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FOSSIL ENERGY

PIPELINE GAS FROM COAL—HYDROGENATION (IGT HYDROGASIFICATION PROCESS)

Project 9000 Quarterly Report No. 8 for April 1—June 30, 1978

March 1979
Date Published

MASTER

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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Chicago, Illinois 60616

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UNITED STATES DEPARTMENT OF ENERGY

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SUMMARY

The objective of this project is to perform the pilot plant operations and related support studies required to obtain data for a commercial/demonstration plant design based on the HYGAS® Process. This quarterly report covers the work done toward achieving these project goals from April 1 through June 30, 1978.

Illinois bituminous coal was used to acquire data for optimizing the HYGAS Process. Three major modifications were made in the pilot plant during March to encourage optimum operating conditions and clinker-free operation. Their effectiveness was assessed by Test 71, which was conducted in April and successfully showed that all three modifications effectively left the steam-oxygen zone entirely clinker-free after the test. A post-run inspection and the necessary turnaround activities were completed after Test 71.

Test 72 was conducted in May as a continued exploration of the conditions for optimizing reactor operation by improving carbon conversion to the 85% to 90% level, increasing the temperature in the steam-oxygen gasifier to 1800°F, and continuing to reduce the steam-to-char ratios to approximately those used in the commercially designed (by Procon, Inc.) HYGAS Process unit. During Test 72, a total of 392 tons of pretreated char was fed to the reactor. Char conversions, determined by quick ash analyses, ranged from 60% to 90% at an operating temperature of 1800°F. The steam feed rate was set at 5000 lb/hr, and the char feed rate was 2 tons/hr. The test was terminated at the end of the month when the vibrating feeder from the pretreated-char storage hopper failed and forced the suspension of char feed to the reactor. Several steady-state periods were selected for detailed study. A post-run inspection of the plant was made to determine the areas where modifications might be made to further increase the operating efficiency of the pilot plant. As a result, the bucket elevator speed was increased to improve its capacity, and the coal-mill crushing speed was lessened to reduce fines generation.

Test 73 was initiated in June to duplicate the conditions of the highly successful Test 71 and to achieve a better steam-to-char ratio and a significantly lower superficial gas velocity in the steam-oxygen gasifier. These test conditions were selected as a part of a program to determine the lower limits of gas velocity in the steam-oxygen gasifier that would give sinter-free operation.

A great deal of data supporting the demonstration plant program was supplied to Procon, Inc., and to the United States Department of Energy (DOE) during this quarter. These data are presented in this report.

Seven different solids-feeding devices for the low-temperature reactor section of the pilot plant have been tested, three of these during this quarter. Detailed results and analyses of these tests are presented in this report, as well as a summary of the results for all seven devices and recommendations based on these results.

INTRODUCTION

This eighth quarterly report covers work conducted between April 1 and June 30, 1978, under DOE Contract No. EF-77-C-01-2434.

Tasks 1 through 6, which demonstrated the feasibility of using lignite, bituminous, and subbituminous coal feedstocks in the HYGAS pilot plant, were completed between July 1, 1976, and June 30, 1977, and are reported in Project 9000 Interim Report No. 1 (DOE Report No. FE-2434-23).

Since July 1, 1977, work has been done on Tasks 7 through 9, which are detailed in Project 9000 Quarterly Report No. 5 (DOE Report No. FE-2434-20), Quarterly Report No. 6 (DOE Report No. FE-2434-25), Quarterly Report No. 7 (DOE Report No. FE-2434-29), and in this report.

PROGRESS REPORT

Task 7. Pilot Plant Experimental Operation

Plant turnaround activities and modifications following Test 70 were completed during the first week of April. The three modifications included installing a new six-nozzle steam-oxygen sparger, replacing and relocating valve 339, and installing double-screening equipment upstream of the pretreater section. The double-screening equipment was tested, and the plant readied for Test 71.

Test 71

The reactor was pressure-tested at 200 psig and was successfully lighted for Test 71 on April 6 at 1530 hours. During a routine check of the reactor on April 10, water was found in the nitrogen jacket; consequently, the reactor had to be cooled and its pressure brought down. The water leak was traced to a crack on the direct, cooling-water-spray line to the slurry dryer section. This leak was fixed, and the reactor was relighted on April 12 at 0733 hours. The reactor temperature and pressure were raised to the required levels, and slurry feed was initiated at 1745 hours on April 13. Operations became self-sustained on April 17 at 1200 hours when oxygen was removed from the start-up burner.

The reactor operated smoothly until April 17 at 1830 hours when a malfunction of the high-pressure feedwater pump forced a temporary shutdown of the high-pressure boiler and interrupted reactor operations for 2 hours. When operations were resumed, a high-pressure drop across the slurry dryer grid made smooth solids feeding to the low-temperature reactor difficult due to the intermittent loss of the seal on the fresh-char feed line. After several unsuccessful attempts to stabilize solids circulation by reducing the pressure drop across the slurry dryer grid using nitrogen blasts, the operating pressure of the reactor was reduced to 500 psig at 1800 hours on April 19. This reduced the high-pressure drop across the slurry dryer grid because of the reduced mass flow rates in the reactor. It also minimized the leaks on manway 0 and on the hot-gas sampler, nozzle 25B, which had been first observed on April 15.

Smooth pretreater operation was hampered by electrical control problems and plugging in the solids conveyors in the double-screening equipment, which was bypassed on April 18 at 1600 hours. The reactor operated very smoothly after the pressure drop across the slurry dryer was reduced by lowering the reactor pressure. Char feed to the reactor was maintained at 2 tons/hr, oxygen feed ranged from 640 to 750 lb/hr, and the total steam feed rate was 5100 lb/hr. The steam-oxygen gasifier temperature was held at a maximum of 1750°F. An 80% char conversion was achieved. The reactor operated smoothly under these conditions except for two interruptions in char feed: the first when the high-pressure slurry pump failed on April 23 at 1030 hours, and the second when the high-pressure boiler shut down momentarily on April 28 at 0230 hours. In both instances, the reactor recovered successfully from these interruptions. Test 71 was voluntarily terminated on April 28 at 1200 hours. Several steady-state periods were selected for material and energy balances and engineering calculations. Preliminary results from Test 71 are given in Table 1.

Test 71 was highly significant for several reasons:

a. This was the first test with the modified steam-oxygen sparger. The new six-nozzle distributor improved mixing and fluidization in the steam-oxygen zone, as was evidenced by an improved temperature profile in the bed and extremely smooth solids flow.

b. The relocation and design change of solids flow valve 339 contributed to improved fluidization and permitted tight shutoff and better solids flow control.

c. The steam flow for achieving the necessary superficial velocity in the steam-oxygen zone was reduced, largely because of the reduced pressure. The flows were much closer to the steam/carbon ratios used in the commercial/demonstration plant design.

d. The entirely clinker-free condition of the steam-oxygen gasification zone strongly indicates that the design goals of steam, carbon, and oxygen flows can be achieved at high carbon conversions and that reliable clinker-free operating limits can be established.

Integrated operation of all pilot plant sections was attained in Test 71. The coal mill section also operated satisfactorily. Pretreater operation began with coal feeding at 1650 hours on April 13 and improved after the double-screening section was bypassed. The pretreater reliably supplied char to the reactor. It operated well with new grid nozzles providing good gas distribution. The slurry preparation section operated satisfactorily, although slurry feed was interrupted for 1 hour when the high-pressure slurry pump failed.

Table 1. PRELIMINARY RESULTS FROM TEST 71
(Note: These Results Are Preliminary and Must Be Confirmed by Additional Studies.)

Date (Hour)	Char Feed[a] Rate, lb/hr	Ash[b] in Spent Char	Ash[b] in Pretreater Char	Char Gasified	lb O_2/ lb Char Feed[a]	lb O_2/ lb Net Char Feed[d]	lb Steam/ lb Char Feed[a]	lb Steam/ lb Net Char Feed[d]	Highest Average SOG Bed Temp, °F
			%						
4/17/78 (1800)	4009	24.1	12.1[c]	57	0.20		2.22		1580
4/21/78 (0000)	4218	41.2	13.0[c]	79	0.15	0.19	1.19	1.47	1690
4/21/78 (0400)	4041	37.3	13.0[c]	75	0.16	0.20	1.24	1.54	1680
4/21/78 (0600)	4200	43.5	13.0	81	0.15	0.19	1.19	1.47	1680
4/21/78 (2200)	3922	49.4	13.0[c]	85	0.17	0.22	1.31	1.65	1702
4/22/78 (0600)	4182	46.4	13.4	82	0.16	0.20	1.23	1.53	1703
4/22/78 (1205)	4296	44.4	13.4[c]	81	0.16	0.19	1.21	1.49	1670
4/23/78 (0400)	4207	49.3	13.4[c]	84	0.16	0.20	1.23	1.52	1693
4/23/78 (0600)	4137	32.1	12.4	70	0.17	0.21	1.26	1.56	1692
4/25/78 (0400)	4124	29.4	11.7[c]	68	0.17	0.21	1.22	1.51	1690
4/25/78 (0600)	4164	30.5	12.7	67	0.16	0.20	1.21	1.49	1700
4/25/78 (1700)	4005	38.9	12.7[c]	77	0.18	0.23	1.29	1.61	1695
4/26/78 (0200)	4119	47.9	12.7[c]	84	0.18	C.23	1.23	1.52	1721
4/26/78 (0600)	4000	34.2	13.3	70	0.18	0.23	1.21	1.52	1742
4/26/78 (1330)	3926	34.0	13.3[c]	70	0.19	0.24	1.30	1.63	1723
4/27/78 (1400)	4066	40.4	10.6	83	0.18	0.23	1.23	1.53	1716

[a] Char feed on a wet basis.

[b] Percent ash in pretreater char and reactor spent char determined by quick ash analysis.

[c] Assumed.

[d] Estimated 800 pounds of overhead fines per hour.

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Reactor operation was generally smooth during Test 71. Two modifications in the steam-oxygen gasifier (the addition of a new six-nozzle steam-oxygen sparger and a new valve 339 relocated 9 feet from the gas distributor) substantially improved its operation and fluid-bed mixing characteristics. Flat temperature profiles were observed in the steam-oxygen bed during Test 71. Temperature profiles recorded in the steam-oxygen gasifier during Tests 66, 70, and 71 are compared in Figures 1 and 2. Two operating pressure levels attempted during Test 71 both showed flat profiles, indicating good solids mixing. In addition, the new valve 339 offered a positive operating seal. A post-run inspection of the reactor confirmed the existence of a completely clinker-free steam-oxygen zone.

Figure 3 presents an overview of Test 71 showing input and output streams. Results of Test 71 are tabulated in Table 2. During Test 71 char conversions in the reactor ranged from 68% to 90%.

The quench and purification sections operated very well. During Test 71, a new antifoam agent, Dow-Corning HB-10, was effectively tried in the diglycolamine solution. Removal of carbon dioxide and sulfur compounds during this test was excellent.

Chem Systems' liquid-phase methanation pilot unit operated for approximately 54 hours, 24 of which were with purified product gas from the HYGAS reactor. The utility section operated well during Test 71, except for two interruptions in the high-pressure boiler operation when the high-pressure feedwater pump pressure was lost temporarily.

The coal mill was inspected after Test 71 and was found to be in good condition. A normal amount of coal had accumulated in the wet-scrubber. The 60-ton raw coal storage hopper was inspected; some packed coal was found along the wall and was cleaned up. The pretreater reactor was clinker-free and in excellent condition. A pinhole leak found in one of its internal cooling coils was repaired. The char cooler was also in good condition after Test 71. The venturi scrubber had a slightly heavier tar accumulation than usual, and the quench tower was full of tarry material. The level indicator and sight-glass on the quench tower were plugged. The char-slurry section was in good condition. The check valves for the high-pressure char-slurry pumps were inspected and were in satisfactory order. The slurry mix tank was emptied in preparation for the next test.

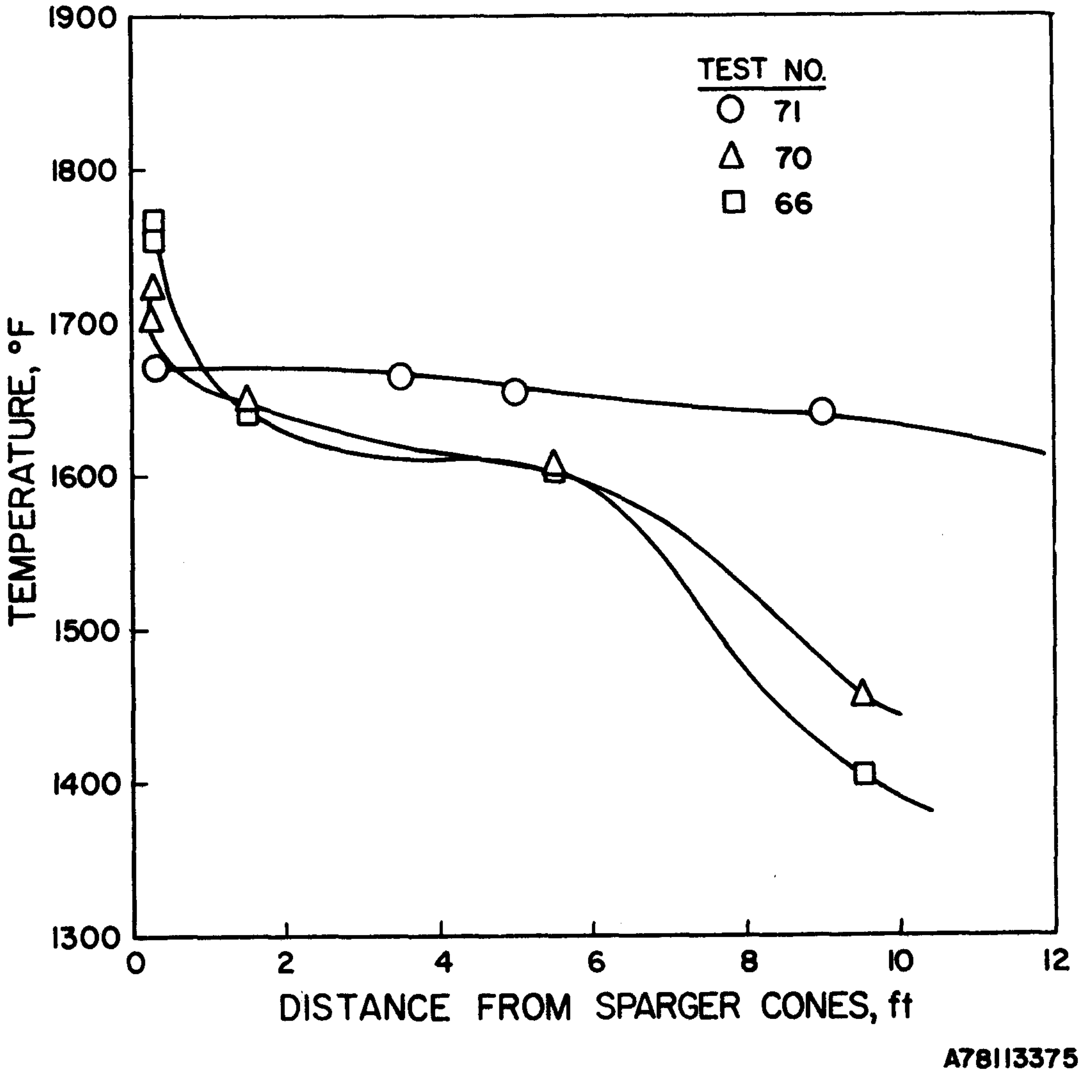

Figure 1. STEAM-OXYGEN GASIFIER TEMPERATURE vs. VERTICAL LOCATION (500 psig)

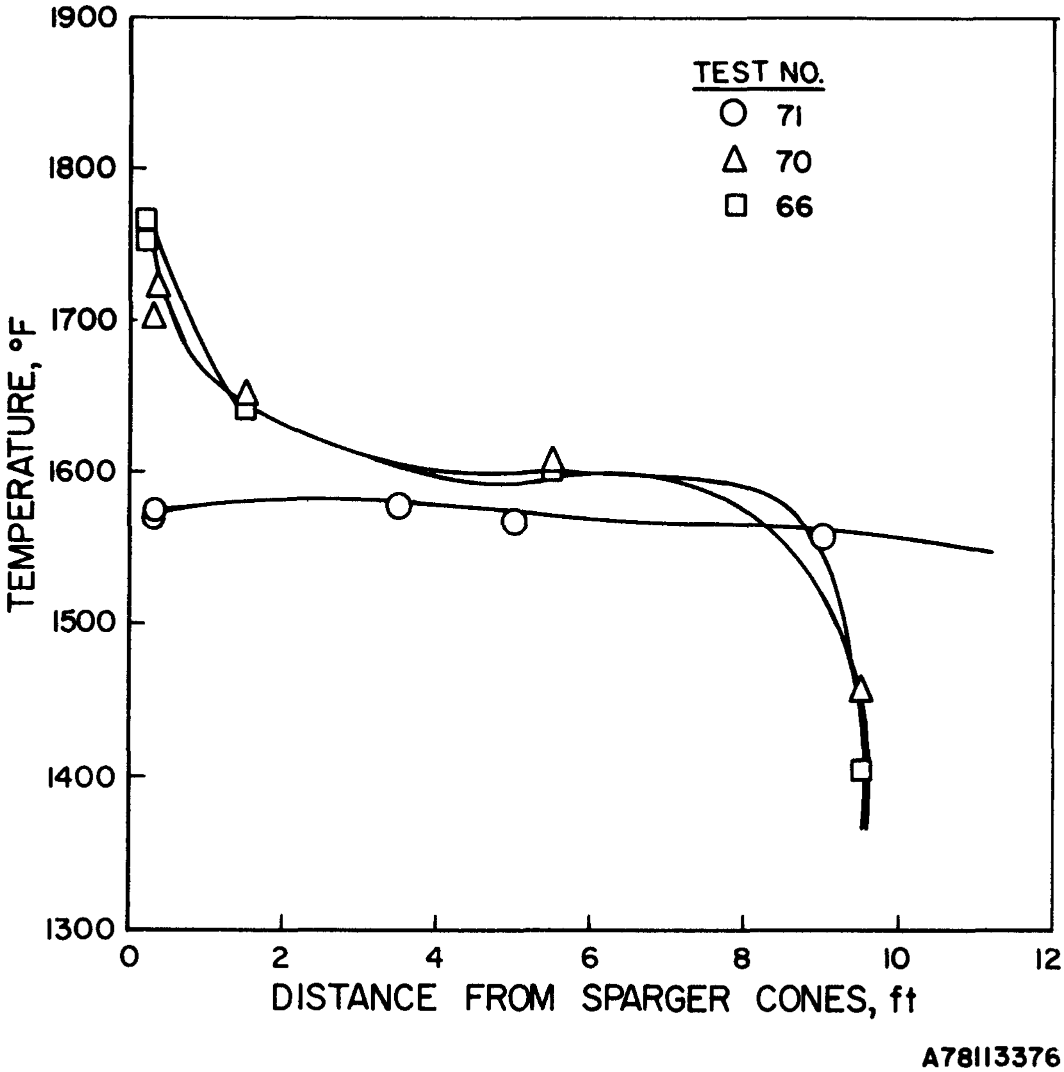

Figure 2. STEAM-OXYGEN GASIFIER TEMPERATURE vs. VERTICAL LOCATION (900 psig)

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Table 2. TEST 71 — PRELIMINARY RESULTS

(Note: These Results Are Preliminary and Must Be Confirmed by Additional Studies.)

Date (Hour)	4/17/78 (1400)-4/17/78 (2100)	4/20/78 (2300)-4/21/78 (1400)	4/20/78 (2300)-4/23/78 (1000)	4/21/78 (1700)-4/23/78 (1000)	4/25/78 (2100)-4/26/78 (2100)	4/26/78 (0100)-4/26/78 (1600)	4/27/78 (1300)-4/28/78 (0200)	4/28/78 (0600)-4/28/78 (1200)
Char to Reactor, dry, lb/hr	3905	3994	4024	4046	4007	4006	3895	3924
Reactor Overhead, lb/hr	591	855	1007	1076	595	676	504	1054
Net Char to Reactor,[a] lb/hr	3314	3139	3017	2970	3412	3330	3391	2870
Oxygen to Sparger, lb/hr	784	639	670	682	744	743	749	745
Steam to Sparger, lb/hr	7405	4143	4140	4141	4068	4046	4024	4026
Steam to Stripping Ring, lb/hr	1050	667	660	656	631	632	619	610
Operating Pressure, psig	847	509	509	510	507	505	519	513
Maximum Steam-Oxygen Gasification Temperature, °F	1601	1747	1706	1692	1717	1714	1727	1735
Steam/Char (Net), lb/lb	2.55	1.53	1.59	1.62	1.37	1.40	1.37	1.62
Oxygen/Char (Net), lb/lb	0.24	0.20	0.22	0.23	0.22	0.22	0.22	0.26
SOG[b] Superficial Velocity, ft/s	1.12	1.15	1.13	1.12	1.12	1.12	1.09	1.11
Char Conversion, %	68	76	78	79	78	77	81	90
Carbon Conversion, %	62	72	73	75	74	73	77	89
Total Product Gas								
lb/hr	4446	3706	4079	4211	4307	4375	4096	4137
mol/hr	193	175	190	196	197	200	188	186
Product Gas Components, mol %								
H_2	33.62	37.14	36.30	35.97	34.25	34.15	34.25	34.60
H_2S	1.04	0.80	0.77	0.67	0.11	0.18	0.08	0.22
C_2H_6	0.39	0.30	0.41	0.47	0.46	0.47	0.48	0.47
CO_2	32.80	27.96	28.67	28.89	29.23	29.32	29.39	31.70
N_2	9.58	10.16	9.65	9.41	9.32	9.29	9.10	8.93
CH_4	14.62	15.00	15.10	15.24	16.62	16.59	17.32	15.61
CO	7.95	8.65	9.10	9.34	10.01	10.00	9.38	8.47
Equivalent Methane Yield/Carbon Gasified, SCF/lb	14.45	16.32	15.98	15.90	15.85	15.81	16.09	15.15
Direct Methane Yield/Carbon Gasified, SCF/lb	8.22	9.07	8.88	8.85	9.24	9.22	9.59	8.69
Superficial Velocity, V_s, in 2-ft diameter distributor,[c]	0.09	0.09	0.09	0.09	0.09	0.09	0.09	0.09
Jet Velocity,[d] ft/s	54.3	48.5	46.0	45.4	47.9	48.1	46.6	47.2
Jet Penetration,[d] in.	32.8	23.6	23.0	22.9	23.4	23.5	23.1	23.2
SOG Nozzle Diameter, in.	1.049	1.049	1.049	1.049	1.049	1.049	1.049	1.049
SOG Nozzle Cones	6	6	6	6	6	6	6	6
HTR Bed Height, ft	10	10	11	11	17	17	18	18
HTR Bed Density, lb/ft^3	24	19	18	17	16	15	15	15
SOG Bed Height, ft	25	19	18	18	17	17	16	16
SOG Bed Density, lb/ft^3	11	10	10	10	12	11	12	13

[a] Net char feed rate = char feed — reactor overhead.

[b] Calculated at maximum steam-oxygen gasifier (SOG) gasification temperature.

[c] Calculated using the temperature of superheated steam.

[d] Calculated using the mixed temperatures of superheated steam and oxygen.

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The HYGAS reactor was inspected. The slurry dryer bed was clean except for some minor plugs of caked coal in the slurry dryer grid, which were evident during the test. The rest of the reactor, all solids transfer lines, the first- and second-stage reactors, and the steam-oxygen gasifier were all in excellent condition. The six-nozzle steam-oxygen sparger and valve 339 were both in good condition. A materials corrosion test coupon assembly had fallen from valve 339 and was found lying on top of the sparger. Two of the thermocouples in the steam-oxygen gasifier were broken after the test. They may have broken while the steam-oxygen gasifier bed was being cleaned. The cyclone and the cyclone dipleg were in good condition. Some solids had accumulated in the cyclone slurry pot plugging the vent line from it. When the reactor internals were pressure-tested, the line 322 expansion joint leaked severely, and was replaced for Test 72. Downstream of the reactor, the quench section was inspected. The quench towers showed normal accumulations, but the quench separator had more solids than usual. The purification section ran well during Test 71 and required no work. Residual oil was drained from the liquid-phase methanation pilot unit, and it was cleaned to ready it for any further internal reactor modifications planned by Chem Systems, Inc. When the light-oil recovery unit was inspected, a large amount of solids was found. The stripper tower was disassembled during the turnaround period. The distributor trays, which had been removed for inspection, were in good condition.

Test 72

The double-screening equipment was tested in preparation for Test 72. The slurry mix tank and recycle oil storage tank were emptied and cleaned. In the methanation section, the ZINGARD catalyst (ZnO for trace sulfur compound removal) was removed, and a new batch was charged prior to beginning the test.

The reactor was closed up and pressure tested on May 14. Light-off for Test 72 occurred at 2000 hours on May 14. Some early problems were encountered in the coal-handling section: plugging of the 60-ton raw coal storage hopper and of the instrumentation lines, screw feeders, lockhoppers and solids waste bins. These delayed char feeding to the reactor until May 19 at 1230 hours. Reactor operation became self-sustaining at 2000 hours on May 20, and continued as such for 9-1/2 days. Continuous char slurry feed to the reactor was logged for over 7 days. Test 72 was terminated on May 30 at 1530 hours. A total of 392 tons of pretreated char was fed to the reactor. Char conversions, determined by quick ash analyses, ranged from 60% to 90%. The char feed rate to the reactor was 2 tons/hr; steam, 5000 lb/hr; and oxygen, a maximum of 750 lb/hr. These conditions in the steam-oxygen gasifier yielded a maximum operating temperature of 1800°F. During Test 72, valve 339 froze in a 30% to 40% open position while transferring solids from the high-temperature reactor to the steam-oxygen gasifier bed. However, satisfactory solids feed was maintained throughout the test.

The slurry preparation section operated satisfactorily during Test 72, although there was a 4-hour interruption on May 20 when suction pressure was lost at the high-pressure slurry pump and two short interruptions occurred due to low seal flush flows. The quench section operated well, but the purification section was not operated because of vibrational problems in the high-pressure amine pump. Consequently, the IGT fixed-bed catalyst methanation unit was not put on-stream. The effluent clean-up section was operated during Test 72, and its operation was significantly improved by double-screening the coal to reduce solids carry-over from the reactor.

Inspection after Test 72 showed that the coal preparation section was in good mechanical condition. The double-screening system, which operated for the duration of Test 72, was in satisfactory condition. As a result of this test, however, some changes in this system were indicated, i.e., an increase in the bucket elevator speed to improve its capacity and a reduction in the coal mill crushing speed to reduce fines generation.

The pretreater reactor was inspected, and a few small clinkers were found on the grid in the southwest corner near the char drain line. These clinkers are believed to have been formed during the 6 hours the pretreater bed was slumped on May 18, when the air compressor supplying fluidizing air to the pretreater was shut down because of a plugged lubrication line. Pressure tests revealed leaks in the northwest cooling bundle inside the pretreater, which were then fixed. The rest of the pretreater area (the char cooler, the lock-hoppers, the venturi scrubber, the quench tower, and all the gas and solids transfer lines) was in satisfactory condition.

The slurry preparation section was inspected, and as suspected, the oil seal flush line to the high- and low-pressure slurry pumps was partially plugged with solids. The line was flushed and cleared.

The HYGAS reactor was opened for inspection. The slurry dryer area was clear. All the internal char transfer lines were clear except line 322, which was plugged. This line, which transfers solids from the first-stage reactor to the second-stage reactor, was not plugged during the test, although a plug did form when it was shut down. The second-stage reactor was clean except for small pieces of refractory, which had spalled off the reactor refractory wall, lying on the second-stage grid. Two clinkers were found in the steam-oxygen gasifier, one, 4 x 9 x 3 inches, lying on top of the sparger, and another, 6 x 3 inches round, encasing thermocouple 3000-16. Valve 339, which had been frozen in a 40% open position during the test, had a galled actuator rod, and char was found packed in the operator assembly. The nitrogen purge to this assembly was relocated to give better protection to this area, and the actuator rod and bushing were polished. Thermocouple 3000-17 in the steam-oxygen gasifier was bent upward about 9 inches. The rest of the steam-oxygen gasifier was in good condition, and the reactor high-pressure cyclone was clean. The usual 1-1/2 feet of solids had accumulated in the cyclone slurry pot.

The quench system, downstream from the reactor, was exceptionally clean because the feed material to the reactor had been double-screened. Similarly, the light-oil stripper section was in good condition. The plant was completely checked and was found to be in good condition prior to beginning Test 73.

Several steady-state periods from Test 72 were selected for detailed study and are presented in Table 3. Figure 4 is an overview of Test 72.

A debriefing session on Test 72 was held at 1330 hours on June 9. Representatives from DOE, Scientific Design, Procon, and IGT attended the meeting.

Test 73

Reactor light-off for Test 73 was achieved on June 12, and char feed to the reactor was begun at 0251 hours on June 14. Test 73 was terminated at 1030 hours, June 16, after temperature excursions in the steam-oxygen gasifier were observed. During Test 73, the char feed rate was maintained at about 3000 lb/hr, the oxygen feed rate was 600 lb/hr, and the steam feed rate was 3500 lb/hr. The temperature profile in the steam-oxygen gasifier was good until June 15 at 0700 hours, when two of the thermocouples in the bed began fluctuating. At about 0500 hours, June 16, temperature excursions to 2300°F were

Table 3. PRELIMINARY RESULTS FOR TEST 72

Date (Hour)	5/23/78 (1300)-5/23/78 (2000)	5/24/78 (0700)-5/25/78 (1300)	5/25/78 (2300)-5/26/78 (2300	5/26/78 (1000)-5/26/78 (2300)	5/27/78 (0800)-5/27/78 (2400)	5/28/78 (2300)-5/29/78 (0600)
Char to Reactor, dry, lb/hr	4090	4112	3989	3968	4084	2962
Reactor Overhead, lb/hr	299	349	370	459	454	297
Net Char to Reactor,[a] lb/hr	3791	3763	3619	3509	3630	2665
Oxygen to Sparger, lb/hr	708	696	667	704	689	517
Steam to Sparger, lb/hr	4330	4224	4125	4158	3808	3871
Steam to Stripping Ring, lb/hr	362	353	354	367	369	371
Operating Pressure, psig	571	539	534	534	553	527
Maximum Steam-Oxygen Gasification Temperature, °F	1764	1790	1787	1794	1862	1677
Steam/Char (Net), lb/lb	1.24	1.22	1.24	1.29	1.15	1.59
Oxygen/Char (Net), lb/lb	0.19	0.18	0.18	0.20	0.19	0.19
SOG[b] Superficial Velocity, ft/s	1.02	1.06	1.05	1.06	1.02	0.94
Char Conversion, %	70	76	66	66	69	79
Carbon Conversion, %	63	72	59	59	62	75
Total Product Gas						
lb/hr	3945	4224	4037	4145	4118	3018
mol/hr	173	185	172	175	174	128
Product Gas Components, mol %						
H_2	30.15	30.90	28.64	28.44	27.62	29.95
H_2S	1.55	0.97	1.42	1.40	1.49	0.91
C_2H_6	0.40	0.43	0.45	0.45	0.48	0.31
CO_2	28.89	29.38	30.91	31.30	30.82	31.96
N_2	7.92	8.26	8.62	8.43	8.00	12.76
CH_4	17.98	16.40	17.88	17.70	18.74	15.03
CO	13.10	13.66	12.08	12.27	12.85	9.08
Equivalent Methane Yield/Carbon Gasified, SCF/lb	15.33	14.82	14.75	14.56	14.80	14.11
Direct Methane Yield/Carbon Gasified, SCF/lb	9.34	8.59	9.14	8.99	9.34	8.37
Superficial Velocity, V_s, in 2-ft Diameter Distributor,[c] ft/s	0.05	0.05	0.05	0.05	0.05	0.05
Jet Velocity,[d] ft/s	45.6	47.0	46.3	46.7	43.0	43.6
Jet Penetration,[d] in.	23.5	23.6	23.1	23.3	21.6	21.4
SOG Nozzle Diameter, in.	1.049	1.049	1.049	1.049	1.049	1.049
SOG Nozzle Cones	6	6	6	6	6	6
HTR Bed Height, ft	18	14	26	27	26	14
HTR Bed Density, lb/ft^3	14	15	11	11	11	20
SOG Bed Height, ft	20	23	24	25	26	21
SOG Bed Density, lb/ft^3	13	10	11	11	10	13

[a] Net char feed rate = char feed − reactor overhead.

[b] Calculated at maximum steam-oxygen gasifier (SOG) gasification temperature.

[c] Calculated using the temperature of superheated steam.

[d] Calculated using the mixed temperatures of superheated steam and oxygen.

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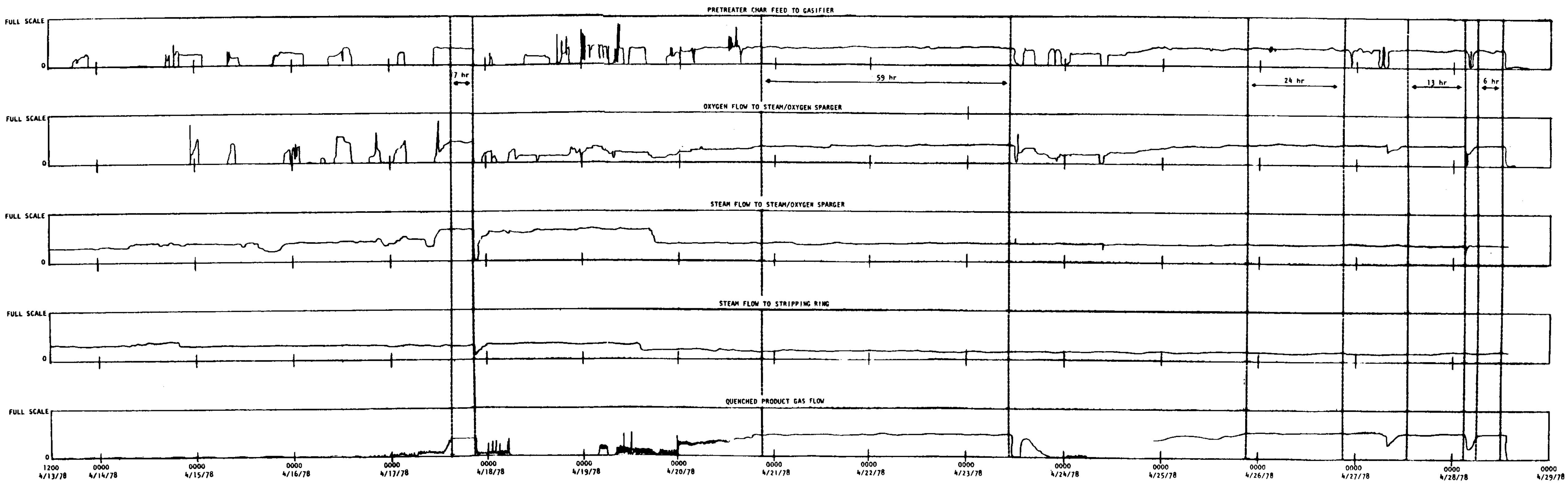

Figure 3. OVERVIEW OF TEST 71

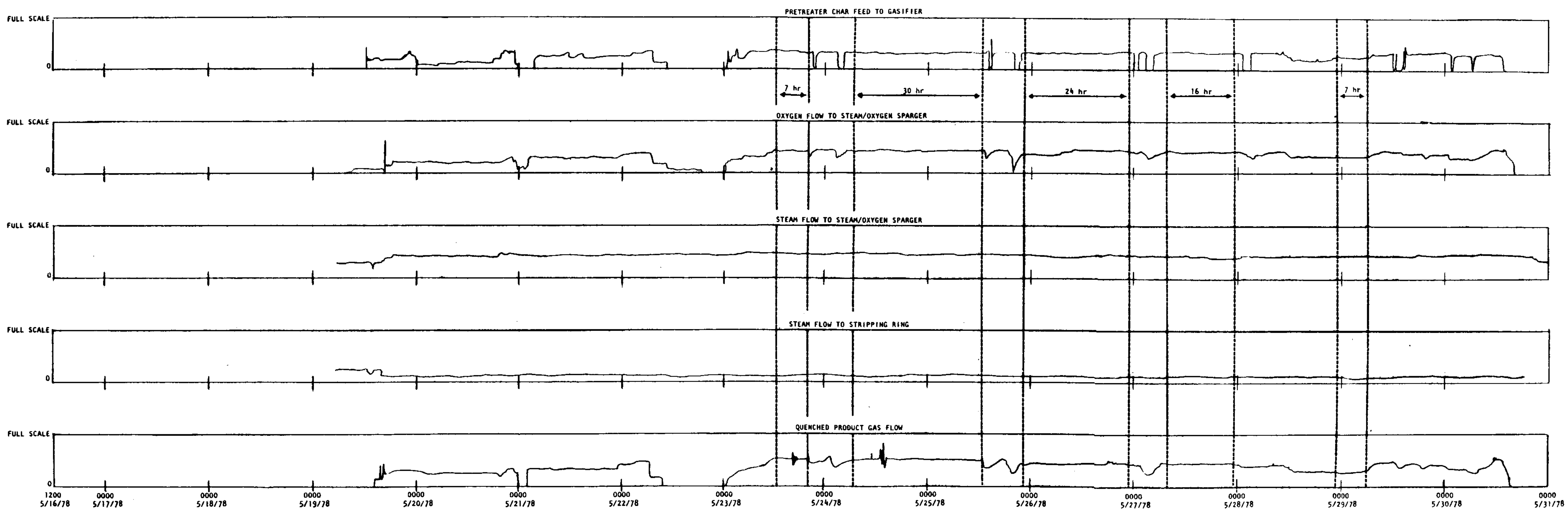

Figure 4. OVERVIEW OF TEST 72

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observed in the lower portion of the steam-oxygen bed. Operation of the reactor continued until 1030 hours when Test 73 was terminated. The objective of this test was to duplicate the conditions of Test 71, and yet achieve a better steam-to-char ratio and a much lower superficial gas velocity in the steam-oxygen gasifier. The superficial velocity used in Tests 66 and 71, which were successful, was approximately 1.2 ft/s. The planned superficial velocity for Test 73 was 0.9 ft/s. These test conditions were desired to determine the lower limits of the gas velocity in the steam-oxygen gasifier that would give sinter-free operation. The six-nozzle steam-oxygen sparger was modified to give the same jet penetration as in Test 71. The six-nozzle steam-oxygen sparger used in Tests 71 and 72 had 1.049-inch-diameter nozzles (1-inch, Schedule 40 pipe size). The nozzle diameter was modified for Test 73, and the diameter of each of the six nozzles was reduced to 0.742 inch (3/4 inch, Schedule 80 pipe size) to yield higher nozzle velocities and, therefore, higher jet penetrations at the reduced fluidizing gas (steam plus oxygen) flows planned for the test. Seventy-four tons of pretreated char were processed through the reactor. Preliminary data for one steady-state period of Test 73 are presented in Table 4 and Figure 5 shows an overview of this test.

The double-screening equipment was used during Test 73, and modifications made on the coal mill and on the bucket elevator speed both improved operation in this section. Pretreater operation began at 1900 hours on June 13. Coal feed to the pretreater was interrupted three times: 1) when the purge lines to the char cooler level control valve plugged, 2) when the bottom discharge of the 60-ton raw coal storage hopper plugged, and 3) when the fines screw conveyor plugged. Seventy-five tons of coal were processed. Char slurry feed to the reactor was interrupted twice, once when a leak developed on the low-pressure slurry return line to the slurry tank and a second time when the motor on the vibrating feeders from the 15-ton pretreater char storage hopper shorted out. The quench section functioned well during Test 73, but the purification section was not put into service due to vibrational problems in the high-pressure amine circulation pump. The light-oil recovery unit was operated, although the Edens separator was bypassed because of a broken chain in the unit.

Following Test 73, the plant was inspected. The coal mill was in good condition; the reduction in mill speed had resulted in a smaller amount of fines being generated in the mill. Some clinkers had formed around the edge of the grid in the pretreater reactor, but the rest of the pretreater section

Table 4. PRELIMINARY RESULTS FOR TEST 73

Date (Hour)	6/15/78 (0800) - 6/16/78 (0000)
Char to Reactor, dry, lb/hr	3088
Reactor Overhead, lb/hr	407
Net Char to Reactor,[a] lb/hr	2631
Oxygen to Sparger, lb/hr	594
Steam to Sparger, lb/hr	2782
Steam to Stripping Ring, lb/hr	715
Operating Pressure, psig	520
Maximum Steam-Oxygen Gasification Temperature, °F	1703
Steam/Char (Net), lb/lb	1.30
Oxygen/Char (Net), lb/lb	0.22
SOG[b] Superficial Velocity, ft/s	0.82
Char Conversion, %	70
Carbon Conversion, %	63
Total Product Gas	
lb/hr	3365
mol/hr	140
Product Gas Components, mol %	
H_2	27.11
H_2S	1.70
C_2H_6	0.36
CO_2	32.16
N_2	7.76
CH_4	18.55
CO	12.36
Equivalent Methane Yield/Carbon Gasified, SCF/lb	14.38
Direct Methane Yield/Carbon Gasified, SCF/lb	9.18
Superficial Velocity, V_s, in 2-ft-diameter distributor,[c] ft/s	0.10
Jet Velocity,[d] ft/s	62.7
Jet Penetration,[d] in.	26.4
SOG Nozzle Diameter, in.	0.742
SOG Nozzle Cones	6
HTR Bed Height, ft	21
HTR Bed Density, lb/ft^3	14
SOG Bed Height, ft	18
SOG Bed Density, lb/ft^3	12

[a]Net char feed rate = char feed − reactor overhead.

[b]Calculated at maximum steam-oxygen gasifier (SOG) gasification temperature.

[c]Calculated using the temperature of superheated steam.

[d]Calculated using the mixed temperatures of superheated steam and oxygen.

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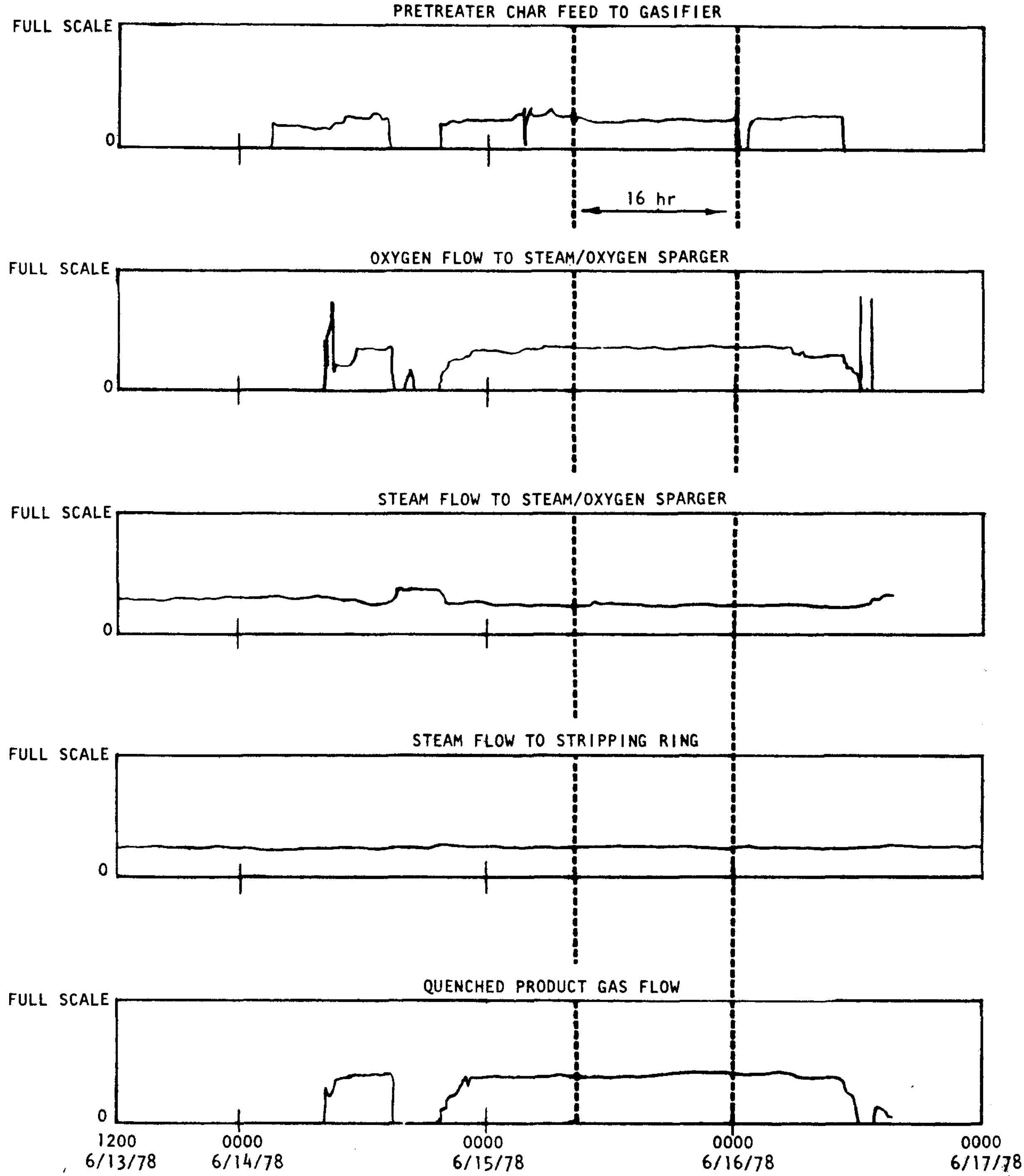


Figure 5. OVERVIEW OF TEST 73

(the char cooler, the lockhoppers, the venturi scrubber, the quench tower, and all the gas and solids transfer lines) was in satisfactory condition. The slurry preparation section received some routine maintenance and no problems were discovered. The reactor was inspected, and the slurry dryer area and all the internal char transfer lines were clean. The second-stage reactor had some pieces of refractory lying on its grid. The steam-oxygen gasifier had a clinker formation covering three of the six sparger nozzles, blocking approximately one-third of the gasifier cross-section. This formation encased five thermocouples and one pressure differential transmitter tap. This clinker was different from previous ones because it was actually in contact with the nozzles of the steam-oxygen sparger, rather than in formation above them. The cyclone, the cyclone dipleg, the cyclone slurry pot, and spent char slurry pots were cleaned.

The quench system had an unusually large amount of solids accumulation for such a short test, but the light-oil recovery system was clean.

Chem Systems' liquid-phase methanation unit was operated with high carbon monoxide-to-hydrogen ratio gas from the HYGAS hydrogen plant. A fire occurred on June 22 in the liquid-phase methanation unit and caused significant damage to the HYGAS pilot plant.

All plant sections were prepared for a complete shutdown for the annual 2-week vacation period, which occurred during the first 2 weeks of July.

Task 8. Demonstration Plant Support

One of the major activities under this task has been the transfer of data to Procon, Inc., for its design of a HYGAS demonstration plant. During the reporting period, the following data were supplied:

a. Spent char slurry water compositions from the HYGAS pilot plant for Tests 60, 61, 63, and 64 that were requested by Procon (Table 5).

   Drawing No. IGT-135-E showing the bundle assembly for the pilot plant pretreater reactor (2-06-01) as it was installed in the pretreater for heat recovery.

Table 5. COMPOSITION OF SPENT-CHAR SLURRY WATER
(Basis: Pretreated Char to Gasifier)

Test No.	60		61		63		64	
Number of Samples	4		5		6		8	
Average and Standard Deviation*	$\overline{X}$	S	$\overline{X}$	S	$\overline{X}$	S	$\overline{X}$	S
	lb/ton char (MAF)							
TDS	6.9	6.6	6.4	2.0	22.7	41.1	5.4	5.0
ϕOH	0.007	0.008	0.034	0.059	0.84	1.8	0.0007	0.0012
CN^-	<0.0001		<0.0001		<0.0001		<0.0001	
TOC	2.2	2.7	2.9	1.7	3.2	1.2	1.6	2.2
SCN^-	0.019	0.006	0.038	0.023	0.037	0.033	0.013	0.015
$S^=$	0.005	0.003	0.15	0.22	0.20	0.24	0.0002	0.00003
NH_3-N	0.12	0.12	2.6	1.5	0.35	0.55	0.033	0.043
Cl^-	0.92	0.52	0.56	0.19	0.42	0.19	0.24	0.10
TSS	129	185	177	178	116	68	87	91
Oil	0.7	0.5	0.31	0.10	0.29	0.32	0.12	0.13
(pH)	(7.3)		(7.4)		(7.8)		(7.2)	

* $$S = \left[\frac{\Sigma(X_i - \overline{X})^2}{M - 1}\right]^{1/2}$$

b. Comments were submitted to Procon on its process flow diagrams for sulfur, water, and steam systems. These should be considered preliminary because complete material balances have not yet been obtained.

1) Sulfur System: There are reservations about the use of the Wellman-Lord system for flue-gas scrubbing. It requires an extremely large amount of energy and significantly reduces overall process efficiency. A double-alkali process might be a preferred alternative. Specifically, the Envirotech-Chemico design should be considered because it was developed by a solids-liquids separation specialist in the pulp and paper industry and incorporates a reasonable extrapolation of a known recaustization process. A large-scale system is now being installed at Central Illinois Power and Light.

   Significant quantities of sulfur are fed to the flue-gas disposal system after incineration. Alternative disposal (either to the acid-gas removal unit or to the acid plant) is recommended to improve overall system efficiency and/or costs. This suggestion is offered whether the Wellman-Lord system is retained or an alternative is chosen. The indicated (and required) efficiency of the flue-gas disposal system is high. The required efficiencies drop if an alternative disposal system is included. It is suggested that efficiencies higher than those required should not be reported.

2) Plant Water System: The following suggestions were made for improving the overall process efficiency by reducing the evaporation package:

   - A hot-lime system on the cooling tower blowdown section will reduce the blowdown requirement to about 2% of the fresh feed.
   - Blowdown requirements for zeolite softeners and mixed-bed demineralizers can be significantly reduced by staging the regeneration processes of these units.
   - The low-pressure steam condensate could be used as high-pressure boiler feedwater makeup.
   - The low-pressure steam blowdown can be directly fed to the cooling tower.
   - The shift methanation condensate could be used as a high-pressure boiler feed makeup.
   - Utilize the cooling tower blowdown for the gasifier quench; any remaining portion of this stream could be sent to solids disposal.
   - Combine the lime-softening and biological sludge before thickening; then filter to recover reasonable-quality water. Three-hundred thousand pounds of high-quality water per hour could be saved.

- Replace the centrifuges and precoat filter with a thickener and belt filter to handle solids from the ash quench and prequench units.

- There are reservations about the oxygen-activated sludge unit, particularly about its ability to handle thiocyanate.

- There are reservations about reverse osmosis because it is not a sufficiently proved industrial process.

- The use of the two stages of lime softening, in series, can be questioned.

- A side-stream retention pond is suggested for use in the biological system to store off-specification sour water.

- The cost effectiveness of the phenol recovery system can be questioned. Other systems available for phenol destruction should be considered for comparison.

3) Steam Balance: The following comments were made on the overall plant steam balance.

- The overall plant efficiency could be improved by 4.4% by using the recommended amount of steam to the gasifier (1 mole of water per mole of carbon) instead of the amount currently being used (1.3 moles of water per mole of carbon).

- A further increase in plant efficiency (about 1%) is possible by generating 1450-psi steam in the methanation unit and acid plant instead of 600-psi steam.

c. Subsieve analyses for both bituminous and lignite coals were supplied to Procon for the design of the product gas cyclone (Tables 6 and 7).

d. Fines Generation in the Crushing Tests: Twelve drums of ROM coal (Peabody No. 10 Mine, Illinois No. 6 coal) were shipped to the T. J. Gundlach Machine Co., Belleville, Illinois, for crushing tests. A trip was made to its plant to discuss the procedure to be used in making the tests. The following test crushing procedure was developed:

1) Obtain a representative sample of the coal, as received, by coning and quartering. (Use a minimum sample of one drum.)

2) Screen one drum of sample for analysis.

3) Save two drums of feed coal for screen manufacturer (Sample A).

4) Screen approximately 1000 pounds of raw feed at 12 mesh.

5) Crush the 1-1/2 inch x 12-mesh fraction of the coal in the Gundlach Model 50-2C4R CAGE-PAKTOR at 200, 300, 400, and 500 rpm. Run each test at 100 pounds.

Table 6. SUBSIEVE ANALYSIS FOR TEST 37 USING MONTANA LIGNITE FROM 7/5/75 (1506 Hours) TO 7/7/75 (0694 Hours)

Sample	Coal Feed	Cyclone Dust
Subsieve Analysis, μ	wt %	
40.3	7.45	32.56
32.0	5.95	31.73
25.4	4.45	29.13
20.2	3.25	25.53
16.0	2.20	21.60
12.7	1.35	17.50
10.1	0.85	14.09
8.0	0.50	10.93
6.4	0.30	8.10
5.0	0.15	5.40
4.0	0.10	2.86
3.2	0.05	1.07

*Cumulative weight percent finer than stated size.

Table 7. SUBSIEVE ANALYSIS FOR TEST 54 USING ILLINOIS NO. 6 BITUMINOUS COAL FROM 7/3/76 (1800 Hours) TO 7/7/76 (0300 Hours)

Sample	Pretreated Char	Cyclone Dust
Subsieve Analysis, μ	wt %	
50.8	6.86	45.93
40.3	4.18	29.73
32.0	2.16	17.46
25.4	0.94	9.48
20.2	0.27	4.78
16.0	0.05	2.43
12.7	0.00	1.39
10.1	0.00	0.80
8.0	0.00	0.40
6.4	0.00	0.13
5.0	0.00	1.09
4.0	0.00	0.00

*Cumulative weight percent finer than stated size.

6) Screen the mill products from step 5, evaluate the results for the generation of −100 mesh fines, and estimate the quantity of the recirculating load.

7) Crush a 300-pound sample at optimum rpm, and make a mixture to approximate the feed in closed-circuit crushing. Then crush this mixture and screen again.

8) Finally, prepare two drums of product crushed at optimum rpm for screen manufacture (Sample B).

The closed-circuit crushing tests were then conducted at T. J. Gundlach Machine Co. The results of the open-circuit crushing tests were as follows:

Crusher Speed, rpm		200	260	300	400	500
Screen Size, wt % solids						
+12 Mesh		62.6	53.6	40.7	24.9	16.2
−12 Mesh x 100 Mesh	Product	30.8	39.1	49.7	60.8	65.5
−100 Mesh (Fines)	Product	6.6	7.3	9.6	14.2	18.3
Total		100.0	100.0	100.0	99.9	100.0
Fines in Product, %		17.6	15.7	16.1	19.0	21.8

First approximations to closed-loop crushing tests were made at 200 and 300 rpm. Closed-circuit crushing tests were not made at 400 and 500 rpm due to the higher production rate of fines. Using an estimated screen efficiency of 90%, the recycle load was established at 200% at 200 rpm, and 80% at 300 rpm. The following results for the closed-circuit crushing test were obtained:

Test		Open Circuit	Closed Circuit	Open Circuit	Closed Circuit
Crusher Speed, rpm		200		300	
Screen Size, wt % solids					
+12 Mesh		62.6	65.4	40.7	39.7
−12 Mesh x 100 Mesh	Product	30.8	30.9	49.7	53.0
−100 Mesh (Fines)	Product	6.6	3.7	9.6	7.5
Total		100.0	100.0	100.0	100.0
Fines in Product, %		17.6	10.7	16.1	12.4

The fines generated during the closed-circuit crushing test at 200 rpm were reduced from 17.6% to 10.7%, as the product (—12 mesh) decreased from 37.4% to 34.6%. At 300 rpm, the product increased from 59.3% to 60.5%, and fines were reduced from 16.1% to 12.4%. It is apparent from these data that fines production will be further reduced during the second and third passes in closed-circuit crushing.

The above data were acquired from a sample of coal that had been screened to remove the product-size material. Approximately 20% of the ROM coal passed through this screen (to go directly to process). This product-size material in the raw coal contained about 6% fines. When the fines content of the first-screen undersize was averaged with the fines content of the crushed product, the quantity of the total fines to process decreased to a range from 10.2% to 11.9%.

The following bases have been suggested after evaluating the above data together with the earlier data from crushing a less friable Illinois coal:

- For cost estimation purposes, it appears that a total fines production (—100 mesh) of less than 9% can be readily obtained for a generic Illinois No. 6 coal.

- For design purposes, a value of 12% fines generation appears reasonable.

e. Low-Pressure Design: Heat and material balances were supplied for the recommended operation of the gasifier at 500 psi. Two designs were included: one for the commercial plant to estimate gas cost and the other specified for lower temperatures and a higher steam/carbon ratio for the demonstration plant. The commercial design is presented in Figure 6 and Tables 8 through 12; the demonstration basis is included in Figure 7 and Tables 13 through 15. Both designs employ the same coal (Table 3) and require the same degree of pretreatment (Table 4) specified for the higher pressure cases.

Cold-Flow Model

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

The procurement status of the various elements of this model is as follows:

a. Compressor: On order. The revised delivery date is July 21.

b. Instrumentation: The instrumentation has been received.

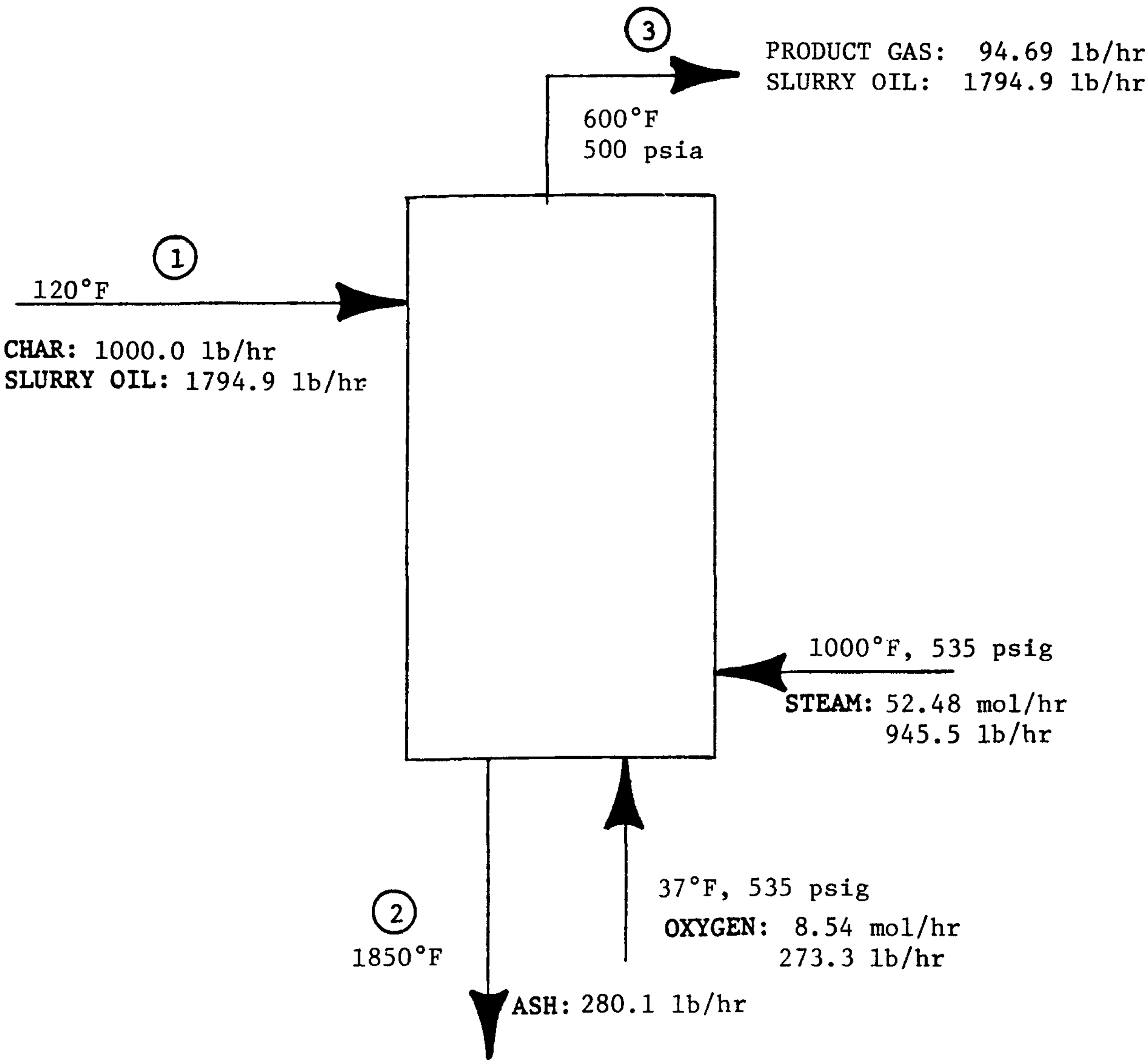


Figure 6. GASIFIER BALANCE (Commercial Design)

Table 8. GASIFIER MATERIAL AND HEAT BALANCE FOR TYPICAL ILLINOIS NO. 6 BITUMINOUS RUN-OF-MINE COAL (500-psia Operation)

Proximate Analysis	wt %
Volatile Matter	32.90
Moisture	12.00
Fixed Carbon	38.21
Ash	16.98
Total	100.00
Ultimate Analysis	
Carbon	62.70
Hydrogen	4.67
Oxygen	7.85
Nitrogen	1.18
Sulfur	4.25
Chloride	0.16
Ash	19.19
Total	100.00

Table 9. ULTIMATE ANALYSIS FOR PRETREATED COAL

Components	wt %
Carbon	63.03
Hydrogen	4.34
Oxygen	7.00
Nitrogen	1.18
Sulfur	3.42
Chloride	0.09
Ash	20.94
Total	100.00

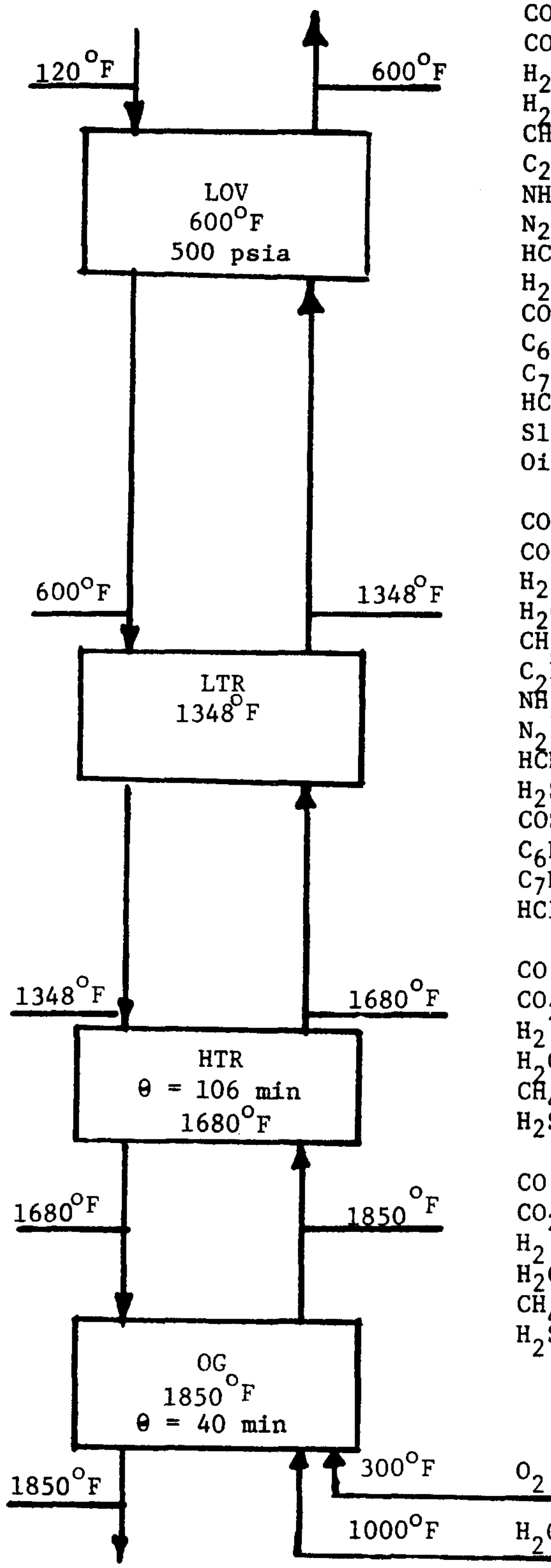
Table 10. GASIFIER BALANCE FOR COAL FEED AND ASH RESIDUE

Stream No.	1			2		
Description	Coal Feed			Ash Residue		
Temperature, °F	120			1800		
Components	lb/hr	wt %	mol/hr	lb/hr	wt %	mol/hr
C	630.3	63.03	52.48	63.0	22.49	5.25
H_2	43.4	4.34	21.53	1.9	0.68	0.95
O	70.0	7.00	4.37	--	--	--
N_2	11.8	1.18	0.42	2.4	0.86	0.08
S	34.2	3.42	1.07	3.4	1.21	0.11
Cl	0.9	0.09	0.02	--	--	--
Ash	209.4	20.94	--	209.4	74.76	--
Total	1000.0	100.00		280.1	100.00	
Slurry Oil	1794.9					

Table 11. GASIFIER BALANCE FOR RAW PRODUCT GAS

Stream No.	3	
Description	Raw Product Gas	
Temperature, °F	600	
Components	mol/hr	mol %
CO	17.11	18.19
CO_2	16.12	17.14
H_2	25.23	26.82
H_2O	24.56	26.11
CH_4	9.19	9.77
C_2H_6	0.26	0.28
NH_3	0.52	0.55
N_2	0.05	0.05
HCN	0.05	0.05
H_2S	0.92	0.98
COS	0.04	0.04
HCl	0.02	0.02
Total (Oil-Free Gas)	94.07	100.00
		wt %
C_6H_6	0.11	15.0
C_7H_8	0.51	85.0
Total (Product Oil)	0.62	100.0
Total (Oil-Free Gas + Product Oil)	94.69	
Slurry Oil, lb/hr	1794.9	

Table 12. GASIFIER BALANCE FOR ILLINOIS BITUMINOUS COAL FOR COMMERCIAL DESIGN CASE (Basis: 1000 lb Char)



Feed to LOV (120°F):

	Moles
C	52.4813
H_2	21.5278
O	4.3750
N_2	0.4212
S	1.0667
Cl	0.0244
Ash*	209.4440
Moisture	0.0
Slurry* Oil	1794.9

Product from LOV (600°F):

	Moles
CO	17.1124
CO_2	16.1172
H_2	25.2250
H_2O	24.5561
CH_4	9.1937
C_2H_6	0.2624
NH_3	0.5247
N_2	0.0505
HCN	0.0481
H_2S	0.9217
COS	0.0384
C_6H_6	0.1050
C_7H_8	0.5098
HCl	0.0244
Slurry* Oil	1794.9

LOV to LTR (600°F):

	Moles
C	52.4813
H_2	21.5278
O	4.3750
N_2	0.4212
S	1.0667
Cl	0.0244
Ash*	209.4440

LTR to LOV (1348°F):

	Moles
CO	17.1124
CO_2	16.1172
H_2	25.2250
H_2O	24.5561
CH_4	9.1937
C_2H_6	0.2624
NH_3	0.5247
N_2	0.0505
HCN	0.0481
H_2S	0.9217
COS	0.0384
C_6H_6	0.1050
C_7H_8	0.5098
HCl	0.0244

LTR to HTR (1348°F):

	Moles
C	42.2870
H_2	2.9261
O	--
N_2	0.0842
S	0.2093
Ash*	209.4440

HTR to LTR (1680°F):

	Moles
CO	17.7546
CO_2	14.1625
H_2	20.6202
H_2O	23.4867
CH_4	5.1218
H_2S	0.1026

HTR to OG (1680°F):

	Moles
C	35.3235
H_2	2.4442
O	--
N_2	0.0842
S	0.2093
Ash*	209.4440

OG to HTR (1850°F):

	Moles
CO	15.9379
CO_2	12.1704
H_2	20.6468
H_2O	29.2876
CH_4	1.9671
H_2S	0.1026

From OG (1850°F):

	Moles
C	5.2481
H_2	0.9543
O	--
N_2	0.0842
S	0.1067
Ash*	209.4440

Feed to OG:

		Moles
300°F	O_2	8.5425
1000°F	H_2O	52.4813

* These quantities are in lbs.

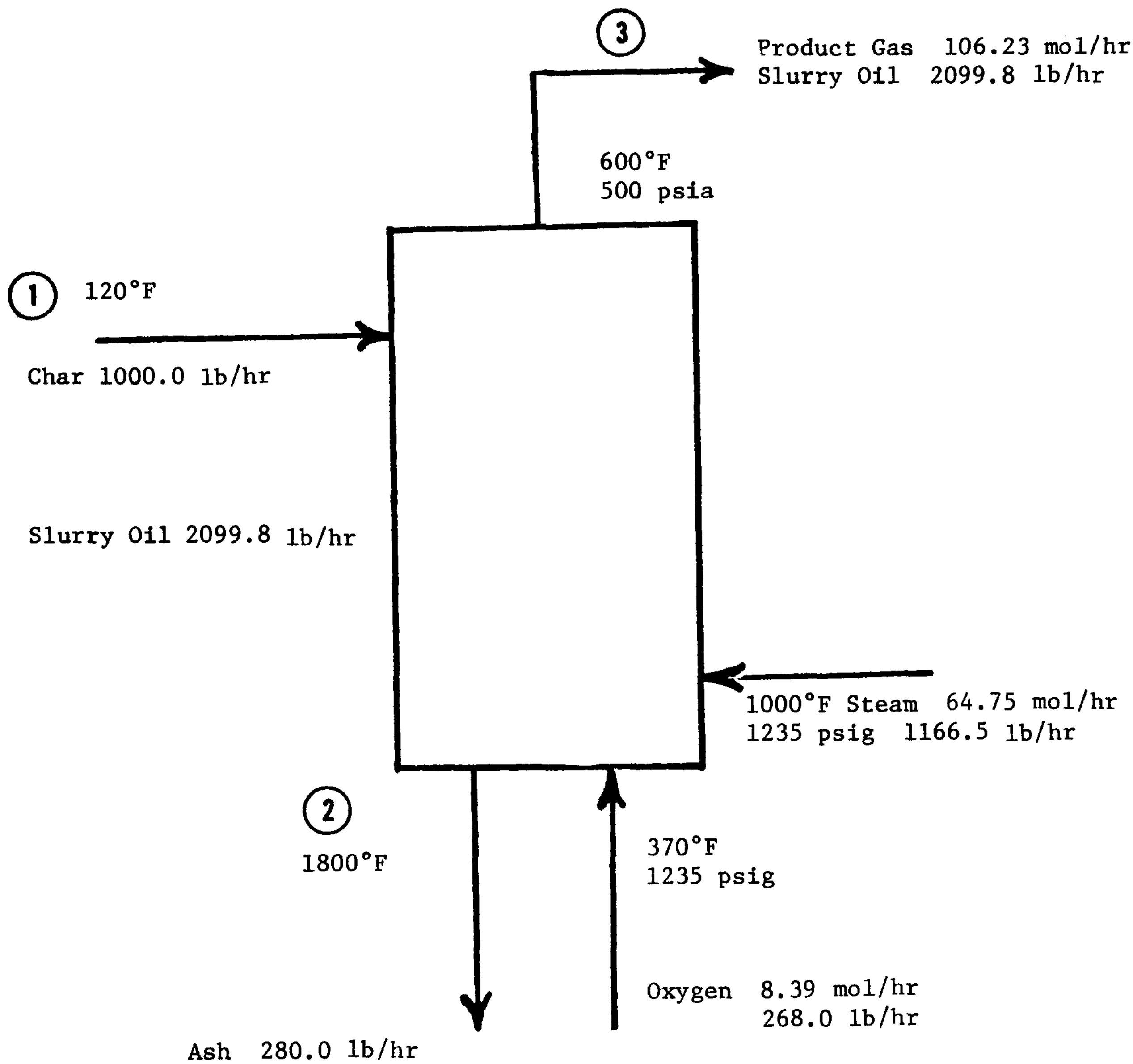


Figure 7. GASIFIER BALANCE

Table 13. GASIFIER BALANCE FOR STREAMS 1 AND 2

Stream No.	1			2		
Description	Coal Feed			Ash Residue		
Temperature, °F	120			1800		
Components	lb/hr	wt %	mol/hr	lb/hr	wt %	mol/hr
C	630.3	63.03	52.48	62.9	22.46	5.24
H_2	43.4	4.34	21.53	1.9	0.68	0.94
O	70.0	7.00	4.37	--	--	--
N_2	11.8	1.18	0.42	2.4	0.86	0.08
S	34.2	3.42	1.07	3.4	1.21	0.11
Cl	0.9	0.09	0.02	--	--	--
Ash	209.4	20.94	--	209.4	74.79	--
Total	1000.0	100.00		280.0	100.00	
Slurry Oil	2099.8					

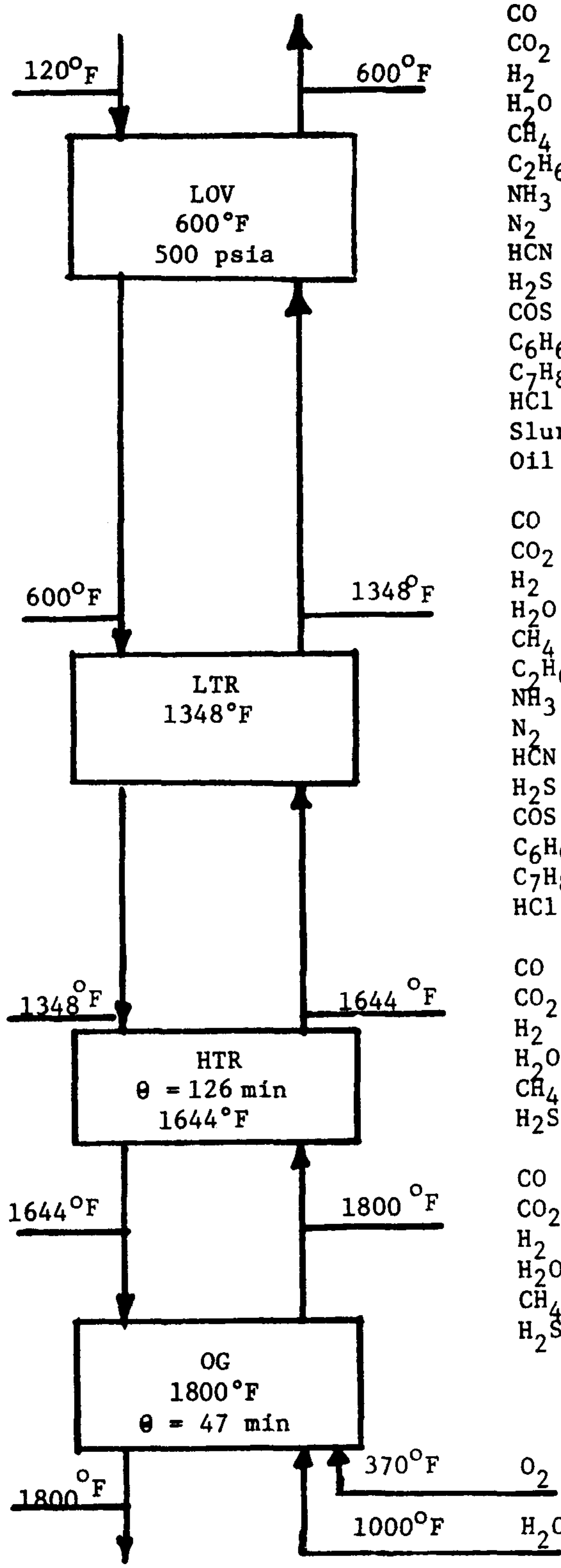
Table 14. GASIFIER BALANCE FOR STREAM 3 AT 600°F

Stream No.	3	
Description	Raw Product Gas	
Temperature, °F	600	
Components	mol/hr	mol %
CO	14.55	18.78
CO_2	18.32	17.35
H_2	26.65	25.23
H_2O	34.67	32.82
CH_4	9.56	9.05
C_2H_6	0.26	0.25
NH_3	0.52	0.49
N_2	0.05	0.05
HCN	0.05	0.05
H_2S	0.92	0.87
COS	0.04	0.04
HCl	0.02	0.02
Total (Oil-Free Gas)	105.61	100.00
		wt %
C_6H_6	0.11	15.0
C_7H_8	0.51	85.0
Total (Product Oil)	0.62	100.0
Total (Oil-Free Gas + Product Oil)	106.23	
Slurry Oil, lb/hr	2099.8	

Table 15. GASIFIER BALANCE FOR ILLINOIS BITUMINOUS COAL
(Basis: 1000 lb Char)

	Moles
C	52.4813
H_2	21.5278
O	4.3750
N_2	0.4212
S	1.0667
Cl	0.0244
Ash*	209.4440
Moisture	0.0
Slurry* Oil	2099.8

	Moles
CO	14.5540
CO_2	18.3170
H_2	26.6535
H_2O	34.6734
CH_4	9.5608
C_2H_6	0.2624
NH_3	0.5247
N_2	0.0505
HCN	0.0481
H_2S	0.9217
COS	0.0384
C_6H_6	0.1050
C_7H_8	0.5098
HCl	0.0244
Slurry* Oil	2099.8



C	52.4813
H_2	21.5278
O	4.3750
N_2	0.4212
S	1.0667
Cl	0.0244
Ash*	209.4440

CO	14.5540
CO_2	18.3170
H_2	26.6535
H_2O	34.6734
CH_4	9.5608
C_2H_6	0.2624
NH_3	0.5247
N_2	0.0505
HCN	0.0481
H_2S	0.9217
COS	0.0384
C_6H_6	0.1050
C_7H_8	0.5098
HCl	0.0244

C	41.3724
H_2	2.9320
O	--
N_2	0.0842
S	0.2006
Ash*	209.4440

CO	14.9060
CO_2	16.6525
H_2	22.1828
H_2O	33.3138
CH_4	5.5743
H_2S	0.0939

C	34.5876
H_2	2.3933
O	--
N_2	0.0842
S	0.2006
Ash*	209.4440

CO	12.9592
CO_2	14.1470
H_2	21.3513
H_2O	40.2716
CH_4	2.2418
H_2S	0.0939

C	5.2396
H_2	0.9429
O	--
N_2	0.0842
S	0.1067
Ash*	209.4440

370°F	O_2	8.3874
1000°F	H_2O	64.7500

* These quantities are in lbs.

c. Building Foundation: The building foundation has been completed.

d. Building Structural Steel: All six decks are up. The siding and insulation remain to be completed.

e. Vessels: The test vessel and the cyclone receiver vessel have been received and erected. The expected delivery date for the solids receiver vessel is now July 21.

f. Cyclone: The cyclone has been received.

g. Pipeline Filters: These have been received and mounted in the structure.

h. Control Valves: The control valves have been received.

Because of the delay in delivery of the solids receiver vessel (due to a defective head), the project completion date was extended by 6 to 7 weeks. Some of this delay was compensated for by piping construction while waiting for the solids receiver vessel. In fact, some of the piping work has already been completed.

Commercial Plant Reactor Design

Seven potential low-temperature reactor (LTR) lift-line feeder configurations have been tested. The lift pot I, lift pot II, L-valve, and reverse-seal (leg) feeding devices were tested last quarter, and detailed test results for these configurations were presented in the Project 9000 Quarterly Report No. 7 (DOE Report No. FE-2434-29). The J-valve, lift pot III, and reverse-seal lift pot configurations were tested during this quarter, and detailed results of these tests follow. The characteristics and relative merits of all the devices tested are also summarized.

J-Valve Tests

The first configuration tested, a J-valve, is shown in Figure 8. In operation, the solids passed down a 2-inch-diameter downcomer in gravity flow to a 180-degree return bend. Aeration gas, added to the J-valve at a point immediately above the bend, caused the solids to flow around the bend and up into the lift line. The flow rate of solids through the J-valve was controlled by varying the amount of aeration gas fed to the J-valve.

The solids were carried into the 2-inch-diameter lift line by the lift gas, which was channelled into the lift line proper by a 3-inch-diameter tube concentric to the vertical portion of the J-valve. (See Figure 8.) The

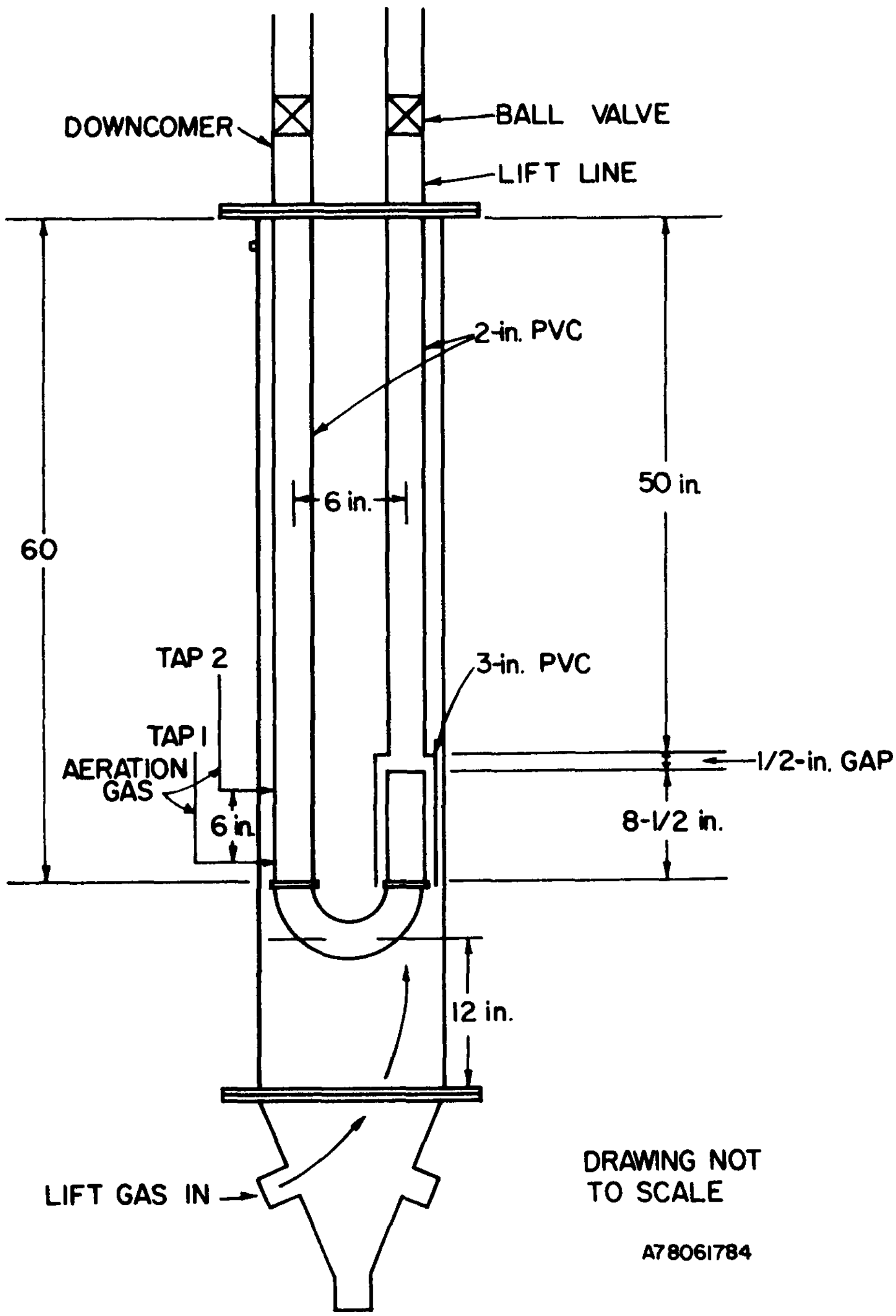


Figure 8. J-VALVE TEST CONFIGURATION

J-valve was constructed of Plexiglas and clear polyvinyl chloride (except for the 180-degree return bend, which was made of metal) so that solids flow through the device could be visually monitored. The dimensions of the J-valve device are also shown in Figure 8 and in a photograph (Figure 9).

In a typical run, the ball valve in the downcomer was fully opened to allow solids to fill up the J-valve, and then the lift-line gas flow rate was set. Aeration gas was added to the J-valve, and solids flow was initiated. Readings were taken at several different solids flow rates, determined by timing individual solids particles as they passed between two marks (12 inches apart) in the clear polyvinyl chloride downcomer, and the results analyzed.

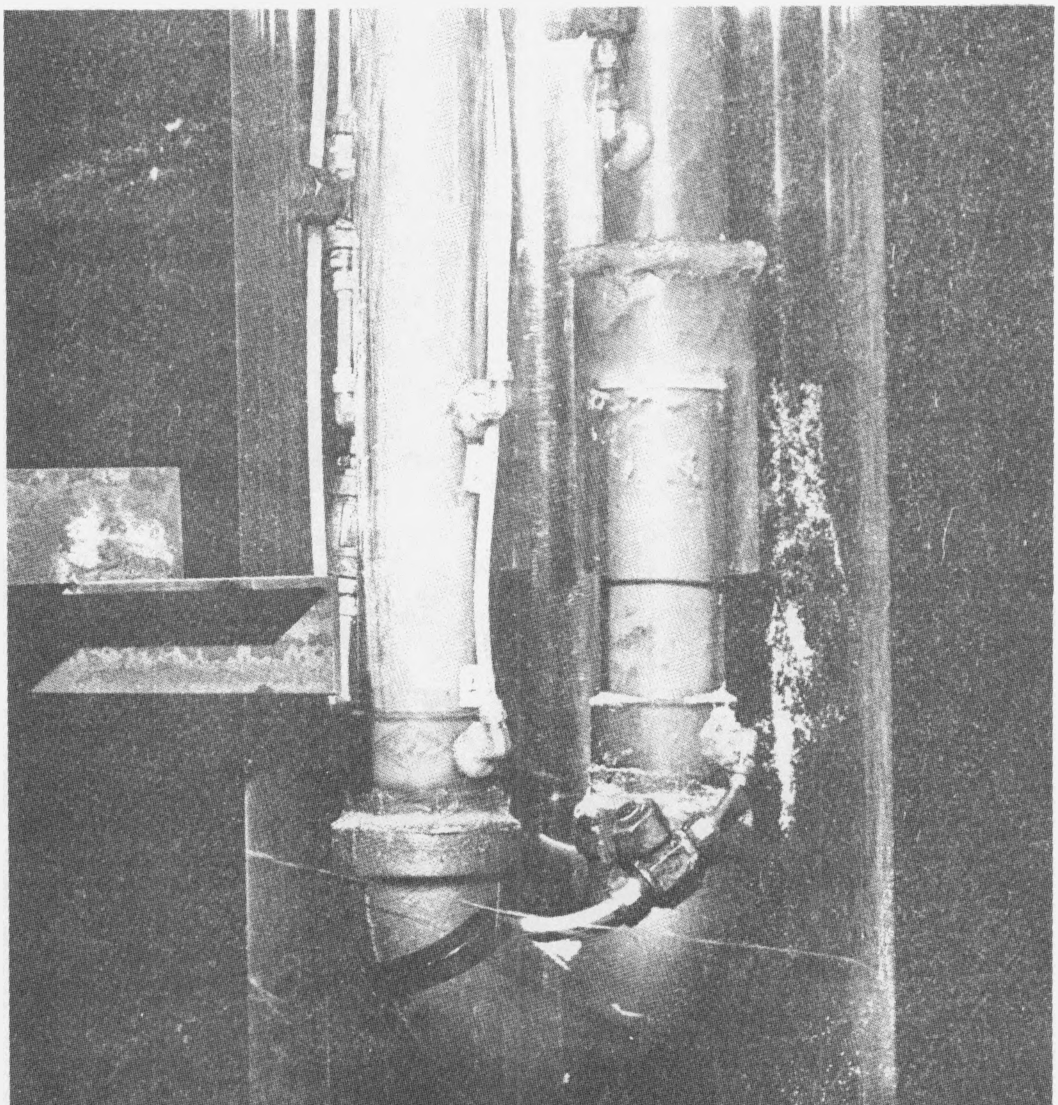
It is important that the solids be injected into the lean-phase lift line smoothly and controllably to prevent slugging and poor conversion in the LTR. Fluctuations in the recorder tracing for the lean-phase lift-line pressure drop were used to analyze the smoothness of the lift-line operation.

The J-valve was tested using −20+80 mesh sand and −20+200 mesh pretreated Illinois No. 6 bituminous coal. When it was first operated (Runs HGD-9A through HGD-9D), the gap between the J-valve exit pipe and the entrance to the lift line was set at 1 inch. With this configuration, a large percentage of the solids spilled down out of the lift-line entrance into the 12-inch Plexiglas tube section. This was clearly unsatisfactory, and the gap was reduced to 0.5 inch, thereby increasing the velocity in this region and eliminating solids spilling.

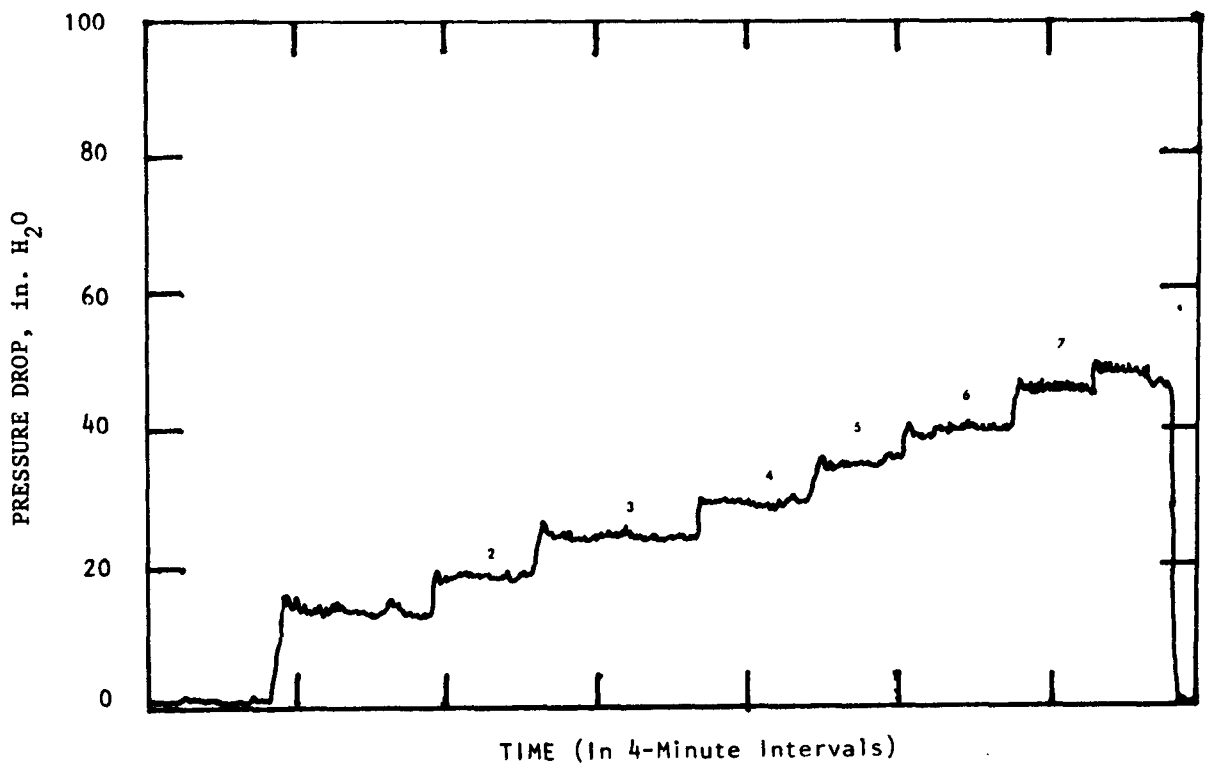
Four tests were conducted with −20+80 mesh sand after the gap had been reduced. In Run HGD-9AA (Figure 10) the total lift-line velocity was set at 50 ft/s, and aeration was added to the J-valve at tap 1 as shown in Figure 8. A maximum solids flow rate of 7400 lb/hr was achieved in this run. The lift-line pressure drop fluctuated ±1 inch of water from the average pressure drop reading. J-valve operation was smooth and controllable.

In Run HGD-9BB (Figure 11) the lift-line velocity was set at 45 ft/s, and aeration gas was still added at tap 1. A maximum solids flow rate of 8050 lb/hr was obtained in this run. Fluctuations in the lift line were the same as in Run HGD-9AA.

Run HGD-9CC (Figure 12) was made with a lift-line velocity of 40 ft/s and with aeration gas again being supplied at tap 1. A maximum solids flow



P78020328
Figure 9. J-VALVE LIFT-LINE FEEDER CONFIGURATION



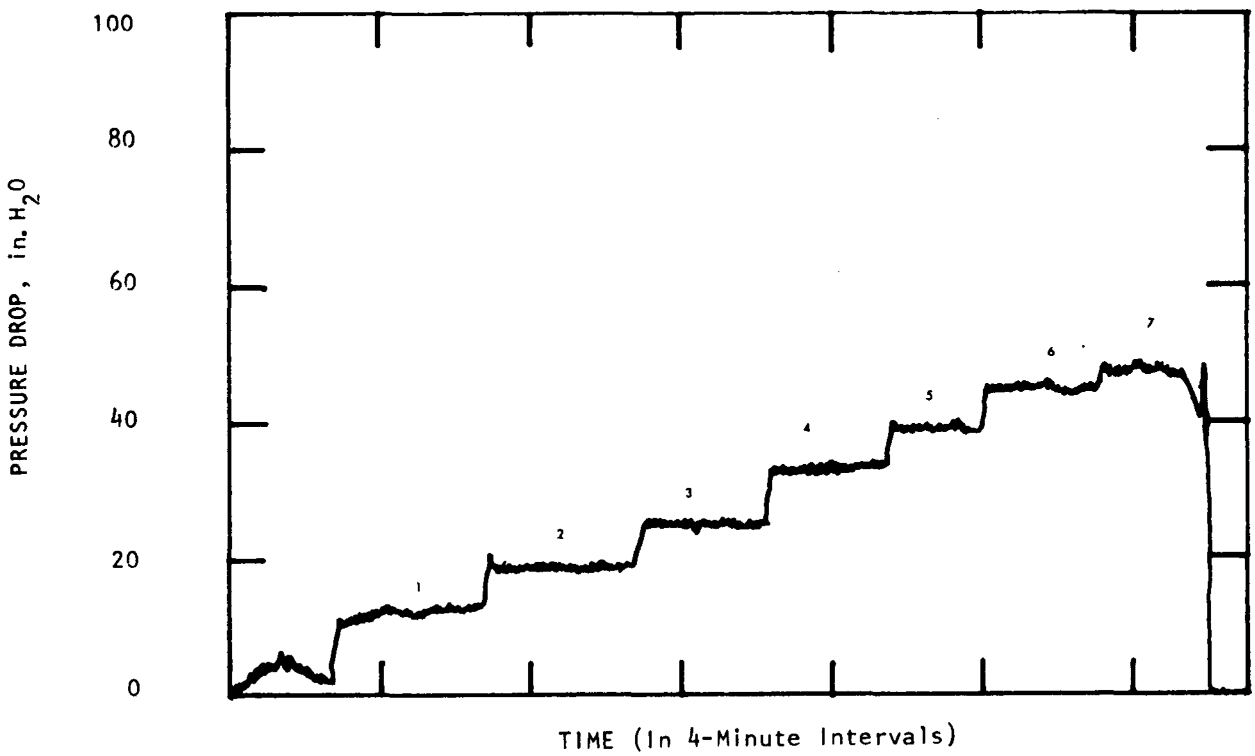
Lift-Line Velocity: 50 ft/s

Solids: —20+80 Mesh Ottawa Sand

Aeration: Point 1

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	2050	1.0379
2	2900	1.2360
3	3800	1.3752
4	4550	1.5569
5	3600	1.7255
6	6400	1.8812
7	7400	2.0758

Figure 10. LIFT-LINE PRESSURE DROPS DURING RUN HGD-9AA



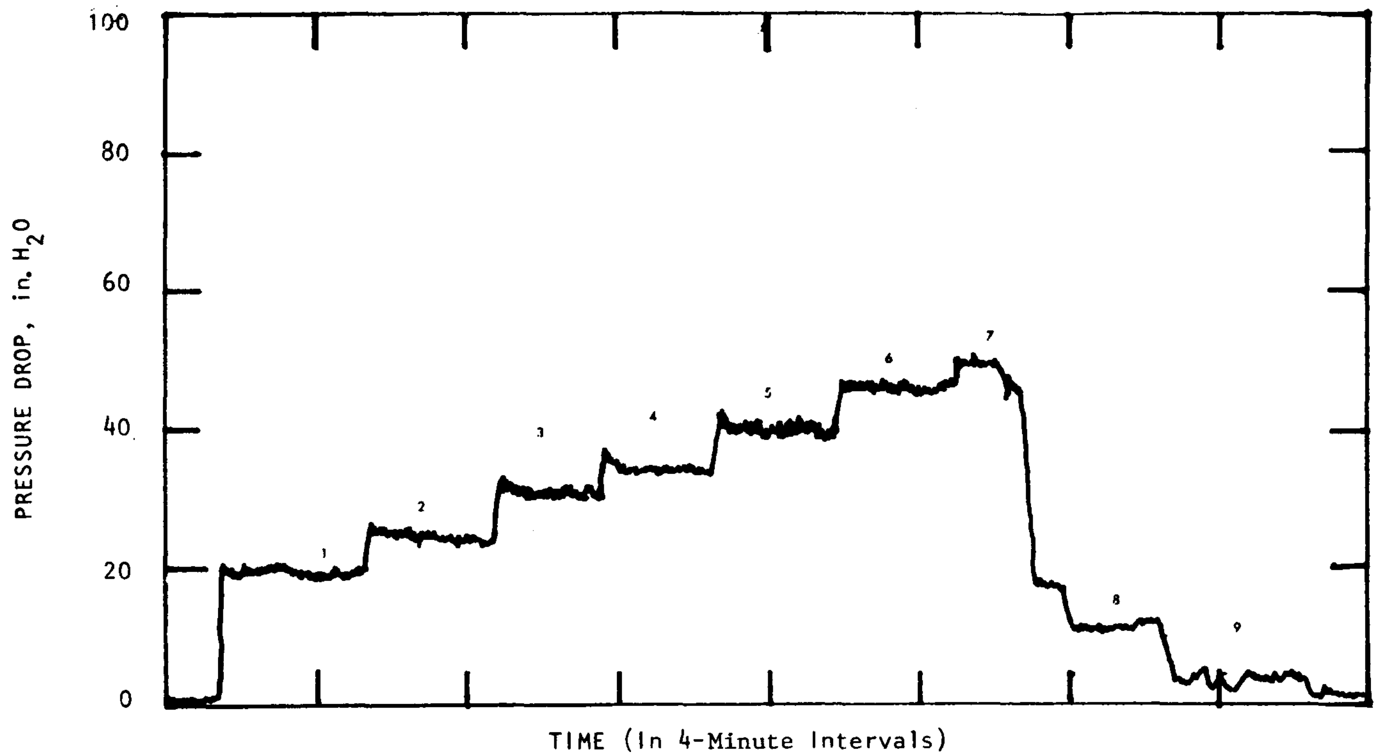
Lift-Line Velocity: 45 ft/s

Solids: −20+80 Mesh Ottawa Sand

Aeration: Point 1

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	--	0.8131
2	1750	0.9757
3	3100	1.2360
4	4250	1.3791
5	5750	1.6263
6	6900	1.8865
7	8050	2.0816

Figure 11. LIFT-LINE PRESSURE DROPS DURING RUN HGD-9BB



Lift-Line Velocity: 40 ft/s

Solids: −20+80 Mesh Ottawa Sand

Aeration: Point 1

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	3530	1.1742
2	4420	1.3047
3	5975	1.5004
4	6635	1.6961
5	7950	1.8865
6	9220	2.0816
7	9880	2.2117
8		0.9757
9		0.8131

Figure 12. LIFT-LINE PRESSURE DROPS DURING RUN HGD-9CC

rate of 9880 lb/hr was achieved in this run. Once again the J-valve operated smoothly and controllably.

Aeration tap 2 was used in Run HGD-9DD (Figure 13) at a lift-line velocity of 40 ft/s. A maximum solids flow rate of 8600 lb/hr was obtained in this run. Lift-line pressure drop fluctuations were again about ±1 inch of water from the average pressure drop reading.

The J-valve configuration was also tested with −20+200 mesh pretreated Illinois No. 6 bituminous coal. In Run HGD-10A, (Figure 14) aeration gas was added at tap 1 at a lift-line velocity of 25 ft/s. As aeration to the J-valve was increased, the solids flow rate up the lift line increased. At reading No. 3, the J-valve pressure drop fell suddenly, but the lift-line pressure drop remained nearly constant despite a threefold increase in aeration. The lowered J-valve pressure drop could possibly have been caused by gas bypassing along the top of the J-bend. The maximum solids flow rate obtained in this run was 1510 lb/hr.

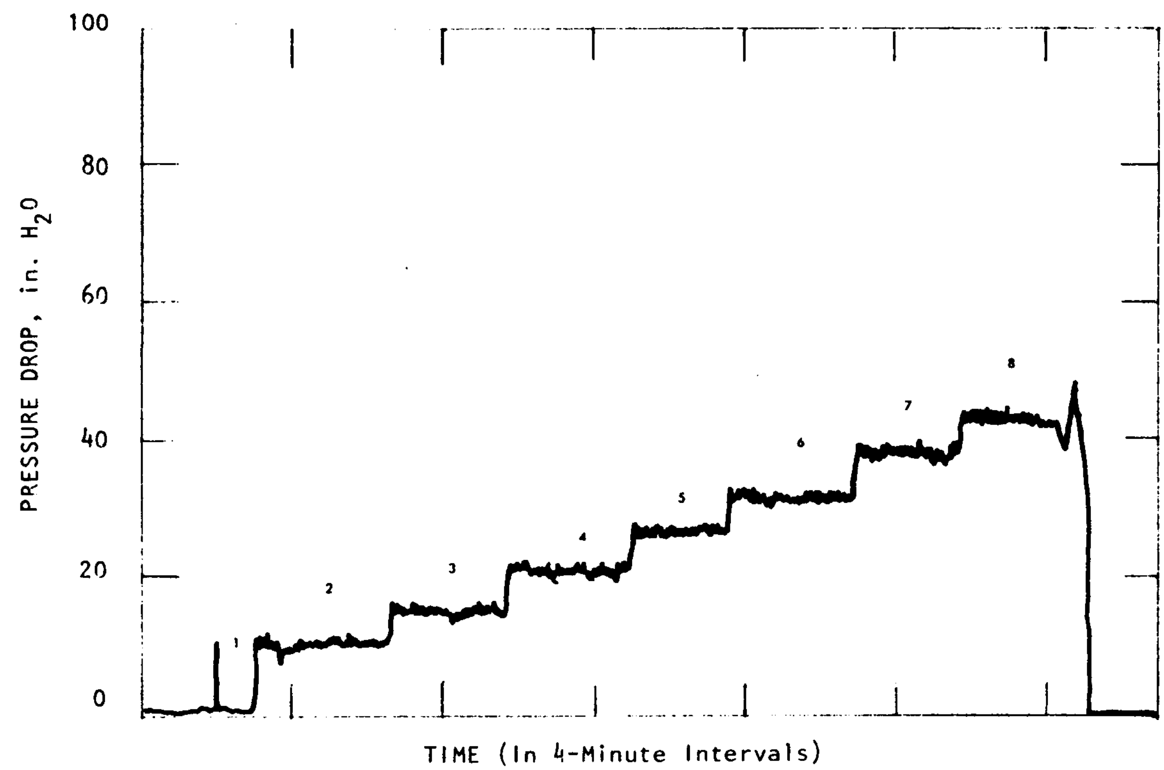
Run HGD-10B (Figure 15) was made with a lift-line velocity of 30 ft/s and with aeration gas being supplied at tap 1. A drop in J-valve pressure was again observed in this run. The maximum solids flow rate obtained in this run was 1860 lb/hr.

Runs HGD-10C (Figure 16) and HGD-10D (Figure 17) were made with aeration gas being added at tap 1 and with lift-line velocities of 35 and 40 ft/s, respectively. The maximum solids flow rate in each run was approximately 1500 lb/hr.

In the final test, Run HGD-10E (Figure 18), the lift-line velocity was set at 30 ft/s, and the aeration point was switched to location 2. The maximum solids flow rate in this run was 1810 lb/hr. In this run, there were no drops in J-valve pressure. Thus, gas bypassing is dependent upon the particular aeration tap used. Fluctuations in the lift-line pressure drop were essentially negligible for all of the runs with coal.

<u>Lift-Pot III</u>

Lift-pot III was also tested during this quarter (Figure 19). This configuration is also shown in Figure 20. In operation, the solids passed down a 2-inch-diameter downcomer in gravity flow to a full-ported ball valve, which controlled the solids flow rate. The solids passed through the ball valve,



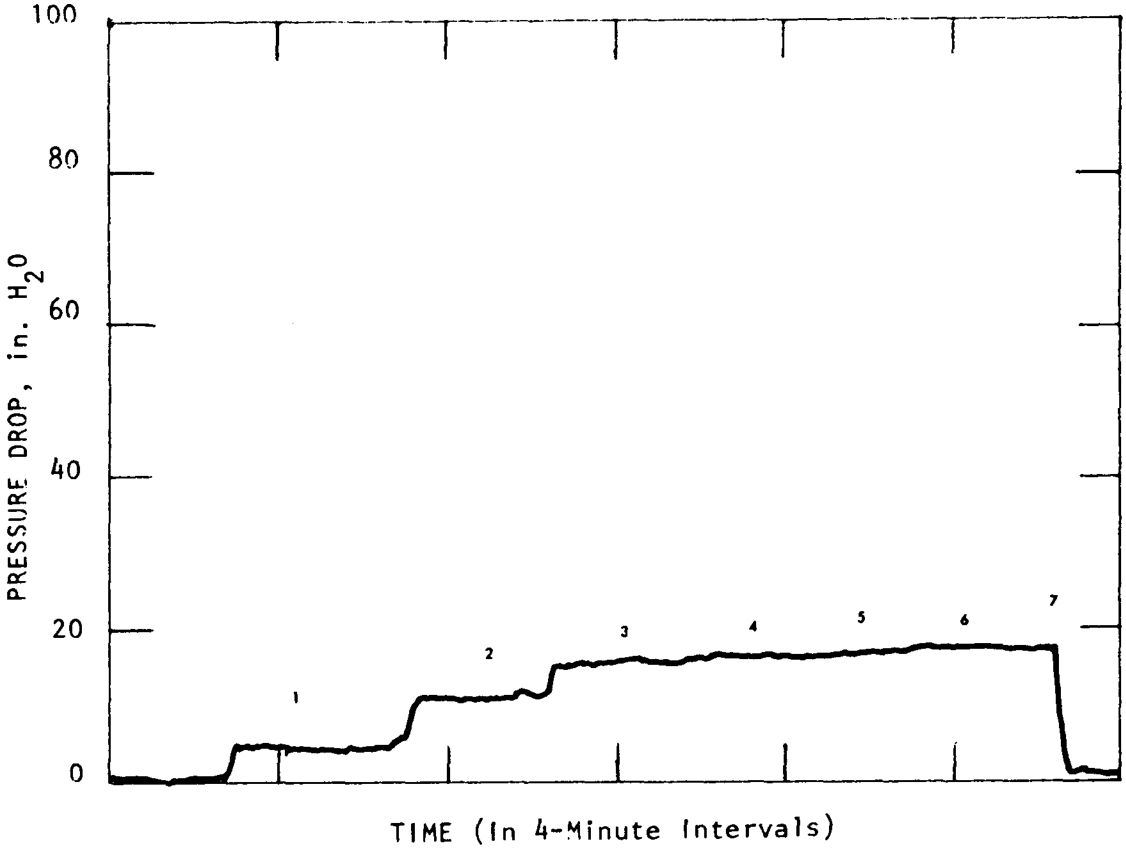
Lift-Line Velocity: 40 ft/s

Solids: −20+80 Mesh Ottawa Sand

Aeration: Point 2

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	0	0.8086
2	1600	0.9703
3	2445	1.1643
4	3765	1.2937
5	5080	1.4878
6	5930	1.6770
7	7530	1.8705
8	8600	2.0640

Figure 13. LIFT-LINE PRESSURE DROPS DURING RUN HGD-9DD



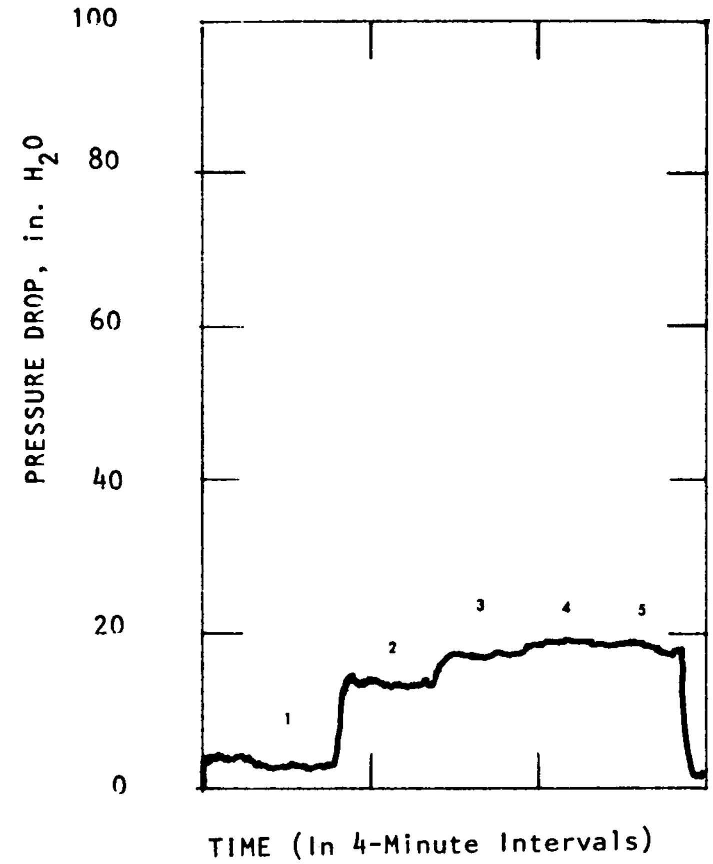
Lift-Line Velocity: 25 ft/s

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

Aeration: Point 1

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	520	0.3394
2	930	0.6788
3	1350	0.8802
4	1350	1.2157
5	1425	1.6209
6	1510	2.2288
7	1510	2.4314

Figure 14. LIFT-LINE PRESSURE DROPS DURING RUN HGD-10A



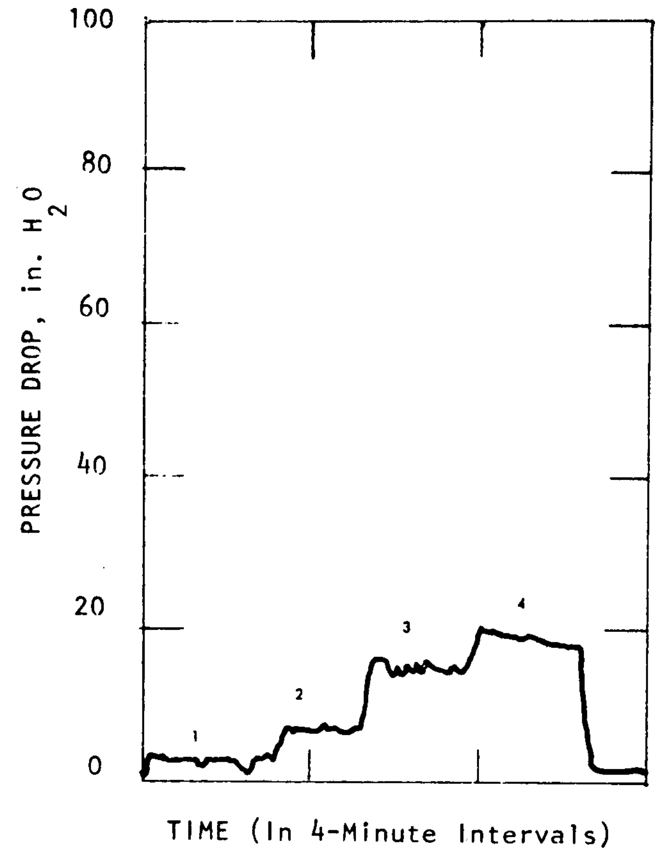
Lift-Line Velocity: 30 ft/s

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

Aeration: Point 1

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	105	0.3377
2	1260	0.7411
3	1650	1.0106
4	1860	1.2127
5	1860	1.4148

Figure 15. LIFT-LINE PRESSURE DROPS DURING RUN HGD-10B



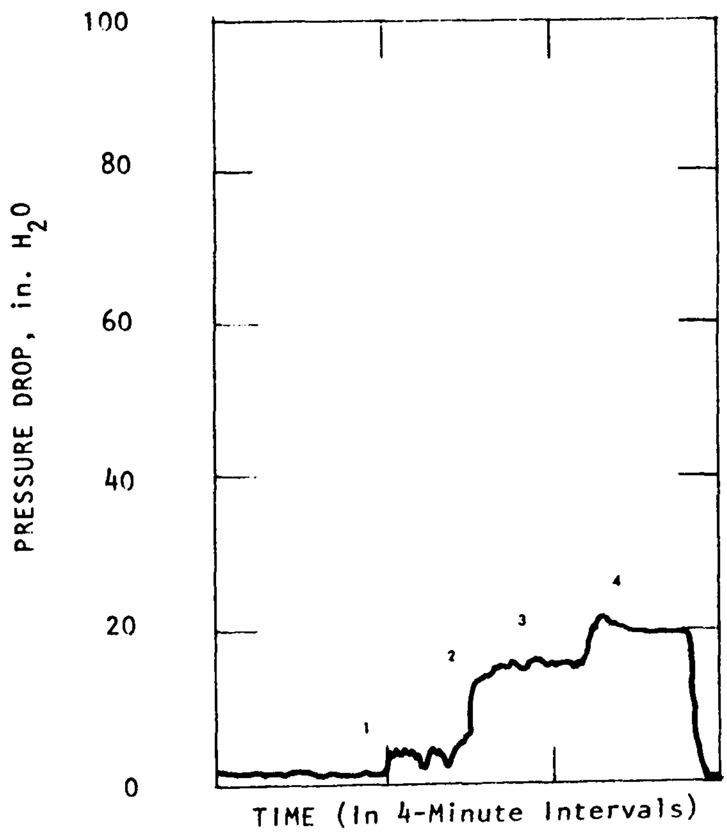
Lift-Line Velocity: 35 ft/s

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

Aeration: Point 1

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	95	0.3360
2	425	0.6720
3	1100	0.8736
4	1515	1.0724

Figure 16. LIFT-LINE PRESSURE DROPS DURING RUN HGD-10C



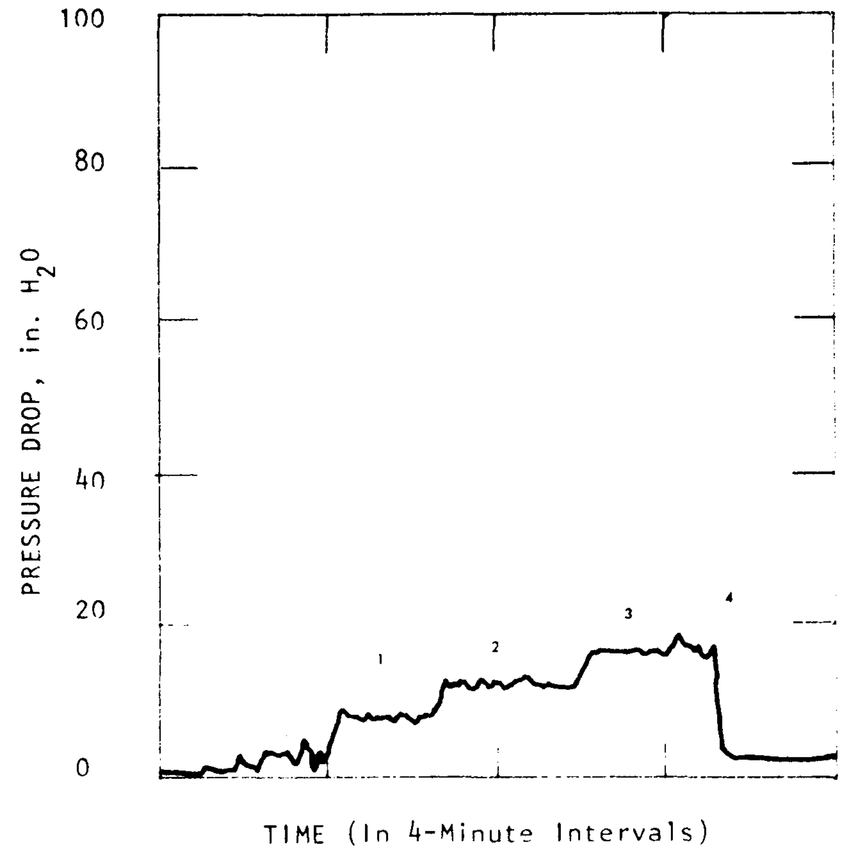
Lift-Line Velocity: 40 ft/s

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

Aeration: Point 1

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	0	0.3203
2	140	0.6369
3	1035	0.8669
4	1490	1.0642

Figure 17. LIFT-LINE PRESSURE DROPS DURING RUN HGD-10D



Lift-Line Velocity: 30 ft/s

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

Aeration: Point 2

Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	650	0.7335
2	1085	0.8669
3	1650	0.9977
4	1810	1.0642

Figure 18. LIFT-LINE PRESSURE DROPS DURING RUN HGD-10E

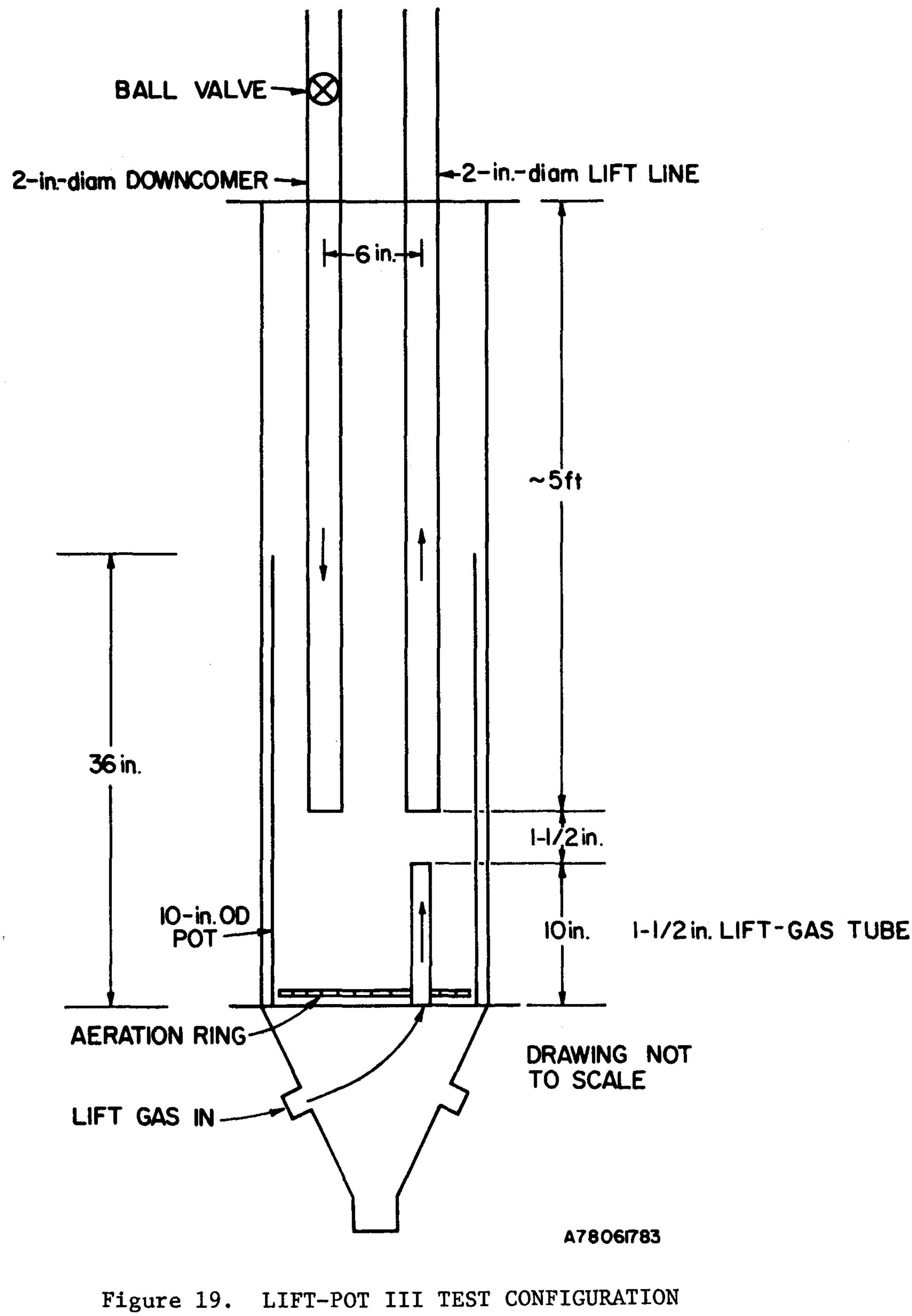


Figure 19. LIFT-POT III TEST CONFIGURATION

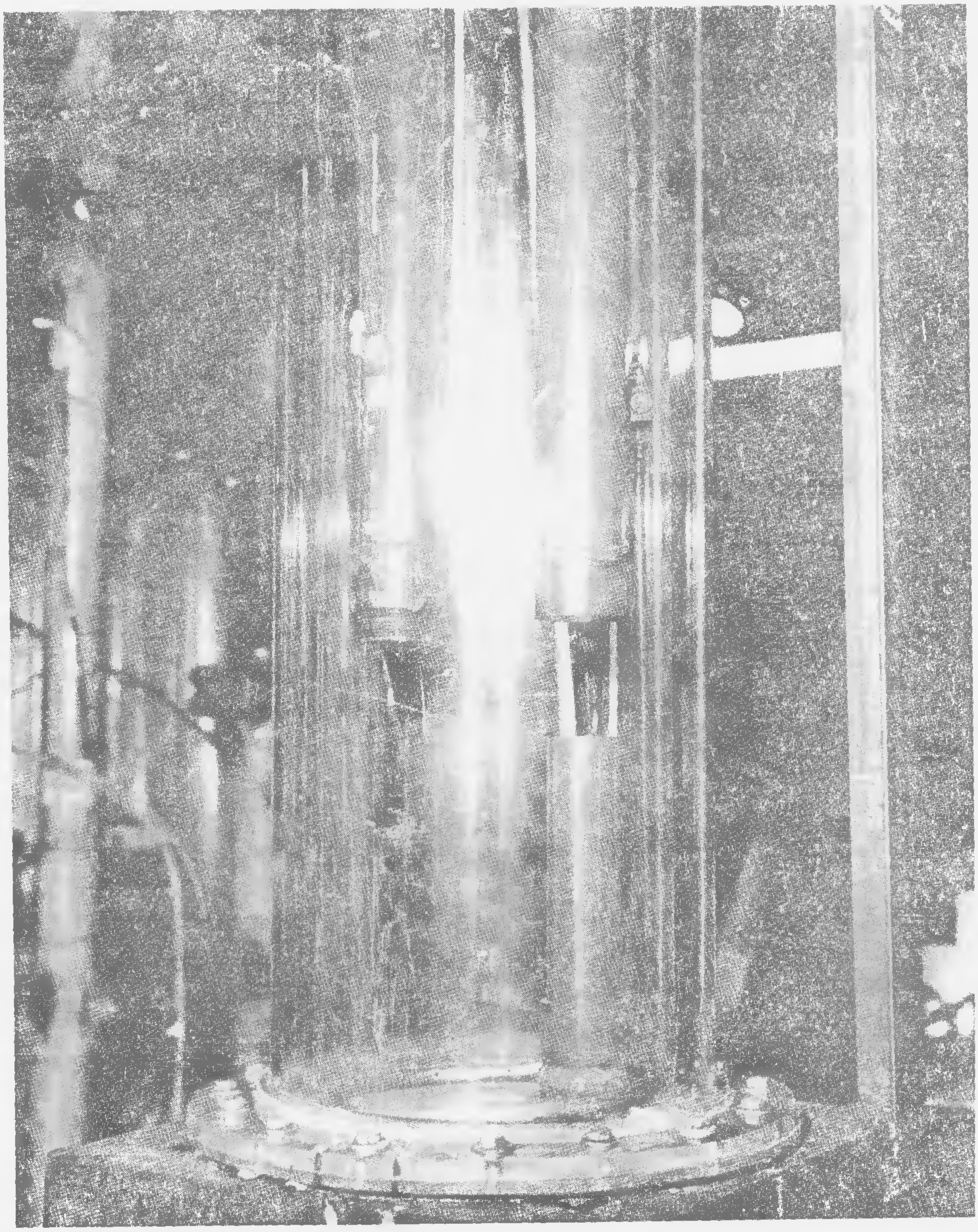


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Figure 20. LIFT-POT III FEEDER CONFIGURATION

dropped into a fluidized bed and passed through it, and were carried into the lift line by the lift gas. The lift gas was introduced into the lift line from a 1.5-inch-diameter tube, which was placed in the bed directly under the lift line to direct the lift gas into the line. The outlet of this tube was placed 1.5 inches below the entrance of the lift line. The entire lift-pot III configuration was constructed of Plexiglas and clear polyvinyl chloride pipe so that the solids flow could be visually monitored.

In a typical run, the ball valve in the downcomer was opened slightly to fill up the lift pot. Aeration gas to fluidize the lift pot was supplied to the lift pot through a ring in the bottom of the pot. The lift gas was then set at the desired velocity. Solids flow rates were controlled with the ball valve in the downcomer.

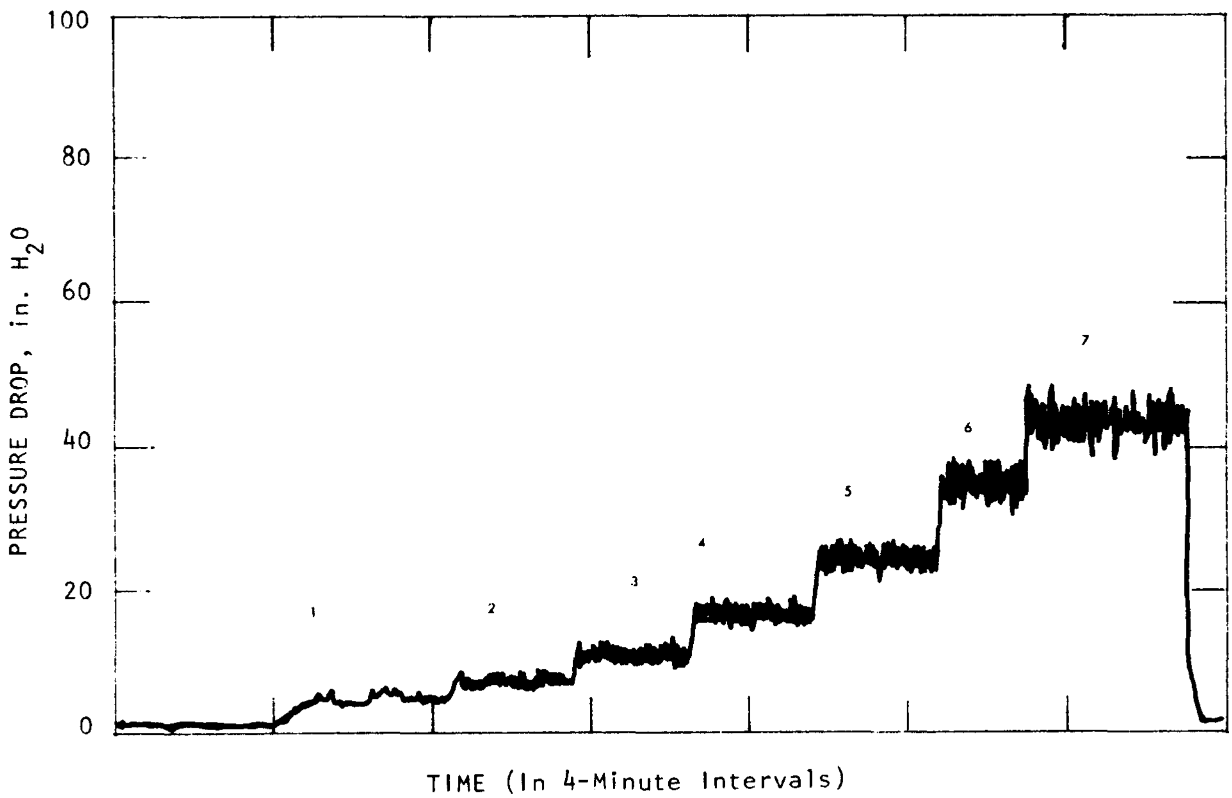
This configuration was first tested (Runs HGD-11A through HDG-11D) with a 3-inch gap between the outlet of the lift-gas routing tube and the entrance to the lift line itself. The device did not operate well, and a smaller gap was tried. With the larger gap, the solids appeared to be "sand blasting" the end of the lift line, and fluctuations in the lift line were much greater.

In the first run with −20+80 mesh Ottawa sand conducted with the smaller gap (Run HGD-11AA shown in Figure 21), the lift-line velocity was set at 40 ft/s and the lift-pot velocity at 0.5 ft/s. The maximum solids flow rate obtained was 9250 lb/hr. Fluctuation in the lift-line pressure drop increased with solids flow and was above ±3.5 inches of water from the average pressure reading at the maximum solids flow rate.

In Run HGD-11BB (Figure 22), the lift-line velocity was kept at 40 ft/s, and pot fluidization was reduced to 0.4 ft/s. The maximum solids flow rate obtained under these conditions was 9250 lb/hr. Fluctuations in the lift-line pressure drop were approximately the same as in Run HGD-11AA.

Run HGD-11CC (Figure 23) was made with a lift-line velocity of 35 ft/s and a lift-pot velocity of 0.4 ft/s. The maximum solids flow rate obtainable under these conditions was only 8700 lb/hr — much lower than that obtained in the previous run. The lift-line pressure drop fluctuations were similar to those observed in the previous runs.

In Runs HGD-11DD (Figure 24) and HGD-11EE (Figure 25), the lift-line velocities were set at 45 and 50 ft/s, respectively. The lift-pot velocity was set at 0.4 ft/s in both runs. The maximum solids flow rates obtained



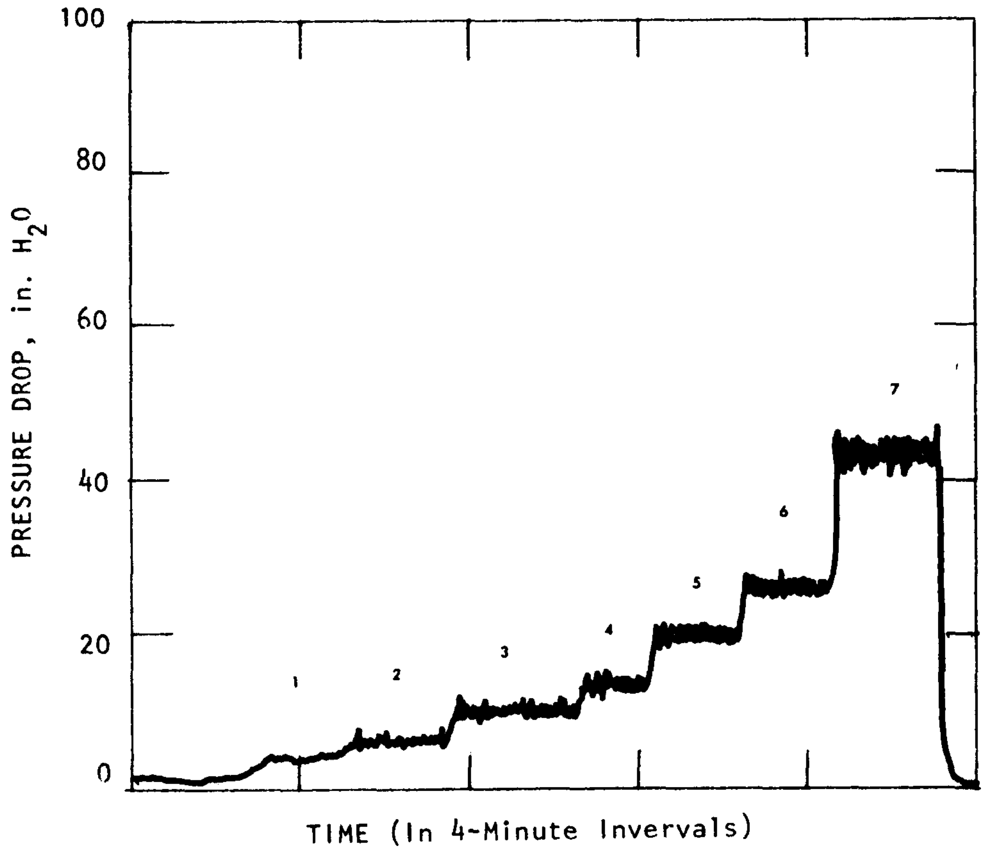
Lift-Line Velocity: 40 ft/s

Solids: −20+80 Mesh Ottawa Sand

Pot Fluidization Velocity: 0.5 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	400
2	750
3	1600
4	3000
5	4600
6	7150
7	9250

Figure 21. LIFT-LINE PRESSURE DROPS DURING RUN HGD-11AA



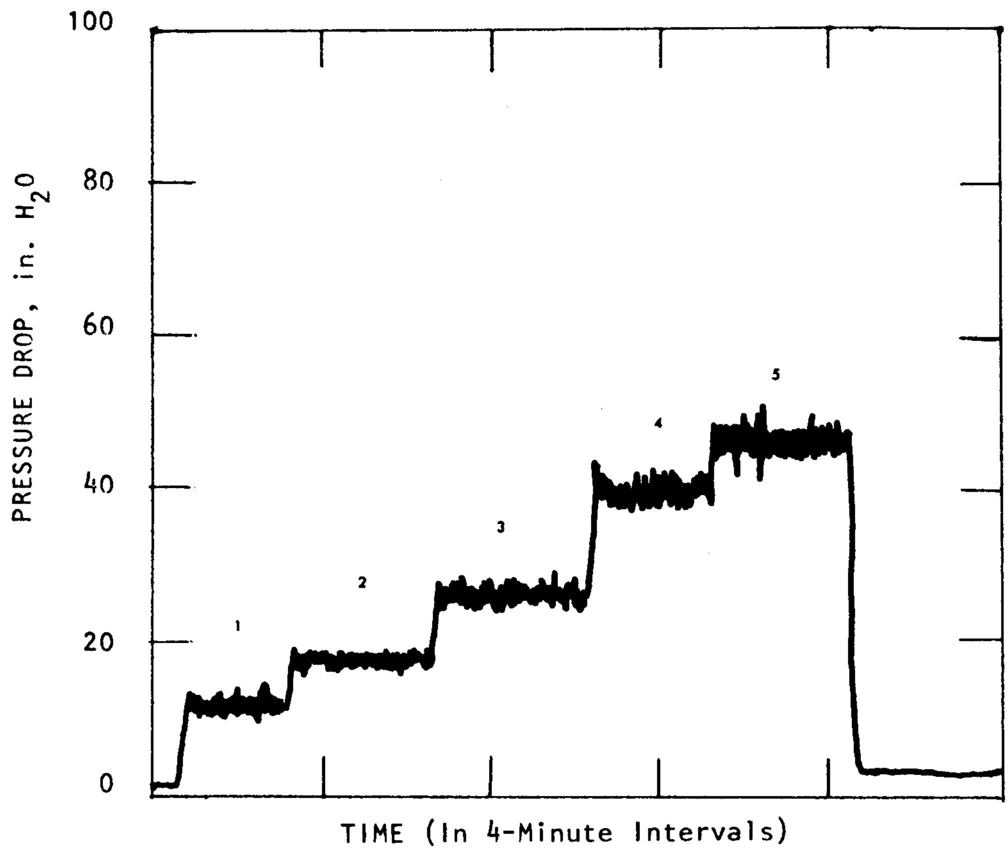
Lift-Line Velocity: 40 ft/s

Solids: −20+80 Mesh Ottawa Sand

Pot Fluidization Velocity: 0.4 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	0
2	450
3	1350
4	2250
5	3700
6	5050
7	9250

Figure 22. LIFT-LINE PRESSURE DROPS DURING RUN HGD-11BB



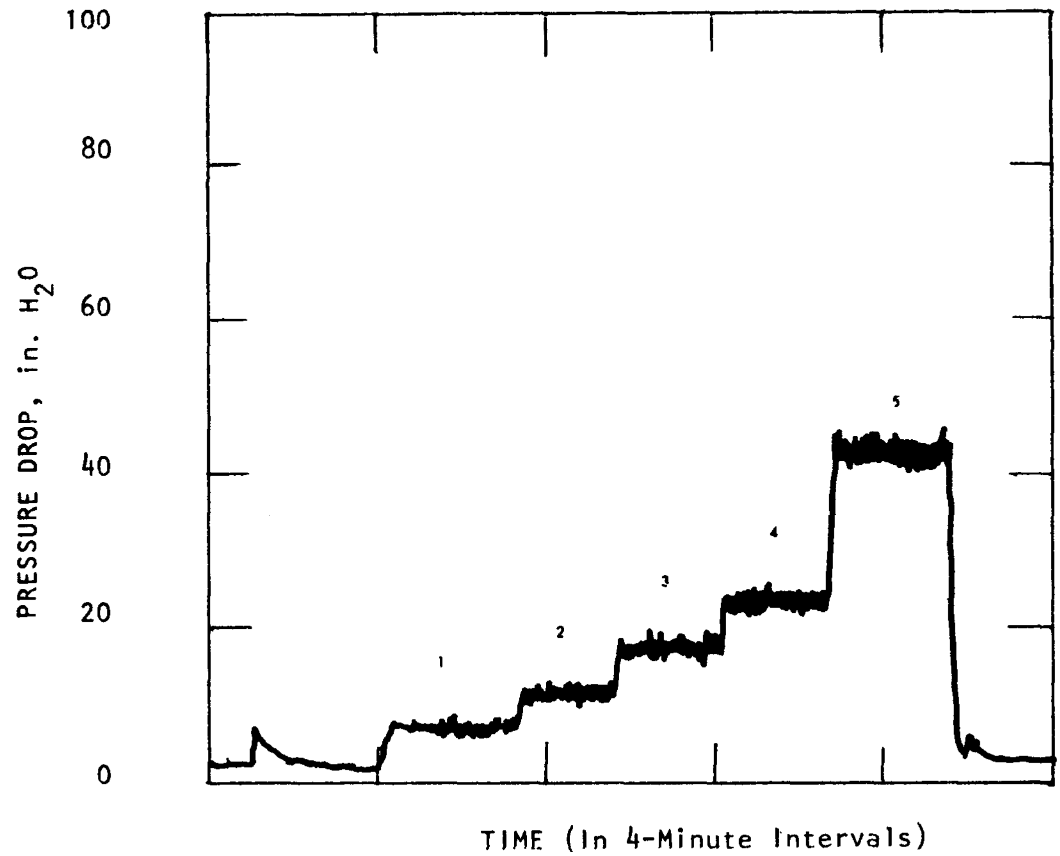
Lift-Line Velocity: 35 ft/s

Solids: −20+80 Mesh Ottawa Sand

Pot Fluidization Velocity: 0.4 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	1400
2	2400
3	4250
4	7000
5	8400

Figure 23. LIFT-LINE PRESSURE DROPS DURING RUN HGD-11CC



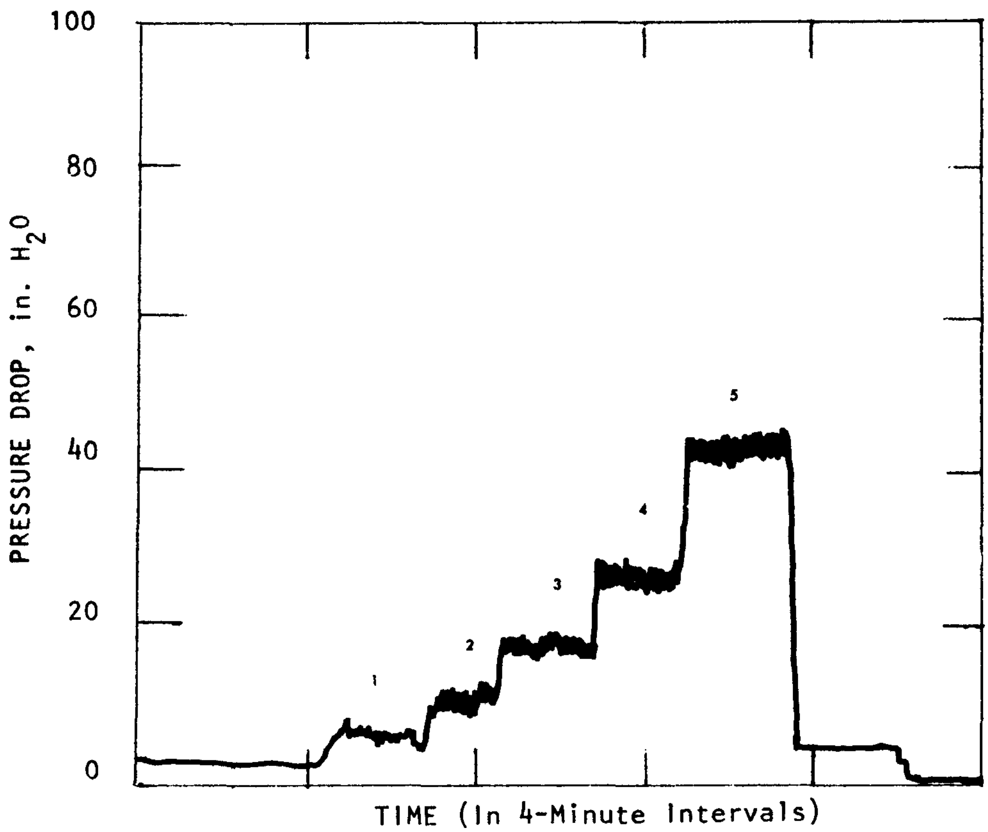
Lift-Line Velocity: 45 ft/s

Solids: −20+80 Mesh Ottawa Sand

Pot Fluidization Velocity: 0.4 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	920
2	1485
3	2765
4	3985
5	7970

Figure 24. LIFT-LINE PRESSURE DROPS DURING RUN HGD-11DD



Lift-Line Velocity: 50 ft/s

Solids: −20+80 Mesh Ottawa Sand

Pot Fluidization Velocity: 0.4 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	640
2	1325
3	2550
4	4145
5	7175

Figure 25. LIFT-LINE PRESSURE DROPS DURING RUN HGD-11EE

were 7970 and 7175 lb/hr, respectively. The fluctuations in the lift-line pressure drop were reduced, however, as the lift-line velocity was increased.

As with the other configurations tested in this study, the lift-pot III device was also operated with —20+200 mesh coal. Five different runs were made.

In Run HGD-12A (Figure 26), the lift-line velocity was set at 30 ft/s, with a lift-pot velocity of 0.3 ft/s. The maximum solids flow rate obtainable in this run was 1970 lb/hr.

Runs HGD-12B (Figure 27) and HGD-12C (Figure 28) were both made at a lift-line velocity of 30 ft/s and with lift-pot fluidization velocities of 0.245 and 0.182 ft/s, respectively. The maximum solids flow rates obtainable in these runs were 800 and 580 lb/hr, respectively.

In Run HGD-12D (Figure 29) the lift-pot velocity was set at 0.245 ft/s, and the lift-line velocity was increased to 35 ft/s. The maximum solids flow rate obtainable was 630 lb/hr.

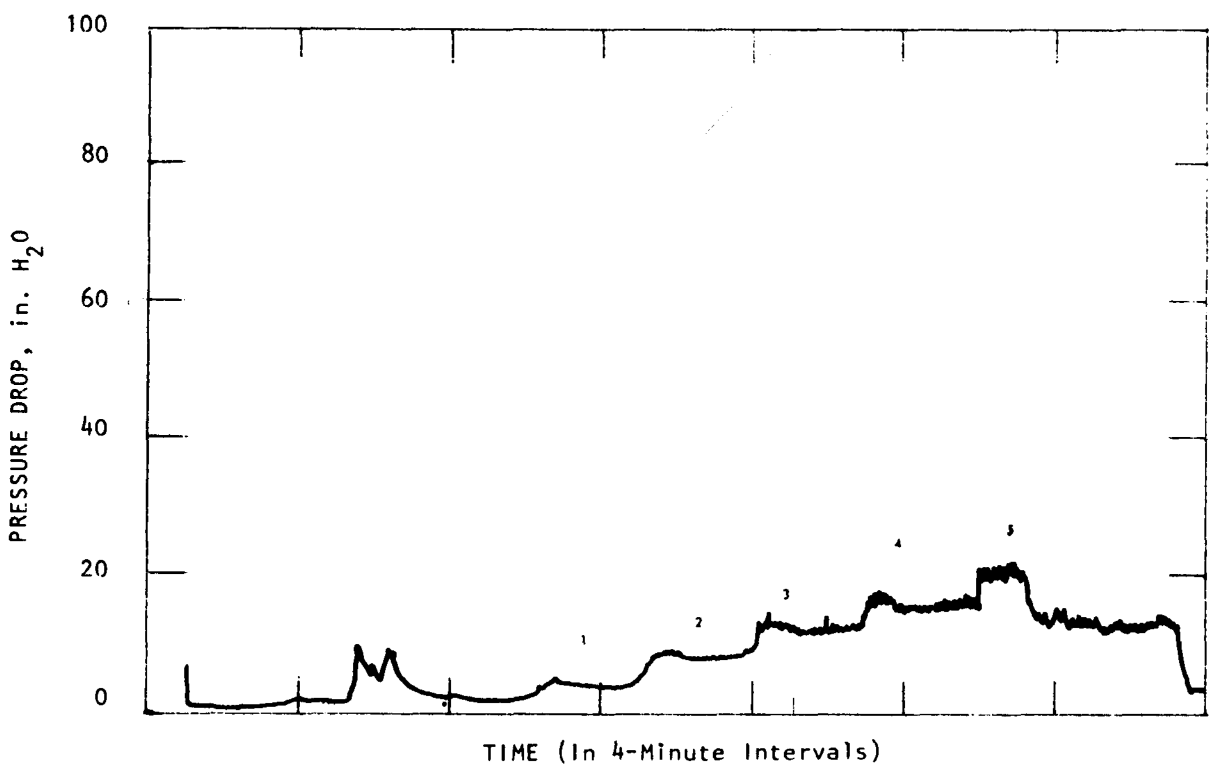
In the final test, Run HGD-12E (Figure 30), the lift-line velocity was set at 40 ft/s and the lift-pot velocity at 0.245 ft/s. The maximum flow rate obtained in this run was 780 lb/hr.

These runs show that the correct lift-pot velocity is important to the successful operation of these devices. Good operation also depends upon lift-line velocity.

Reverse-Seal Lift Pot

A reverse-seal lift-pot configuration was tested as another possible lift-line feeder configuration for the LTR section of the hydrogasifier in the HYGAS demonstration plant. It is important that the solids be injected into the lean-phase lift line smoothly and controllably to prevent slugging and poor conversion in the LTR. Fluctuations in the recorder tracings of the lean-phase lift-line pressure drop were again used to analyze the smoothness of the lift-line operation in this test.

The intial configuration of the reverse-seal lift pot is shown in Figure 31. During operation, the solids passed down a 2-inch-diameter downcomer into a 3-inch-diameter open seal well, which was part of the lift-pot fluidized bed. The lift-pot bed was fluidized using air introduced through



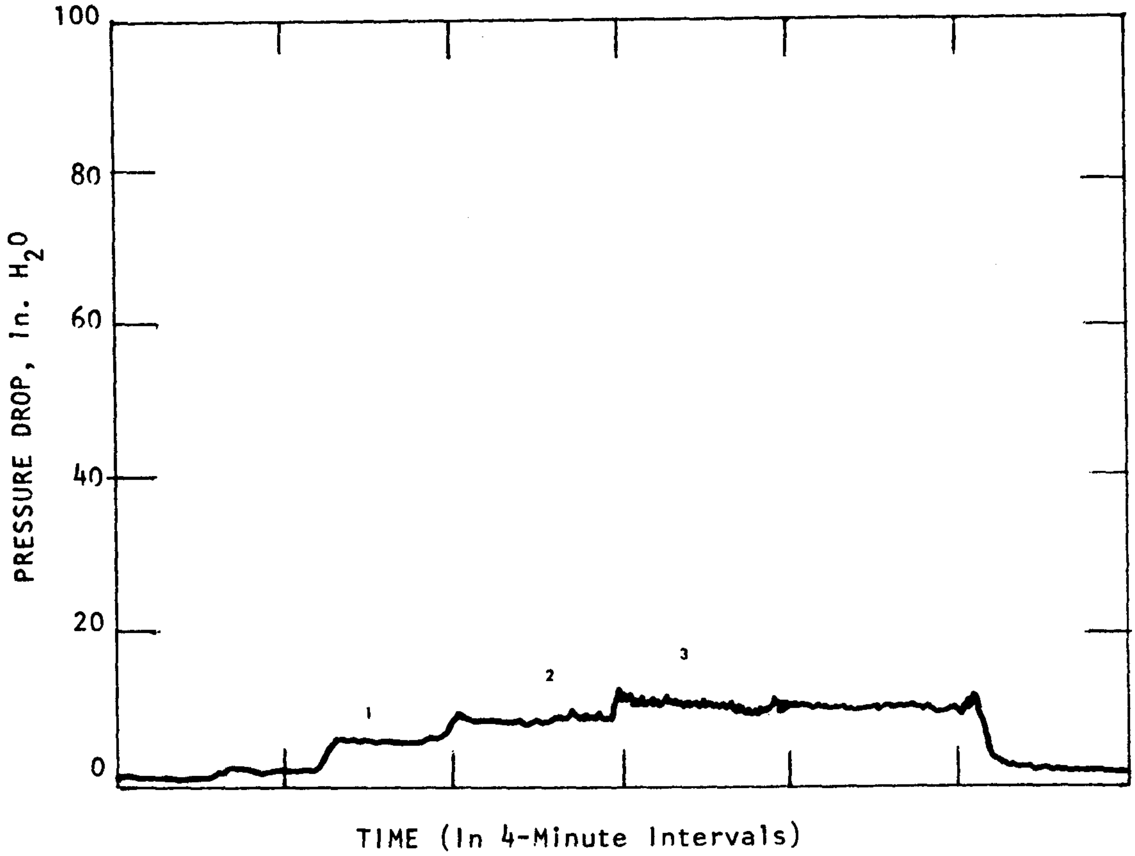
Lift-Line Velocity: 30 ft/s

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

Pot Fluidization Velocity: 0.3 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	100
2	570
3	1040
4	1740
5	1970

Figure 26. LIFT-LINE PRESSURE DROPS DURING RUN HGD-12A



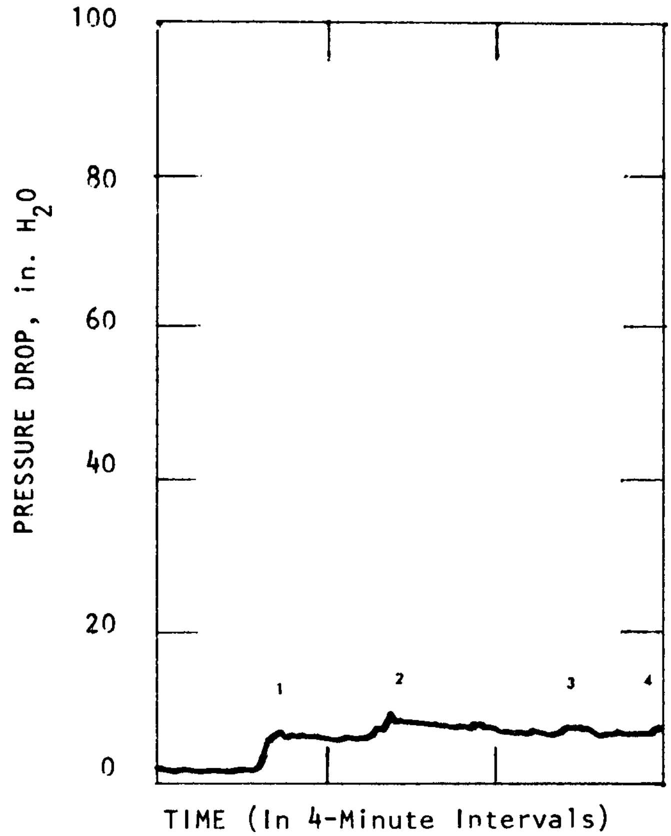
Lift-Line Velocity: 30 ft/s

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

Pot Fluidization Velocity: 0.245 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	340
2	690
3	920
4	800

Figure 27. LIFT-LINE PRESSURE DROPS DURING RUN HGD-12B



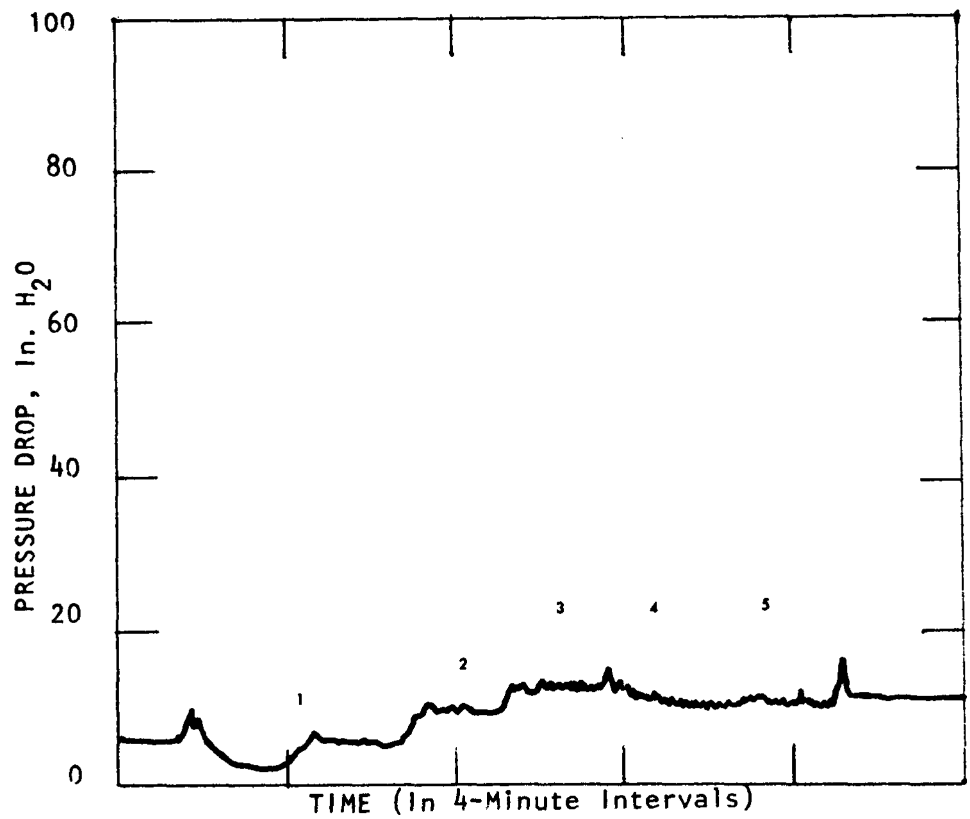
Lift-Line Velocity: 30 ft/s

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

Pot Fluidization Velocity: 0.182 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	340
2	580
3	460
4	580

Figure 28. LIFT-LINE PRESSURE DROPS DURING RUN HGD-12C



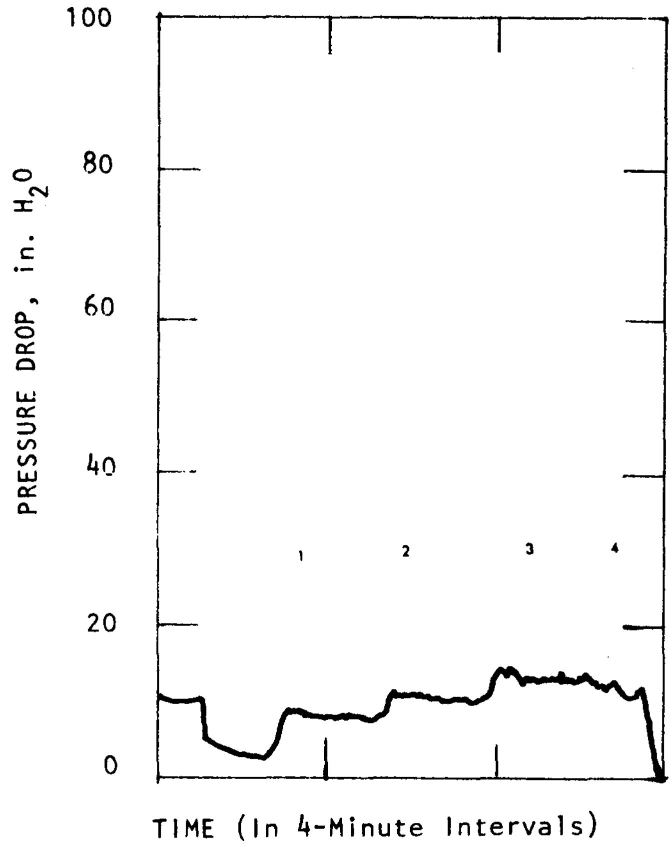
Lift-Line Velocity: 35 ft/s

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

Pot Fluidization Velocity: 0.245 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	280
2	630
3	800
4	720
5	630

Figure 29. LIFT-LINE PRESSURE DROPS DURING RUN HGD-12D



Lift-Line Velocity: 40 ft/s

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

Pot Fluidization Velocity: 0.245 ft/s

Reading No.	Solids Flow Rate, lb/hr
1	390
2	550
3	780
4	dropped

Figure 30. LIFT-LINE PRESSURE DROPS DURING RUN HGD-12E

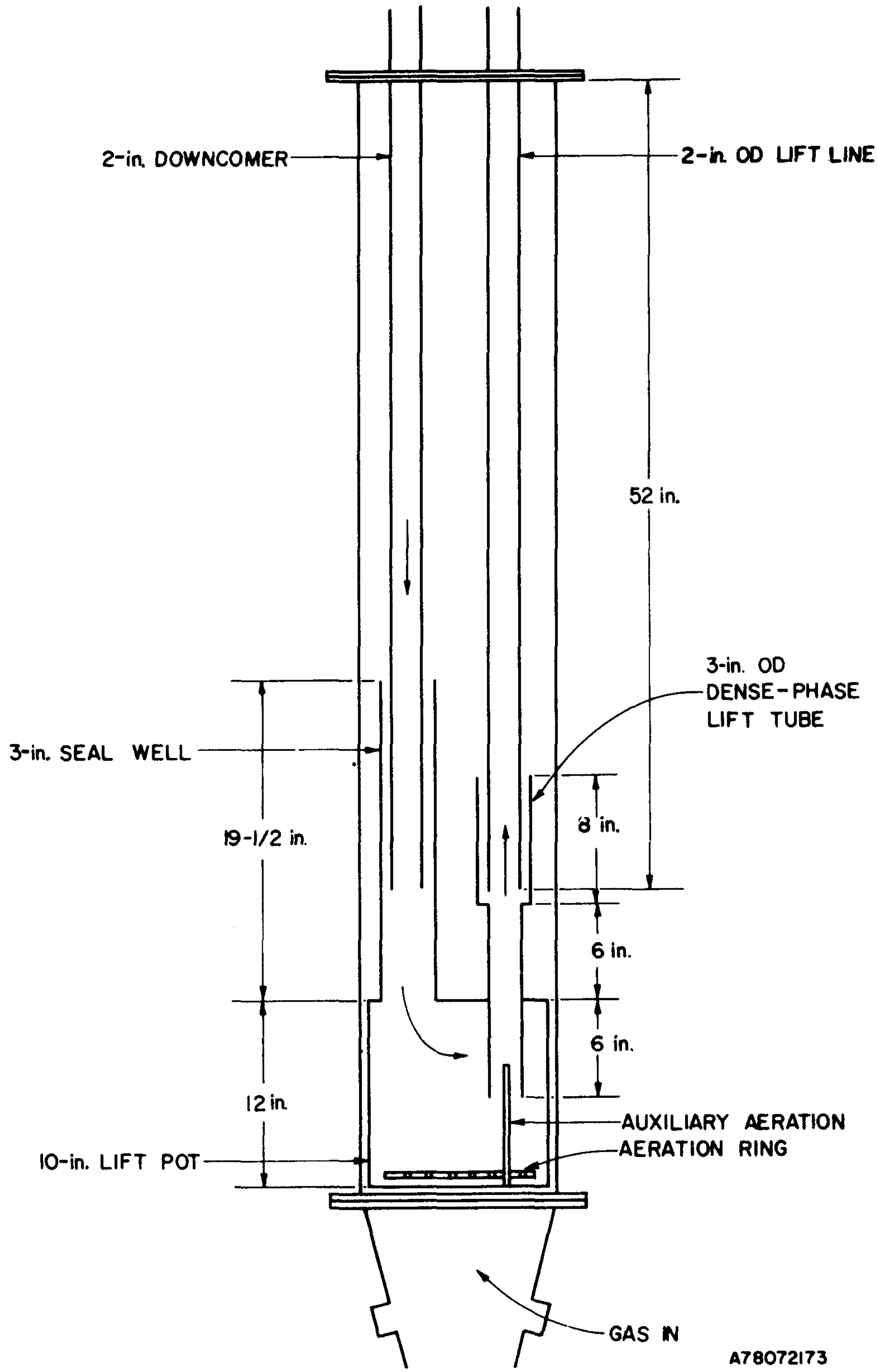


Figure 31. REVERSE-SEAL LIFT POT

a fluidization ring in the bottom of the pot. The solids flow rate was controlled by a ball valve located in the lower part of the downcomer.

The solids were routed into the 2-inch-diameter lift line through a short 2-inch-diameter dense-phase lift section, Figure 31. The entire reverse-seal lift pot was constructed of Plexiglas and clear polyvinyl chloride so that the solids flow through the device could be visually monitored.

In a typical run with this device, the desired fluidization velocity in the lift pot is set first, and then the lift-line gas flow rate is 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) on the clear polyvinyl chloride downcomer.

The operation of the reverse-seal pot using the configuration shown in Figure 31 was clearly unsatisfactory. At all but the lowest solid flow rates, slugs (approximately 1 foot in length) formed in the lowest section of the lift line. The reverse-seal pot fluidizing gas would then bypass the dense-phase lift section and pass up through the open seal well, blowing solids out of it in the process. The pressure in the vessel enclosing the device would also rise as each slug formed.

The device was modified by increasing the height of the seal well and enlarging the diameter of the dense-phase lift section, Figure 32. This configuration operated much better.

This modified reverse-seal pot was operated using −20+80 mesh Ottawa sand and −20+200 mesh pretreated Illinois No. 6 bituminous coal. Five tests were conducted with the sand. In Run HGD-13A (Figure 33), the lift-line velocity was set at 35 ft/s and the reverse-seal pot fluidization velocity at 0.32 foot of air per second, using gas as the fluidizing medium. In addition, 0.7 actual cubic foot of nitrogen per minute was added to the dense-phase lift section. Solids circulation could be controlled, but at high solids flow rates, the pressure-drop fluctuations in the lift line were excessively high (±9 to 10 inches of water). The maximum solids flow rate obtainable under these conditions was 8750 lb/hr.

Runs HGD-13B (Figure 34) and HGD-13C (Figure 35) lift-line velocities were set at 40 and 45 ft/s, respectively. Aeration in the lift pot and in the

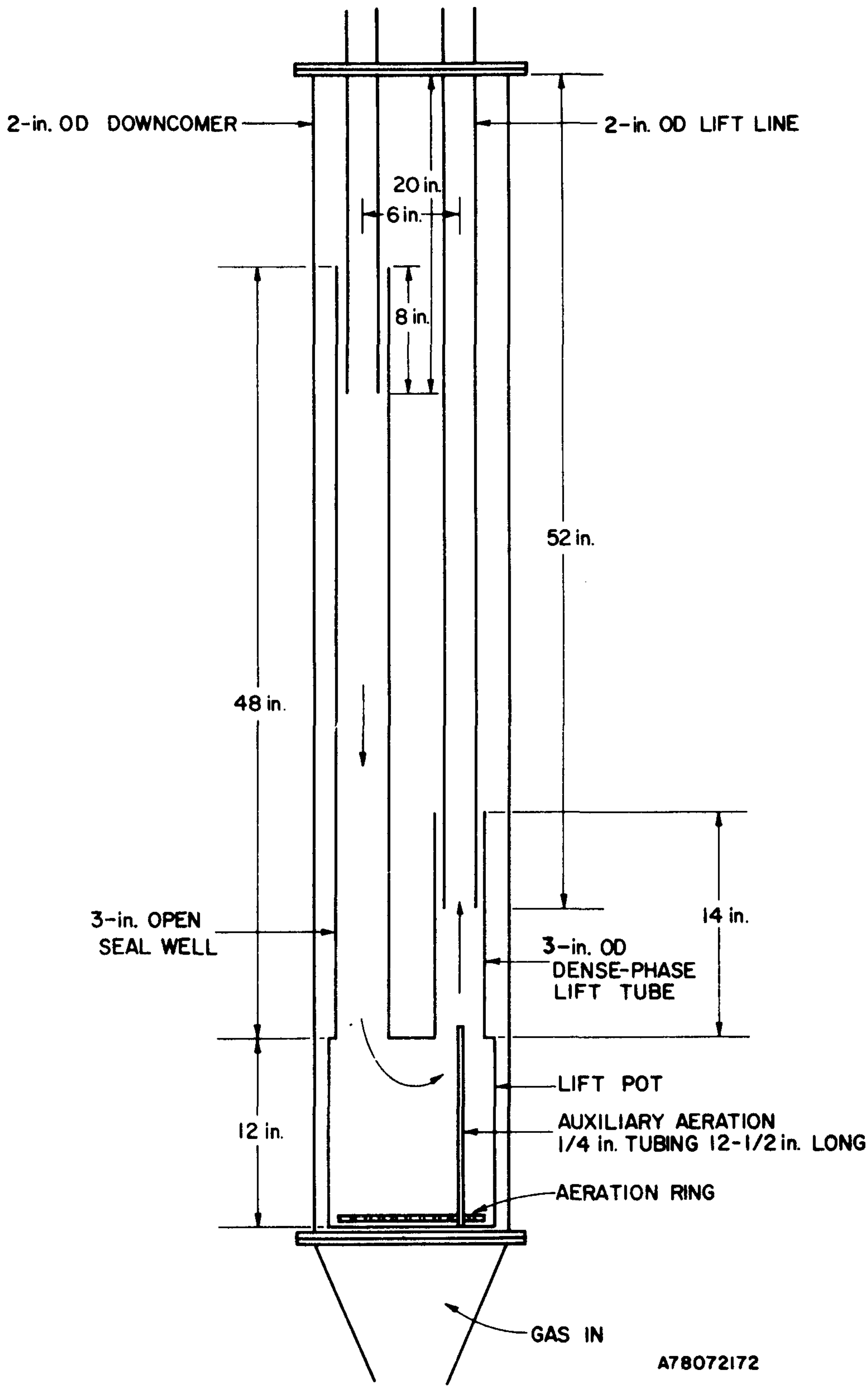
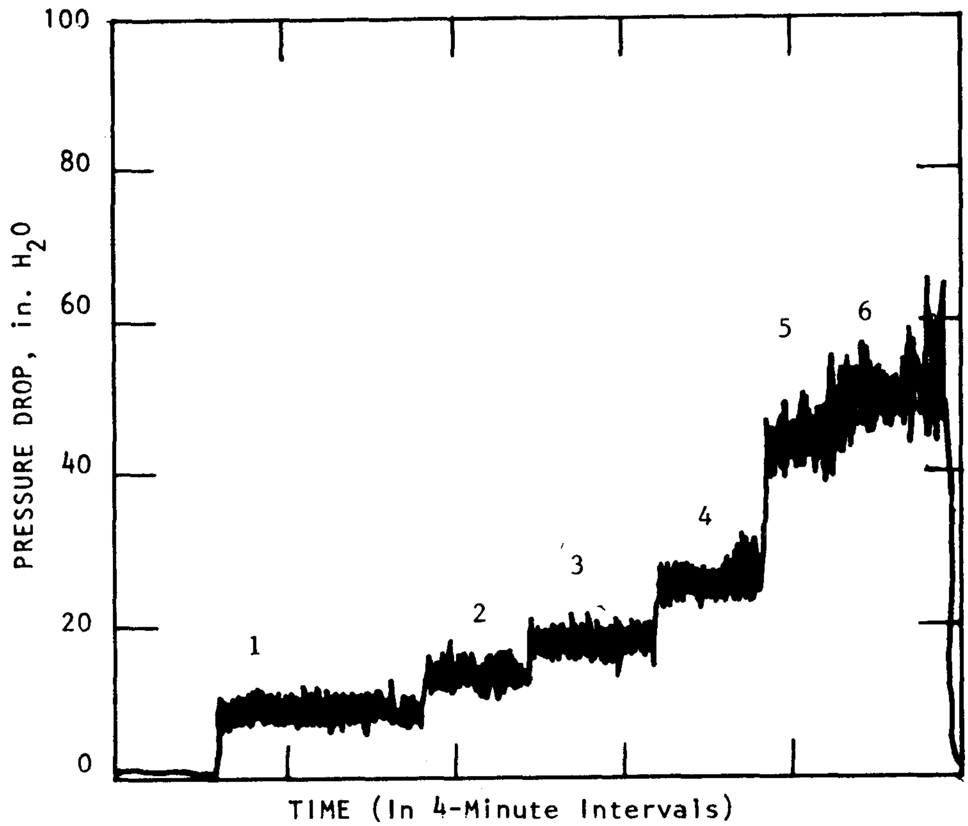


Figure 32. MODIFIED REVERSE-SEAL LIFT POT



Reading No.	Solids Flow Rate, lb/hr
1	1320
2	2170
3	2870
4	4475
5	7800
6	8750

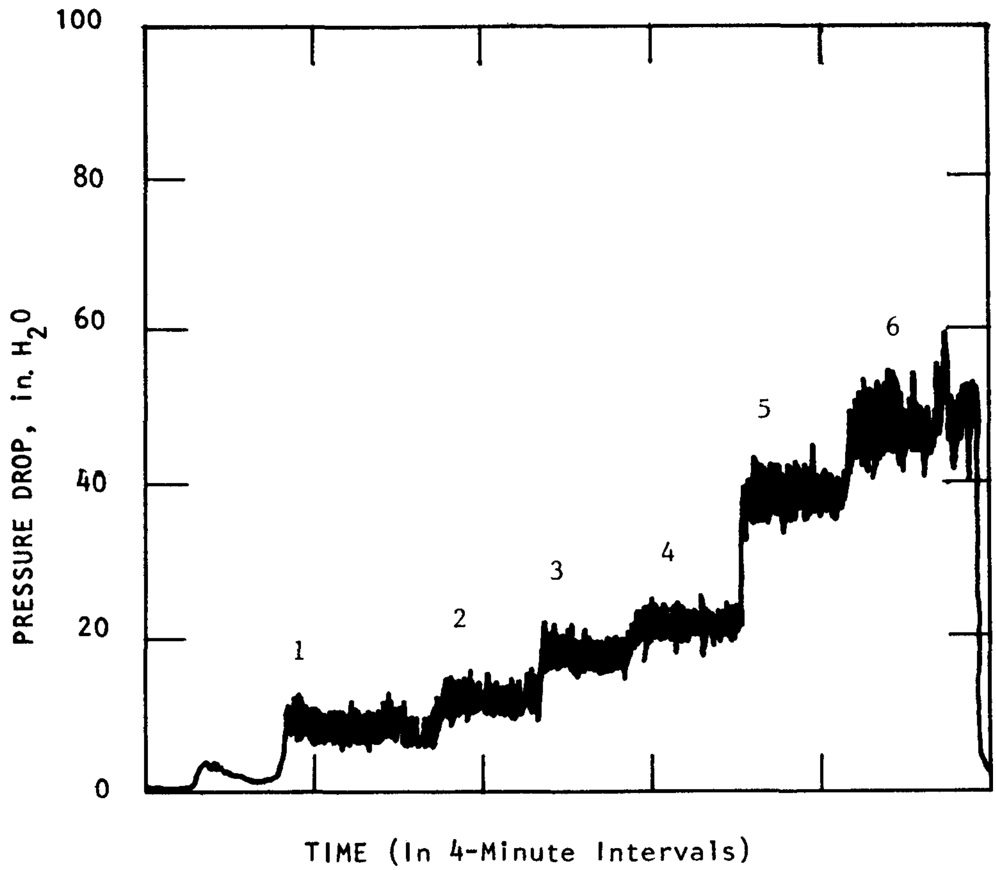
Lift-Line Velocity: 35 ft/s

Solids: –20+80 Mesh Ottawa Sand

Aeration: 0.7028 ACFM

Pot Fluidization Velocity: 0.32 ft/s

Figure 33. LIFT-LINE PRESSURE DROPS DURING RUN HGD-13A



Reading No.	Solids Flow Rate, lb/hr
1	1095
2	1835
3	3105
4	3620
5	7000
6	8550

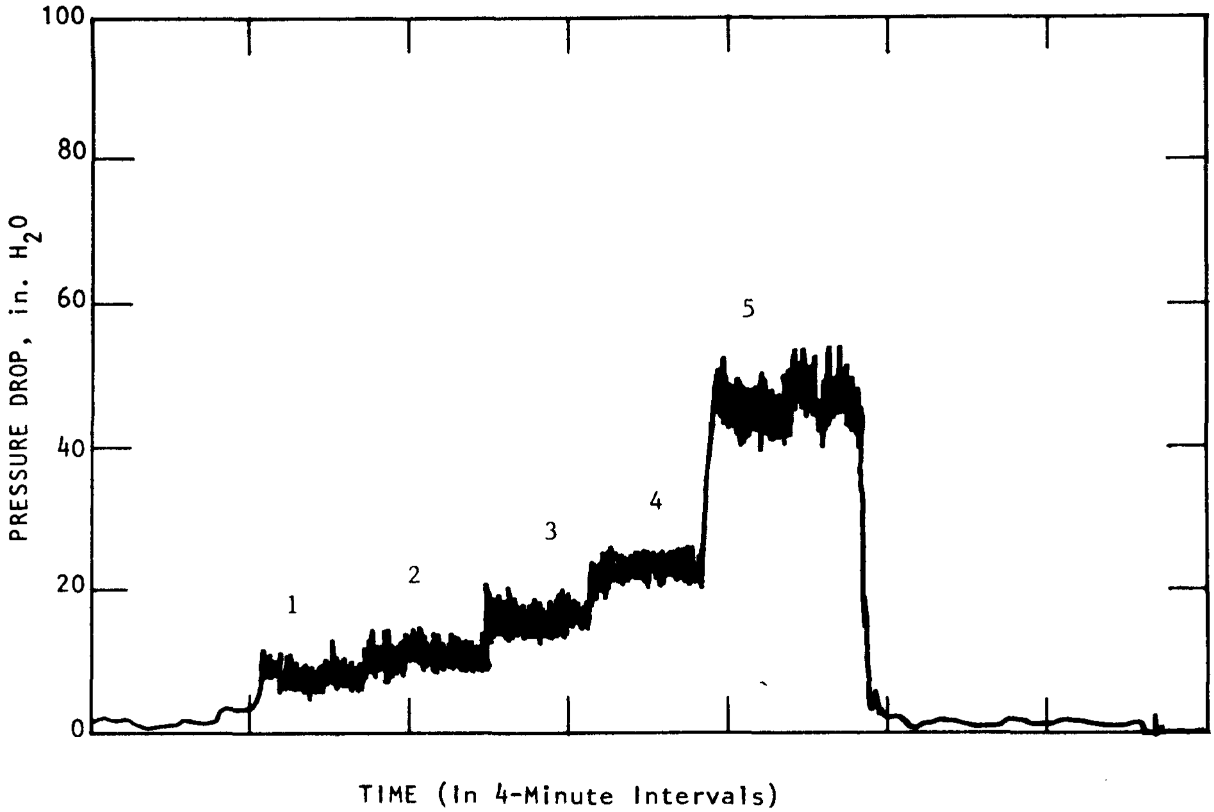
Lift-Line Velocity: 40 ft/s

Solids: —20+80 Mesh Ottaw Sand

Aeration: 0.7028 ACFM

Pot Fluidization Velocity: 0.32 ft/s

Figure 34. LIFT-LINE PRESSURE DROPS DURING RUN HGD-13B



Reading No.	Solids Flow Rate, lb/hr
1	885
2	1395
3	2375
4	3990
5	8000

Lift-Line Velocity: 45 ft/s

Solids: −20+80 Mesh Ottawa Sand

Aeration: 0.7028 ACFM

Pot Fluidization Velocity: 0.32 ft/s

Figure 35. LIFT-LINE PRESSURE DROPS DURING RUNS HGD-13C

dense-phase lift was the same as in Run HGD-13A. Maximum solids flow rates were 8550 lb/hr in Run HGD-13B and 8000 lb/hr in Run HGD-13C. Lift-line pressure-drop fluctuations were about ±8 inches of water in both runs.

Run HGD-13D (Figure 36) was made to determine the effect of pot fluidization velocity on the operation of the reverse-seal pot. In this run, the lift-line velocity was set at 35 ft/s, and the dense-phase lift aeration was set at 0.7 actual ft^3/min. The pot fluidization velocity was increased to 0.5 ft/s. The maximum solids flow rate was 8950 lb/hr. However, fluctuations in the lift-line pressure drop were much too high.

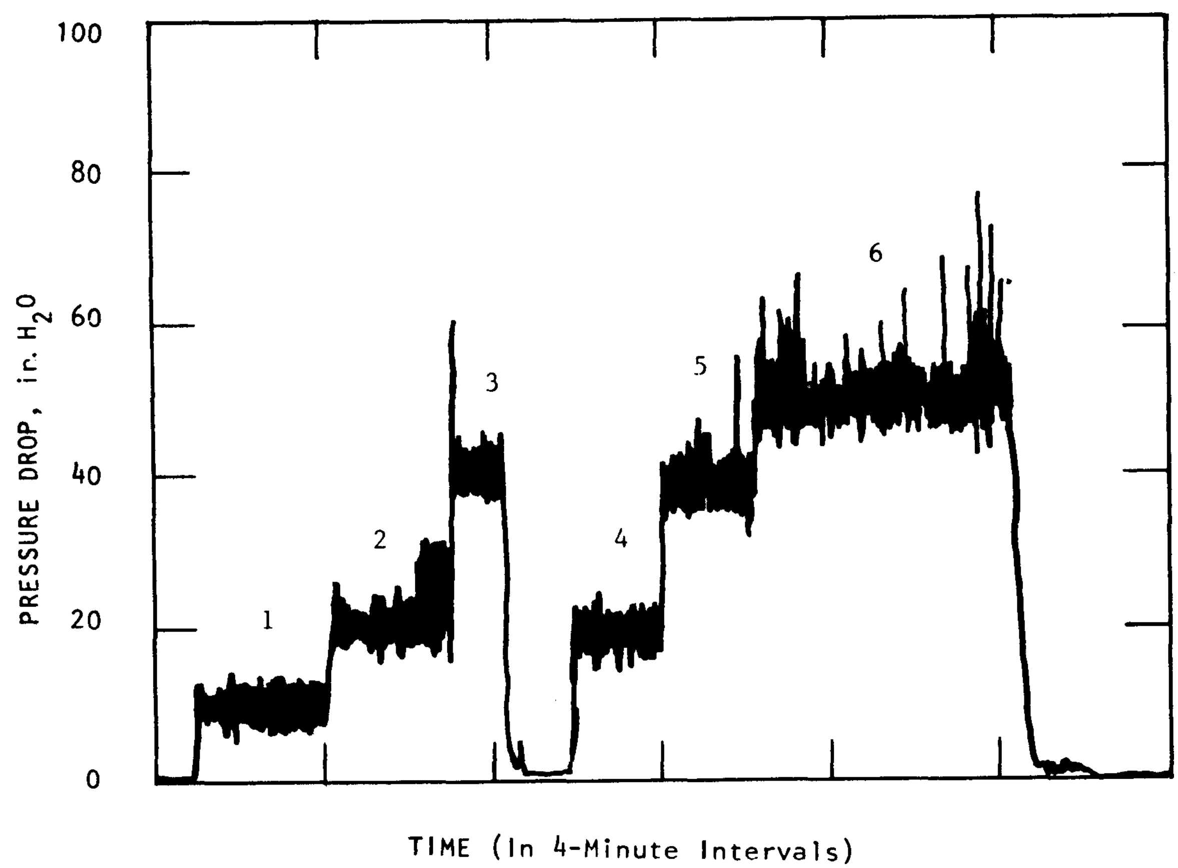
In Run HGD-13E (Figure 37), the lift-line and pot fluidization velocities were set at 35 ft/s and 0.32 ft/s, respectively. Aeration in the dense-phase lift section was increased from 0.7 to 2.2 actual ft^3/min. Under these conditions, operation was rough, and circulation was stopped at 6800 lb/hr when the solids continued to rise up the open seal well.

From these tests, it was evident that extra dense-phase lift aeration was detrimental to the operation of the reverse sea . Five tests were also made with the reverse seal pot using −20+200 mesh pretreated Illinois No. 6 bituminous coal. The testing procedure used with this material was the same as that used for the Ottawa sand. Lower lift-line and pot fluidization velocities were used with this material because it is lighter than sand.

In the first three of these runs, HGD-14A (Figure 38), HGD-14B (Figure 39), and HGD-14C (Figure 40), the pot fluidization velocity was kept constant at 0.245 ft/s, while the lift-line velocities were set at 30, 35, and 45 ft/s, respectively. The aeration in the dense-phase lift section was also kept constant at 0.425 actual ft^3/min in these three runs. There was good control of the solids flow rate in these runs; however, lift-line pressure drop fluctuations were relatively high for coal. The maximum solids flow rates obtainable were 1230 (Run HGD-14A), 2150 (Run HGD-14B), and 2000 lb/hr (Run HGD-14C).

Run HGD-14D (Figure 41) was made with a reduced pot fluctuation velocity of 0.182 ft/s. The lift-line and dense-phase lift velocities were set at 35 and 0.425 ft/s, respectively. This lower pot fluidization did not significantly alter the operation of the device. The maximum solids flow rate in this run was 2000 lb/hr.

Run HGD-14E (Figure 42) was made at a lift-line velocity of 35 ft/s and a pot fluidization velocity of 0.182 ft/s. The dense-phase lift aeration



Reading No.	Solids Flow Rate, lb/hr
1	1450
2	3500
3	7250
4	3150
5	6900
6	8950

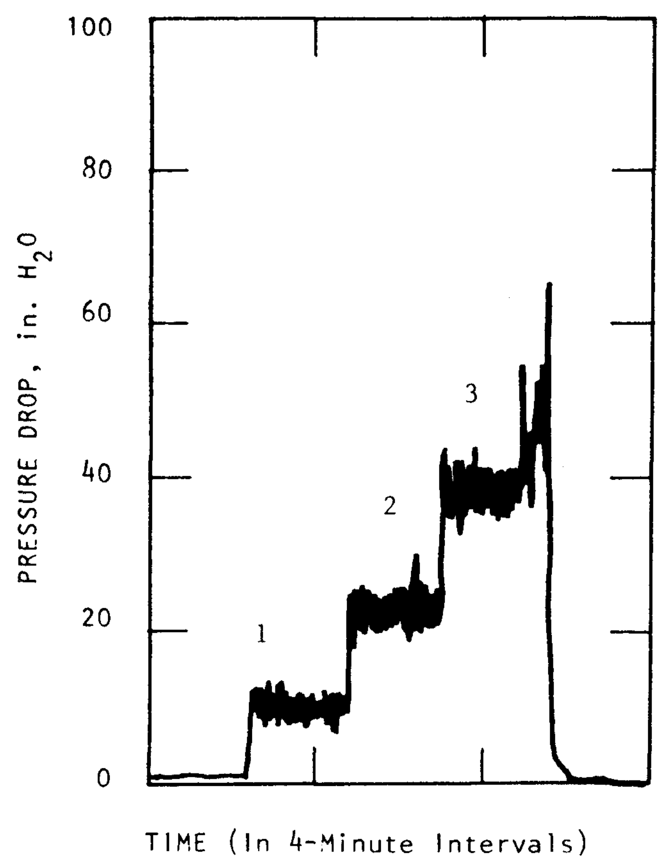
Lift-Line Velocity: 35 ft/s

Solids: –20+80 Mesh Ottawa Sand

Aeration: 0.7028 ACFM

Pot Fluidization Velocity: 0.5 ft/s

Figure 36. LIFT-LINE PRESSURE DROPS DURING RUN HGD-13D



Reading No.	Solids Flow Rate, lb/hr
1	1450
2	3900
3	6800

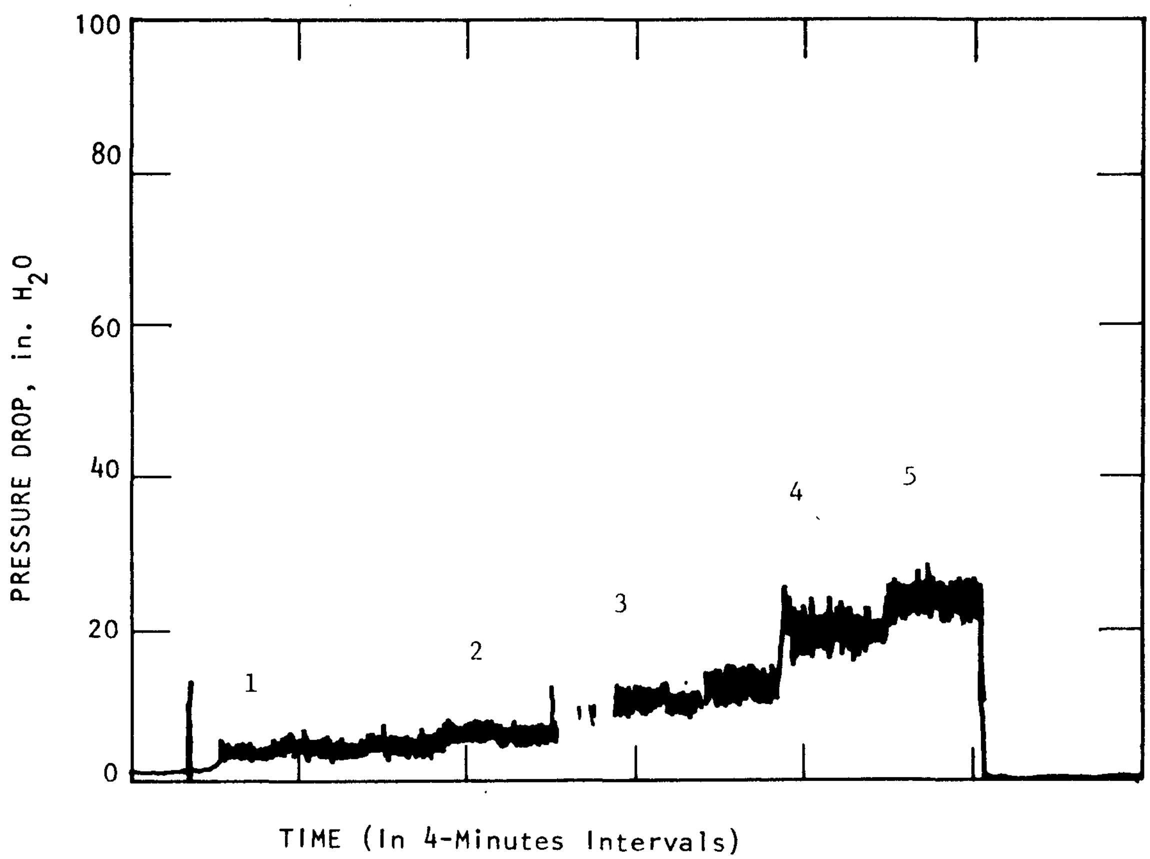
Lift-Line Velocity: 35 ft/s

Solids: −20+80 Mesh Ottawa Sand

Aeration: 2.2138 ACFM

Pot Fluidization Velocity: 0.32 ft/s

Figure 37. LIFT-LINE PRESSURE DROPS DURING RUN HGD-13E



Reading No.	Solids Flow Rate, lb/hr
1	232
2	400
3	732
4	1066
5	1230

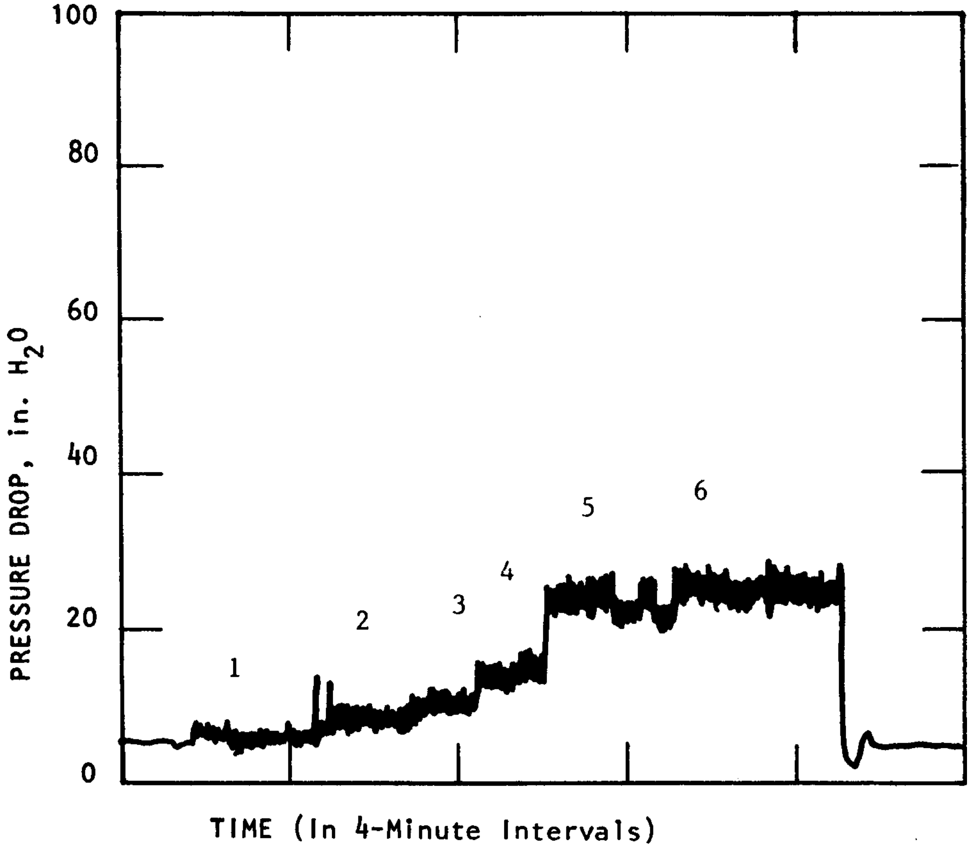
Lift-Line Velocity: 30 ft/sec

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

Aeration: 0.425 ACFM

Pot Fluidization Velocity: 0.245 ft/s

Figure 38. LIFT-LINE PRESSURE DROPS DURING RUN HGD-14A



Reading No.	Solids Flow Rate, lb/hr
1	216
2	460
3	588
4	1050
5	2000
6	2150

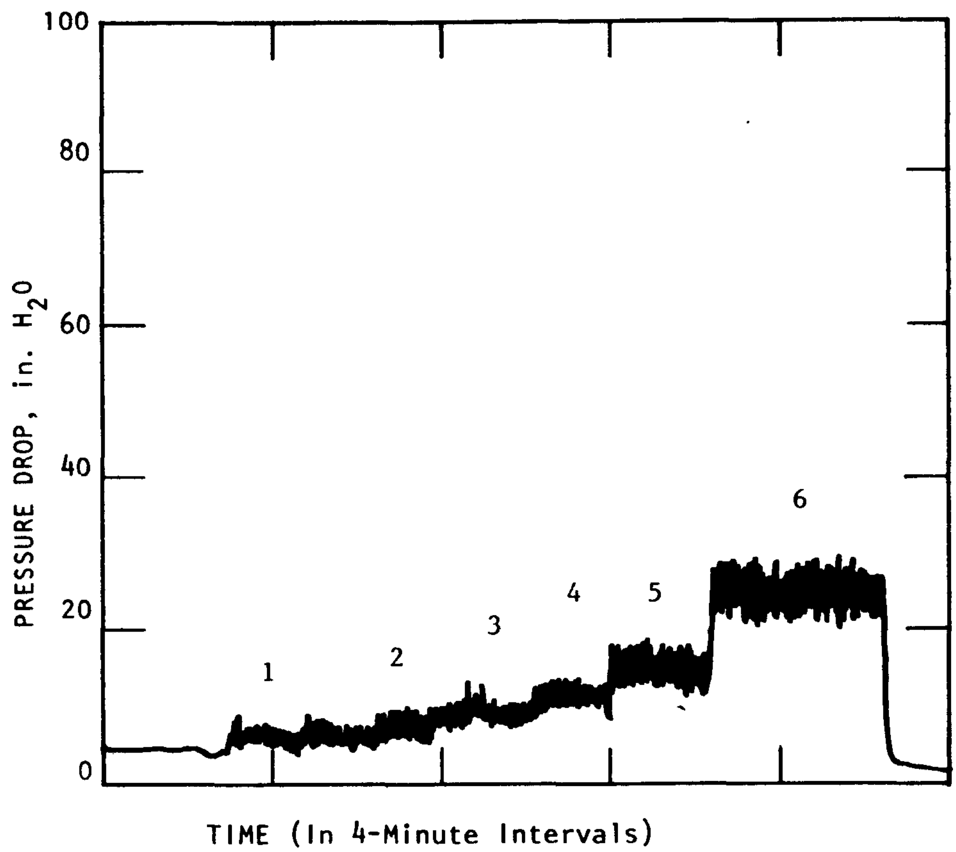
Lift-Line Velocity: 35 ft/sec

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

Aeration: 0.425 ACFM

Pot Fluidization Velocity: 0.245 ft/sec

Figure 39. LIFT-LINE PRESSURE DROPS DURING RUN HGD-14B



Reading No.	Solids Flow Rate, lb/hr
1	260
2	388
3	507
4	722
5	1050
6	2000

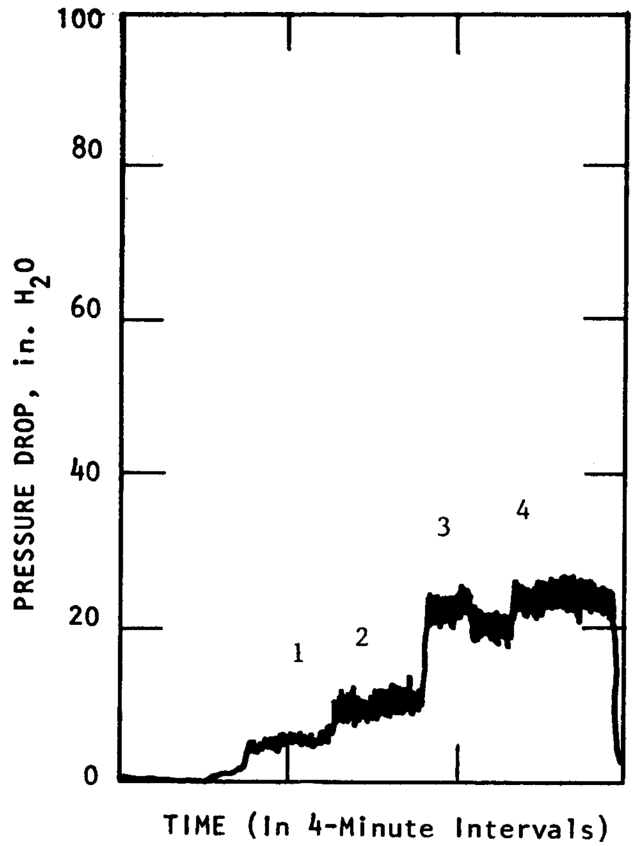
Lift-Line Velocity: 40 ft/sec

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

Aeration: 0.425 ACFM

Pot Fluidization Velocity: 0.245 ft/s

Figure 40. LIFT-LINE PRESSURE DROPS DURING RUN HGD-14C



Reading No.	Solids Flow Rate, lb/hr
1	130
2	630
3	1770
4	2000

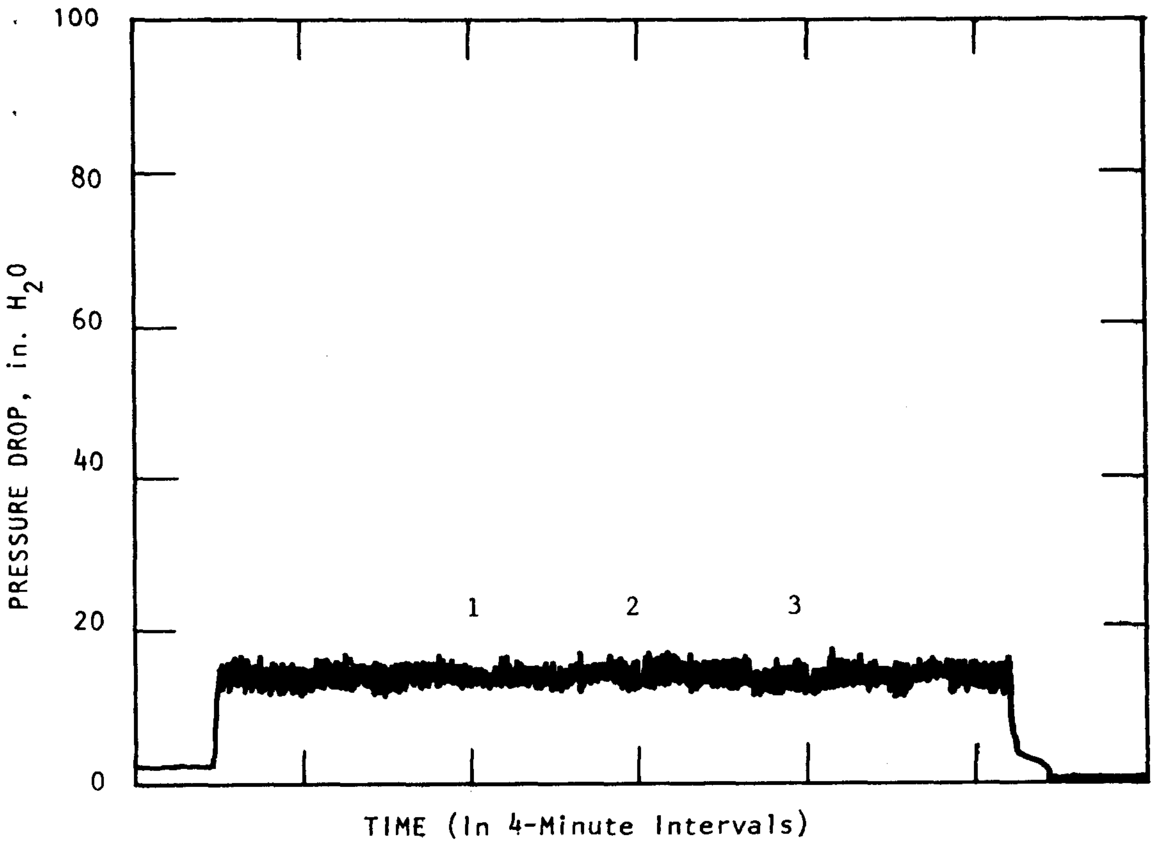
Lift-Line Velocity: 35 ft/s

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

Pot Fluidization Velocity: 0.182 ft/s

Aeration: 0.425 ACFM

Figure 41. LIFT-LINE PRESSURE DROPS DURING RUN HGD-14D



Reading No.	Solids Flow Rate, lb/hr	Aeration, ACF/min
1	1020 (not max.)	0.425
2	1020 (not max.)	1.701
3	1020 (not max.)	0

Lift-Line Velocity: 35 ft/s

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

Pot Fluidization Velocity: 0.182 ft/s

Figure 42. LIFT-LINE PRESSURE DROPS DURING RUN HGD-14E

velocity was varied from 0 to 1.7 actual ft^3/min. As in the Ottawa sand run, aeration in the dense-phase lift section was unnecessary as long as the pot was sufficiently fluidized to keep the solids moving.

This configuration operated no better than the other three lift pots tested. Maximum solids flow rates were higher than those with the L-valve, J-valve, and reverse seal, but the lift-line fluctuations were also significantly higher.

Summary For Lift-Line Feeder Configurations

Seven different possible feeder configurations were studied in the plastic model. These seven configurations are shown schematically in Figures 43 and 44. The seven configurations can be grouped into two general categories: lift-pot feeders (Figure 43) and nonmechanical feeders (Figure 44).

The lift-pot devices control solids flow by a mechanical valve in the downcomer. The solids pass through the control valve and fall onto the top of the fluidized bed. They are transferred into the fluidized bed and are then transported into the lift line by the lift gas. The primary difference between each lift-pot device is the method by which the lift gas is introduced into the lift line, as can be seen in Figure 43.

The other types of feeder devices tested were nonmechanical valve feeders. In these devices, the solids flow rate is not controlled by a mechanical valve, but by the amount of aeration gas fed to the valve. Thus, these devices have an advantage over the lift-pot devices in that they do not depend on a mechanical valve, which may seize or wear rapidly in the severe operating environment of coal gasifiers. (Note that the reverse seal tested did not require a mechanical valve. In the commercial concept, the valve controls the solids flow in this unit and, therefore, the device could also be classified as mechanically controlled.)

The operations of each device have been detailed in Project 9000 Quarterly Report No. 7 (DOE Report No. FE-2434-29) and in this report. All seven devices were tested using two solids: −20+80 mesh Ottawa sand and −20+200 mesh pretreated Illinois No. 6 bituminous coal.

Each device was analyzed by determining how smoothly it fed solids into the lift line. This "smoothness" was analyzed by looking at the magnitude of

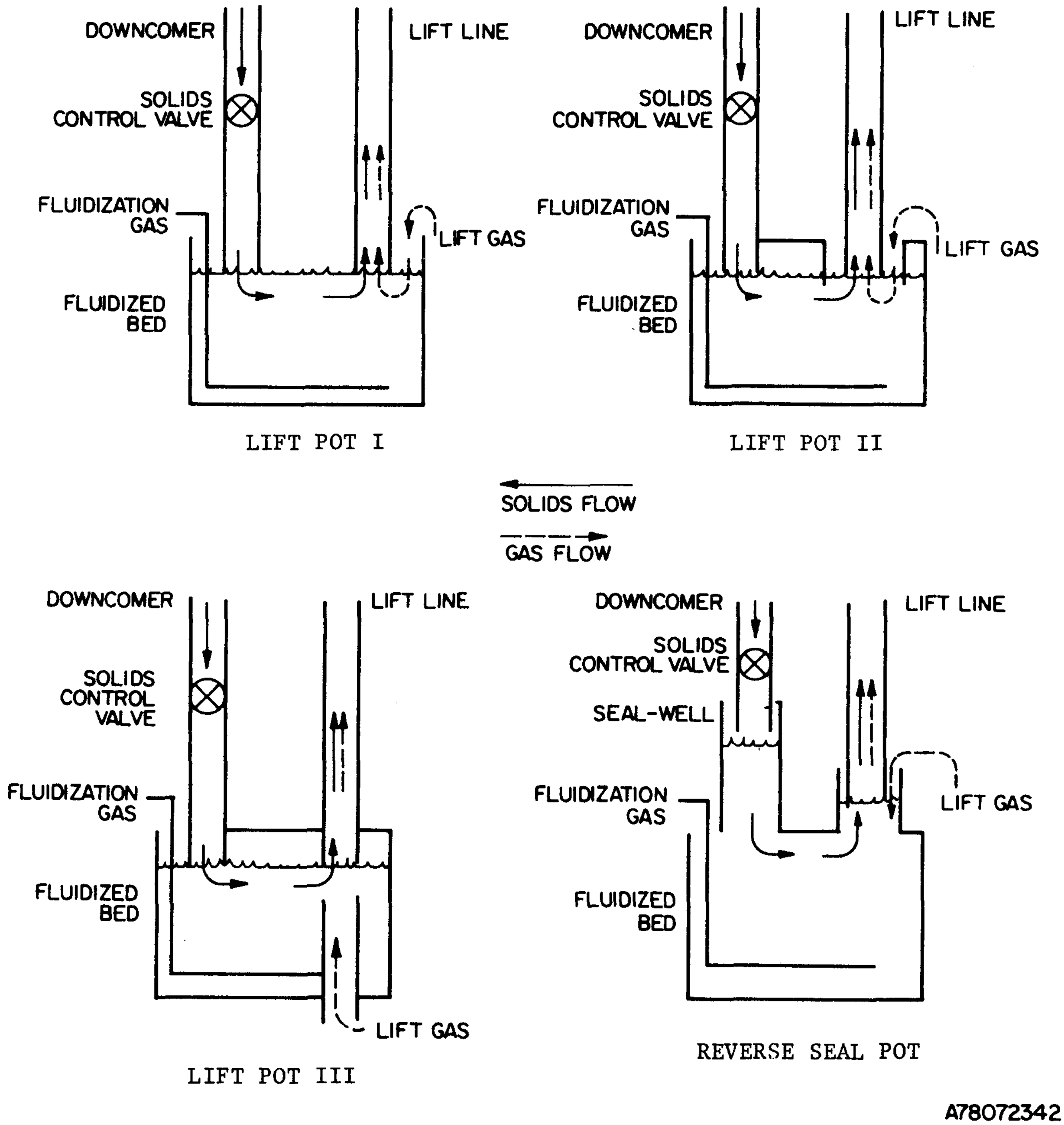


Figure 43. LIFT-POT LOW TEMPERATURE REACTOR FEEDER CONFIGURATIONS

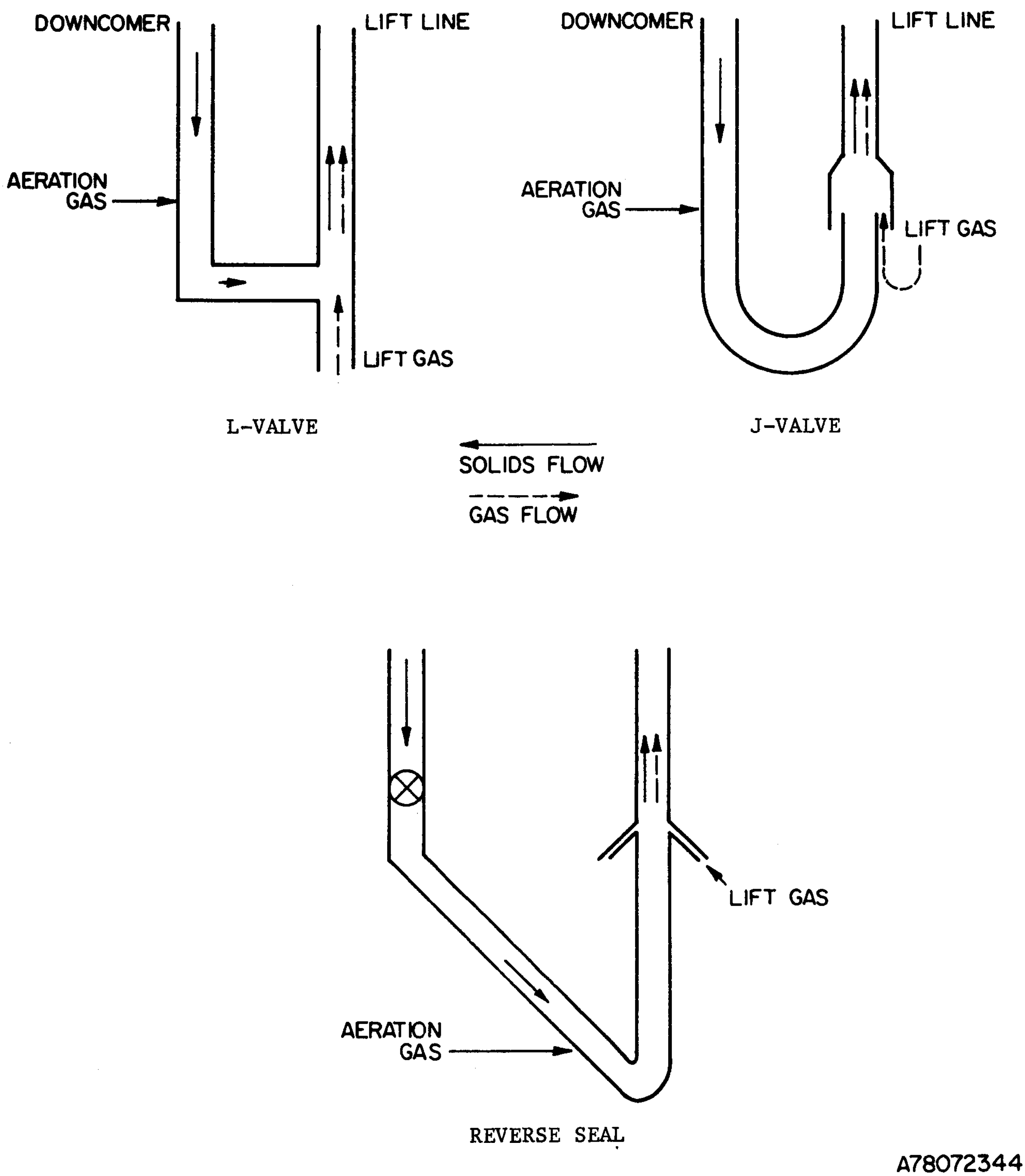


Figure 44. NONMECHANICAL LOW TEMPERATURE REACTOR FEEDER CONFIGURATIONS

the pressure-drop fluctuations in the lift-line pressure-drop trace. Each configuration was also checked for maximum solids flow rate and the amount of gas (aeration or fluidization gas) needed for it to operate well.

All of the configurations worked, i.e., they fed solids into the lift line. All of the configurations also gave approximately the same maximum solids flow rate: 9000 to 10,000 pounds of sand per hour and 1500 to 2000 pounds of coal per hour (for the small test unit). Among the lift-pot devices, the one which gave the worst performance was the reverse seal pot. This device fed solids into the lift line unevenly, resulting in high fluctuations in the lift-line pressure drop. Thus, this device was unacceptable.

The other three lift-pots fed solids reasonably well. The smoothest lift-line operation was obtained when the pot fluidization velocity was at its lowest possible value, and the lift-line velocity at a relatively high value (i.e., away from the choking regime). It was, however, necessary to keep the pot velocity well above the minimum fluidization velocity. This was because the solids flow rate was dependent upon pot fluidization velocity, and at low pot fluidization velocities, the solids flow rates were significantly limited.

Maximum values for the lift-line pressure-drop fluctuations for the lift-pot devices were approximately 3 inches of water from the average pressure-drop reading at the maximum solids flow rate for coal.

One advantage of the lift-pot I and II configurations is that no expansion joints are required for these devices. The downcomer and lift line are not attached to the lift pot, and can "grow" as the temperature is changed from ambient to the operating level.

The nonmechanical valve feeders also worked well. These feeders were simpler in design than the lift-pot feeders. Also, they did not require a mechanical valve to control the solids flow rate (although the reverse seal configuration could be used with the mechanical control valve, as in the commercial concept).

As the aeration to these devices was increased, the solids flow rate through them increased. The amount of aeration needed to produce a particular solids flow rate varied for the three nonmechanical valves (Figure 45). The most efficient device was the L-valve, which was just slightly more efficient than the J-valve. The reverse seal required much more aeration gas

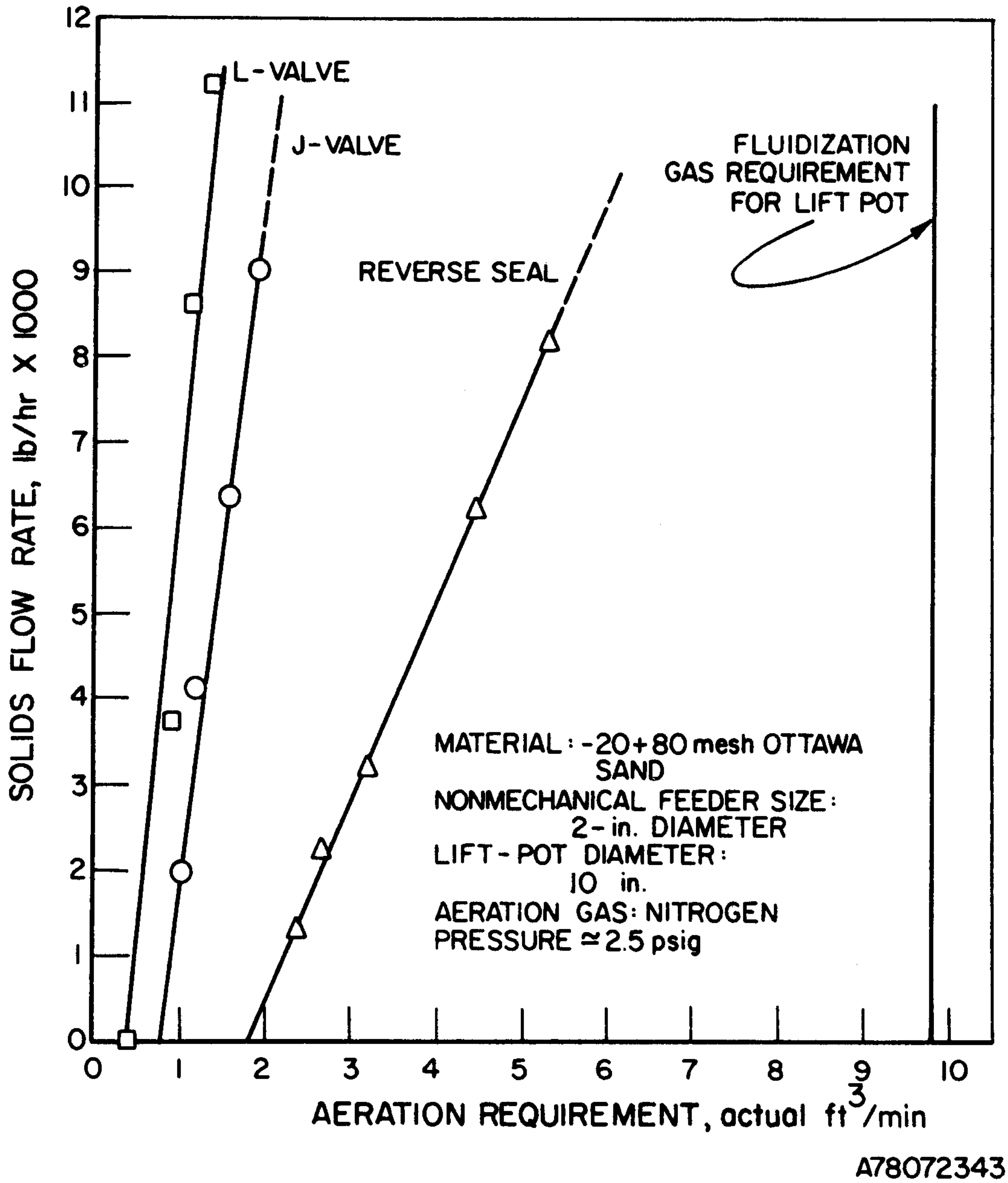


Figure 45. AERATION REQUIREMENTS FOR NONMECHANICAL VALVE FEEDERS

than the L- and J-valves to produce the same solids flow rate. (Note that in the configurations tested, the major difference between the J-valve and the reverse seal is the length of the return section. With the longer return length, the reverse-seal design requires more gas and greater pressure drop.) Also shown on the plot is the aeration that is required to fluidize the lift pot for good operation. This gas requirement was much more than that needed for nonmechanical valves.

The nonmechanical valves fed the lift line more smoothly than the lift-pot devices. For example, maximum lift-line pressure-drop fluctuations for the nonmechnaical valve feeders were approximately ±3 inches of water from the average lift-line pressure-drop reading for coal.
from the average lift-line pressure-drop reading for coal.

In conclusion, the results of the low-pressure, small-scale LTR feeder study indicate that the nonmechanical valve feeders:

- Gave smoother lift-line operation
- Needed less aeration gas to operate
- Were simpler to construct and operate
- Were better lift-line feeders than the lift-pot feeders.

Of the three nonmechanical valve feeders tested, the L- and J-valves required less aeration than the reverse seal. The order of reverse sealing ability was reverse seal > J-valve > L-valve; the value of this ability remains debatable.

Thus, based on the small-scale model tests, the best LTR feeder configurations are the nonmechanical L-valve, J-valve, and reverse-seal devices. (There is no substantial order of preference in the operability of these devices.)

Task 9. Support Studies

Plant Effluent Processing

The plant effluent cleanup section was operated during Test 71, and periodic upsets occurred in the light-oil recovery unit. Post-run inspection of the unit revealed a large accumulation of char in the tower and the separator. The unit's operation was much improved during Test 72 when double-screening equipment was used to screen out fines from the reactor. The unit was exceptionally clean after Test 72, and double screening the reactor feed is thought to have been responsible.

The plant effluent cleanup section was in operation for Test 73, but the Edens solids separator was not because of a broken chain that was not repaired in time for the test.

Test Methanation Systems and Catalysts

The IGT fixed-bed methanation section was on standby during this quarter. The Chem Systems' liquid-phase methanation pilot unit was used in Test 71. It operated for a total of 54 hours: 30 hours with CO-rich reformed natural gas from the hydrogen plant and the last 24 hours with purified gas from the HYGAS plant. The unit was then drained of its residual oil, cleaned, and readied for reactor modification by Chem Systems, Inc. The modifications were completed, and the liquid-phase methanation unit was operated using high carbon monoxide-to-hydrogen ratio gas from the HYGAS hydrogen plant. On June 22 at 2345 hours, an oil fire occurred in the liquid-phase methanation unit. It required the presence of the Chicago Fire Department, and resulted in substantial damage to the HYGAS pilot plant. DOE has set up a fire investigation committee to discern the cause of the fire.

Hot-Oil Quench System

This system was not worked on during this quarter.

Materials Testing

Metal Properties Council (MPC) corrosion-erosion coupons were exposed during Tests 71 through 73.

Engineering Service

Routine engineering services were conducted during this quarter. IGT personnel visited Butler Manufacturing Company in Salina, Kansas, to witness pneumatic conveying tests on raw and pretreated bituminous coals. Butler is supplying IGT with 400-ton storage silos and pneumatic conveying systems to increase the reliability of its reactor operations.

Engineering specifications and designs of the four approved modifications were forwarded to Scientific Design Corporation for review. These include the 400-ton storage units, the additional demineralized water unit, the new high-pressure cyclone slurry vessel, and a new wet-scrubber unit. Details of the proposed HYGAS modifications were also released to Scientific Design Corporation for review.

The study of pretreatment minimization of Illinois No. 6 Peabody No. 10 Mine coal was resumed using a 1-inch batch-type unit.

CONCLUSIONS

Tests 71 through 73 were completed during this quarter. Tests 71 and 72 were highly successful in obtaining smooth gasifier operation at char conversions ranging from 70% to 90% and at reactant ratios (steam/char and oxygen/char) near the commercial/demonstration plant design base used by Procon. These two tests, operated at superficial velocities in the steam-oxygen gasifier zone of 1.2 ft/s, were essentially sinter-free. (Test 71 was entirely sinter-free, while Test 72 had some small sinter formations near the steam-oxygen distributor nozzles.)

Test 73, conducted with a superficial gas velocity in the steam-oxygen gasifier zone of approximately 0.8 ft/s, but at otherwise similar conditions to Tests 71 and 72, had a rapid and significant sinter formation in the steam-oxygen gasifier zone.

Based on these three tests, sinter-free operation of the HYGAS steam-oxygen gasifier zone requires superficial velocities higher than 0.8 ft/s. Excellent operation was obtained at a velocity of 1.2 ft/s, but it is not known if this can be reduced to a somewhat lower level.

Mechanically, the modifications to the steam-oxygen gasifier distributor and the relocation of valve 339 added to the overall operability of the HYGAS reactor and improved the temperature profiles (reduced the overall steam-oxygen gasifier bed temperature differential) in the steam-oxygen gasifier zone.

Cold-flow model tests of the low temperature reactor section in the HYGAS reactor indicate that a nonmechanical valve feeding system (L-valve, J-valve, and reverse seal) have an operational advantage over mechanical (lift-pot devices) types.

FUTURE WORK

The annual HYGAS pilot plant turnaround activites are planned for July as well as repairs made necessary by the fire in the liquid-phase methanation unit.

Test 74 is planned with the objective of testing an alternative design for the steam-oxygen gasification zone in the pilot plant. Two mechanical changes in the steam-oxygen gasification zone distributor and stripping-steam distributor are fundamental to this objective and will be incorporated into the pilot plant before the test. These are —

1. The 3-inch-diameter, 60-degree angle to the vertical coned sections at the end of individual nozzles in the steam-oxygen distributor will be removed, and the nozzles will be beveled at an angle 30 degrees from the vertical. (Thermocouples, which have in the past been attached to the coned portion of the steam-oxygen distributor, will be removed from this area and located at the wall of the reactor. One thermocouple will be located between the stripping-steam and the steam-oxygen distributors.)

2. The stripping steam introduced to the bottom of the reactor below the steam-oxygen distributor will be increased to give the velocity required for complete fluidization in this zone. A new steam distributor will be installed to distribute steam over the 24-inch reactor cross section.

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