FOR RUN HGD-8D



Lift-Line Velocity: 40 ft/s

Solids: Pretreated Illinois No. 6 Bituminous Coal
-20 + 200 mesh

Lift-Pot Fluidization Velocity: 0.245 ft/s

Reading Number	Solids Flow Rate, lb/hr	Pressure Drop	Scale, in. H_2O
1	100		
2	390	Across Lift Line	0 to 50
3	580		
4	1,060	Across Downcomer	0 to 25
5*	1,780		

* Valve fully open

Figure 53. RESULTS AND CONDITIONS FOR RUN HGD-8E

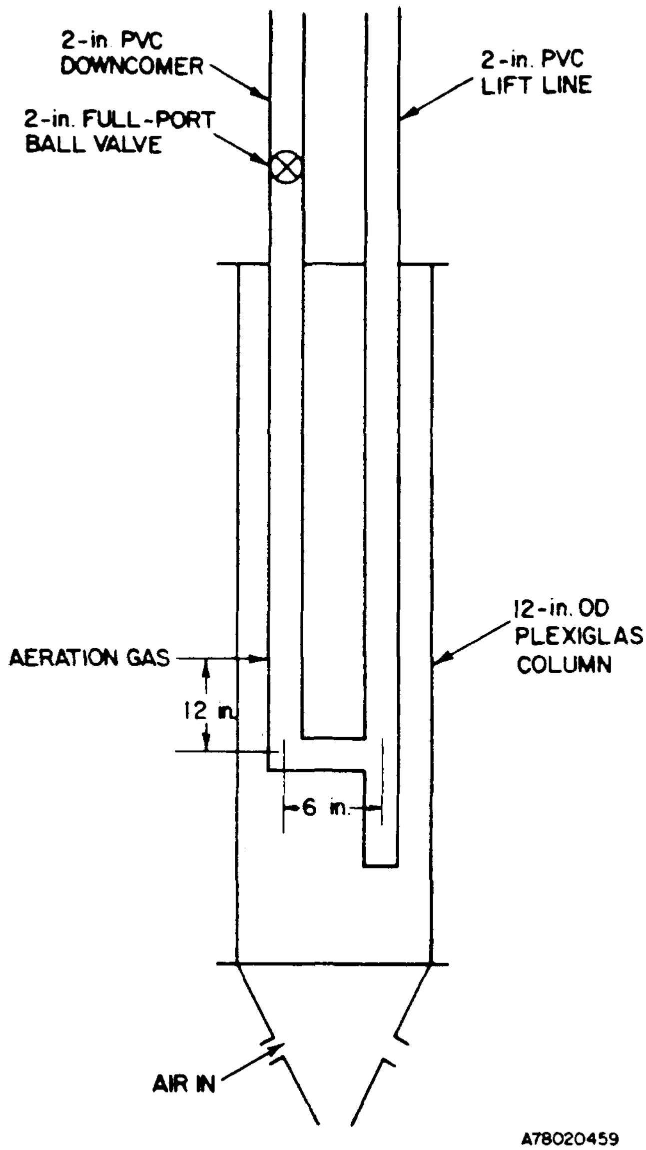


Figure 54. L-VALVE TEST CONFIGURATION

aeration gas supplied to the valve at a point 12-inches above the centerline of the horizontal sections.

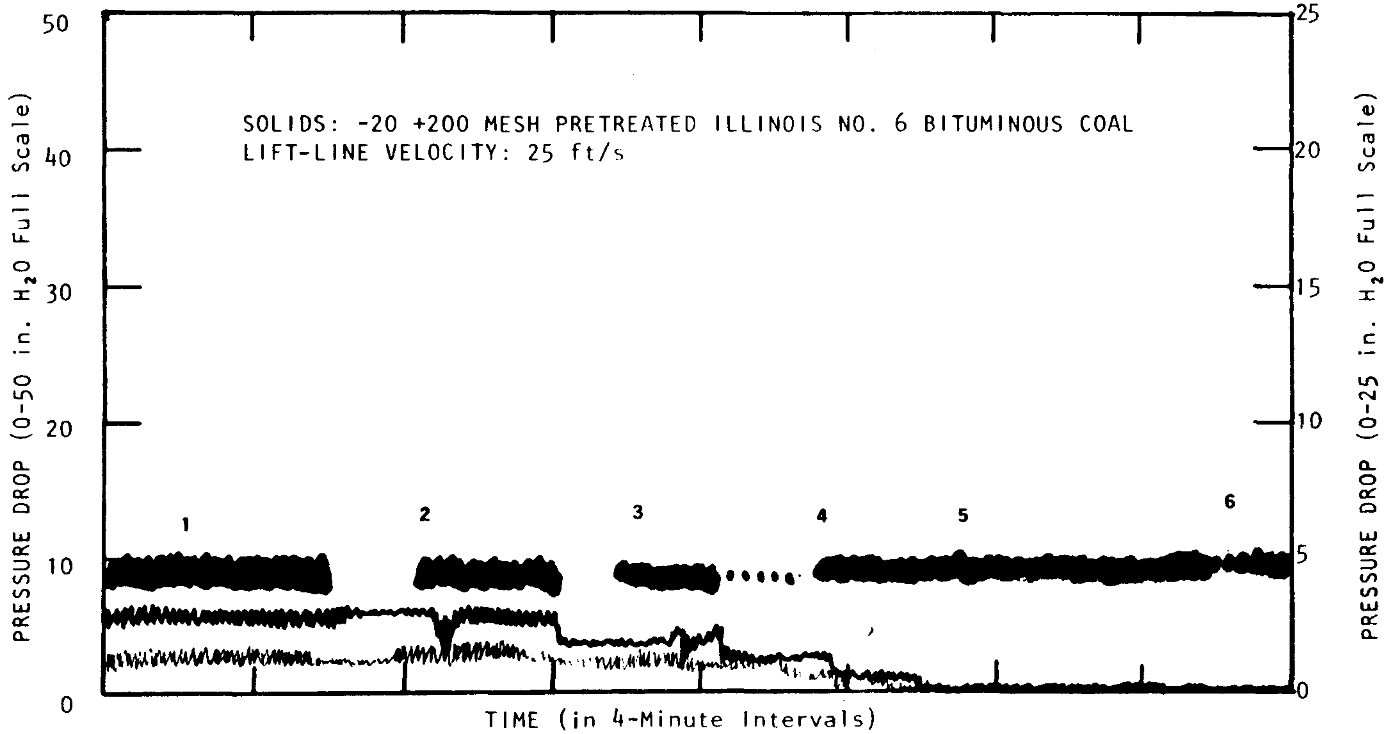
In a typical run, the upper bed of solids is first fluidized. The ball valve in the downcomer is then fully opened, and the solids flow rate into the lift-line is metered by controlling the amount of aeration gas fed to the L-valve. The solids flow rate is determined by timing particles as they pass between two points, 12-inches apart, on the downcomer. Lift-line pressure-drop readings are taken at several solids flow rates. The first series of tests involving the L-valve was conducted using −20+200 mesh pretreated Illinois No. 6 bituminous coal.

In Run HGD-3A, the lift-line velocity was set at 25 ft/s. The lift-line pressure-drop recorder traces and the run conditions used are shown in Figure 55. In this run the solids flow rate into the lift line was increased incrementally by increasing the aeration gas flow to the L-valve. Downcomer operation was relatively smooth, and the maximum lift-line pressure-drop fluctuations were approximately ±0.1 inch of water from the average pressure-drop reading.

In Runs HGD-3B through HGD-3D, the lift-line velocity was set at 30, 35, and 40 ft/s, respectively. The results are shown in Figures 56, 57, and 58. In all of these runs, the solids flow rate was increased to approximately 1100 to 1200 lb/hr using only L-valve aeration. Attempts to further increase the solids flow rate only diluted it in the downcomer, and also increased the lift-line pressure-drop fluctuations. The L-valve was also tested using −20+80 mesh Ottawa sand as the solids. The procedure used for these tests was identical to that used for the coal tests.

Initially, the lift velocity was set at 30 ft/s, and sand was fed through the L-valve to the lift line. At this velocity, however, some of the sand "dropped" through the short lift-line section immediately below the L-valve. This also occurred at a lift velocity of 35 ft/s.

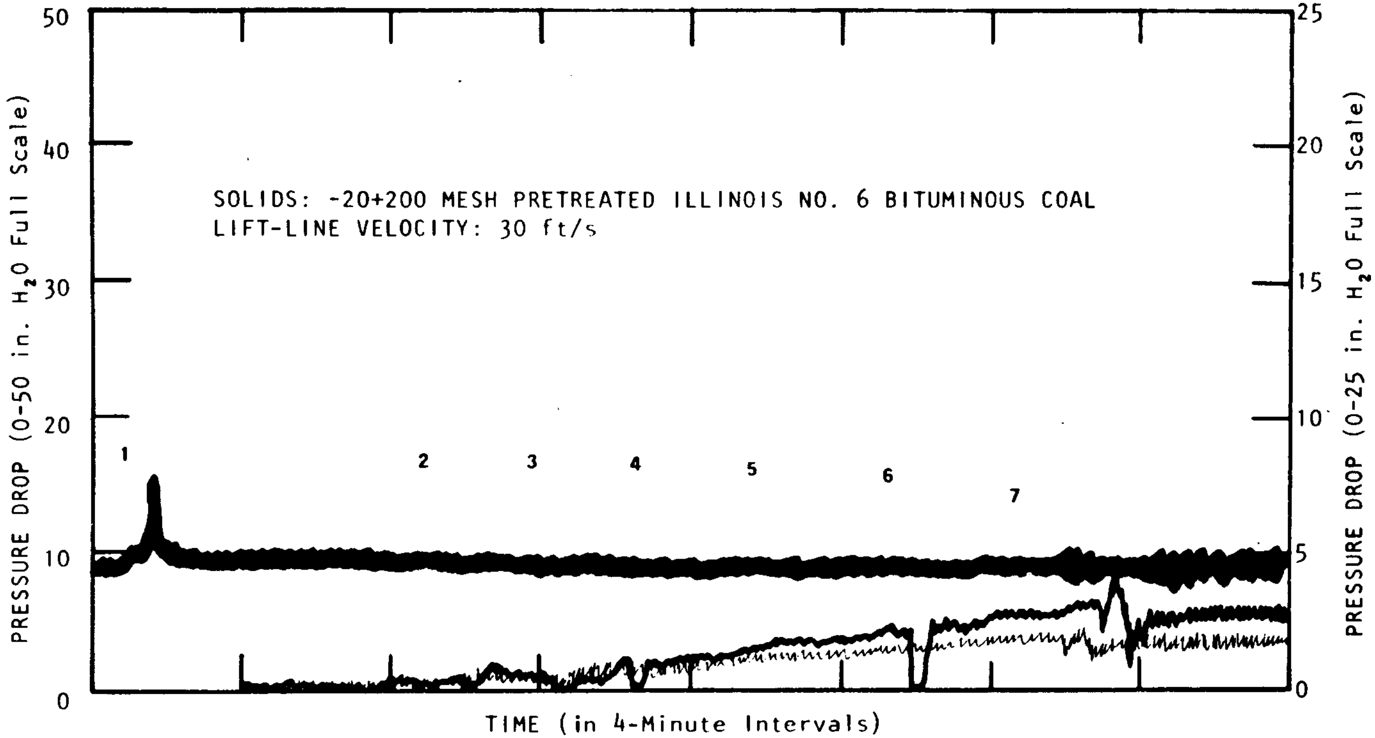
The first L-valve test using sand (Run HGD-4A) was made with a lift-line velocity of 40 ft/s. The results of this run are shown in Figure 59. L-valve operation was controllable up to a solids flow rate of about 7600 lb/hr. The solids flow rate could be increased beyond this rate, but the downcomer flow eventually became dilute. The maximum pressure-drop fluctuations in the



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min
1	0	0.1988
2	235	0.3959
3	760	0.6375
4	830	0.7073
5	1010	0.7815
6	1250	0.9737

PRESSURE DROP	SCALE, in. H_2O
ACROSS DOWNCOMER	0-50
ACROSS LOWER SECTION OF LIFT LINE	0-50
ACROSS UPPER SECTION OF LIFT LINE	0-25

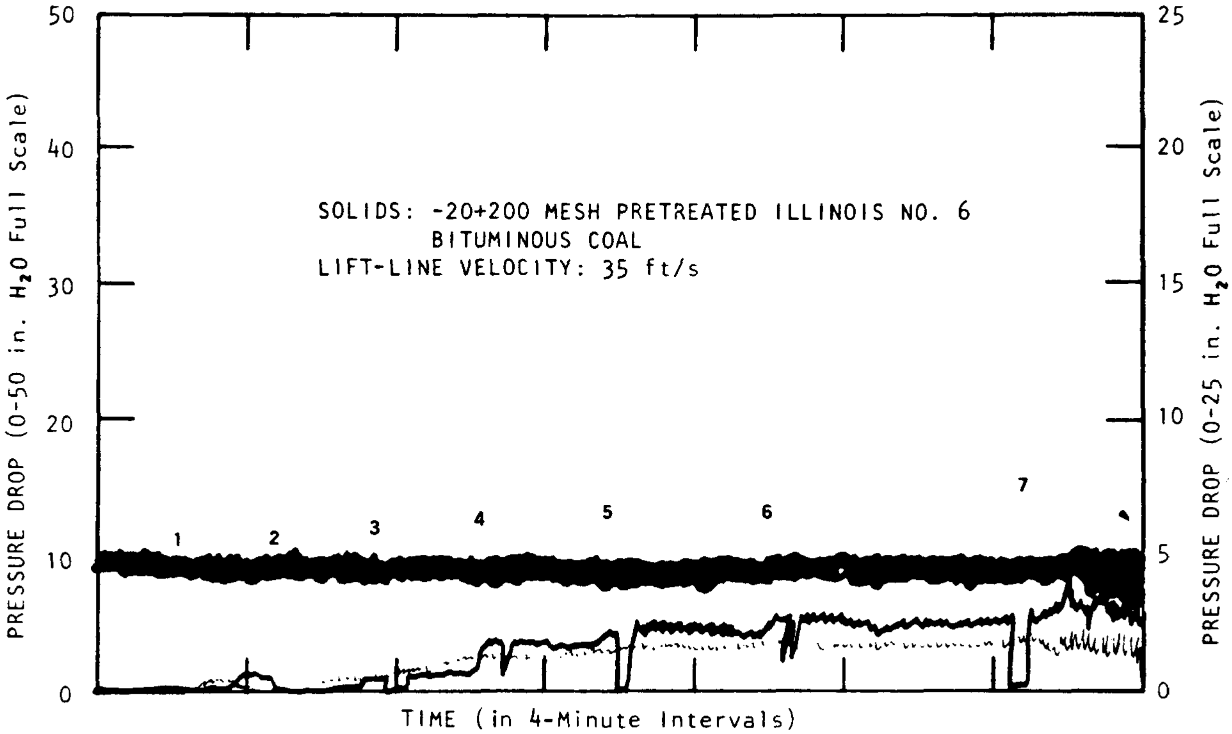
Figure 55. LIFT-LINE PRESSURE DROP FOR RUN HGD-3A



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min
1	0	0.1971
2	0	0.3197
3	230	0.4402
4	460	0.5443
5	920	0.7041
6	1140	0.7753
7	1510	0.9666

PRESSURE DROP	SCALE, in. H_2O
ACROSS DOWNCOMER	0-50
ACROSS LOWER SECTION OF LIFT LINE	0-50
ACROSS UPPER SECTION OF LIFT LINE	0-25

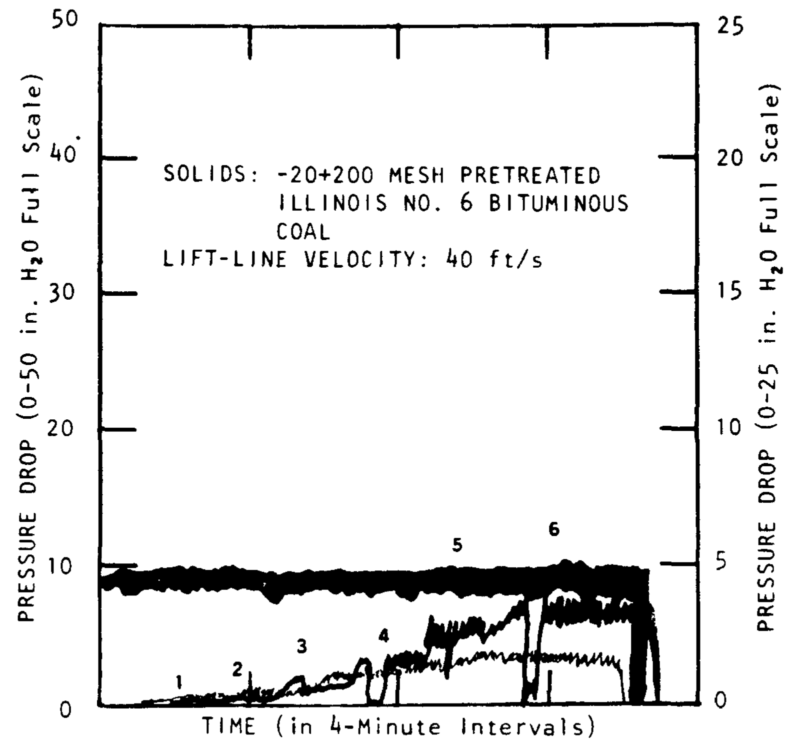
Figure 56. LIFT-LINE PRESSURE DROP FOR RUN HGD-3B



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min
1	0	0.1985
2	110	0.3414
3	450	0.5129
4	800	0.6775
5	1100	0.8436
6	1210	1.0117
7	1380	1.1770

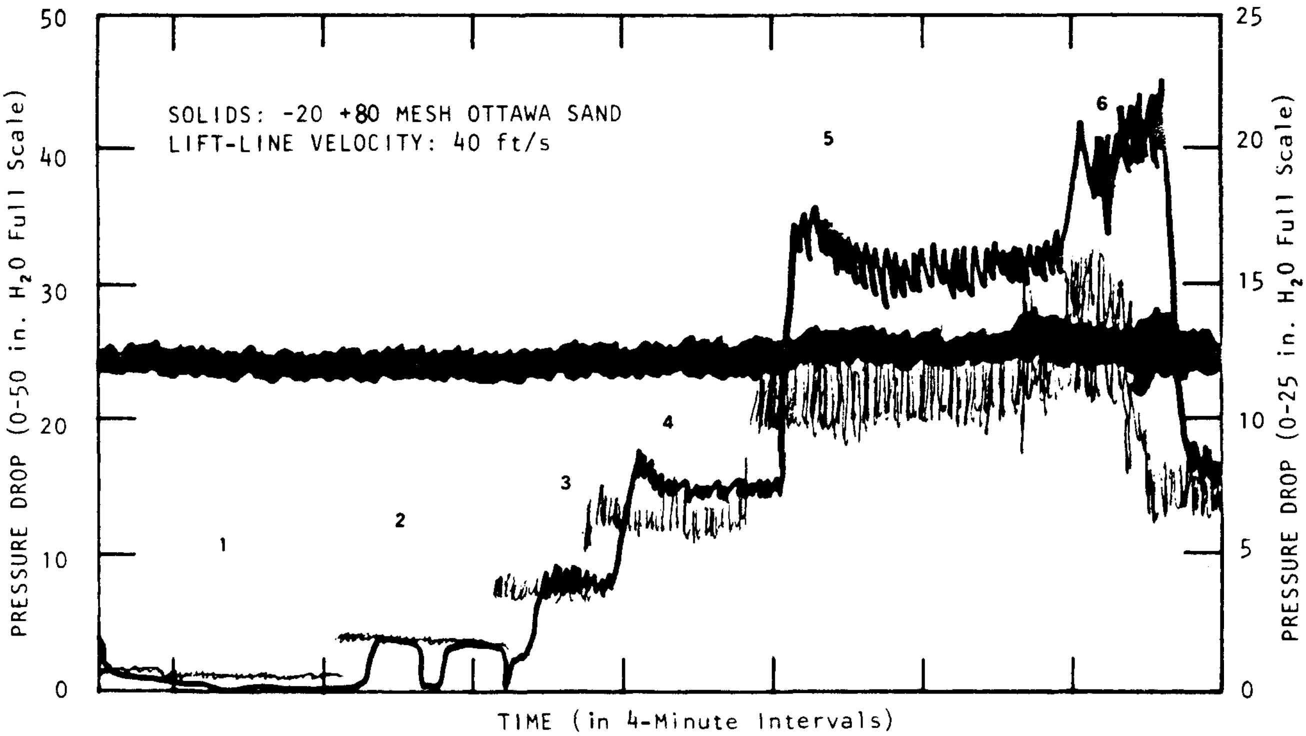
PRESSURE DROP	SCALE, in. H_2O
ACROSS DOWNCOMER	0-50
ACROSS LOWER SECTION OF LIFT LINE	0-50
ACROSS UPPER SECTION OF LIFT LINE	0-25

Figure 57. LIFT-LINE PRESSURE DROP FOR RUN HGD-3C



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min	PRESSURE DROP	SCALE, in. H_2O
1	0	0.1950	ACROSS DOWNCOMER	0-50
2	100	0.3399	ACROSS LOWER SECTION OF LIFT LINE	0-50
3	425	0.5075		
4	700	0.6495	ACROSS UPPER SECTION OF LIFT LINE	0-25
5	1020	0.8409		
6	1280	1.0032		

Figure 58. LIFT-LINE PRESSURE DROP FOR RUN HGD-3D



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min	PRESSURE DROP	SCALE in. H_2O
1	0	0.3367	ACROSS DOWNCOMER	0-50
2	200	0.4290	ACROSS LOWER SECTION OF LIFT LINE	0-50
3	1,500	0.6609		
4	3,250	0.8890	ACROSS UPPER SECTION OF LIFT LINE	0-25
5	7,600	1.3530		
6	10,000	1.3530		

Figure 59. LIFT-LINE PRESSURE DROP FOR RUN HGD-4A

lift-line were approximately ± 1.25 inches of water above the average pressure-drop readings. This was approximately one-half the fluctuation obtained with the lift-pot. Thus, the L-valve feeder configuration resulted in smoother lift-line operation than did the lift-pot feeder.

Runs HGD-4B and HGD-4C were made at lift-line velocities of 45 and 50 ft/s, respectively. As with the first run, both of these runs gave excellent solids flow control up to a solids flow rate of about 8000 lb/hr before the downcomer flow became dilute. Fluctuations in the lift-line were also approximately the same as in Run HGD-4A. The results of these two runs are shown in Figures 60 and 61.

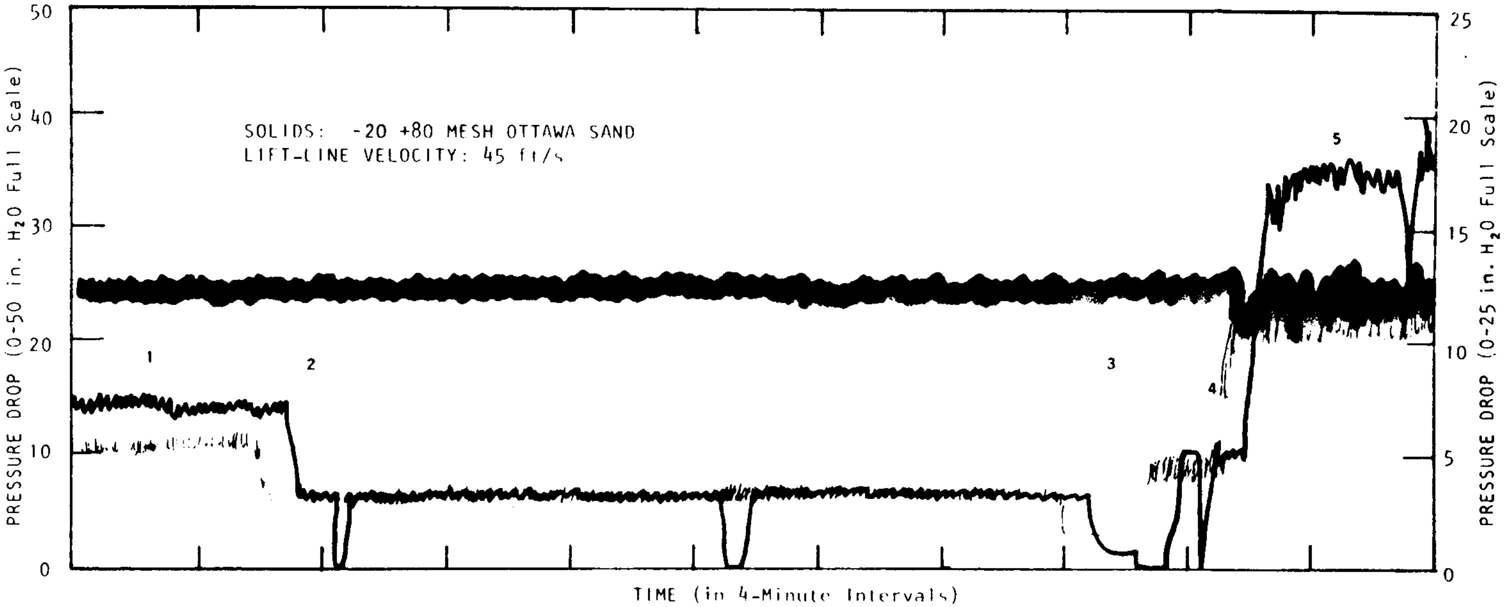
The L-valve controlled the solids flow rate extremely well. The lift-line pressure-drop fluctuations were smoother than those observed with the lift pot. Also, the reversion of the downcomer to dilute-phase operation at high flow rates is not a characteristic of the L-valve itself. This type of downcomer flow occurred because the solids could be made to flow through the L-valve faster than they could pass through the opening at the top of the downcomer. This resulted in streaming, or dilute-phase flow, in the downcomer.

The L-valve is also a much simpler device than the lift pot. It has no moving parts, and only aeration is used to control the solids, whereas, the lift pot requires a mechanical valve to control the solids flow rate out of the downcomer into the lift pot. The amount of gas needed to fluidize the lift-pot area is 4 to 10 times greater than that needed to fluidize the L-valve. The lift-pot device has an important advantage: It can be constructed without an expansion joint.

Reverse-Seal Leg LTR Feeder Device

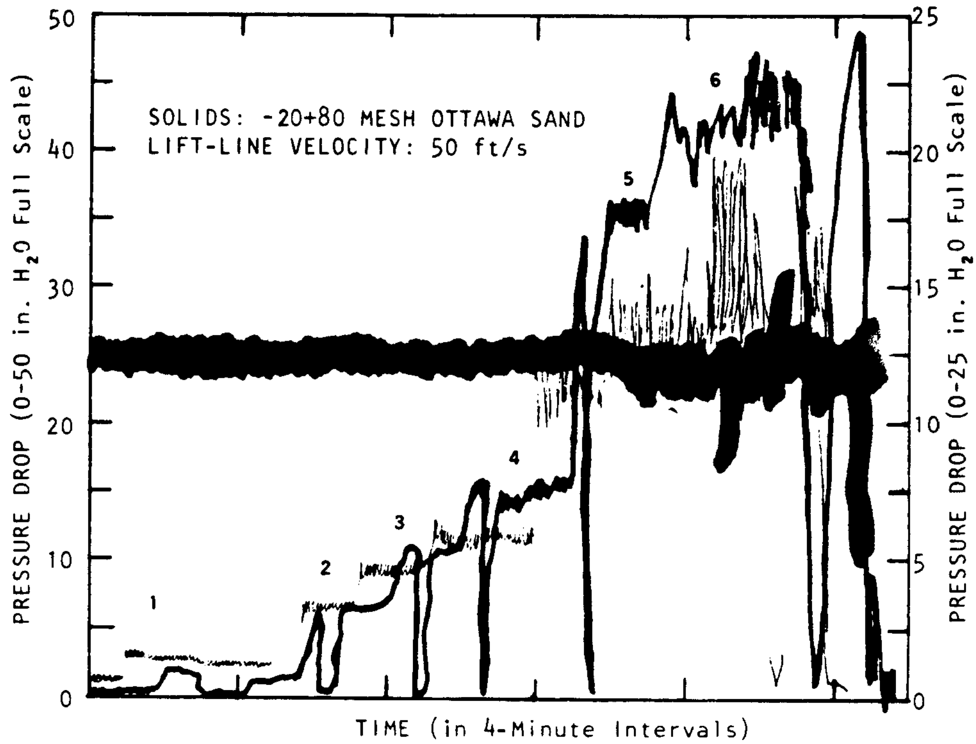
A typical reverse seal and its dimensions are shown in Figures 62 and 63, respectively. In the reverse seal, solids pass down a downcomer containing a solids control valve. They then pass through a 45-degree lateral section and around a sharp 135-degree return bend into a dense-phase lift (DPL) section before being injected into a lean-phase lift section.

The reverse seal can be operated in two modes: 1) aeration-control and 2) valve-control. When aeration is the means of control, aeration gas is fed to the reverse seal before the 135-degree return bend. The aeration causes the solids to flow along the bend, and controls the solids flow rate. In



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min	PRESSURE DROP	SCALE in. H_2O
1	3540	0.8687	▬ ACROSS DOWNCOMER	0-50
2	1350	0.5615	—— ACROSS LOWER SECTION OF LIFT LINE	0-50
3	350	0.3328	—— ACROSS LOWER SECTION OF LIFT LINE	0-25
4	2265	0.6985		
5	8450	1.1031		

Figure 60. LIFT-LINE PRESSURE DROP FOR RUN HGD-4B



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min	PRESSURE DROP	SCALE, in. H_2O
1	245	0.3305	ACROSS DOWNCOMER	0-50
2	1000	0.5150	ACROSS LOWER SECTION OF LIFT LINE	0-50
3	1325	0.5523		
4	2340	0.6844	ACROSS UPPER SECTION OF LIFT LINE	0-25
5	3690	0.8472		
6	7550	1.0667		

Figure 61. LIFT-LINE PRESSURE DROP FOR RUN HGD-4C

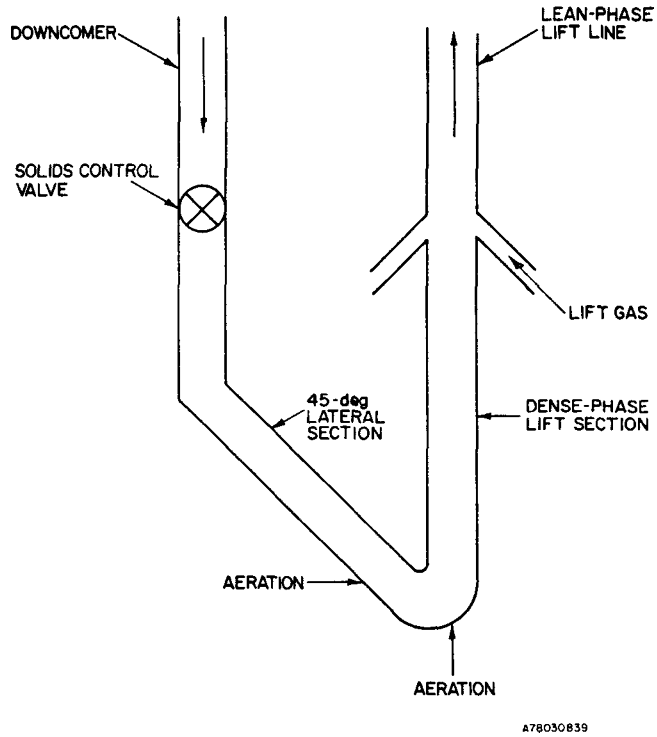


Figure 62. REVERSE SEAL

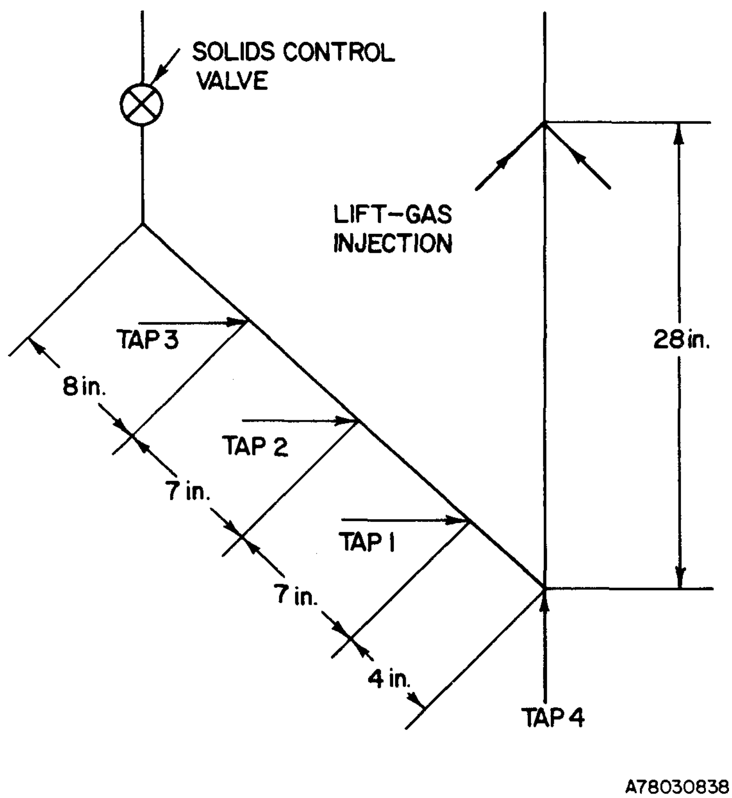


Figure 63. DIMENSIONS OF REVERSE SEAL

this mode, the solids control valve is fully open. The 45-degree lateral section runs full of solids in this mode.

When the valve is the means of control, aeration to the 45-degree lateral section is set at a high rate so that it does not control the solids flow. The solids flow rate is then controlled by the opening of the valve. In this mode, the solids in the 45-degree lateral section flow only along the bottom of the pipe.

The small, low-pressure model of the entire reverse seal was constructed of 2-inch clear PVC pipe. A 2-inch full-port ball valve was used as the solids control valve. Three aeration tap locations were placed in the 45-degree lateral section to test the effectiveness of each tap. The dense-phase lift was also constructed of 2-inch clear PVC pipe.

The low-pressure test unit used for testing the reverse seal is shown, as modified, in Figure 64. In a typical run, the upper bed is fluidized and the desired lean-phase lift-line velocity is set. Readings are taken at several different solids flow rates. The fluctuations in the recorder trace for the lean-phase lift-line pressure drop are used to analyze the smoothness of the lift-line operation.

As with the other configurations which were studied, the reverse seal was tested using −20+80 mesh sand and −20+200 mesh pretreated Illinois No. 6 bituminous coal. Both materials were tested with aeration-control and then with valve-control. Each aeration tap location was tested under aeration-control. Solids flow rates in each test were determined by timing individual particles as they passed between two marks, 12-inches apart, on the downcomer.

Eight tests were conducted with the reverse seal using the −20+80 mesh sand material. In run HGD-5A, the lean-phase lift (LPL) velocity was set at 35 ft/s, and aeration was added at Tap 1. (See Figure 63). The reverse seal was operated in the aeration-control mode in this run. As aeration to the 45-degree lateral section was increased, solids flow through the reverse seal increased. Pressure-drop readings across the lateral section and the LPL were made at each solids flow rate (Figure 65).

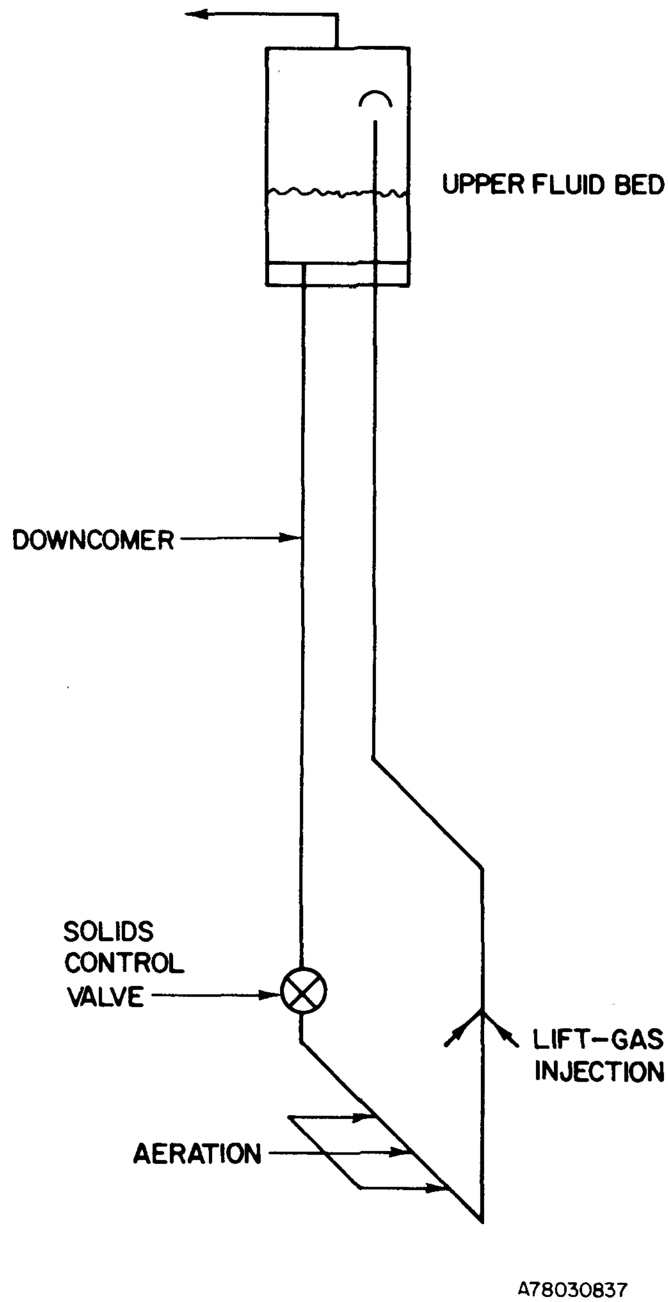
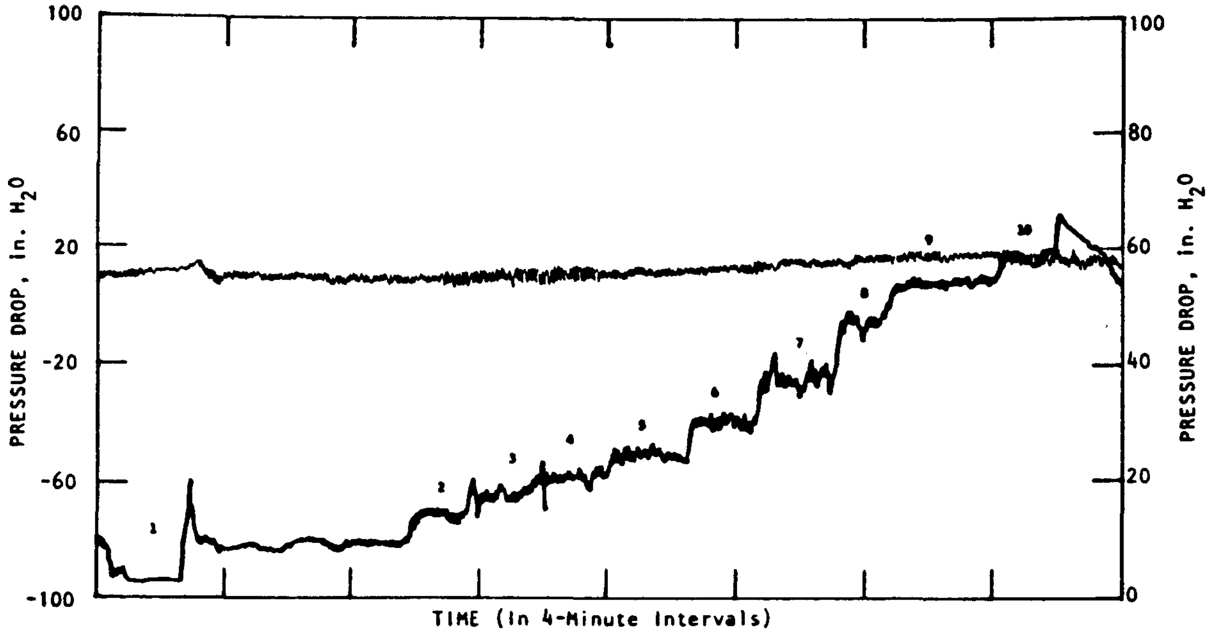


Figure 64. LOW-PRESSURE REVERSE SEAL TEST LOOP



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	820	1.80
2	1310	2.00
3	1630	2.22
4	2140	2.42
5	3050	2.74
6	3300	2.95
7	4100	3.41
8	5100	4.05
9	5900	4.47
10	6350	5.36

	Pressure Drop	Scale, in. H_2O
—— (thin)	Across Lateral Section	–100 to 100
—— (thick)	Across Lift Line	0 to 100

Lift-Line Velocity: 35 ft/s

Solids: –20 + 80 Mesh Ottawa Sand

Aeration: Tap 1

Figure 65. RESULTS AND CONDITIONS FOR RUN HGD-5A

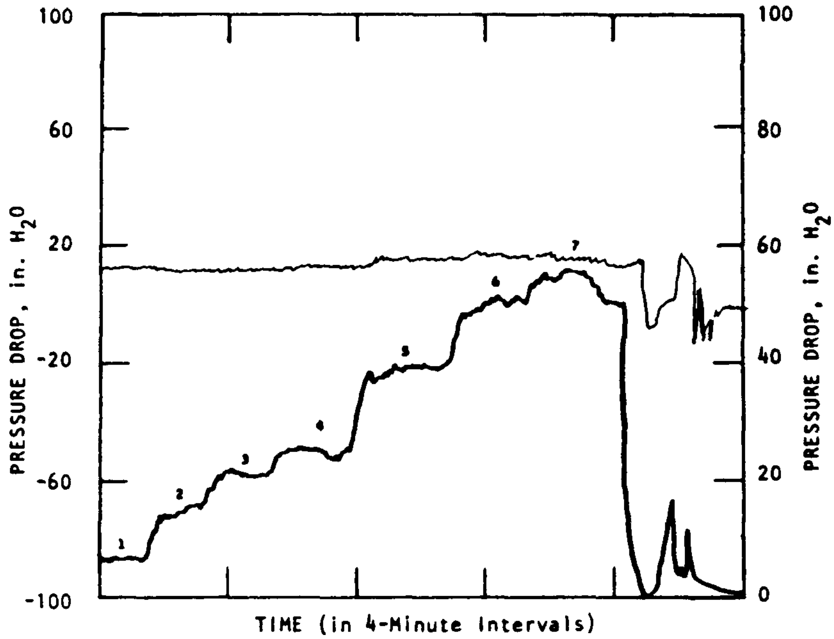
The pressure-drop across the lateral section changed very little with increased solids flow rate. The LPL pressure-drop increased with increased solids flow rate. At 35 ft/s, the pressure-drop fluctuations in the LPL were approximately ±2 inches of water from the average pressure-drop reading. A maximum sand flow rate of 6350 lb/hr of sand was achieved in this run.

The conditions of Runs HGD-5B, HGD-5C, and HGD-5D were identical to those in Run HGD-5A except for the LPL velocity. LPL velocities of 40, 45, and 50 ft/s were used in Runs HGD-5B, HGD-5C, and HGD-5D, respectively. The results of each of these runs were also similar to the results of Run HGD-5A. The results of these runs are shown in Figures 66, 67, and 68. First of all, lateral pressure-drop changed very little with increased solids flow rate. Also, the LPL pressure-drop increased with increased solids flow rate. However, with the higher LPL velocities, the PLP pressure-drop fluctuations decreased, indicating that the 35 ft/s velocity in Run HGD-5A was probably too close to choking. In addition, maximum solids flow rates increased to 7500 lb/hr in Run HGD-5D.

In Runs HGD-5E, HGD-5F, and HGD-5G, aeration Taps 2, 3, and 4, respectively, were used. The results of these runs are shown in Figures 69, 70, and 71. An LPL velocity of 40 ft/s was used in these tests. For Taps 2 and 3, the lateral pressure drop (the low-pressure tap being located at the 135-degree bend) was more sensitive to the solids flow rate. This was because the aeration gas had to pass through more solids to reach the 135-degree bend. The lateral pressure drop for Tap 3 was extremely sensitive to the aeration rate. Tap 2 gave a much higher maximum solids flow rate than Tap 3 (8150 lb/hr versus 5900 lb/hr, respectively). The maximum solids flow rate through Tap 2 was also greater than the maximum solids flow rate through Tap 1, making Tap 2 the most effective aeration tap for sand.

In Run HGD-5G, Tap 1 was tested. However, no solids flow was achieved at any aeration flow rate using this tap. This result is not surprising, because it is necessary to aerate any non-mechanical valve before the constricting bend so that the aeration can assist the solids through the constrictor.

In Run HGD-5H, the valve in the downcomer section was used to control the solids flow rate. In this run, aeration Tap 1 was set at a high value (5.27 ACFM), and the valve was used to control the solids flow rate. The LPL flow rate was constant at 40 ft/s. If aeration is not added to the lateral section,



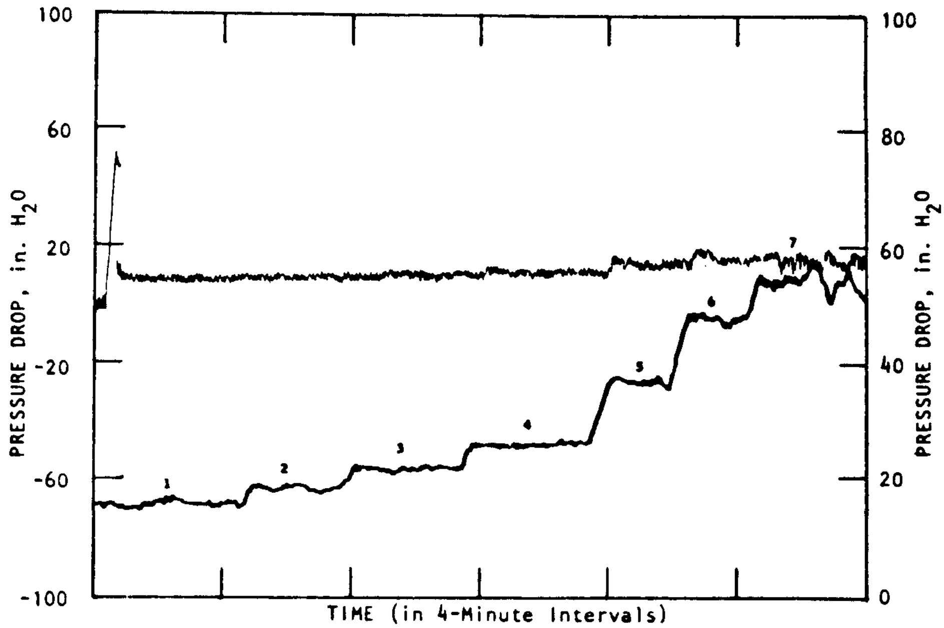
Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min		Pressure Drop	Scale, in. H_2O
1	300	2.05	——	Across Lateral Section	–100 to 100
2	1400	2.36			
3	2230	2.63	▬▬	Across Lift Line	0 to 100
4	3050	2.89			
5	4650	3.72			
6	6125	4.46			
7	6950	5.33			

Lift-Line Velocity: 40 ft/s

Solids: –20 + 80 Mesh Ottawa Sand

Aeration: Tap 1

Figure 66. RESULTS AND CONDITIONS FOR RUN HGD-5B



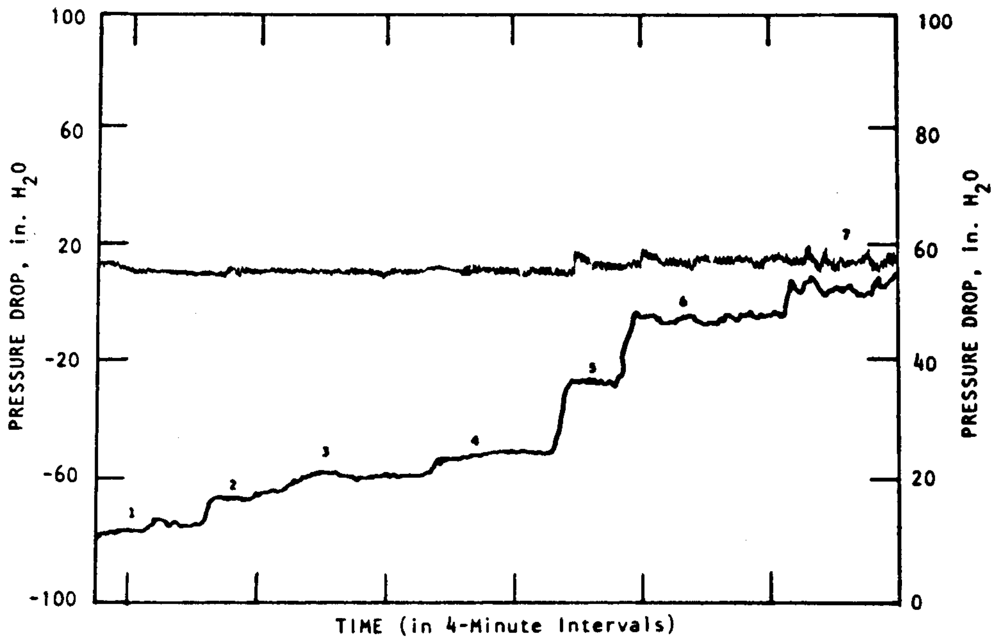
Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min		Pressure Drop	Scale, in. H_2O
1	1570	2.11	——	Across Lateral Section	–100 to 100
2	2040	2.37			
3	2665	2.63	━━	Across Lift Line	0 to 100
4	3035	2.94			
5	4850	3.71			
6	6125	4.45			
7	6950	5.31			

Lift-Line Velocity: 45 ft/s

Solids: –20 + 80 Mesh Ottawa Sand

Aeration: Tap 1

Figure 67. RESULTS AND CONDITIONS FOR RUN HGD-5C



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	1015	2.27
2	1725	2.47
3	2450	2.68
4	2875	2.94
5	4950	3.71
6	6550	4.44
7	7500	5.31

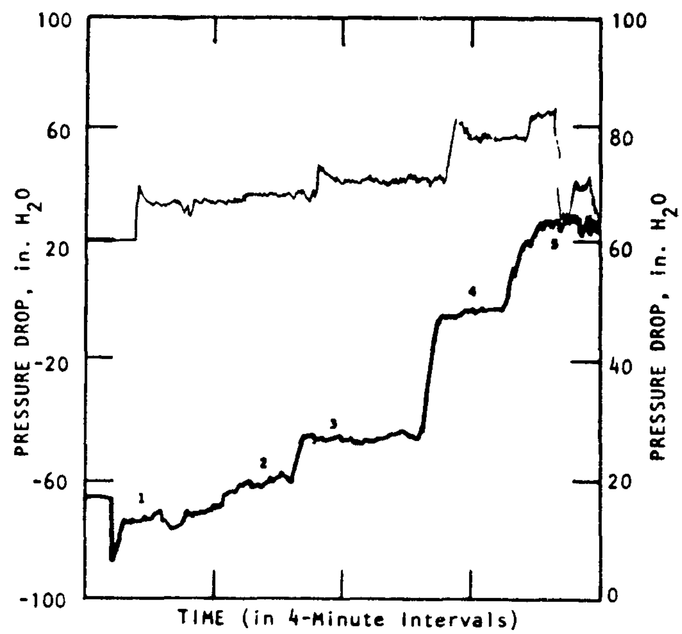
	Pressure Drop	Scale, in. H_2O
—— (thin line)	Across Lateral Section	–100 to 100
—— (thick line)	Across Lift Line	0 to 100

Lift-Line Velocity: 50 ft/s

Solids: –20 + 80 Mesh Ottawa Sand

Aeration: Tap 1

Figure 68. RESULTS AND CONDITIONS FOR RUN HGD-5D



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	1300	2.36
2	2250	2.63
3	3250	3.20
4	6250	4.44
5	8150	5.31

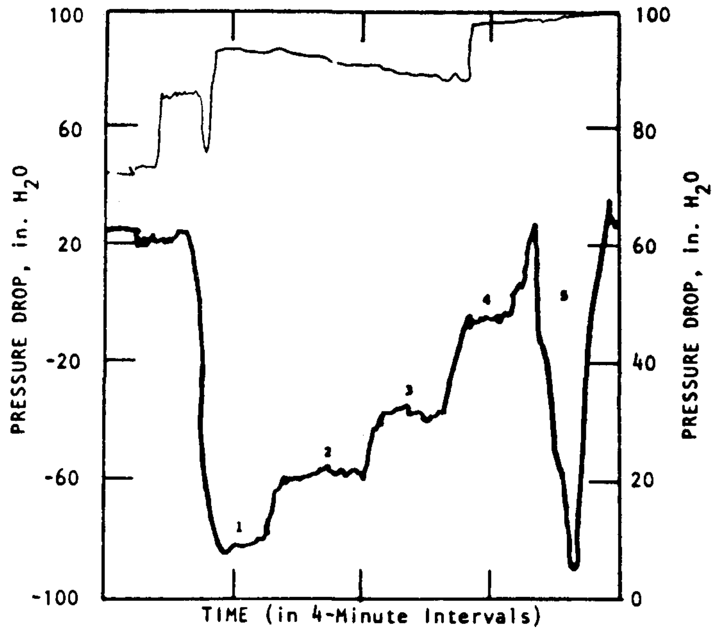
	Pressure Drop	Scale, in. H_2O
———	Across Lateral Section	–100 to 100
━━━	Across Lift Line	0 to 100

Lift-Line Velocity: 40 ft/s

Solids: –20 + 80 Mesh Ottawa Sand

Aeration: Tap 2

Figure 69. RESULTS AND CONDITICNS FOR RUN HGD-5E



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	700	2.36
2	2325	2.63
3	3700	3.20
4	5900	4.44

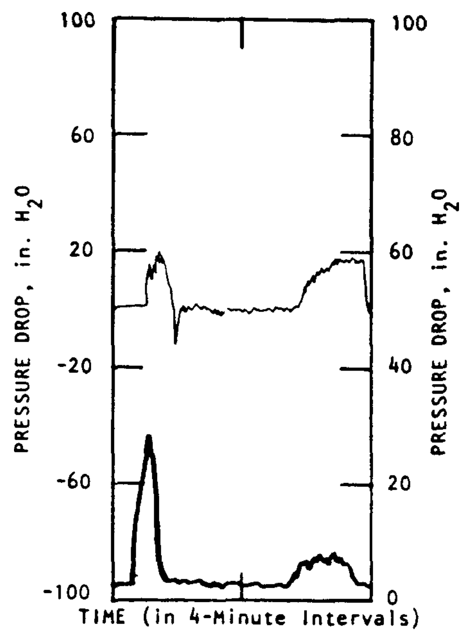
	Pressure Drop	Scale, in. H_2O
——	Across Lateral Section	–100 to 100
▬▬	Across Lift Line	0 to 100

Lift-Line Velocity: 40 ft/s

Solids: –20 + 80 Mesh Ottawa Sand

Aeration: Tap 3

Figure 70. RESULTS AND CONDITIONS FOR RUN HGD-5F



Pressure Drop	Scale, in. H_2O
—— Across Lateral Section	−100 to 100
▬▬ Across Lift Line	0 to 100

Lift-Line Velocity: 40 ft/s

Solids: −20 + 80 Mesh Ottawa Sand

Aeration: Tap 4

No Solids Flow Obtainable

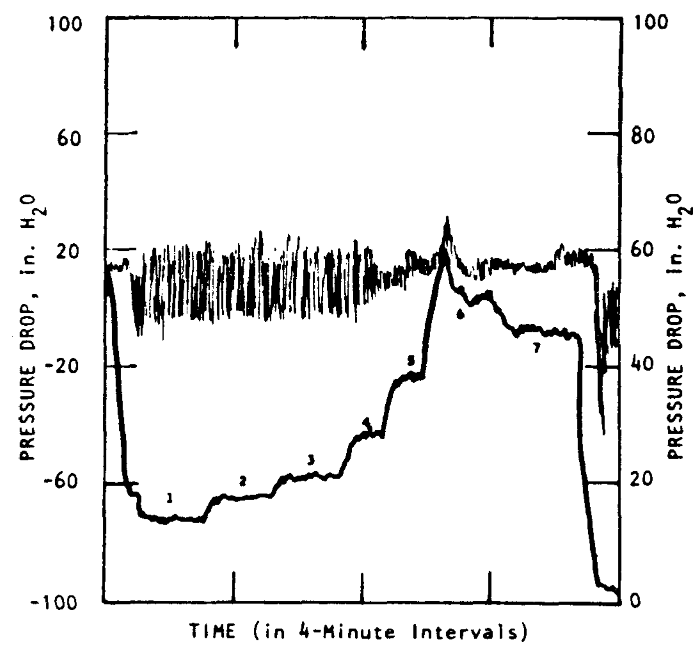
Figure 71. RESULTS AND CONDITIONS FOR RUN HGD-5G

the solids would not flow and, therefore, the valve would not control. Also, if the valve was opened to a position which would allow solids to flow through it faster than the aeration rate could flow them around the bend, the reverse seal operation reverted to the aeration-control model. Good solids flow control was obtained in this mode. However, the lateral pressure-drop fluctuations were extremely large (Figure 72).

The reverse seal configuration was also tested with —20+200 mesh pretreated Illinois No. 6 bituminous coal. In the first test, Run HGD-6A, a LPL velocity of 25 ft/s was used. The run was made in the aeration-control mode, with aeration being added at Tap 1. With aeration at this tap, a solids flow rate of only 430 lb/hr (Figure 73) could be achieved, and solids control was extremely poor. The reason for this is that the aeration jet penetrated the lateral diameter, and gas bypassed the top of the 135-degree bend. A lift velocity of 30 ft/s (Run HGD-6B) resulted in a maximum solids flow rate of 860 lb/hr (Figure 74), but an increase of aeration gas beyond that condition caused the solids flow rate to drop — again due to jet penetration and bypassing.

In Run HGD-6C, aeration Tap 2 was used to control the solids flow rate with a LPL velocity of 25 ft/s. The maximum solids flow rate under these conditions was approximately 1100 lb/hr (Figure 75). In Run HGD-6D, the aeration tap was again used, but at a lift-line velocity of 30 ft/s. The maximum solids flow rate under these conditions increased to 1285 lb/hr (Figure 76). In each of these first four runs, the LPL pressure drop was very steady. The lateral pressure-drop for Tap 1 did not change. For the runs with Tap 2, the lateral pressure-drop decreased at the point where an increase in aeration failed to increase the solids flow rate. This decline in lateral pressure-drop also seemed to be due to gas bypassing the top of the 135-degree bend.

In Runs HGD-6E and HGD-6F, aeration tap 3 was tested at LPL velocities of 35 and 40 ft/s, respectively. The maximum solids flow rate attainable using this tap was 1450 lb/hr on both runs. As with the other runs, the solids flow rate increased with an increase in the aeration rate, but reached a limiting value at which further aeration failed to increase solids flow. At this limiting value, the lateral pressure-drop decreased (Figures 77 and 78). The pressure-drop in the LPL, however, was very smooth at each solids flow rate with minimal fluctuations.



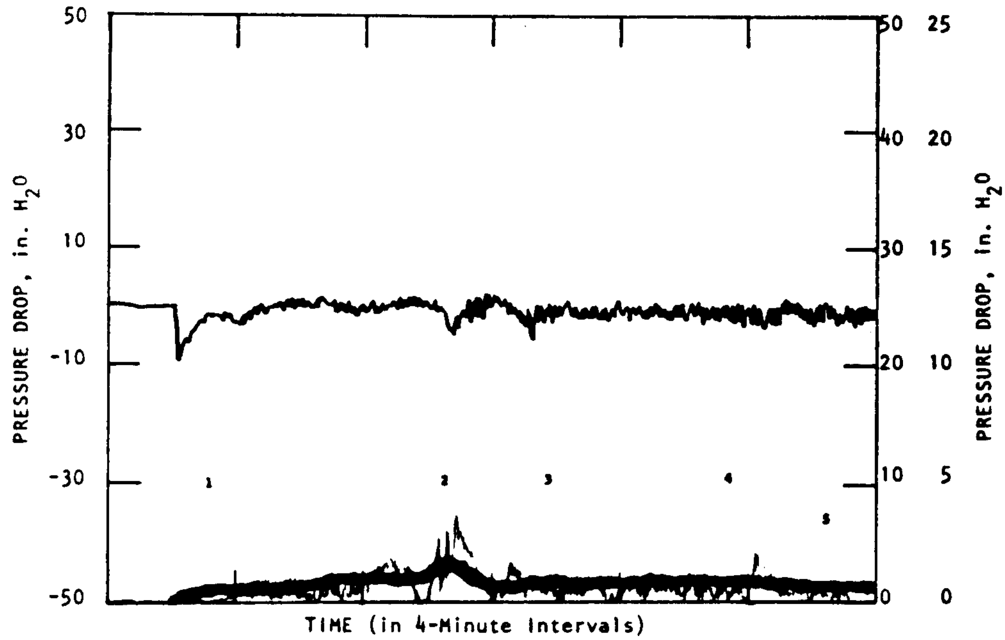
Reading No.	Solids Flow Rate, lb/hr		Pressure Drop	Scale, in. H_2O
1	1400	——	Across Lateral Section	–100 to 100
2	1900			
3	2400	▬▬	Across Lift Line	0 to 100
4	2650			
5	4650			
6	6550			
7	5750			

Lift-Line Velocity: 40 ft/s

Solids: –20 + 80 Mesh Ottawa Sand

Aeration: at Tap 1, but Ball Valve Controlling

Figure 72. RESULTS AND CONDITIONS FOR RUN HGD-5H



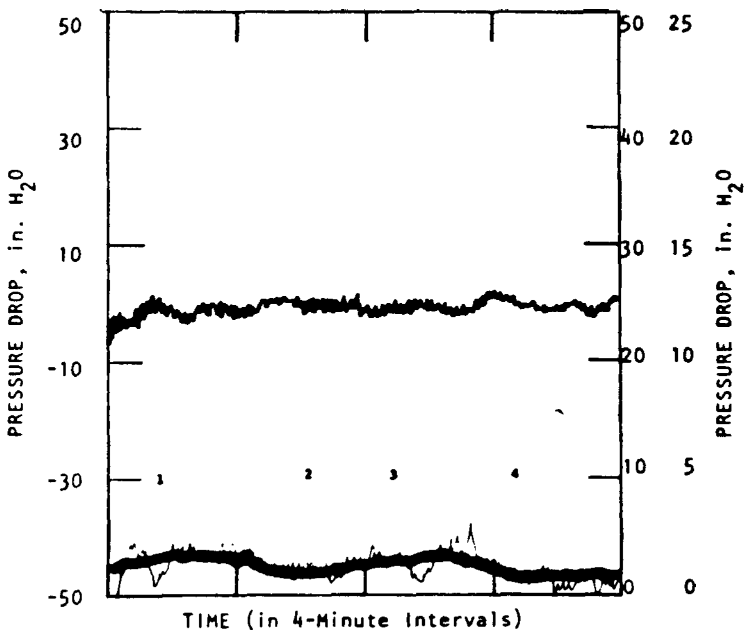
Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min	Pressure Drop	Scale, in. H_2O
1	345	1.4248	Across Lateral Section	−50 to 50
2	430	1.7942		
3	345	2.3115	Across Lift Line	0 to 50
4	345	2.8894		
5	345	4.2553	Across Ball Valve	0 to 25

Lift-Line Velocity: 25 ft/s

Solids: −20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 1

Figure 73. RESULTS AND CONDITIONS FOR RUN HGD-6A



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	860	1.2137
2	560	0.9011
3	760	2.3115
4	460	2.8894

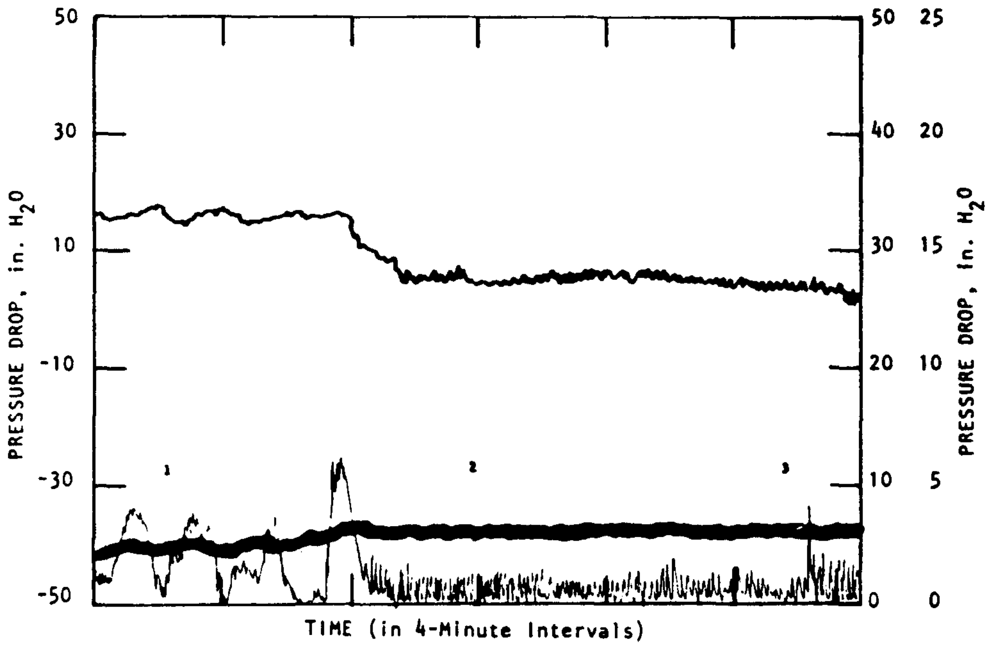
Pressure Drop	Scale, in. H_2O
Across Lateral Section	–50 to 50
Across Lift Line	0 to 50
Across Ball Valve	0 to 25

Lift-Line Velocity: 30 ft/s

Solids: –20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 1

Figure 74. RESULTS AND CONDITIONS FOR RUN HGD-6B



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	945	1.0602
2	1100	1.3193
3	1106	1.7942

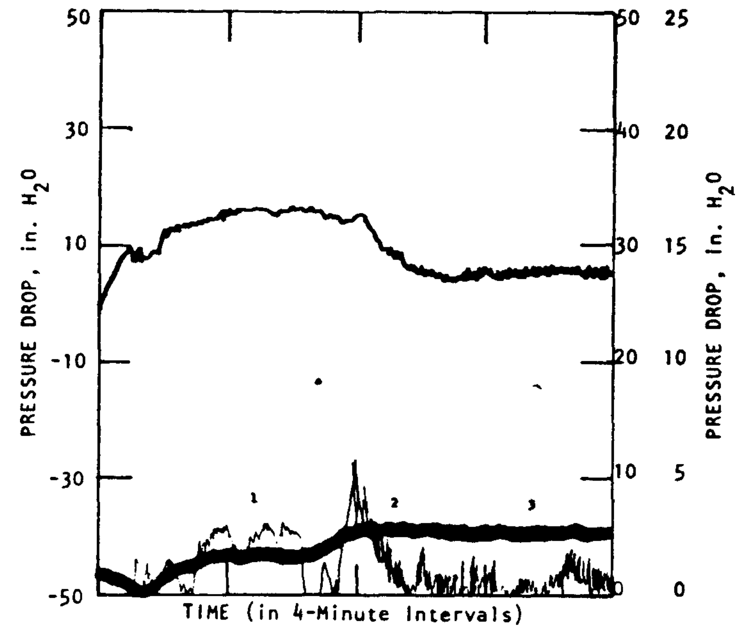
Pressure Drop	Scale, in. H_2O
Across Lateral Section	−50 to 50
Across Lift Line	0 to 50
Across Ball Valve	0 to 25

Lift-Line Velocity: 25 ft/s

Solids: −20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 3

Figure 75. RESULTS AND CONDITIONS FOR RUN HGD-6C



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	860	1.0554
2	1285	1.3193
3	1285	1.7942

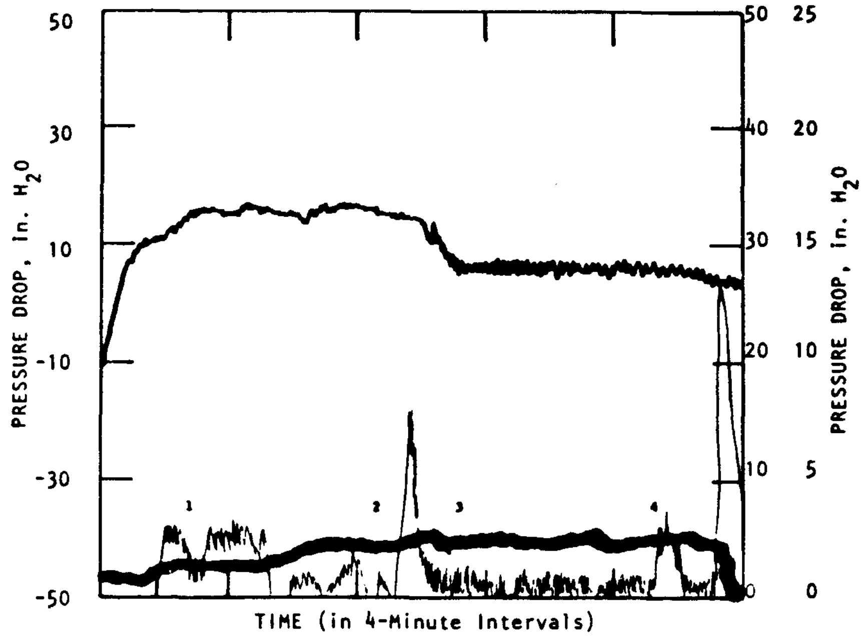
Pressure Drop	Scale, in.H_2O
▬ Across Lateral Section	−50 to 50
■ Across Lift Line	0 to 50
— Across Ball Valve	0 to 25

Lift-Line Velocity: 30 ft/s

Solids: −20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 3

Figure 76. RESULTS AND CONDITIONS FOR RUN HGD-6D



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	800	1.0554
2	1290	1.3193
3	1450	1.5304
4	1450	1.7942

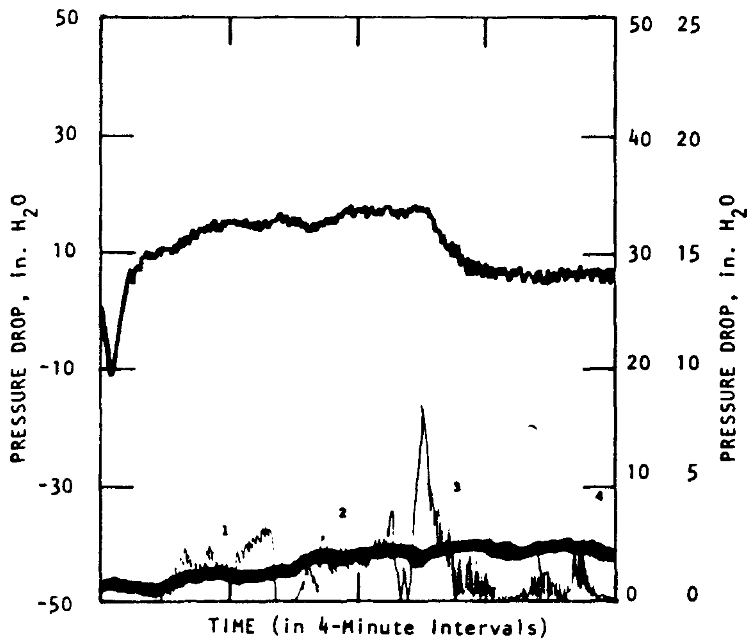
Pressure Drop	Scale, in. H_2O
Across Lateral Section	–50 to 50
Across Lift Line	0 to 50
Across Ball Valve	0 to 25

Lift-Line Velocity: 35 ft/s

Solids: –20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 3

Figure 77. RESULTS AND CONDITIONS FOR RUN HGD-6E



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min	Pressure Drop	Scale, in. H_2O
1	850	1.0554	Across Lateral Section	–50 to 50
2	1180	1.3193		
3	1450	1.5304		
4	1450	1.7942	Across Lift Line	0 to 50
			Across Ball Valve	0 to 25

Lift-Line Velocity: 40 ft/s

Solids: –20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 3

Figure 78. RESULTS AND CONDITIONS FOR RUN HGD-6F

In Run HGD-6G, Tap 4 was tried. As in the sand run, no solids could be made to flow using Tap 4 only (Figure 79).

In Run HGD-6H, Tap 2 was again tested at a LPL velocity of 35 ft/s. A maximum solids flow rate of 1610 lb/hr was achieved under these conditions. The results of this run are shown in Figure 80.

The valve in the downcomer was used to control the solids flow rate in Run HGD-6I. Aeration was added to the lateral section at Tap 2. As the valve opening was increased, the solids flow rate increased. A maximum value of 1610 lb/hr of solids was achieved. The LPL pressure-drop was smooth, with little fluctuation (Figure 81). As with the sand test in the valve-control mode, the lateral pressure-drop fluctuated more at low solids flow rates than at the higher flow rates.

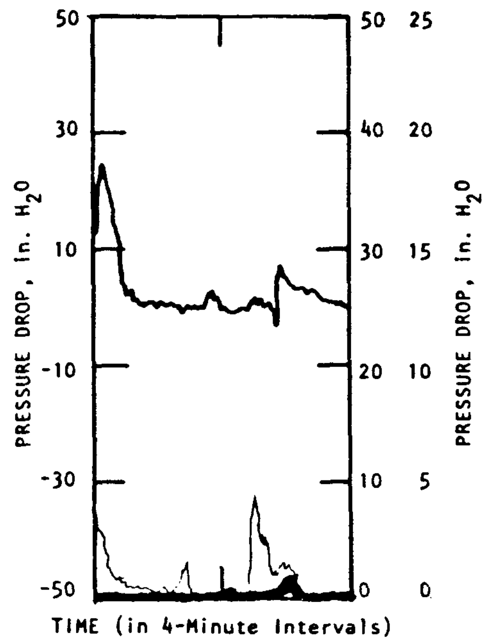
A photograph of the reverse seal is shown in Figure 82. This picture clearly shows the 4 aeration taps tested, the solids control valve, and the LPL aeration gas injection point. The photograph was taken during a sand test.

When visually observing the operation of the reverse seal, perhaps the most noticeable feature is the slugging of the dense-phase-lift (DPL) section of the device. In this section the sand is fluidized. The gas from the aeration taps passes up the DPL in the form of bubbles. The bubbles grow in diameter to a size equal to that of the diameter of the DPL section and slugging results. This slugging action is extremely "jerky"; the bubbles cause the solids to be "burped" up into the LPL in a very unsmooth manner. After watching the DPL operation, it was surprising that the LPL pressure drop did not fluctuate more than it did.

Task 9. Support Studies

Plant Effluent Processing

The plant effluent clean-up section was in operation during Tests 68, 69, and 70. The section was inspected during the turnaround activities conducted in March, and the refractory in the high-capacity incinerator was patched.



Pressure Drop	Scale, in. H_2O
Across Lateral Section	-50 to 50
Across Lift Line	0 to 50
Across Ball Valve	0 to 25

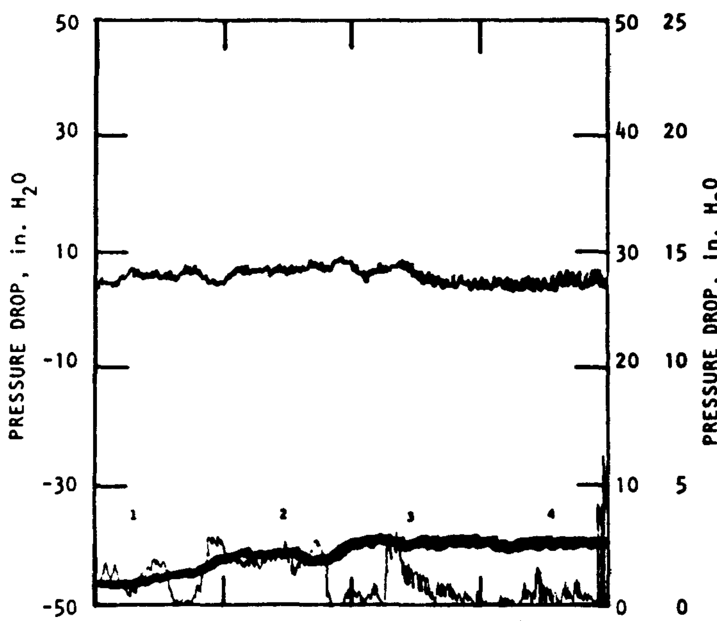
Lift-Line Velocity: 35 ft/s

Solids: -20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 4

No Solids Flow Obtainable

Figure 79. RESULTS AND CONDITIONS FOR RUN HGD-6G



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	640	1.05
2	1290	1.32
3	1610	1.53
4	1610	1.84

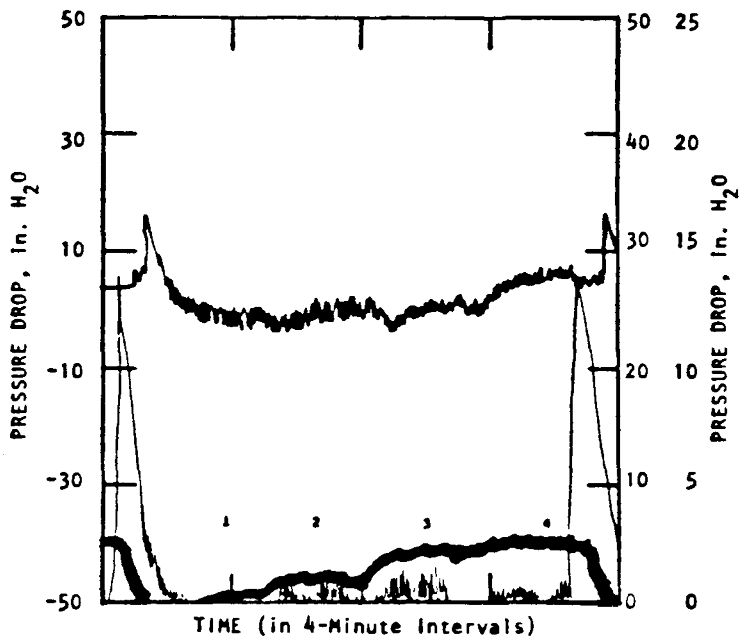
	Pressure Drop	Scale, in. H_2O
▬	Across Lateral Section	−50 to 50
■	Across Lift Line	0 to 50
—	Across Ball Valve	0 to 25

Lift-Line Velocity: 35 ft/s

Solids: −20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 2

Figure 80. RESULTS AND CONDITIONS FOR RUN HGD-6H



Reading No.	Solids Flow Rate, lb/hr	Pressure Drop	Scale, in. H_2O
1	70	Across Lateral Section	–50 to 50
2	470		
3	1290	Across Lift Line	0 to 50
4	1610		
		Across Ball Valve	0 to 25

Lift-Line Velocity: 35 ft/s

Solids: –20 + 200 Mesh Pretreated
Illinois No. 6 Bituminous Coal

Aeration: Tap 2, Ball Valve Controlling
1.8997 ACF/min

Figure 81. RESULTS AND CONDITIONS FOR RUN HGD-6I

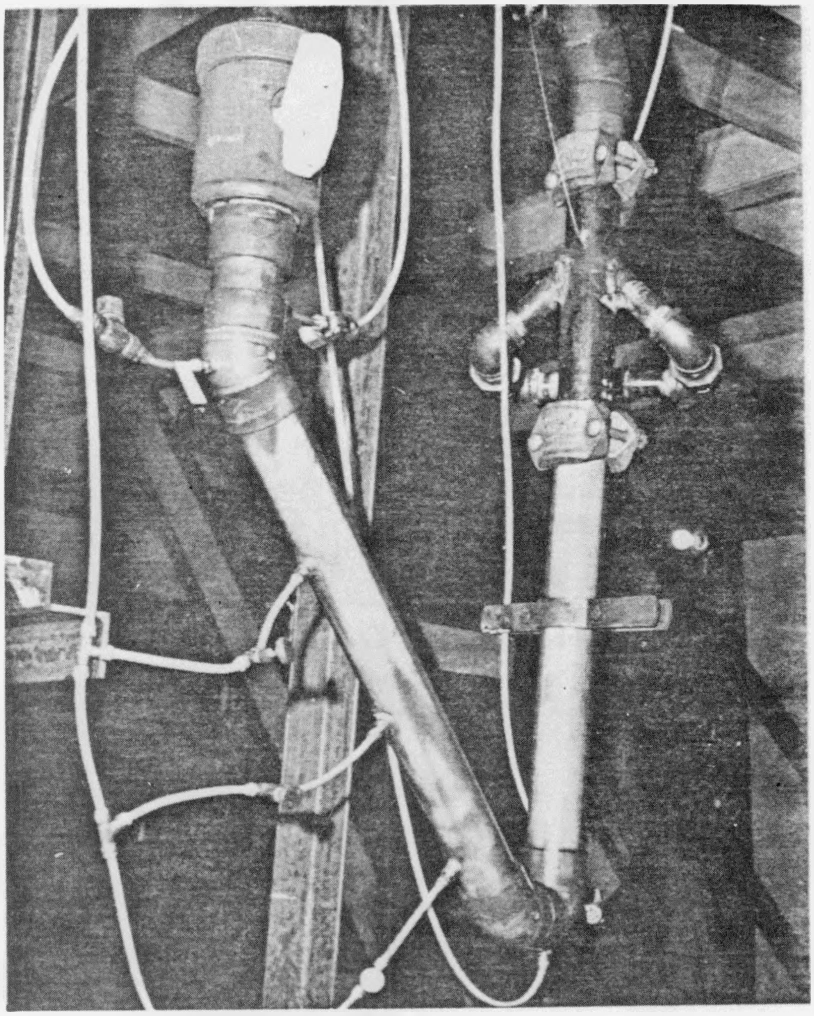


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Figure 82. REVERSE SEAL

Test Methanation Systems and Catalysts

IGT Fixed-Bed Catalyst Methanation Section

The IGT fixed-bed catalyst methanation section was put on standby for Tests 68 and 69, but was not put on-line because of the early termination of these tests.

Chem Systems' Liquid-Phase Methanation (LPM) Unit

Modifications to the Chem Systems' Liquid-Phase Methanation unit were completed during the quarter. These modifications were made to minimize the amount of elutriation of catalyst fines from the reactor. Although the unit was not ready for operation during Tests 68 and 69, it was put on-line during Test 70. The LPM unit accepted purified gas from the HYGAS reactor for 25 hours beginning at 0100 hours on February 23.

Investigate Hot-Oil Quench System

A preliminary engineering study was completed for this quarter for the design of a hot-liquid quench unit which may be added to the HYGAS pilot plant. The objective was to determine the most suitable liquid medium to wet and remove the solids carried by the gasifier product gas that would not be removed by the existing cyclone system and would not significantly affect the gas composition. It was concluded that a hot-water scrubbing system operating at, or above, the product-gas dew point is the most promising alternative that can be implemented within the present program schedule.

Objective

The existing oil-water separation system has had intermittent operational problems caused by the formation of an oil-solids-water layer in the quench system separation vessel. The three-phase layer may be formed by solids that get through the existing cyclone system during upset conditions and/or because the cyclone efficiency during steady operation allows a small amount of particulate matter (which builds up in the oil-water separation system) to pass through. The hot-water quench is expected to provide effective downstream capacities for the removal of particulate matter that gets through the cyclone system during any mode of gasifier operation, thereby allowing smoother operation of the oil-water separation vessel during an entire gasification run. The most important consideration is to design a scrubbing system that would maintain a product gas composition that would be essentially unchanged as it passes from the gasifier through the scrubber vessel.

Procedure

Water and toluene were studied as possible candidates for the scrubbing medium. They were chosen because 1) water is inherently the easiest and safest medium and 2) previous observation has revealed that particulate matter is better removed, or attracted to, the toluene layer in the separation process.

The existing prequench tower vessel was designated as the scrubbing vessel. Product-gas quench and water/oil cooling and condensation will then be affected in the quench tower system or other system that is designed to meet the necessary heat-duty requirements of the scrubber contactor vessel off-gas.

Because it is most desirable that the contactor off-gas composition be essentially the same as the inlet gas composition, the dictated mode of operation requires that the contactor off-gas temperature be the same as that of the gasifier product-gas dew point, or slightly higher, if possible.

Using the SSI 100 Process simulation program, the dew point curve was obtained for the gasifier off-gas as a function of gas composition at pressures between 900 and 1000 psig. The computer program generates dew point data using Chao-Seider thermodynamic correlations for equilibria calculations. This program can readily generate the necessary dew point based on gasifier off-gas composition. Because the generated data assumed an adiabatic system, it was also necessary to calculate the prequench tower heat losses to estimate the magnitude of heat loss in the contactor vessel. The results indicated that tower losses were minimal within the desired operating range of 400° to 450°F.

Removal of the prequench vessel necessitated an evaluation of the possible modifications to the downstream system which are required to effectively quench the gasifier product gas to the 100°F level. It was again intended to minimize equipment requirements. Two approaches were used: 1) modification of the quench tower as necessary for quenching the gas, and 2) installation of an in-line cooler to reduce heat duty into the quench system and thereby minimize quench tower modifications.

To establish the quench system material balance, the SSI 100 Process simulation program was used, after making modifications to satisfy the process requirements. Once again, the computer used the Chao-Seider thermodynamic correlations to generate steam composition data.

Results

Scrubber Vessel

Calculated dew points indicate that the scrubber vessel should operate with an off-gas temperature between 400° and 450°F. Dew points were calculated using data generated during HYGAS Run 61. Operating temperatures may be modified as necessary, based on more recent test data, but significant changes are not expected.

Based on computer results, the following is a comparison of the expected advantages and disadvantages of operation with water or toluene.

a. Toluene Advantages

- Observed affinity to solids particulate matter
- Available liquid stream
- Lower heat capacity than water.

b. Toluene Disadvantages

- Dew point lower than water, i.e., water will condense before oil will
- Safety problems involved in case of pump seal failure
- Separation problem not solved with operation below the dew point
- Gas humidification expected to be significant during operation at, or above, the gas dew point
- Increased scrubbing liquid rates requiring more pump horsepower
- Depends on stripper performance and capacity for efficient and proper solids-liquid separation of the liquid waste stream.

c. Water Advantages

- Available liquid stream
- Noncombustible liquid
- Solids/liquid separation method easily handled by cooling and filtration
- Possible to operate at the dew point temperature with essentially unchanged gas composition

- At temperatures higher than the dew point, water partial-pressure in product gas is higher than the saturation pressure. Therefore, humidification problems that may occur are minimized.

d. Water Disadvantages

- Solids scrubbing efficiency may be lower than that of toluene
- Higher heat capacities.

Downstream Quench Results

The removal of approximately 16 million Btu/hr is estimated to be required of the quench downstream of the scrubber contact vessel. Forty-seven plates and 300 gpm of 100°F water would be required if the quench tower was going to do quench duty alone. Without major modifications, the present system cannot handle these requirements.

The gas can be quenched if an in-line cooling device is put upstream of the quench system. The in-line cooling device will reduce gas temperatures to approximately 380°F, with partial condensation of the off-gas stream. Then the quench tower modifications can be managed, since only 20 trays of baffles are necessary for the system.

Problem Areas

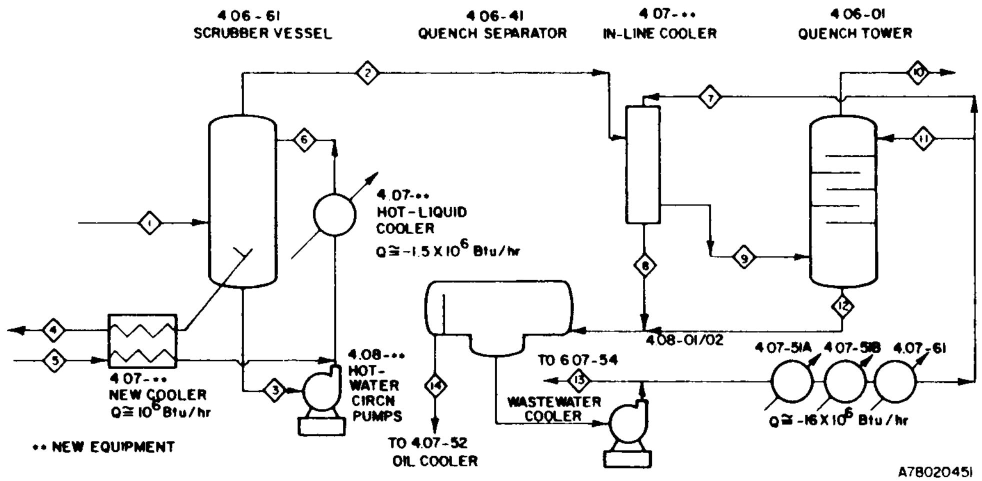
The exact configuration of the required in-line cooler is not known. Problems with the flow regime can be expected. This cooler might possibly be purchased from an outside vendor. Water-scaling tendencies, if raw water is used, may present a problem in the contact vessel.

Conclusions

IGT recommends that a hot-liquid scrubber be designed using water as the scrubbing medium. Figure 83 shows the conceptual process flow diagram for the bituminous case (90% carbon conversion) using water as the scrubbing medium. Minimal plant modifications are required. IGT anticipates that the specification and procurement of the hot-liquid circulation pumps will be the factors critical to the completion of the project.

Materials Testing

Information on materials testing was gathered through exposure of MPC corrosion and erosion test coupons during Tests 68, 69, and 70.



Stream No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Pressure, psig							900							
Temperature, °F	600	443	461	100	85	430	85	358	358	100	85	170	241	241
Stream Name	Gasifr Prod Gas	Scrub. Prod Gas	Scrub. Liq Btms	Scrub. Waste-water	H_2O Makeup	Hot H_2O to Scrub. Vessel	Quench H_2O to In-Line Cooler	In-Line Cooler Btms	In-Line Cooler Off-Gas	Quench Tower Off-Gas	Quench H_2O to Quench Tower	Quench Tower Btms	Excess Quench H_2O	Oil to Storage
Component							lb-mol/hr							
H_2O	406	434	2834.5	137.5	165.5	3000	1390	1728.4	95.6	0.30	2780	2875.3	433.7	--
H_2	136.5	136.5	--	--	--	--	--	1.2	135.3	134.3	--	1.0	--	2.2
CO	39.7	39.7	--	--	--	--	--	0.8	38.9	38.3	--	0.6	--	1.4
CO_2	164	164	--	--	--	--	--	6.8	157.2	143.1	--	14.1	--	20.9
H_2S	2.14	2.14	--	--	--	--	--	0.14	2.0	1.6	--	0.4	--	0.54
CH_4	84.8	84.8	--	--	--	--	--	1.6	83.2	80.6	--	2.6	--	4.2
C_2H_6	3.1	3.1	--	--	--	--	--	0.1	3.0	2.7	--	0.3	--	0.4
Oil	121.6	121.6	--	--	--	--	--	56.6	65.0	0.70	--	64.3	--	120.9
							lb/hr							
Dust	25	--	515	25	--	515	--	--	--	--	--	--	--	--

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Figure 83. CONCEPTUAL HOT-WATER SCRUB PROCESS FLOW DIAGRAM

Engineering Services

Routine engineering services were conducted during the quarter. In addition, the new steam-oxygen sparger was designed, and installed in March. Evaluation and construction of the double-screen equipment in the coal preparation was conducted.

In order to study the relocation of the solids transfer 339 valve from the high-temperature reactor to the steam-oxygen reactor, a scaled-down model of the valve was built. Its operation was tested and found to be satisfactory.

During Tests 69 and 70, Argonne National Laboratory's personnel collected operating data on the low-pressure slurry line using their two flow test meters.

A reliability study of the HYGAS plant was begun. Details of this study will be report later.

FUTURE WORK

At the end of this quarter, the HYGAS plant was ready to begin Test 71. The objective of this test is to conduct an extended run at reasonably high char conversions and char throughput without clinker formation in the reactor. Operations with the 6-nozzle sparger and relocated 339 valve will be evaluated. Double-screening of the raw coal feed will also be studied.

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