FE-2434-21

PIPELINE GAS FROM COAL-HYDROGENATION (IGT HYDROGASIFICATION PROCESS)

Project 9000 Monthly Status Report

For the Period October 1 Through October 31, 1977

Prepared by

Institute of Gas Technology

IIT Center, 3424 S. State Street

Chicago, Illinois 60616

Date Published — December 1977

Prepared for the

UNITED STATES DEPARTMENT OF ENERGY

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

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REA

INSTITUTE OF GAS TECHNOLOGY · IIT CENTER · CHICAGO 60616

Project Status Report
for
DEPARTMENT OF ENERGY

AND

AMERICAN GAS ASSOCIATION

Report for

October 1977

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

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

I. PROJECT OBJECTIVE

The objective of this project is to perform the necessary pilot plant operations and related support studies required to acquire data for a commercial/demonstration plant design based on the HYGAS® Process. To this end, we are conducting tests with Illinois bituminous coal to acquire data on optimizing the HYGAS Process. We conducted Test 66 during October. It was significant because the reactor was operated for 9 days at coal conversions up to 85% and also because the reactor was found to be exceptionally clean after the run.

II. SUMMARY

We made a post-run inspection for Test 65 this month, after which Test 66 was conducted. Over 9 days of reactor operation with hydrogen and char feed were logged for Test 66. Steady-state, self-sustained operations were achieved at a char feed rate of about 2 tons/hr and at coal conversions ranging from 78% to 85%.

A large amount of data was supplied to Procon, Inc., to aid in its design of a commercial/demonstration HYGAS plant.

Results of Argonne National Laboratory's cyclone inspection are given in this report. Methods for calculating coal and carbon conversions in the HYGAS reactor are also presented.

III. ACHIEVEMENTS

Task 7. Pilot Plant Experimental Operation

Test 65 was conducted during September. It was very short, and was involuntarily terminated due to an electrical power failure caused by an equipment malfunction on Commonwealth Edison's main power supply to the plant.

Our post-run inspection after Test 65 revealed that the pretreater reactor and the pretreater char cooler were both free of clinkers. The condition of the latter was the result of using 100% nitrogen for fluidizing the hot char in the char cooler. The pretreater quench tower had a small accumulation of coal fines and tar in the bottom, and there was a slight buildup of tar in the venturi scrubber. We discovered a small leak at the top of one of the cooling coils inside the pretreater and repaired it. A broken distributor nozzle was also found on the pretreater grid and was fixed. The pretreater section was then cleaned up for Test 66.

Our inspection of the reactor revealed a large quantity of caked coal on the slurry dryer wall above and below the slurry inlet nozzle. This material was free-flowing by the IGT agglomeration boat test. During the start-up of Test 65, a full slurry concentration was introduced to the slurry dryer bed. Normally, clean recycle oil feed is first established and then pretreated char is added to the mix tank so that the slurry concentration is slowly built up over several hours. In Test 65, we attempted to feed an oil-slurry mixture at a normal, steady-state running concentration. The temperatures in the slurry dryer bed fell below the vaporization level (due to the increased heat load of the slurry) and wet char accumulated on the walls of the slurry dryer. We will revert to the standard method of slowly introducing slurry feed in all future operations. The rest of the reactor was completely clean.

Due to the poor operation of the slurry dryer bed during Test 65, an unusually large amount of char carry-over was found in the quench separator, and the liquid lines from the quench separator to the prequench tower were partially plugged. We cleaned the entire quench system and prepared it for Test 66. We also inspected the effluent cleanup section and found a small amount of char fines in the light-oil recovery quench separator unit. The effluent cleanup section was readied for Test 66.

A mechanical seal on the high-pressure amine pump in the purification section was replaced, and circulation tests were carried out. The purification section was readied for operation. The IGT fixed-bed catalyst methanation system was activated for Test 66.

The reactor heat-up cycle for Test 66 was started on October 1. However, the gasifier pressure had to be brought down to repair a leak, which developed on the water inlet line to the water jacket for the start-up heater. The reactor was reignited on October 4, and by October 5 at 1430 hours, self-sustained operation was achieved. The reactor had reached steady-state conditions at a char feed rate of 2 tons/hr, when the spent-char discharge line plugged on October 6. Oxygen feed to the reactor was stopped, and the spent-char discharge line was cleared. Char feed to the reactor was restarted at 1000 hours on October 7. The chokes plugged because small pieces of refractory had passed the bottom discharge valve of the reactor but could not go through the restriction orifice in the chokes. On October 8 at 0900 hours, the screw feeder motor on the pretreater char feed to the slurry tank burned out. This interrupted the char feed to the reactor, and stopped the oxygen flow to the steam-oxygen gasifier. A higher speed motor was the only one available over the weekend, but operating the screw feeder with it proved to be unsatisfactory. A new motor was installed on this feeder on October 11, and char feed to the reactor was reinitiated at 0200 hours that same day. On October 12, a pressure upset in the absorber system again interrupted solids flow to the reactor. The gas purification section was taken out of service, and char feed to the reactor was reestablished at 0700 hours on October 12. Steady-state, self-sustained operations were again achieved at a char feed rate of approximately 2 tons/hr and at coal conversions ranging from 78% to 85%.

On October 17 at 0700 hours, city water pressure to the HYGAS pilot plant was suddenly reduced. Later we found that a truck had backed into and knocked over a fire hydrant upstream of the HYGAS pilot plant city water supply. This situation was not corrected until 1000 hours on October 17. This sudden loss of city water pressure to the plant forced the termination of Test 66, which had operated for a total of 290 hours, starting from the beginning of char feed to the reactor at 0500 hours on October 5. A total of 424 tons of pretreated char was fed to the reactor at an average rate of about 2 tons/hr. Over 9 days of reactor operation with oxygen and char feed were logged for Test 66.

Pretreater operation during Test 66 was excellent. Eleven and one-half days of operation were recorded at an average coal feed rate of 1.9 tons/hr. A total of 534 tons of raw coal was processed through the pretreater at a temperature of about 760°F, producing nonagglomerating char for the HYGAS reactor.

The quench section operated satisfactorily during Test 66. The purification section was put on-stream for Test 66. Foaming continued to be a problem in the absorber tower. The use of antifoam additives offered only temporary relief. On October 12, a pressure upset in the absorber tower stopped char feed to the reactor, and the purification section was taken off-stream. Due to the unsteady operation of the purification section, the methanation section was not put on-line for Test 66. The effluent cleanup section operated satisfactorily for Test 66.

Several steady-state periods for both the pretreater and the reactor were selected for heat and material balances and engineering information. Table 1 presents some preliminary quick ash results for Test 66.

An inspection of the plant following Test 66 indicated that the pretreater reactor and the char cooler were both in excellent condition and free of clinkers. A slight buildup of fines and tar was found on the bottom of the pretreater quench tower, and fines and tar had accumulated at the inlet of the venturi scrubber. Five distribution nozzles on the pretreater grid were broken. We fixed the pretreater section and cleaned it for Test 67. The slurry preparation section was in good condition following Test 66, and the slurry mix tank was emptied in preparation for Test 67.

We inspected the reactor after Test 66, and found it in excellent condition. The slurry dryer area and the first- and second-stage gasifiers were all clean, with no plugs in any of the transfer lines between them. A small piece of clinker was found clinging to the refractory wall behind line 339, 6 feet above the steam-oxygen sparger ring. This clinker could have been left over from previous tests, because the rest of the steam-oxygen gasifier was completely clean and in good condition.

When we inspected the quench section, we found that some solids had accumulated in the quench separator and the prequench tower. These vessels and the quench section lines were cleaned and readied for Test 67. The purification section's foaming problems are still being studied. A consultant from Benfield Corp. was invited to the plant, and he suggested various antifoam agents for use in the next test.

Table 1. PRELIMINARY QUICK ASH RESULTS[a] FOR TEST 66

Date (Time)	Char Feed[c] Rate, lb/hr	Ash in Spent Char	Ash in Pretreated Char	Char Gasified	lb O_2/lb Char Fed	lb Steam/lb Char Fed	SOG Bed Temp, °F
		%	%	%			
10/6 (0600)	4411	25.9	14.7	50.7	0.18	1.9	1590
10/6 (1430)	3822	35.2	13.9	70.3	0.23	2.3	1600
10/8 (0600)	2959	38.7	14.1	74.0	0.32	3.1	1623
10/11 (0615)[b]	3513	51.9	13.3	85.8	--	--	--
10/12 (0405)[b]	1498	26.7	13.4	57.5	--	--	--
10/12 (2005-0005)	4333	50.6	13.4[d]	84.9	0.25	2.3	1625
10/13 (1015)	4127	40.4	13.0	78.0	0.25	2.3	1636
10/14 (0600)	4157	61.4	13.1	90.5	0.25	2.3	1633
10/15 (0600)	4262	47.6	13.7	82.5	0.23	2.2	1604
10/16 (0600)	4290	55.2	12.1	88.8	0.24	2.2	1638
10/17 (0600)	4092	61.7	13.1	90.6	0.24	2.3	1602

[a] Note: These char gasification results must be confirmed by additional analytical tests.

[b] These samples were taken during process upsets.

[c] Wet basis.

[d] Assumed.

The effluent cleanup section was also in good condition after Test 66. A small amount of solids was found in the light-oil recovery unit's quench separator. The cleanup section was prepared for the next test.

The results of some steady-state periods for Test 64 have been completed and are presented in Tables 2 through 4 and Figures 1 through 9. The mechanical status of the HYGAS pilot plant for October is shown in Figure 10.

Task 8. Demonstration Plant Support

One of the major activities under this task has been the transfer of data to Procon, Inc. During the reporting period, the following data were supplied:

a. Current piping and instrument drawings for the HYGAS pilot plant and drawings of the HYGAS reactor

b. Monthly work reports for the HYGAS Environmental Assessment Program

c. Drawings and operating data for the high-pressure cyclone

d. Fluidization data at high and low pressures for lignite, coal, and chars

e. Pilot plant data on lignite and bituminous coals that give a basis for cyclone efficiency, bed densities, and line densities

f. Suggested design bed densities for lignite and bituminous coals in the various reactor zones

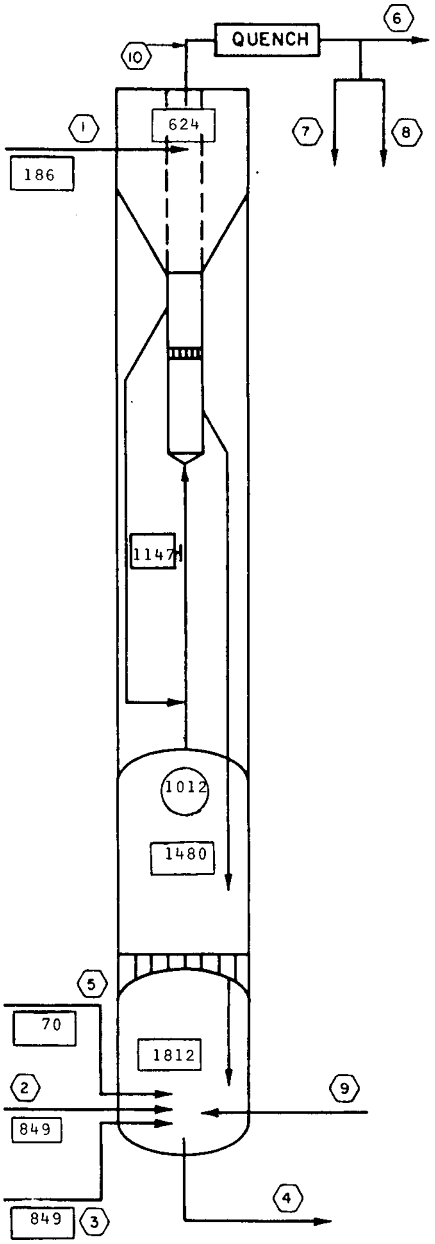
g. Suggested design gas velocities in the various beds.

IGT has also supplied heat and material balances for gasifiers operating at pressures of 800, 1000, and 1200 psig when using lignite feed. These balances were the results of initial gasifier cost-evaluation studies. For these balances, the cost of coal, steam, oxygen, CO_2 removal, and gasifier construction (based upon "Braun Factored Estimates for Western Coal Commercial Concepts") were evaluated for minimization of overall system cost. Additional variables such as L/D ratios in the gasifier beds, carbon conversion, and maximum temperature were also considered. The resulting balances, Figures 11 through 16 and Tables 5 through 8, appear to be reasonable design bases for the lignite study. These balances are based upon the results of several dozen computer runs with the IGT HYGAS Process for gasification. When compared with the preliminary balance supplied earlier, Figure 11, these later computer calculations result in reduced steam and oxygen consumption but in larger gasifier size for an overall reduced system cost. IGT does not represent these

Table 2. MATERIAL BALANCE SUMMARY FOR THE HYGAS GASIFIER FOR TEST 64 FROM 8/24/77 (1500 Hours) TO 8/26/77 (0700 Hours)

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

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



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

MATERIAL BALANCE, %

C	100
H	104
O	104
ASH	94

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

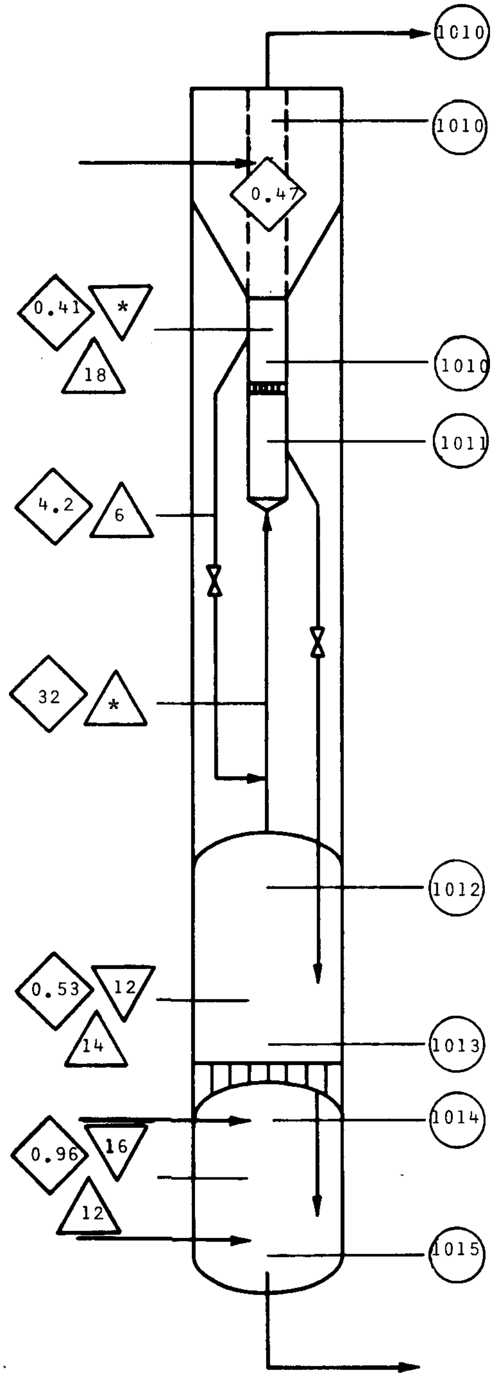
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	69.49	74.85	59.20
H	3.46	2.59	0.87
N	1.41	1.19	0.28
Cl	--	--	--
S	4.22	2.83	0.84
ASH	12.62	14.04	38.43
O	8.80	4.50	0.38
lb/hr (dry)	4286	579	1115
MOISTURE, wt %	2.5		
SLURRY CONCN, wt %	34		

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

GAS FEED STREAMS			
2	3	5	9
mol/hr			
348	83	30	0

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

Figure 1. HYGAS REACTOR DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/24/77 (1500 Hours) TO 8/26/77 (0700 Hours)



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

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

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

BY ASH BALANCE

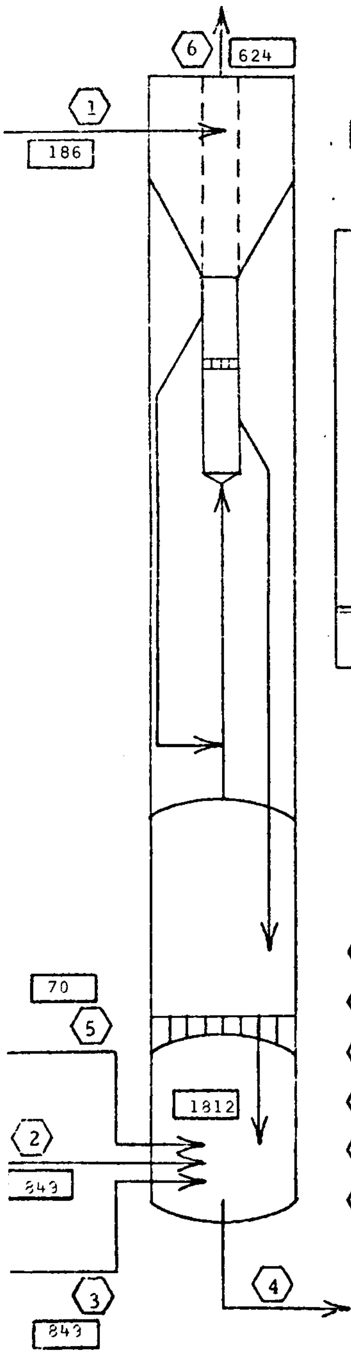
MAF† COAL GASIFIED, % = 77
CARBON GASIFIED, % = 72
METHANE YIELD SCF/lb COAL FED = 4.0
EQUIVALENT METHANE YIELD, SCF/ lb COAL FED = 6.3

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 10
SOG = 14

†MOISTURE ASH FREE.

Figure 2. HYGAS REACTOR ENGINEERING DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/24/77 (1500 Hours) TO 8/26/77 (0700 Hours)



⬡ Stream No.

▭ Temperature, °F

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

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

(1) Feed Slurry

(2) High-Pressure Steam to Steam-Oxygen Sparger

(3) High-Pressure Steam to Stripping Ring

(4) Spent Char

(5) High-Pressure Oxygen to Steam-Oxygen Sparger

(6) Reactor Overhead

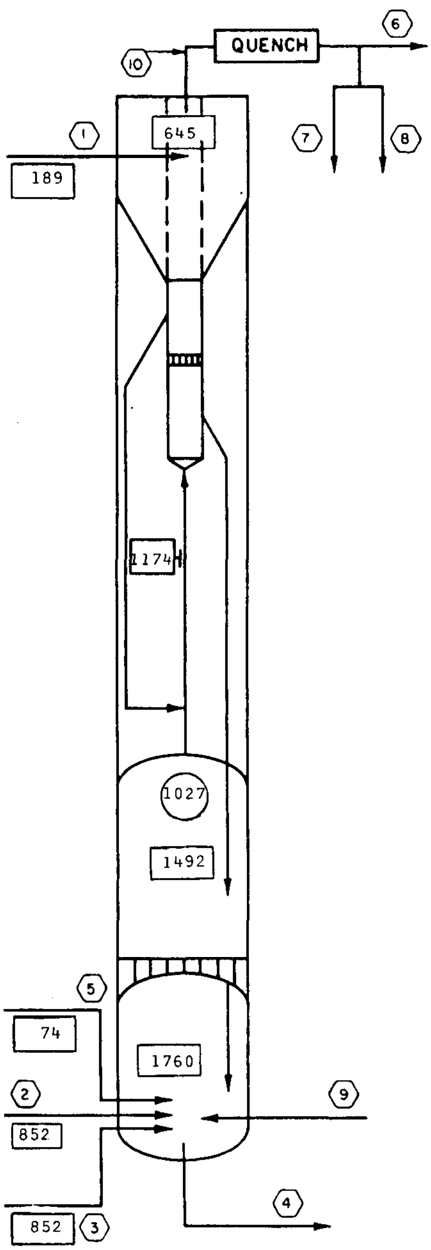
* High heating value.

Figure 3. HYGAS REACTOR HEAT BALANCE DATA SHEET FOR TEST 64 FOR STEADY PERIOD FROM 8/24/77 (1500 Hours) TO 8/26/77 (0700 Hours)

Table 3. MATERIAL BALANCE SUMMARY FOR THE HYGAS GASIFIER FOR TEST 64 FROM 8/22/77 (1500 Hours) TO 8/23/77 (0700 Hours)

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

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



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

MATERIAL BALANCE, %	
C	100
H	101
O	102
ASH	96

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

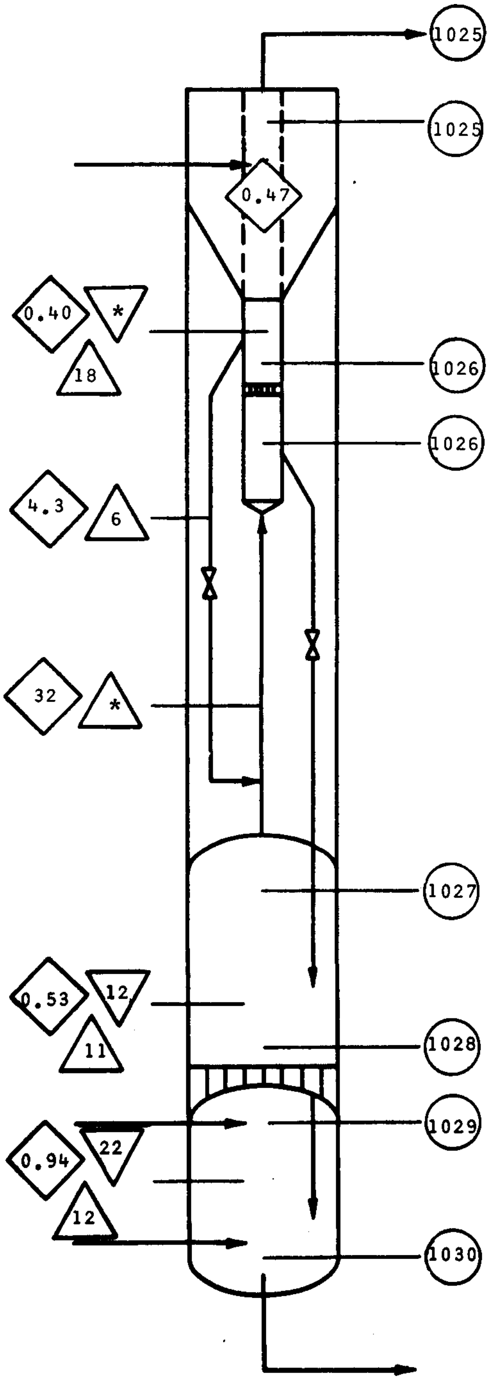
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	69.00	74.78	45.65
H	3.58	2.69	0.69
N	1.44	1.20	0.19
Cl	--	--	--
S	4.77	2.96	0.63
ASH	13.49	14.04	52.50
O	7.72	4.33	0.34
lb/hr (dry)	4263	900	812
MOISTURE, wt %	2.2		
SLURRY CONCN, wt %	33		

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

GAS FEED STREAMS			
2	3	5	9
mol/hr			
349	83	33	0

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

Figure 4. HYGAS REACTOR DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/22/77 (1500 Hours) TO 8/23/77 (0700 Hours)



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

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

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

BY ASH BALANCE

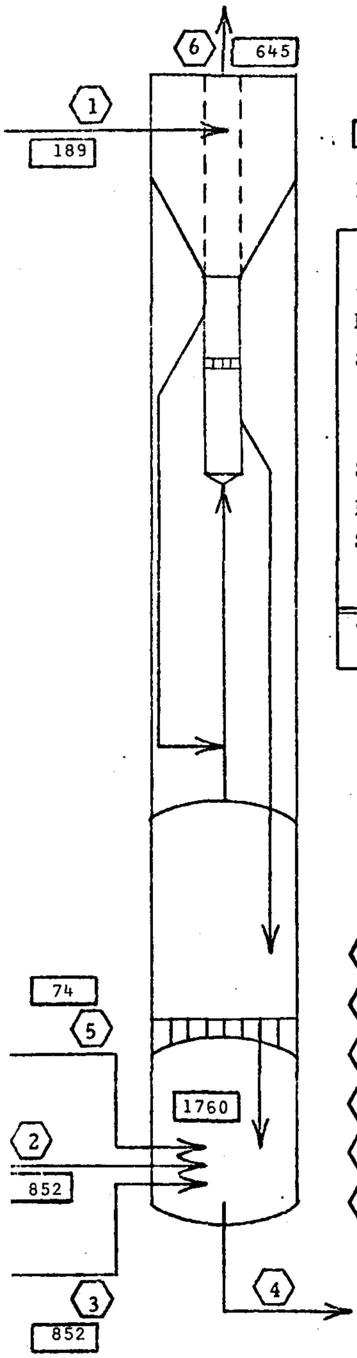
MAF† COAL GASIFIED, % = 86
CARBON GASIFIED, % = 84
METHANE YIELD SCF/lb COAL FED = 4.4
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 7.0

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 10
SOG = 18

†MOISTURE ASH FREE.

Figure 5. HYGAS REACTOR ENGINEERING DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/22/77 (1500 Hours) TO 8/23/77 (0700 Hours)



Stream No.

Temperature, °F

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

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

1 Feed Slurry

2 High-Pressure Steam to Steam-Oxygen Sparger

3 High-Pressure Steam to Stripping Ring

4 Spent Char

5 High-Pressure Oxygen to Steam-Oxygen Sparger

6 Reactor Overhead

* High heating value.

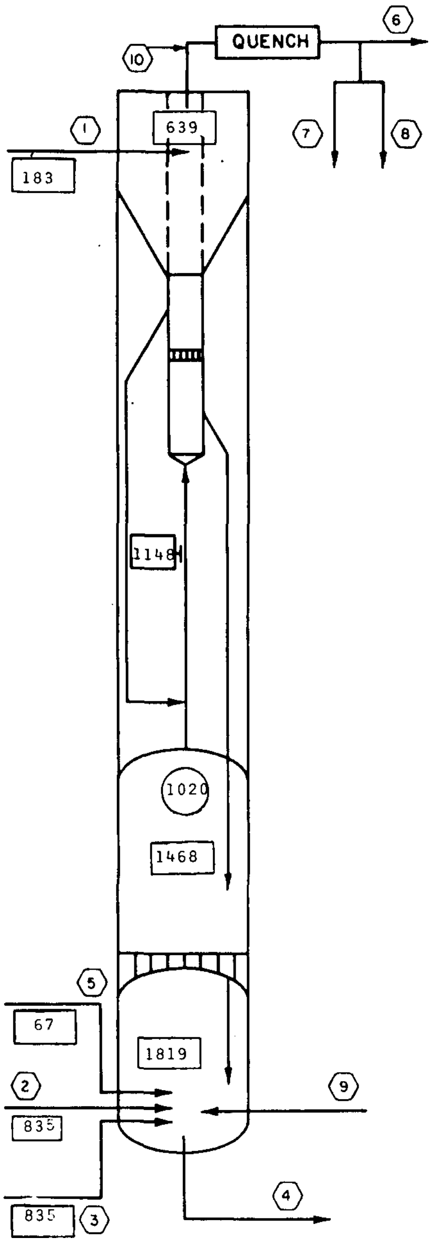
Figure 6. HYGAS REACTOR HEAT BALANCE DATA SHEET FOR TEST 64 FOR STEADY PERIOD FROM 8/22/77 (1500 Hours) TO 8/23/77 (0700 Hours)

Table 4. MATERIAL BALANCE SUMMARY FOR THE HYGAS GASIFIER FOR TEST 64 FROM 8/23/77 (1600 Hours) TO 8/24/77 (0500 Hours)

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

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

*Coal feed analysis taken from a previous time period.



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

MATERIAL BALANCE, %

C	97
H	101
O	101
ASH	92

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

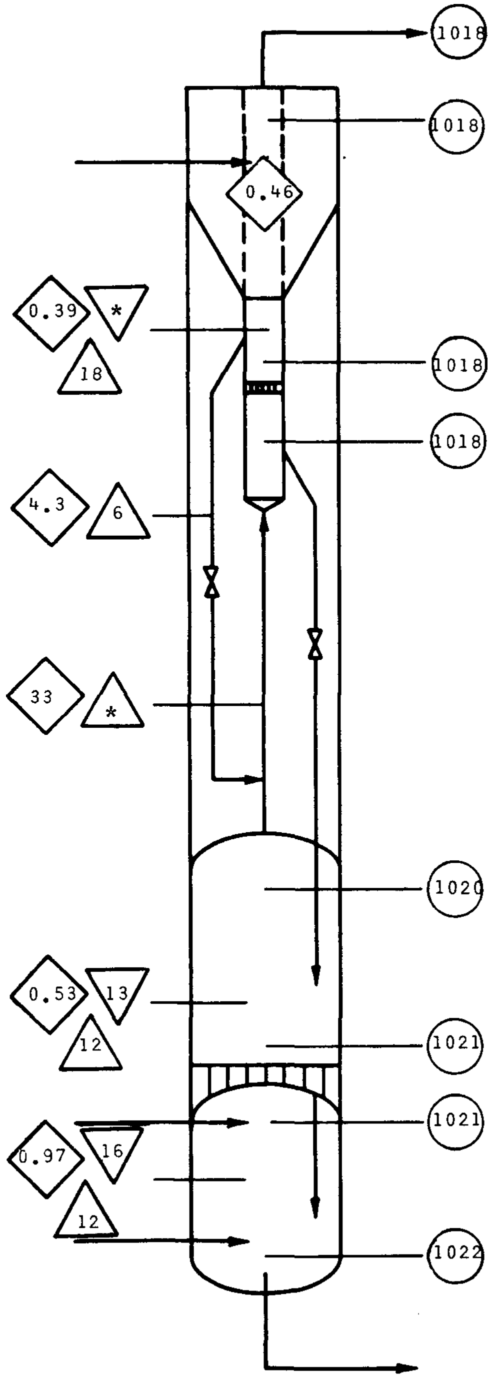
	SOLIDS STREAMS		
	1	10	4
wt % (dry)			
C	69.49	76.63	58.44
H	3.46	2.75	0.83
N	1.41	1.19	0.26
Cl	--	--	--
S	4.22	2.88	0.93
ASH	12.62	12.29	38.97
O	8.80	4.26	0.57
lb/hr (dry)	4539	833	1095
MOISTURE, wt %	1.8		
SLURRY CONCN, wt %	36		

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

GAS FEED STREAMS			
2	3	5	9
mol/hr			
352	84	30	0

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

Figure 7. HYGAS REACTOR DATA FOR TEST 64 FOR STEADY PERIOD FROM 8/23/77 (1600 Hours) TO 8/24/77 (0500 Hours)



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

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

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

BY ASH BALANCE

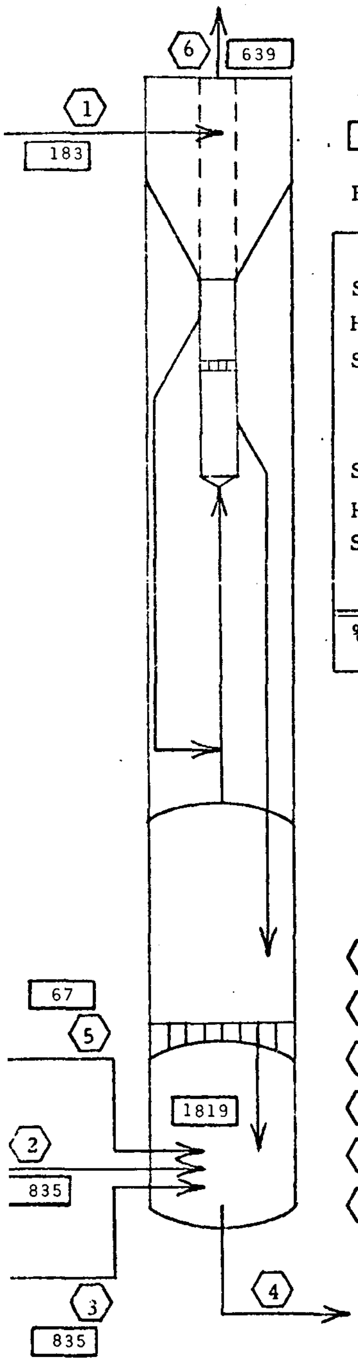
MAF† COAL GASIFIED, % = 77
CARBON GASIFIED, % = 73
METHANE YIELD SCF/lb COAL FED = 3.9
EQUILVALENT METHANE YIELD, SCF/lb COAL FED = 5.9

BED HEIGHT, ft

SLURRY DRYER = 2
HTR = 11
SOG = 14

†MOISTURE ASH FREE.

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



⬡ Stream No.

▭ Temperature, °F

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

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

1. Feed Slurry
2. High-Pressure Steam to Steam-Oxygen Sparger
3. High-Pressure Steam to Stripping Ring
4. Spent Char
5. High-Pressure Oxygen to Steam-Oxygen Sparger
6. Reactor Overhead

* High heating value.

Figure 9. HYGAS REACTOR HEAT BALANCE DATA SHEET FOR TEST 64 FOR STEADY PERIOD FROM 8/23/77 (1600 Hours) TO 8/24/77 (0500 Hours)

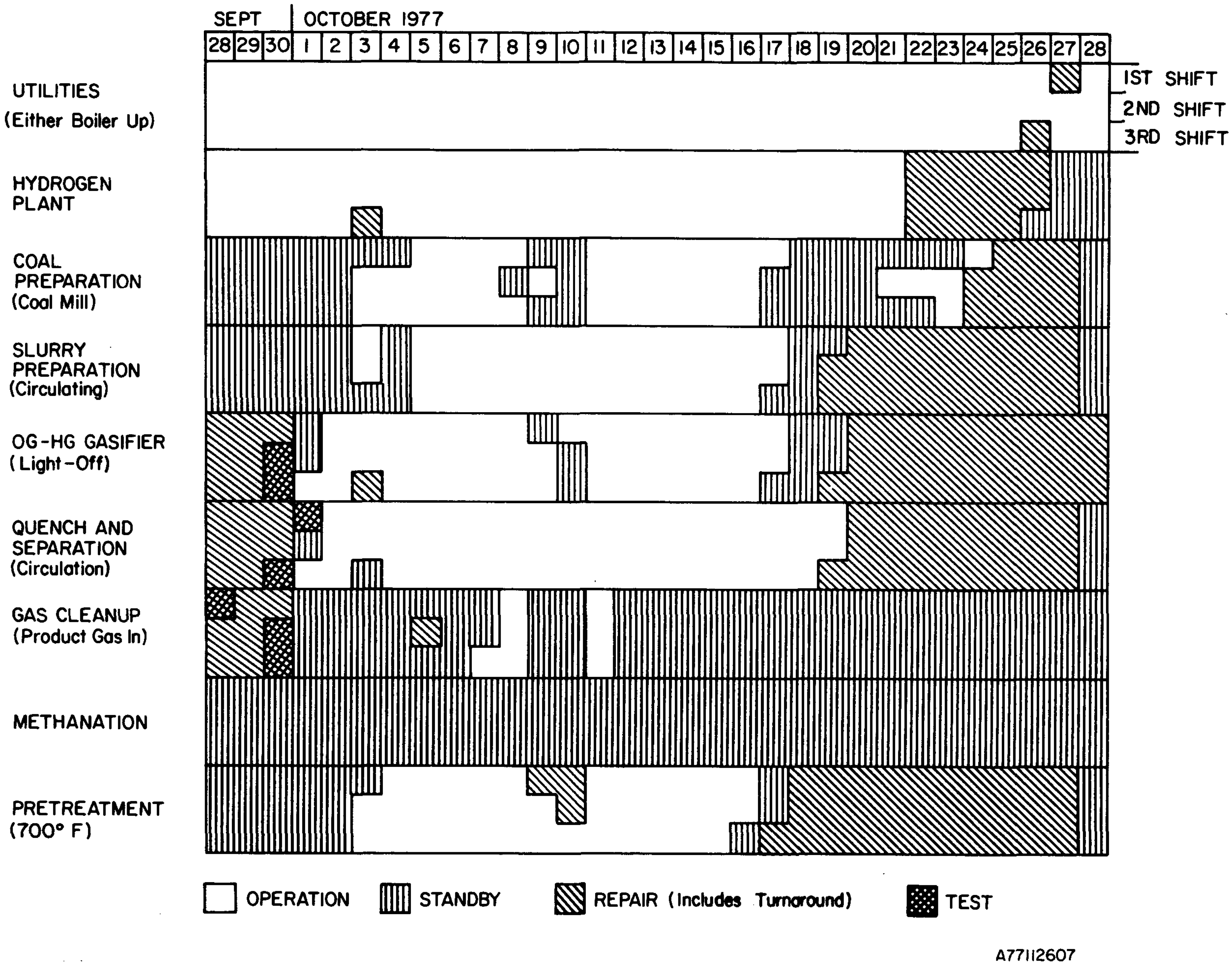


Figure 10. MECHANICAL STATUS OF THE HYGAS PLANT FOR OCTOBER 1977

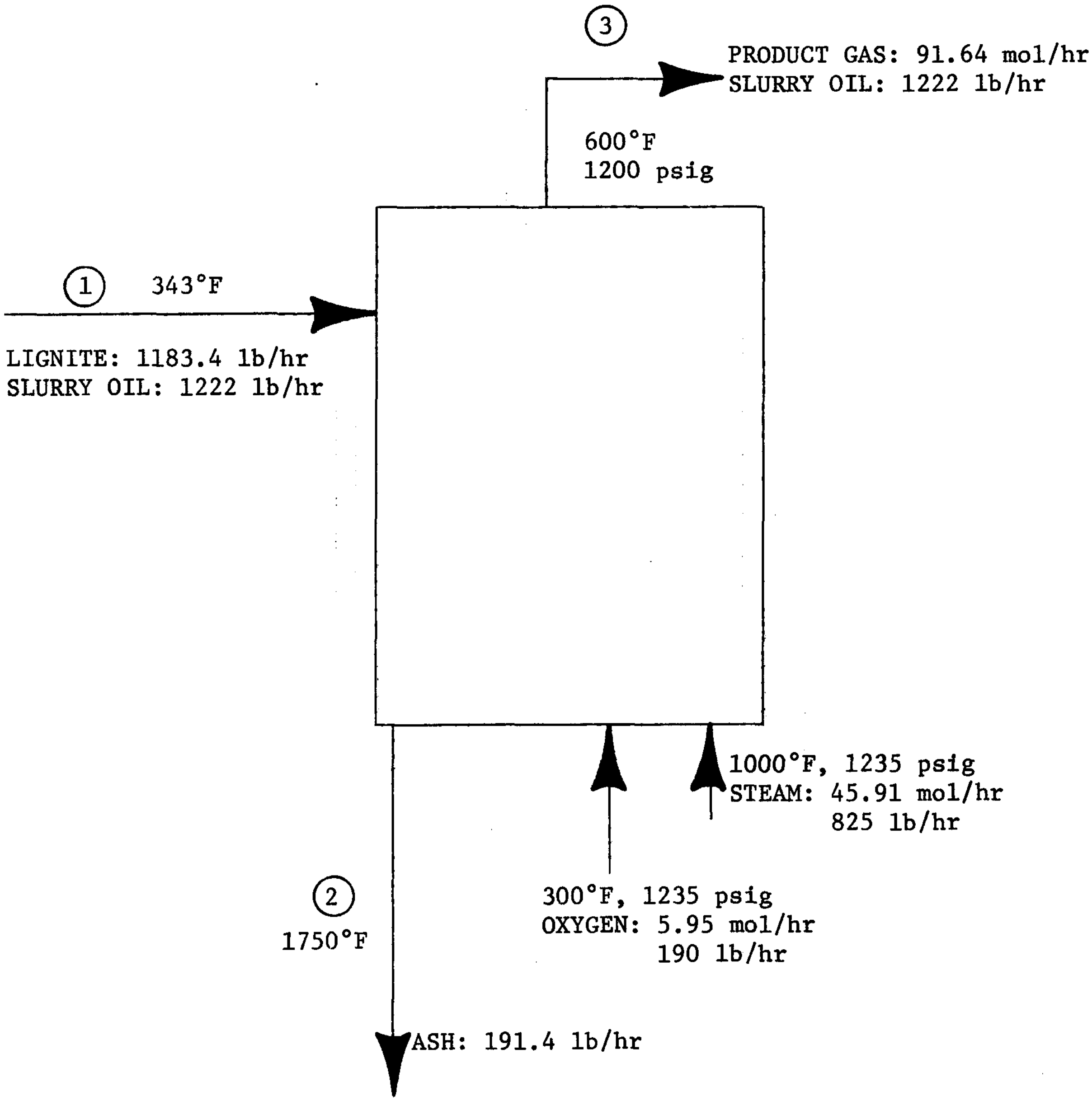


Figure 11. CALCULATED OVERALL GASIFIER MATERIAL BALANCE FOR MONTANA LIGNITE COAL (1200-psig Case)

Table 5. CALCULATED GASIFIER MATERIAL AND HEAT BALANCE FOR MONTANA LIGNITE COAL (1200-psig Case)

Stream No.	1			2		
Description	Lignite Feed			Ash Residue		
Temperature, °F	343			1750		
Components	lb/hr	wt%	mol/hr	lb/hr	wt%	mol/hr
C	612.7	61.27	51.02	61.3	32.03	5.10
H_2	42.0	4.20	20.83	1.5	0.78	0.76
O	201.0	20.10	12.56	--	--	--
N_2	9.7	0.97	0.35	1.9	0.99	0.07
S	8.8	0.88	0.27	0.9	0.47	0.03
Ash	125.8	12.58	--	125.8	65.73	--
Total	1000.0	100.00		191.4	100.00	
Moisture	183.4					
Slurry Oil	1222					

Stream No.	3	
Description	Raw Product Gas	
Temperature, °F	600	
Components	mol/hr	mol%
CO	12.05	13.25
CO_2	17.35	19.07
H_2	15.86	17.44
H_2O	33.81	37.18
CH_4	10.36	11.39
C_2H_6	0.77	0.85
NH_3	0.44	0.48
N_2	0.04	0.04
HCN	0.03	0.03
H_2S	0.24	0.26
COS	0.01	0.01
Total (Oil-Free Gas)	90.96	100.00
	mol/hr	wt%
C_6H_6	0.12	15.0
C_7H_8	0.56	85.0
Total (Product Oil)	0.68	100.0
Total (Oil-Free Gas + Product Oil)	91.64	
Slurry Oil, lb/hr	1222	

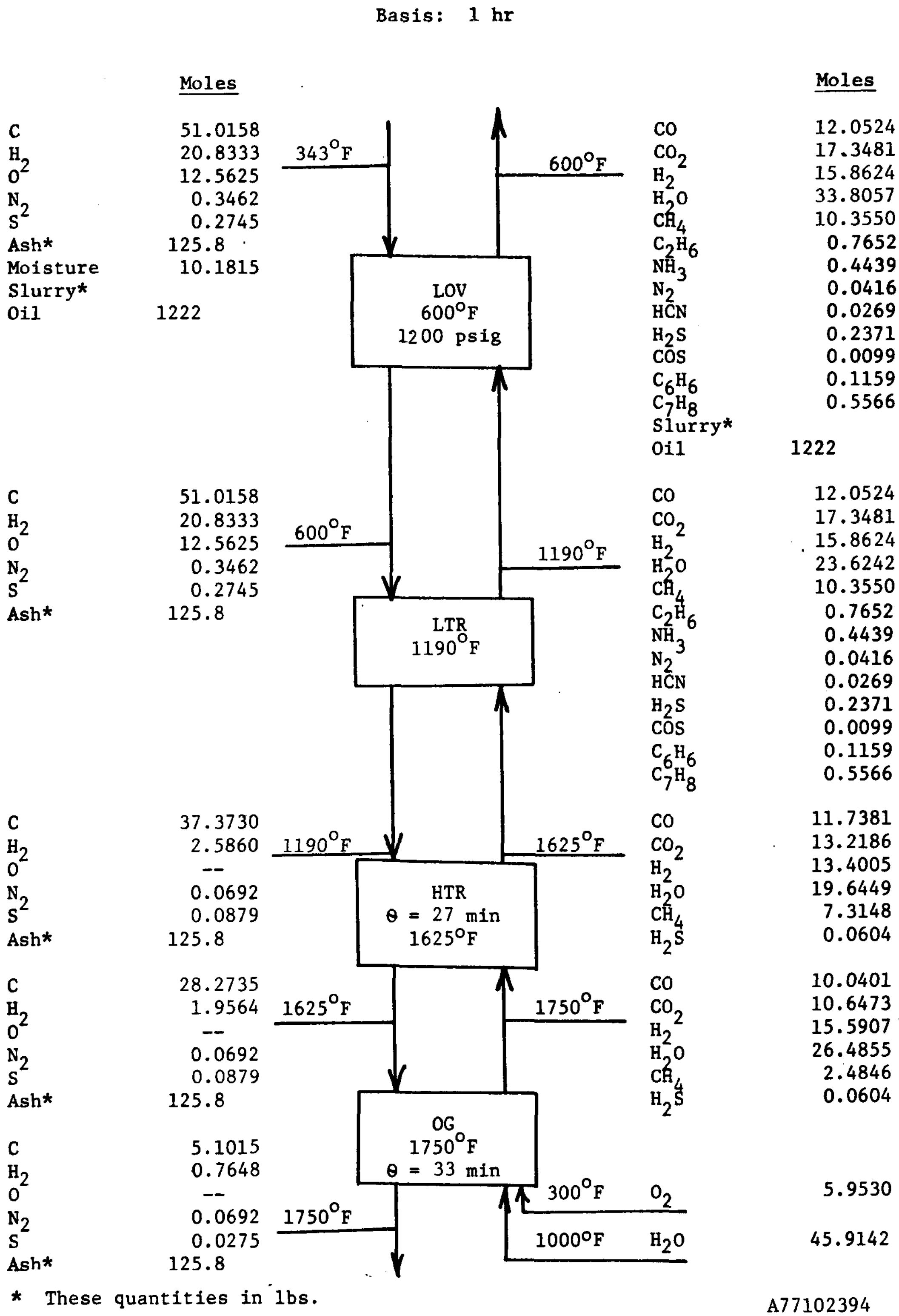


Figure 12. CALCULATED GASIFIER MATERIAL BALANCE FOR MONTANA LIGNITE COAL SHOWING INTERSTAGE COMPOSITION (1200-psig Case)

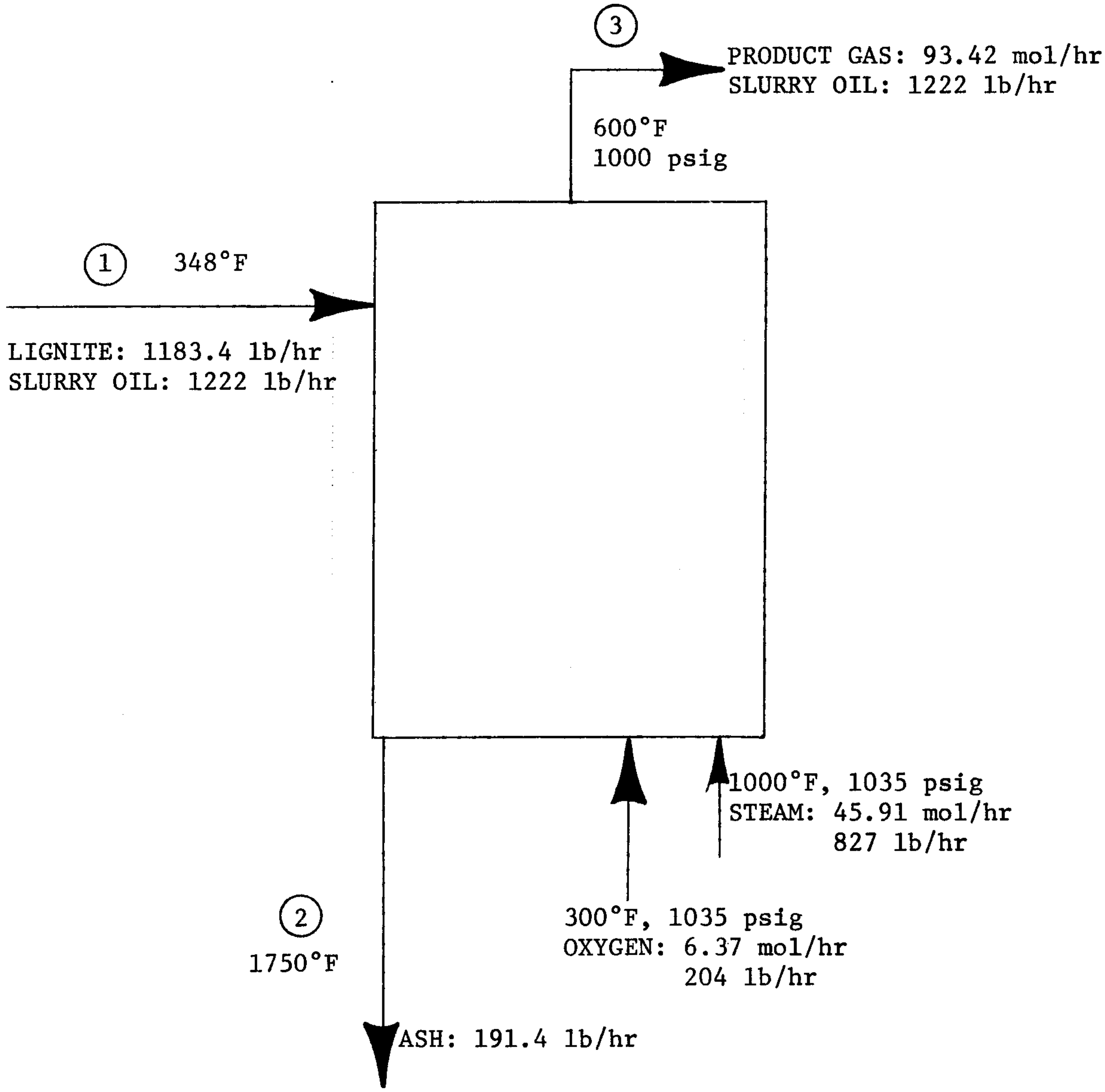


Figure 13. CALCULATED OVERALL GASIFIER MATERIAL BALANCE FOR MONTANA LIGNITE COAL (1000-psig Case)

Table 6. CALCULATED GASIFIER MATERIAL AND HEAT BALANCE FOR MONTANA LIGNITE COAL (1000-psig Case)

Stream No.	1			2		
Description	Lignite Feed			Ash Residue		
Temperature, °F	348			1750		
Components	lb/hr	wt%	mol/hr	lb/hr	wt%	mol/hr
C	612.7	61.27	51.02	61.3	32.01	5.10
H_2	42.0	4.20	20.83	1.6	0.84	0.79
O	201.0	20.10	12.56	--	--	--
N_2	9.7	0.97	0.35	1.9	0.99	0.07
S	8.8	0.88	0.27	0.9	0.47	0.03
Ash	125.8	12.58	--	125.8	65.69	--
Total	1000.0	100.00		191.5	100.00	
Moisture	183.4					
Slurry Oil	1222					

Stream No.	3	
Description	Raw Product Gas	
Temperature, °F	600	
Components	mol/hr	mol%
CO	13.14	14.17
CO_2	17.17	18.51
H_2	17.55	18.93
H_2O	33.91	36.57
CH_4	9.44	10.18
C_2H_6	0.77	0.83
NH_3	0.44	0.47
N_2	0.04	0.04
HCN	0.03	0.03
H_2S	0.24	0.26
COS	0.01	0.01
Total (Oil-Free Gas)	92.74	100.00
	mol/hr	wt%
C_6H_6	0.12	15.0
C_7H_8	0.56	85.0
Total (Product Oil)	0.68	100.0
Total (Oil-Free Gas + Product Oil)	93.42	
Slurry Oil, lb/hr	1222	

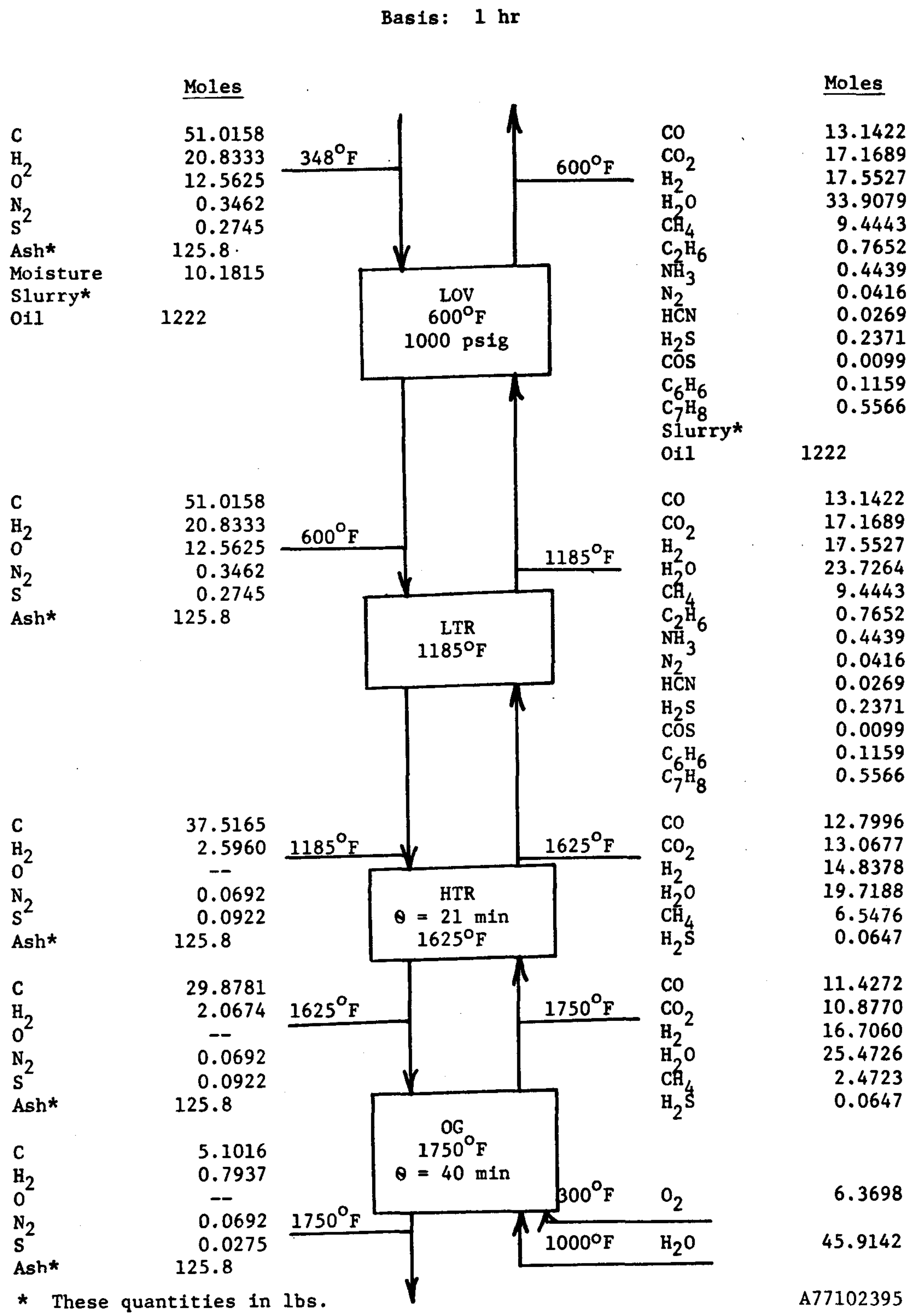


Figure 14. CALCULATED GASIFIER MATERIAL BALANCE FOR MONTANA LIGNITE COAL SHOWING INTERSTAGE COMPOSITION (1000-psig Case)

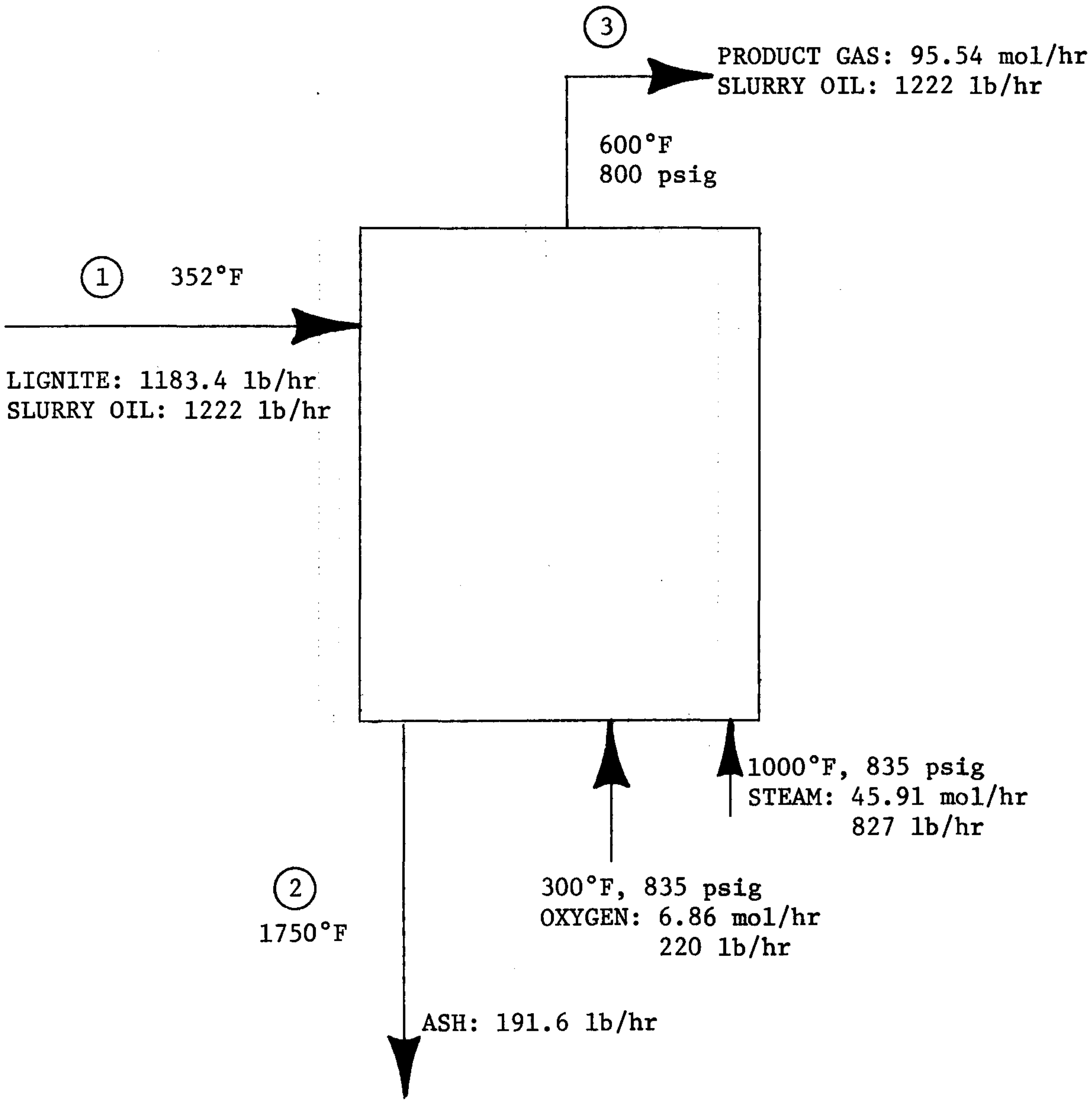


Figure 15. CALCULATED OVERALL GASIFIER MATERIAL BALANCE FOR MONTANA LIGNITE COAL (800-psig Case)

Table 7. CALCULATED GASIFIER MATERIAL AND HEAT BALANCE FOR MONTANA LIGNITE COAL (800-psig Case)

Stream No.	1			2		
Description	Lignite Feed			Ash Residue		
Temperature, °F	352			1750		
Components	lb/hr	wt%	mol/hr	lb/hr	wt%	mol/hr
C	612.7	61.27	51.02	61.3	31.99	5.11
H_2	42.0	4.20	20.83	1.7	0.89	0.83
O	201.0	20.10	12.56	--	--	--
N_2	9.7	0.97	0.35	1.9	0.99	0.07
S	8.8	0.88	0.27	0.9	0.47	0.03
Ash	125.8	12.58	--	125.8	65.66	--
Total	1000.0	100.00		191.6	100.00	
Moisture	183.4					
Slurry Oil	1222					

Stream No.	3	
Description	Raw Product Gas	
Temperature, °F	600	
Components	mol/hr	mol%
CO	14.41	15.19
CO_2	16.97	17.89
H_2	19.56	20.63
H_2O	34.02	35.87
CH_4	8.37	8.82
C_2H_6	0.77	0.81
NH_3	0.44	0.46
N_2	0.04	0.04
HCN	0.03	0.03
H_2S	0.24	0.25
COS	0.01	0.01
Total (Oil-Free Gas)	94.86	100.00
	mol/hr	wt%
C_6H_6	0.12	15.0
C_7H_8	0.56	85.0
Total (Product Oil)	0.68	100.0
Total (Oil-Free Gas + Product Oil)	95.54	
Slurry Oil, lb/hr	1222	

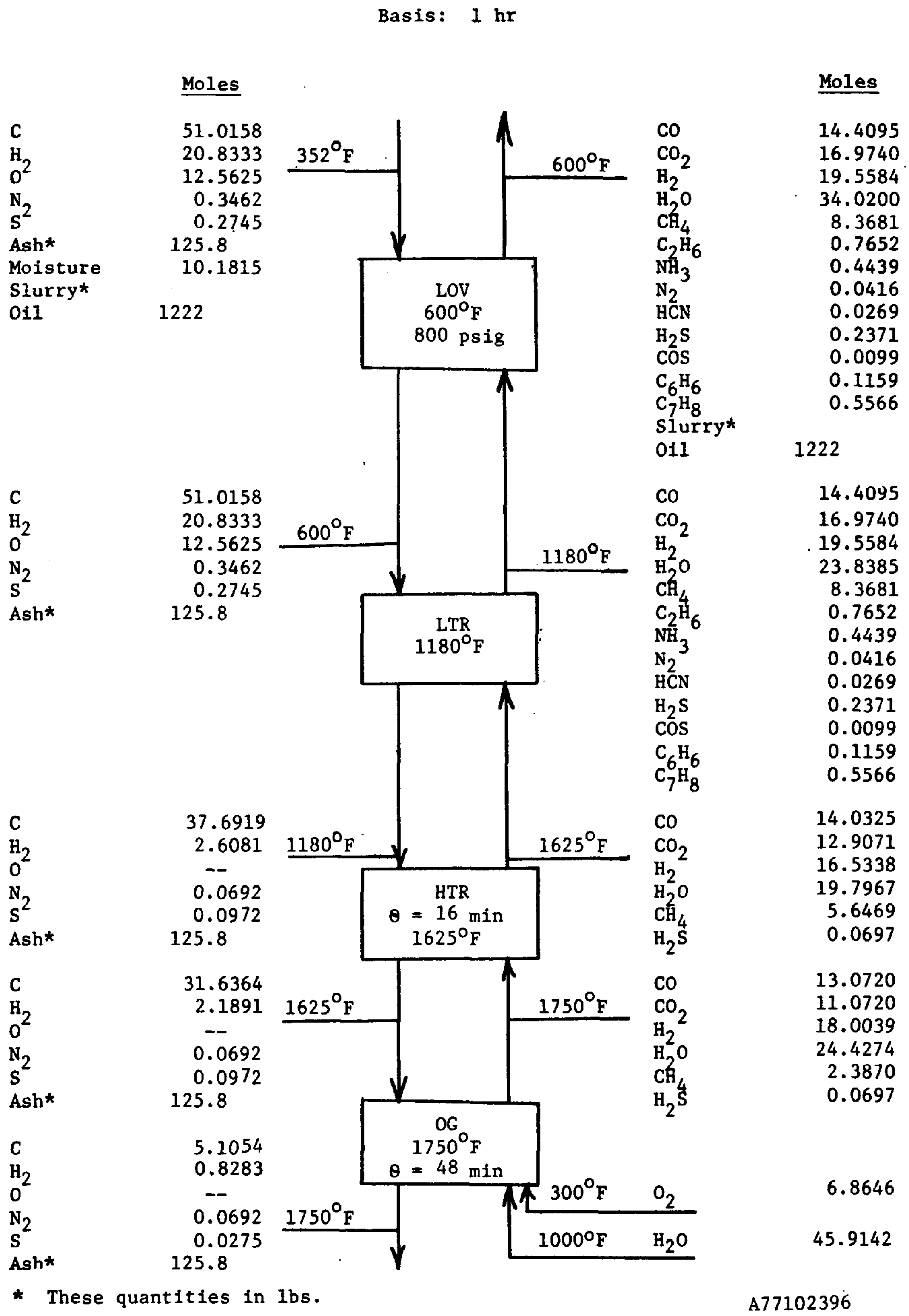


Figure 16. CALCULATED GASIFIER MATERIAL BALANCE FOR MONTANA LIGNITE COAL SHOWING INTERSTAGE COMPOSITION (800-psig Case)

balances as the absolute optimum, but rather as good guidelines for a process design study to reduce cost. Part of this reluctance to claim an optimum is due to the basic cost data. Relative to other information, the basic cost of steam (and, therefore, of oxygen) appears to be high in this data base. Another factor is the selection of the design point. Further cost reduction may be possible at increased temperatures, bed L/D ratios, or carbon conversions, but confirmatory testing would be desirable before specifying these design bases.

Table 8. PROXIMATE AND ULTIMATE ANALYSIS FOR TYPICAL MONTANA LIGNITE COAL

Proximate Analysis, wt%	
Volatile Matter	35.47
Moisture	15.50
Fixed Carbon	38.40
Ash	10.63
Total	100.00
Ultimate Analysis, dry, wt%	
Carbon	61.27
Hydrogen	4.20
Oxygen	20.10
Nitrogen	0.97
Sulfur	0.88
Ash	12.58
Total	100.00

High Heating Value = 10,182 Btu/lb (dry)

Task 9. Support Studies

A. Plant Effluent Processing

The effluent cleanup section operated during both Tests 65 and 66. When we inspected the section after both tests, we found only a small amount of solids in the quench separator of the light-oil recovery unit. The high-capacity incinerator was started, but problems were encountered in maintaining steady pilot flames in the unit. Steps are being taken to correct the situation.

B. Test Methanation Systems and Catalysts

The IGT fixed-bed catalyst methanation section was on standby for Test 66. Because of the intermittent operation of the purification section, it was not put on-stream. It will be ready for Test 67.

Following Test 64, Chem Systems' liquid-phase methanation (LPM) pilot unit was put on-line using synthesis gas from the HYGAS hydrogen plant. Some problems in elutriating the catalyst from the methanator reactor were experienced in the LPM pilot unit, and the test was terminated. A study is currently under way by the Chem Systems group to modify the methanation reactor to minimize the amount of elutriation of catalyst fines from the reactor. The LPM unit will not be ready for Test 67.

C. Investigate Hot-Oil Quench System

Design work for the hot-oil quench system is continuing.

D. Materials Testing

MPC corrosion and erosion test coupons were exposed during Test 66.

The cyclone was taken to Argonne National Laboratory's Material Science Division during the HYGAS plant's annual July shutdown. It was inspected by Mark I Pulser-Receiver Sonic Tests, as well as with a boroscope, which allowed a visual inspection of the cyclone's internals to be made.

The view through the boroscope revealed gouging caused by the solids vortex at the cyclone solids discharge nozzle of the cyclone. This wear was also evident in the wall thickness measurements taken by ultrasonics. One of the latest readings taken at the cyclone solids discharge nozzle indicated that the wall thickness had been reduced as much as 30% since the cyclone outlet was first inspected in February 1976. The rate of erosion is approximately 7.7 mils per 100 hours of feeding.

The gouging of the solids discharge line was rapid enough to warrant our modifying the existing cyclone by installing a slip liner, which was made of Incoloy 800 and had an inside diameter of 2.30 inches and a wall thickness of 0.275 inch, in the outlet discharge.

Ultrasonic readings were also taken around the inlet region of the stellite overlay. The most rapid erosion appears to be taking place in the unlined region of the stellite, about 205 degrees clockwise of the solids inlet tangent. At this point the wall thickness has been worn to 0.756 inch. Since the minimum wall thickness allowed is 0.378 inch, there still remains twice the minimum wall thickness required. The rate of erosion, thus far, is equivalent to approximately 7.8 mils per 100 hours; therefore, giving the unstellited region 4800 hours of service life before the critical wall thickness is reached.

In the stellite-lined region of the cyclone, the wall thickness has been worn to 0.799 inch after 2300 hours of service. The average rate of erosion indicates that there are about 6000 hours of service remaining before the 0.125-inch-thick stellite overlay is worn through. A wear rate of 1.8 mils per 100 hours is occurring on the stellited surface.

The stellite overlay has superior erosion resistance (by a factor of more than 4) compared with the base metal, Type-316 stainless steel. In future cyclone designs, we recommend using the overlay over a wider area.

E. Engineering Services

The results of the 6-inch simulation of the steam-oxygen gasifier have been received and are presented in Table 9. This test was carried out to study the effect of residence time on the degree of conversion achievable in the steam-oxygen gasifier. Results indicate that high char conversion can be achieved in the steam-oxygen gasifier as it is presently being operated in the HYGAS reactor.

Many questions have been raised on the method used to calculate coal and carbon conversions in the HYGAS reactor. The following derivations indicate the origin of the equations.

Coal Conversion Equation

% Ash Balance

X_{coal} is defined as the conversion (%) of dry, ash-free coal fed minus the cyclone fines:

$$X_{coal} = \frac{\text{ash-free coal in} - \text{ash-free coal out}}{\text{ash-free coal in}} \cdot 100$$

Table 9. RESULTS OF PROCESS DEVELOPMENT UNIT STEAM-OXYGEN GASIFICATION OF PEABODY NO. 10 MINE BITUMINOUS CHAR

Month	March		October
Test No.	1	2	3
Spent-Char Analysis, wt %			
Ash	92.22	77.84	77.03
Carbon	7.18	20.58	23.5
Hydrogen	0.12	0.27	0.29
Sulfur	0.40	1.15	0.45
Nitrogen	0.08	0.16	0.07
Oxygen (by Difference)	0	0	0
Mean Residence Time, min	22	25	11
Superficial Gas Velocity, ft/s	0.88	0.75	0.77
lb O_2/lb Char in Steam-Oxygen Gasifier	0.47	0.44	0.45
lb Steam/lb Char in Steam-Oxygen Gasifier	2.45	2.78	2.18
Carbon Gasified in Steam-Oxygen Gasifier (by Gas Balance), %	88	86	82.5
High Bed Temperature, °F	1805	1825	1840
Pressure, psig	399	398	403

Let W_f represent the flow rate of dry coal after subtracting the amount of dry cyclone fines from the actual amount of dry feed coal. Then —

A_f = weight fraction (dry basis) of ash in coal,

W_{sc} = flow rate of dry, spent char, and

A_{sc} = weight fraction (dry basis) of ash in spent char.

An ash balance is assumed:

$$\text{Ash in in feed} = \text{ash out in spent char}$$

$$W_f \cdot A_f = W_{sc} \cdot A_{sc}$$

Therefore —

$$W_f = W_{sc} \cdot \frac{A_{sc}}{A_f}$$

$$\text{Ash-free coal in} = W_f - W_f \cdot A_f = W_f(1 - A_f)$$

$$\text{Ash-free coal out} = W_{sc}(1 - A_{sc})$$

Substitute $W_f = W_{sc}\ A_{sc}/A_f$ in the above, and —

$$\text{Ash-free coal in} = W_{sc} \cdot A_{sc}/A_f(1 - A_f)$$

Substituting these quantities into the conversion equation yields —

$$X_{coal} = \frac{W_{sc} \cdot \frac{A_{sc}}{A_f}(1 - A_f) - W_{sc}(1 - A_{sc})}{W_{sc}\frac{A_{sc}}{A_f}(1 - A_f)} \cdot 100$$

$$= 1 - \frac{W_{sc}(1 - A_{sc})}{W_{sc}\frac{A_{sc}}{A_f}(1 - A_f)} \cdot 100 = \left(1 - \frac{A_f(1 - A_{sc})}{A_{sc}(1 - A_f)}\right) \cdot 100$$

<u>% Weight Balance</u>

In this case, the amount of dry, ash-free coal in the cyclone fines is subtracted from the feed, using explicit knowledge of their ash content and amount. We define the additional three quantities as follows:

W_d = flow rate of dry cyclone fines (dust), lb/hr

A_d = weight fraction (dry basis) of ash in the cyclone fines

$$X'_{coal} = 100 \cdot \frac{\text{ash-free coal in} - \text{ash-free coal in fines} - \text{ash-free coal in spent char}}{\text{ash-free coal in} - \text{ash-free coal in fines}}$$

and

Ash-free coal in = $W_f(1 - A_f)$

Ash-free coal in cyclone fines = $W_d(1 - A_d)$

Ash-free coal in spent char = $W_{sc}(1 - A_{sc})$

Substituting the above into the definition gives the equation—

$$X'_{coal} = \frac{W_f(1 - A_f) - W_d(1 - A_d) - W_{sc}(1 - A_{sc})}{W_f(1 - A_f) - W_d(1 - A_d)}$$

Carbon Conversion Equation

% Ash Balance

$$X_{carbon} = \frac{\text{carbon in} - \text{carbon out}}{\text{carbon in}} \cdot 100$$

The carbon in the coal is the only source of carbon to the gasifier. Only solid carbon out need be accounted for. We assume that W_f has been adjusted by subtracting the cyclone fines from the actual dry feed amount.

C_f = weight fraction carbon (dry basis) in feed coal

C_{sc} = weight fraction carbon (dry basis) in spent char

C_d = weight fraction carbon (dry basis) in cyclone fines (dust)

Carbon in = $W_f \cdot C_f$

Carbon out = $W_{sc} \cdot C_{sc}$

Ash Balance

$$\text{Ash in} = \text{ash out}$$

$$\text{Ash in} = W_f \cdot A_f$$

$$\text{Ash out} = W_{sc} \cdot A_{sc}$$

$$W_f \cdot A_f = W_{sc} \cdot A_{sc}$$

Therefore —

$$W_f = W_{sc}\frac{A_{sc}}{A_f}$$

Substituting into the conversion equation and simplifying yields —

$$X_{carbon} = \frac{\frac{W_{sc}A_{sc}}{A_f} \cdot C_f - W_{sc}C_{sc}}{\frac{W_{sc}A_{sc}}{A_f} \cdot C_f} \cdot 100$$

$$X_{carbon} = \left(1 - \frac{W_{sc}C_{sc}}{\frac{W_{sc}A_{sc}}{A_f}\ C_f}\right) \cdot 100 = 1 - \frac{A_f C_{sc}}{A_{sc}C_f} \cdot 100$$

This is in the form desired.

% Weight Balance

$$X'_{carbon} = \frac{\text{carbon in} - \text{carbon out}}{\text{carbon in}} \cdot 100$$

Carbon in = $W_fC_f - W_dC_d$ = carbon in in feed coal — carbon out in cyclone fines (dust)

Carbon out = $W_{sc}C_{sc}$

Plugging these in to the expression for X'_{carbon} yields —

$$X'_{carbon} = \frac{(W_fC - W_dC_d) - (W_{sc}C_{sc})}{W_fC_f - W_dC_d} \cdot 100$$

$$X'_{carbon} = \frac{W_fC_f - W_dC_d - W_{sc}C_{sc}}{W_fC_f - W_dC_d} \cdot 100$$

To recapitulate, the ash balance method uses the fact that ash is conserved (which is a fairly accurate approximation in most cases) to eliminate any flow rate measurements from being required to compute the coal or carbon conversion. The assumption that the cyclone fines are of the same composition as the feed is also made, and the conversion is made on the basis of the coal or carbon fed, excluding the contribution of fines. The weight basis method is devoid of such assumptions.

Contribution of Gasified Char to Cyclone (Overhead) Fines

If the overhead fines are a mixture of feed and spent char, then let —

X = weight fraction of fines from feed
y = weight fraction of fines from spent char

Carbon Balance

$$C_d = X\ C_f + y\ C_{sc} \qquad (1)$$

Ash Balance

$$A_d = X\ A_f + y\ A_{sc} \qquad (2)$$

From Equation 1 —

$$X = \frac{C_d - y\ C_{sc}}{C_f}$$

From Equation 2 —

$$X = \frac{A_d - y\ A_{sc}}{A_f}$$

but X + y = 1, and —

$$\frac{C_d}{C_f} - y\ \frac{C_{sc}}{C_f} = \frac{A_d}{A_f} - y\ \frac{A_{sc}}{A_f}$$

$$\frac{C_d}{C_f} - (1 - X)\ \frac{C_{sc}}{C_f} = \frac{A_d}{A_f} - (1 - X)\ \frac{A_{sc}}{A_f}$$

Rearranging gives —

$$y = (1 - X) = \left(\frac{C_d}{C_f} - \frac{A_d}{A_f}\right) / \left(\frac{C_{sc}}{C_f} - \frac{A_{sc}}{A_f}\right)$$

$$X = \left[1 - \left(\frac{C_d}{C_f} - \frac{A_d}{A_f}\right) \Big/ \left(\frac{C_{sc}}{C_f} - \frac{A_{sc}}{A_f}\right)\right].$$

IV. PROBLEMS

No problems were encountered in October.

V. PATENT STATUS

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

VI. WORK PLAN AND SCHEDULE

A work schedule of the HYGAS Process is presented in Figure 17.

Approved [signature]
Bernard S. Lee
Executive Vice President

Signed [signature]
Wilford G. Bair
Director

TASK 7. PILOT PLANT EXPERIMENTAL OPERATION

A. PEABODY NO. 10 MINE WASHED BITUMINOUS COAL*

1. INVESTIGATE HIGH CARBON CONVERSION

2. INVESTIGATE MAXIMUM PILOT PLANT CAPACITY

B. INVESTIGATE MINIMUM PRETREATMENT CONDITIONS

C. INVESTIGATE FINES UTILIZATION*

1. TEST DOUBLE SCREENED FEED

2. TEST FINES REINJECTION

D. RUN-OF-MINE BITUMINOUS COAL

1. INVESTIGATE HIGH CARBON CONVERSION

2. INVESTIGATE MAXIMUM PILOT PLANT CAPACITY

TASK 8. DEMONSTRATION PLANT DESIGN SUPPORT

A. EXPERIMENTAL PROCESS STUDIES AND COLD FLOW MODELING OF REACTOR

B. ASSIST ERDA-MFPM IN PLANT DESIGN TASKS

TASK 9. OPERATIONS SUPPORT STUDIES

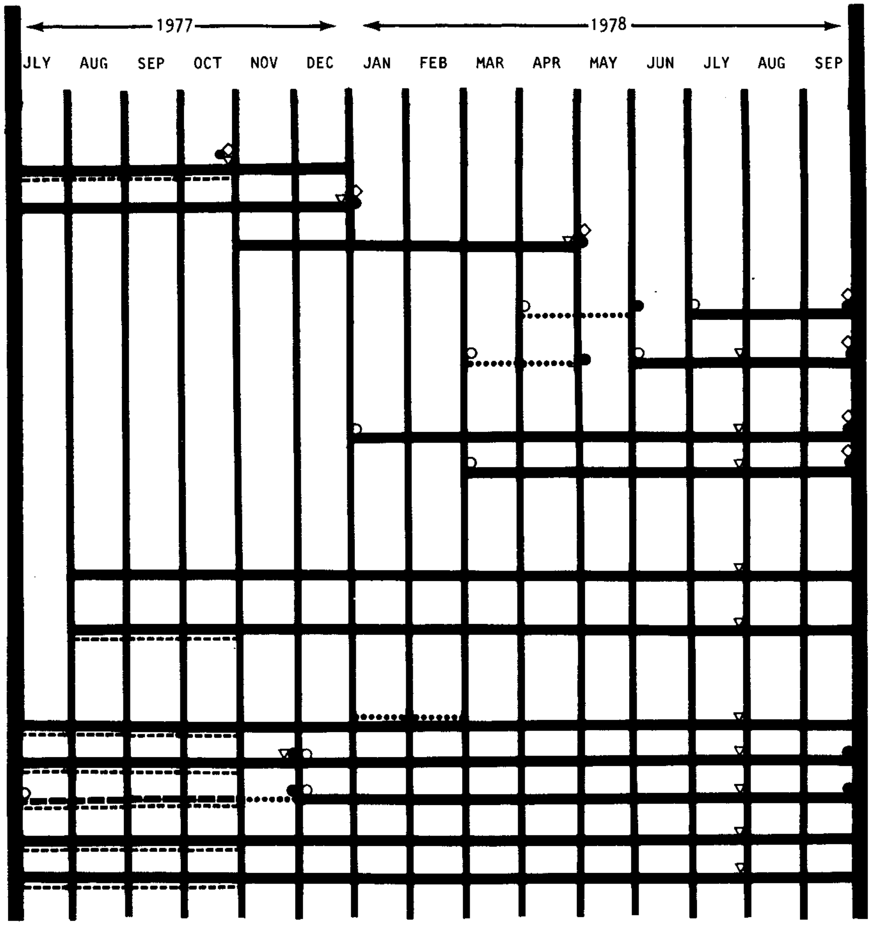
A. PLANT EFFLUENT PROCESSING*

B. TEST METHANATION SYSTEMS AND CATALYSTS*

C. INVESTIGATE HOT OIL QUENCH SYSTEM

D. MATERIALS TESTING*

E. ENGINEERING SERVICES*



*THIS TASK INITIATED UNDER PRIOR CONTRACT

LEGEND

○ INITIATE
● COMPLETE
◇ MILESTONE
▽ DECISION POINT
DESIGN EFFORT
TEST OPERATION
INSTALL EQUIPMENT
ACTUAL PROGRESS

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Figure 17. WORK SCHEDULE