FE-2434-26

PIPELINE GAS FROM COAL-HYDROGENATION (IGT HYDROGASIFICATION PROCESS)

Project 9000 Monthly Status Report

For the Period January 1 Through January 31, 1978

Prepared by

Institute of Gas Technology

IIT Center, 3424 S. State Street

Chicago, Illinois 60616

Date Published — March 1978

Prepared for the

UNITED STATES DEPARTMENT OF ENERGY

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INSTITUTE OF GAS TECHNOLOGY · IIT CENTER · CHICAGO 60616

Project Status Report

for

DEPARTMENT OF ENERGY

AND

AMERICAN GAS ASSOCIATION

Report for

January 1978

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

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

I. PROJECT OBJECTIVE

The objective of this project is to perform the necessary pilot plant operations and related support studies required to acquire data for a commercial/demonstration plant design based on the HYGAS® Process. To this end, we are conducting tests with Illinois bituminous coal to acquire data on optimizing the process. Eleven such tests have been made, including Tests 68 and 69 which were conducted during January. This series of tests has provided data on the operating conditions necessary for high carbon conversion. We are currently working on achieving high char throughput at high carbon conversion rates.

II. SUMMARY

Two tests were conducted during January. Test 68 was terminated after 20 hours of char feed to the reactor due to a leak in a reactor manway. We started Test 69 after conducting a turnaround. Stable conditions were established in the reactor at a char feed rate exceeding 2 tons/hr, when a liquid nitrogen shortage, caused by a severe winter storm, forced us to terminate the test.

Further information was supplied to Procon, Inc., to aid its staff in their design of a commercial/demonstration HYGAS plant. Additional test work was done on the lift-pot, lift-line feeder device.

Results of a preliminary engineering study on the design of a hot-liquid quench system for the HYGAS plant are presented here.

III. ACHIEVEMENTS

Task 7. Pilot Plant Experimental Operation

The initial light-off for Test 68 occurred on December 22. A hot pressure test on December 24 revealed leaks in the reactor, which were fixed and the reactor relit on December 25. Char feed to the reactor began on December 31 at 1345 hours, and the reactor operation became self-sustained at 0445 hours on January 1. Test 68 was terminated at 1330 hours on January 3 due to a leak on manway 0 of the reactor. During Test 68, we experienced a problem in getting proper mixing of the pretreater char-oil slurry in the slurry mix tank. High-density slurry plugged the low-pressure, slurry circulation loop and temporarily interrupted char feed to the reactor. During this test, 45 tons of char were fed to the reactor over a 20-hour period.

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

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

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

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

In Test 69 the pretreater began operating at 2130 hours on January 18 and satisfactorily provided nonagglomerating char for the reactor feed. The pretreater feed system was interrupted several times by problems in operating the ball valve in the lockhopper feeding system. The slurry preparation section operated satisfactorily for Test 69. Argonne National Laboratory's personnel operated their test slurry flowmeters during this test. The quench section operated well; however, the purification and methanation sections were not put on-stream. The effluent cleanup section was in service for Test 69 and will be inspected following it. The utilities operated satisfactorily during the test. The hydrogen plant operated well, supplying reactor cooldown gases at the end of the test. We will conduct a routine post-run inspection throughout the entire plant before beginning Test 70. Because Tests 68 and 69 were terminated prematurely, Test 70 will have the same objective of establishing high throughput of char through the reactor at high conversions. The mechanical status of the HYGAS pilot plant for January is presented in Figure 1.

A meeting was held at the U.S. Department of Energy (DOE) headquarters on January 10 to lay the ground rules for the transfer of environmental information from IGT to Procon, Inc., and DOE personnel.

A debriefing session for Test 68 was held on the morning of January 18. Representatives from DOE; Darcom; Procon; Scientific Design Company, Inc.; C. F. Braun & Co.; and IGT attended. This same day, these participants met to consider IGT's available data on the operating requirements necessary for sinter-free operation in the steam-oxygen gasifier.

After an initial discussion took place and recommendations were made, the first review of Procon's commercial HYGAS reactor design was presented.

Task 8. Demonstration Plant Support

IGT continued to provide Procon (through DOE) with specific material balances and selected process design information. IGT personnel attended two meetings, related to the demonstration plant design support, during the month. A special meeting, relating to Procon's need for environmental data to formulate the effluent treatment sections of the design, was held at DOE headquarters on January 10. On January 18, IGT met with Procon and DOE personnel to outline conditions for sinter-free operations in the reactor. IGT-Procon representatives held initial discussions on the proposed reactor vessel design.

We continued our evaluation of various configurations of the low-temperature reactor (LTR) transport zone using an ambient temperature, low-pressure, scale model. Two LTR solids-feeding options, a lift-pot design and an L-valve feeder, were tested.

Evaluation of a Lift-Pot and an L-Valve LTR Feeder Device

Lift Pot

The lift-pot lift-line feeder device (Figure 2), tested in December with sand, was retested this month using –20+200 mesh pretreated Illinois No. 6 bituminous coal. The purpose of the test was to evaluate the operation of the device using a material similar to that which would be used in the LTR section of the HYGAS demonstration plant.

Figure 1. MECHANICAL STATUS OF THE HYGAS PLANT FOR JANUARY 1978

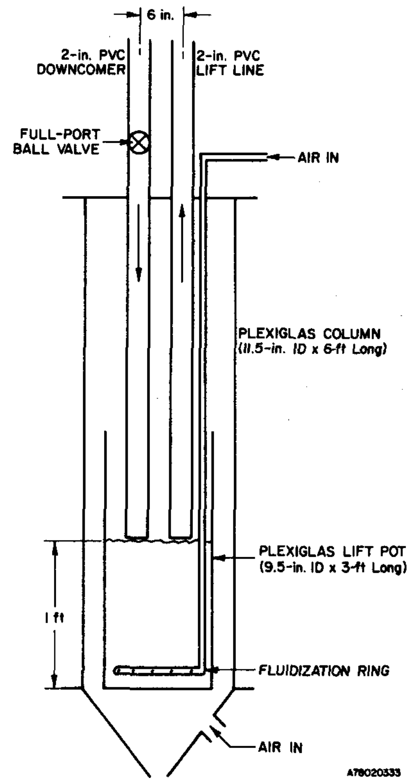


Figure 2. LIFT-POT TEST CONFIGURATION

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

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

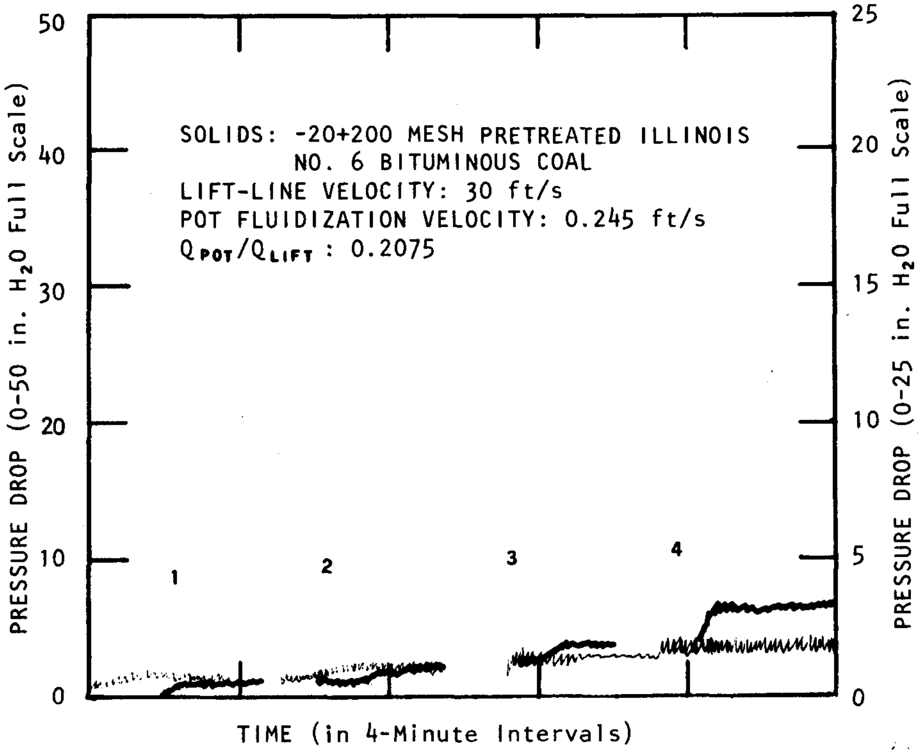
In a typical run, the upper bed was first fluidized. The desired lift-pot fluidization velocity and the lift-line velocity were then set. We took readings at several different solids flow rates and then analyzed the results.

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

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

The first test, Run HGD-2A, was made at a lift-pot velocity of 0.245 ft/s and at a lift-line velocity of 30 ft/s. The recorder tracings and the conditions used for this run are shown in Figure 3. As the ball valve in the downcomer was successively opened wider, the solids flow rate to the lift-pot bed increased as did the solids flow into the lift line, thus causing a rise in the lift-line pressure drop. Reading 4 was taken with the downcomer ball valve wide open. At this reading the lift-line pressure drop fluctuated approximately ±0.1 inch of water from the average lift-line pressure drop reading. This fluctuation was approximately one-tenth that observed in the tests with sand.

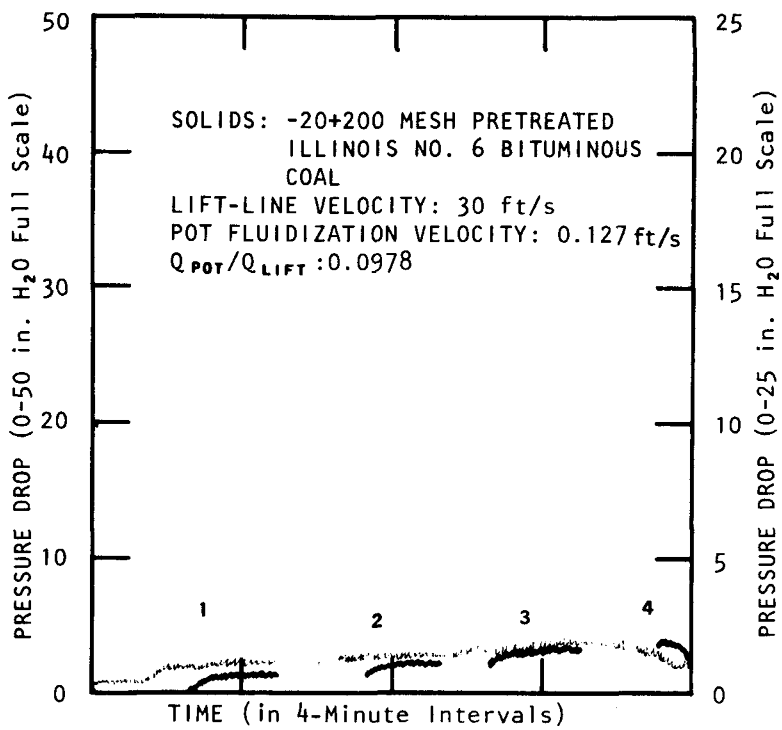
In Run HGD-2B, the lift-line velocity was kept at 30 ft/s, but the lift-pot velocity was reduced to 0.127 ft/s. The results obtained for this run are shown in Figure 4. For the first three readings, as the valve in the downcomer was successively opened wider, the solids flow rate to the lift pot and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



READING NO.	SOLIDS FLOW RATE, lb/hr
1	285
2	580
3	1100
4	1685

	PRESSURE DROP	SCALE, in. H_2O
▬ (heavy line)	ACROSS LOWER SECTION OF LIFT LINE	0-50
— (light line)	ACROSS UPPER SECTION OF LIFT LINE	0-25

Figure 3. LIFT-LINE PRESSURE DROP FOR RUN HGD-2A



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

Figure 4. LIFT-LINE PRESSURE DROP FOR RUN HGD-2B

In Run HGD-2C, a lift-line velocity of 30 ft/s was used once again, but this time with a lift-pot velocity of 0.3 ft/s. The results (Figure 5) were similar to those from Run HGD-2A; however, the amplitude of the lift-line fluctuations was somewhat higher than that in Run HGD-2A.

The lift-line velocity in Run HGD-2D (Figure 6) was increased to 35 ft/s, and the lift-pot velocity was set at 0.245 ft/s. At the maximum solids flow rate in this run, the entire downcomer was in streaming flow, and the lift-line fluctuations were twice as large as those in Run HGD-2A, which had the same lift-pot velocity but a lower lift-line velocity.

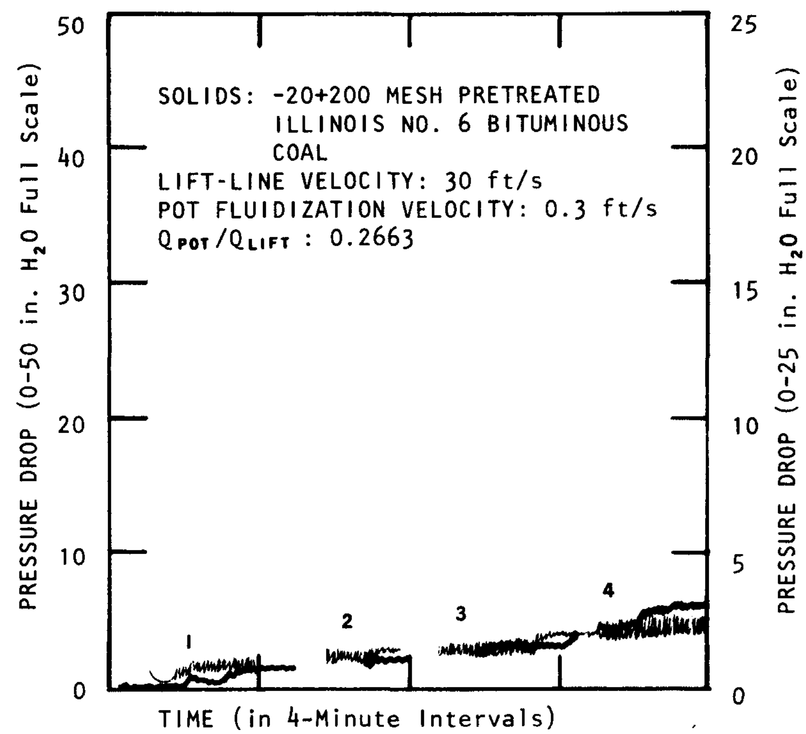
Run HGD-2E (Figure 7) was made at a lift-line velocity of 40 ft/s, while maintaining the lift-pot velocity at 0.245 ft/s. Lift-line fluctuations in this run were about the same as those in Run HGD-2D. When the ball-valve in the downcomer was fully opened, the entire downcomer became dilute and the solids flow rate fell sharply.

In Run HGD-2F (Figure 8), the lift-line velocity was set at 30 ft/s, and the lift-pot velocity was set at 0.182 ft/s. This combination of lift-line and lift-pot velocities gave results similar to those of Run HGD-2A. Lift-line fluctuations were very small (±0.1 inch of water), and a stable downcomer flow pattern was maintained even at the wide-open ball valve position.

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

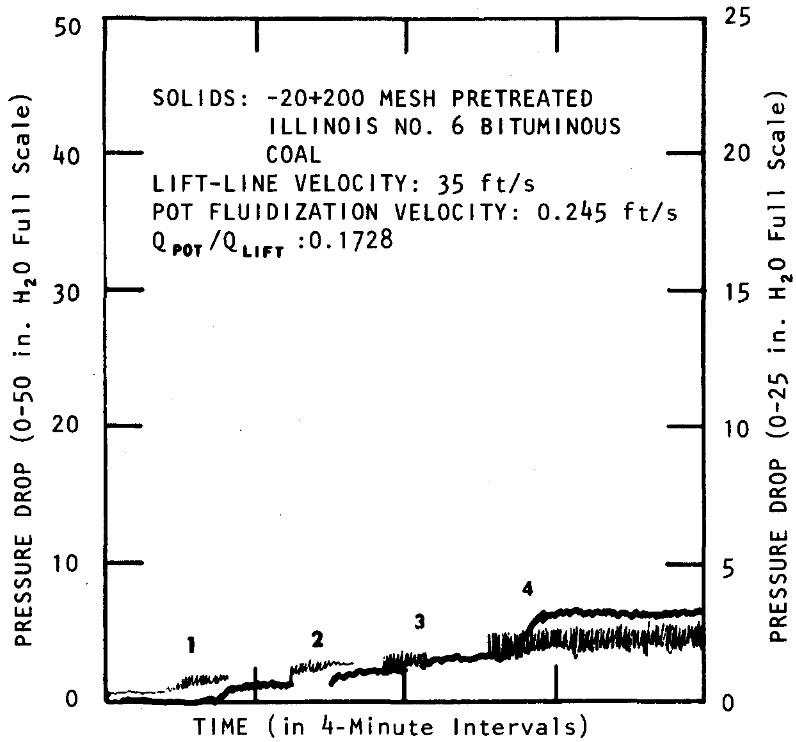
Since coal is much lighter than sand, the lift velocities used with coal were further away from choking. The results probably mean that there is an optimum lift-line velocity that will minimize the lift-line, pressure-drop fluctuations — one that is not too far from (nor too close to) choking.

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



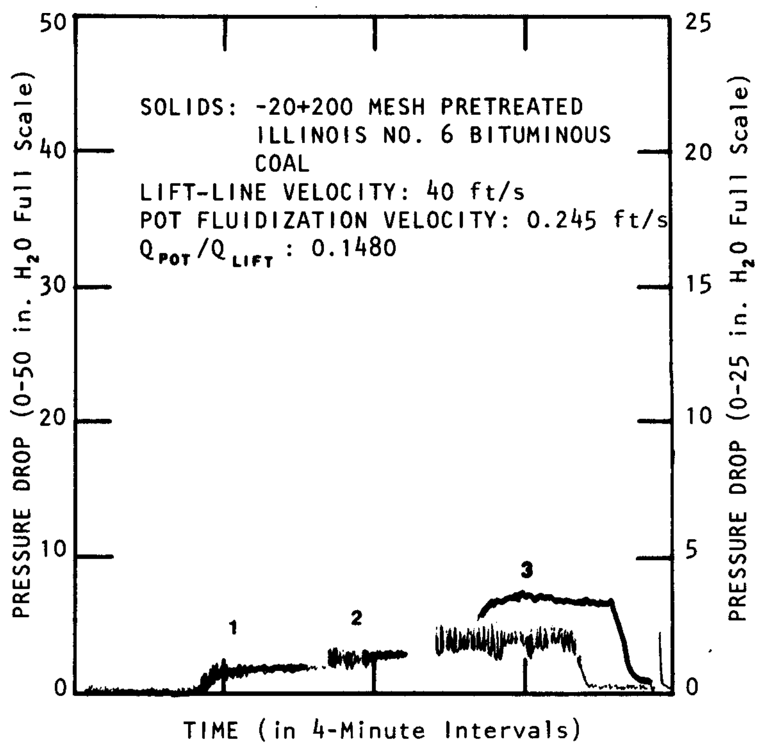
READING NO.	SOLIDS FLOW RATE, lb/hr		PRESSURE DROP	SCALE, in. H_2O
1	380	——— (heavy)	ACROSS LOWER SECTION OF LIFT LINE	0-50
2	575			
3	895	——— (light)	ACROSS UPPER SECTION OF LIFT LINE	0-25
4	1750			

Figure 5. LIFT-LINE PRESSURE DROP FOR RUN HGD-2C



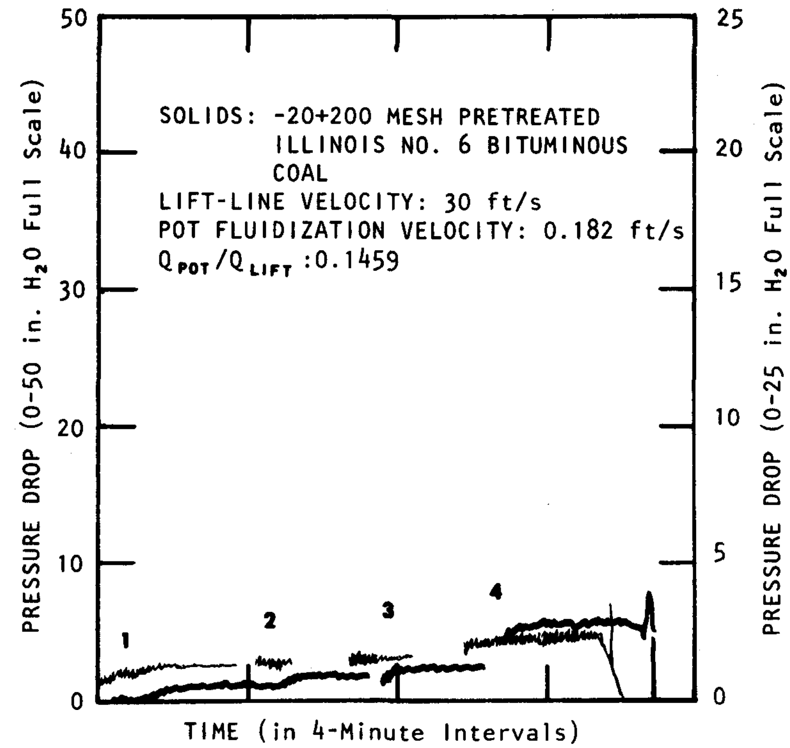
READING NO.	SOLIDS FLOW RATE, lb/hr		PRESSURE DROP	SCALE, in. H_2O
1	275	——— (thick)	ACROSS LOWER SECTION OF LIFT LINE	0-50
2	510			
3	805	——— (thin)	ACROSS UPPER SECTION OF LIFT LINE	0-25
4	1530			

Figure 6. LIFT-LINE PRESSURE DROP FOR RUN HGD-2D



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

Figure 7. LIFT-LINE PRESSURE DROP FOR RUN HGD-2E



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

Figure 8. LIFT-LINE PRESSURE DROP FOR RUN HGD-2F

L-Valve

A second lift-line feeder device, the L-valve, was also tested during January. A sketch of this device is shown in Figure 9.

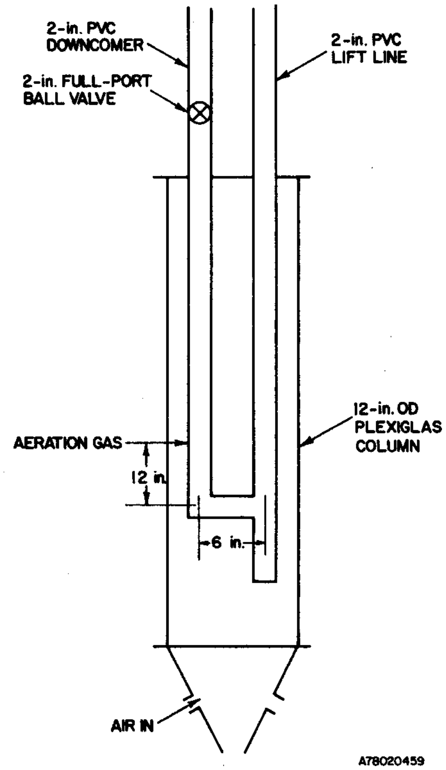


Figure 9. L-VALVE TEST CONFIGURATION

The L-valve was constructed of 2-inch-diameter PVC pipe so that we could observe the flow of solids through it. The flow was controlled using only aeration gas supplied to the valve at a point 12 inches above the centerline of the horizontal sections.

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

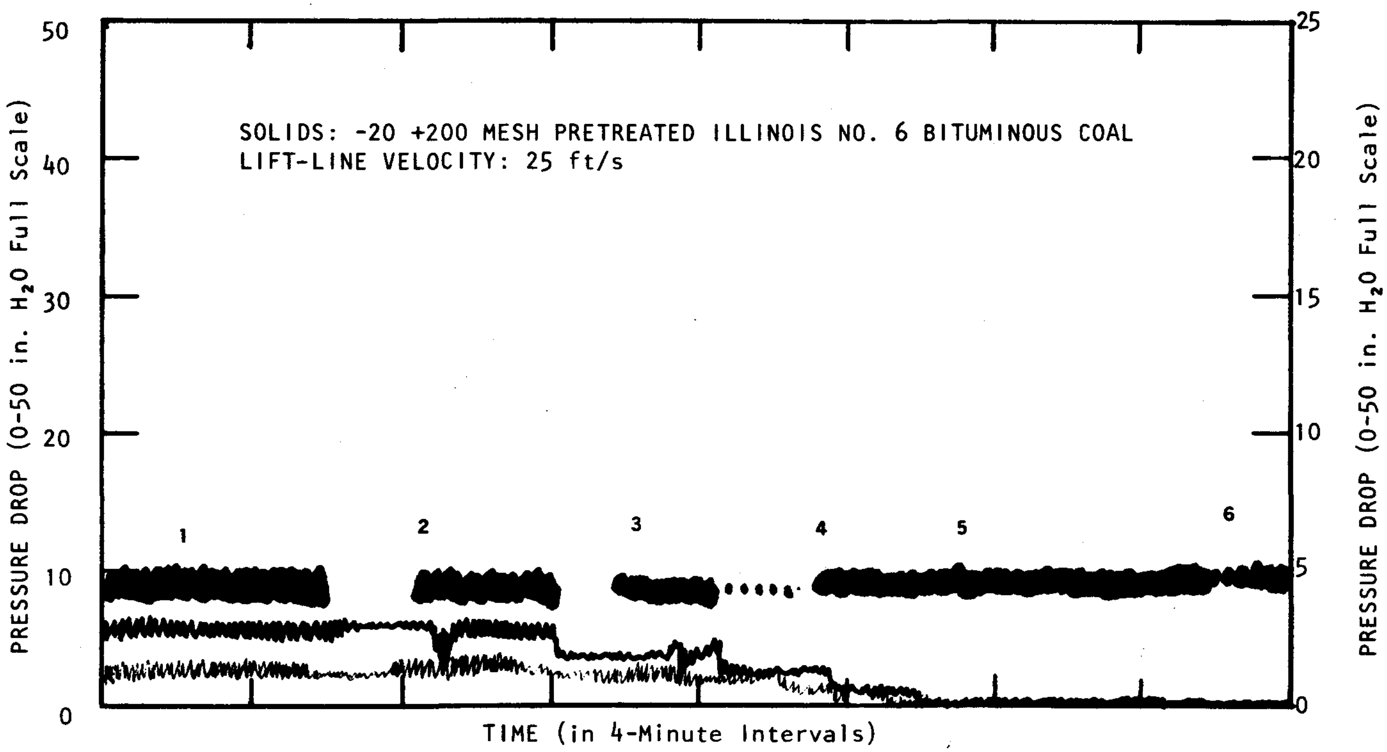
In Run HGD-3A, the lift-line velocity was set at 25 ft/s. The lift-line, pressure-drop recorder traces and the run conditions used are shown in Figure 10. In this run, the solids flow rate into the lift line was increased in steps by increasing the aeration gas flow to the L-valve. Downcomer operation was relatively smooth, and the maximum lift-line, pressure-drop fluctuations were approximately ±0.1 inch of water from the average pressure-drop reading.

In Runs HGD-3B through HGD-3D, the lift-line velocity was set at 30, 35, and 40 ft/s, respectively. The results are shown in Figures 11, 12, and 13. In all of these runs, the solids flow rate was increased up to approximately 1100 to 1200 lb/hr using only L-valve aeration. Our attempts to increase the solids flow rate only diluted it in the downcomer and also increased the lift-line, pressure drop fluctuations. We also tested the L-valve using −20+80 mesh Ottawa sand as the solids. The procedure used for these tests was identical to that used for the coal tests.

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

The first L-valve test using sand (Run HGD-4A) was made with a lift-line velocity of 40 ft/s. The results of this run are shown in Figure 14. L-valve operation was controllable up to a solids flow rate of about 7600 lb/hr. The solids flow rate could be increased beyond this rate, but the downcomer flow eventually became dilute. The maximum pressure-drop fluctuations in the lift line were approximately ±1.25 inches of water above the average pressure-drop readings. This was approximately half the fluctuation obtained with the lift pot. Thus, the L-valve feeder configuration resulted in smoother lift-line operation than did the lift-pot feeder.

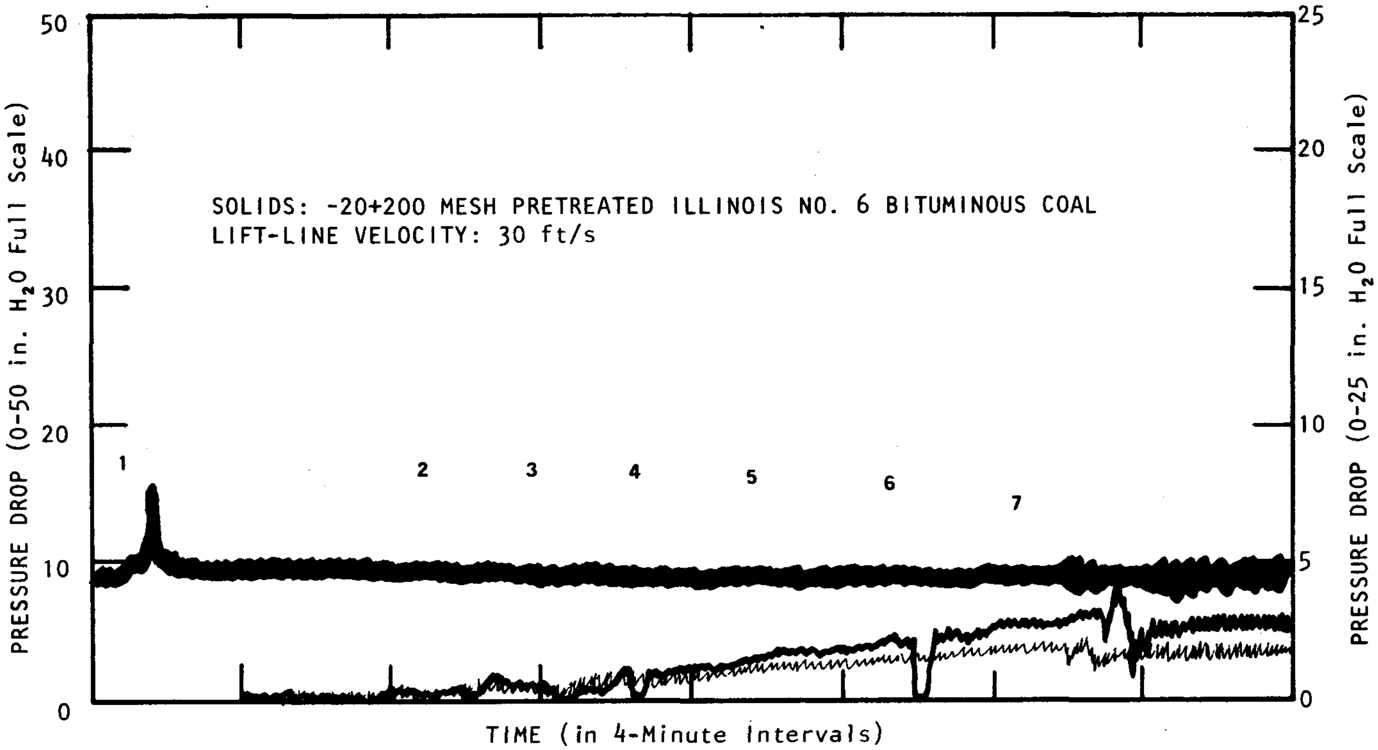
Runs HGD-4B and HGD-4C were made at lift-line velocities of 45 and 50 ft/s, respectively. As with the first run, both of these runs gave excellent solids flow control up to a solids flow rate of about 8000 lb/hr before the downcomer flow became dilute. Fluctualtions in the lift line were also approximately the same as in Run HGD-4A. The results of these two runs are shown in Figures 15 and 16.



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

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

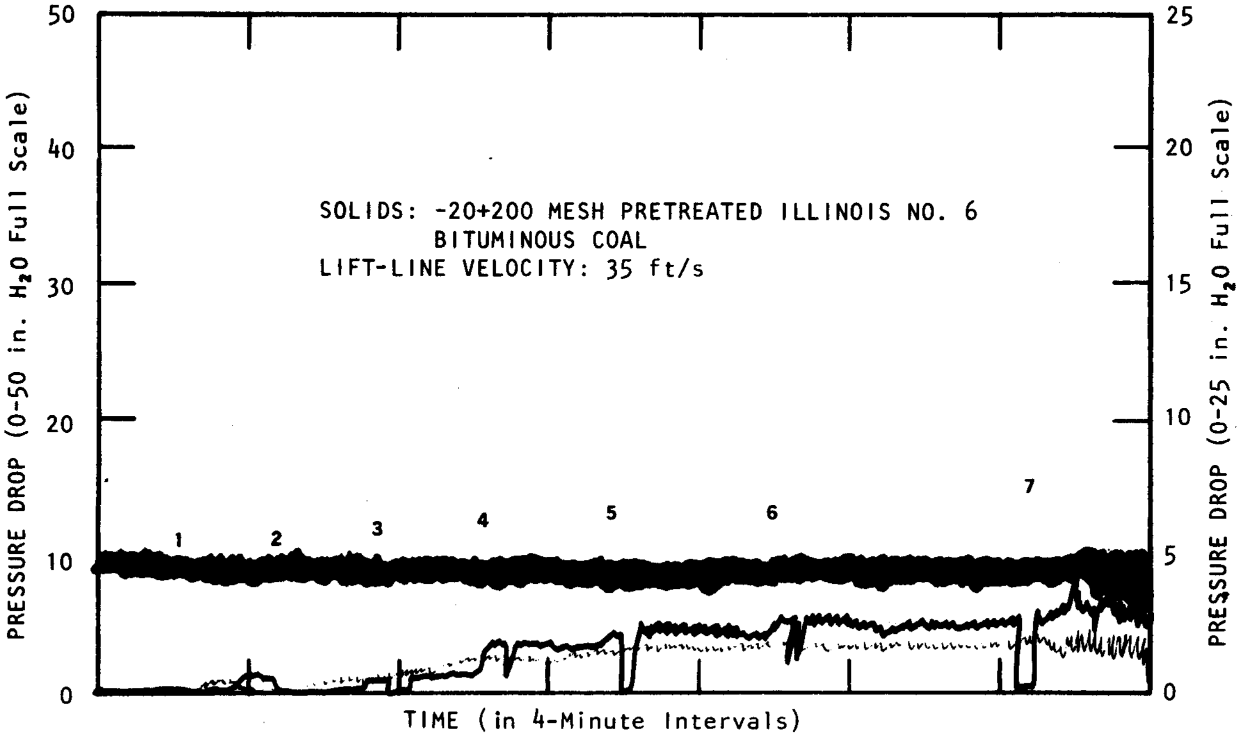
Figure 10. LIFT-LINE PRESSURE DROP FOR RUN HGD-3A



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

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

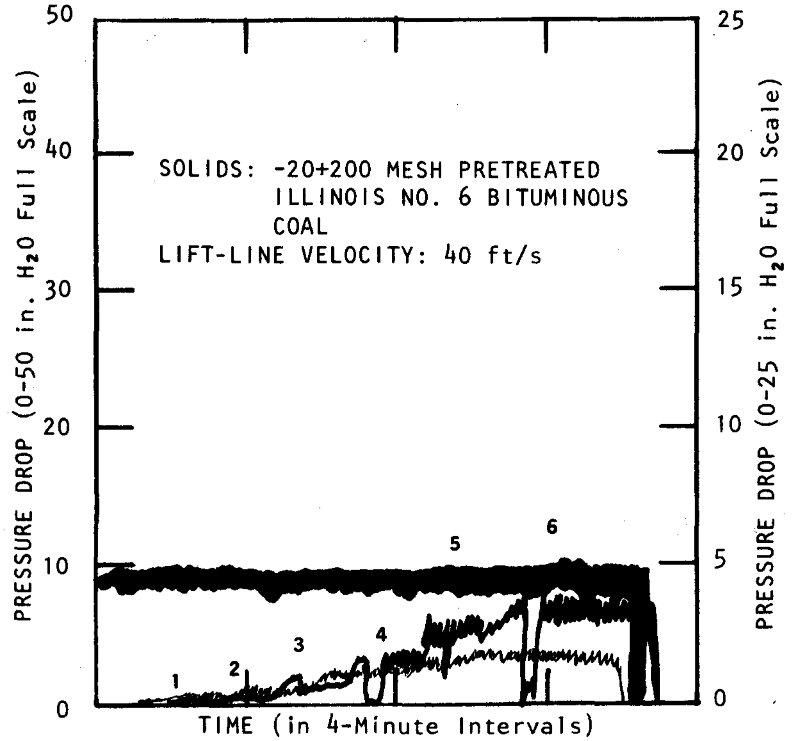
Figure 11. LIFT-LINE PRESSURE DROP FOR RUN HGD-3B



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

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

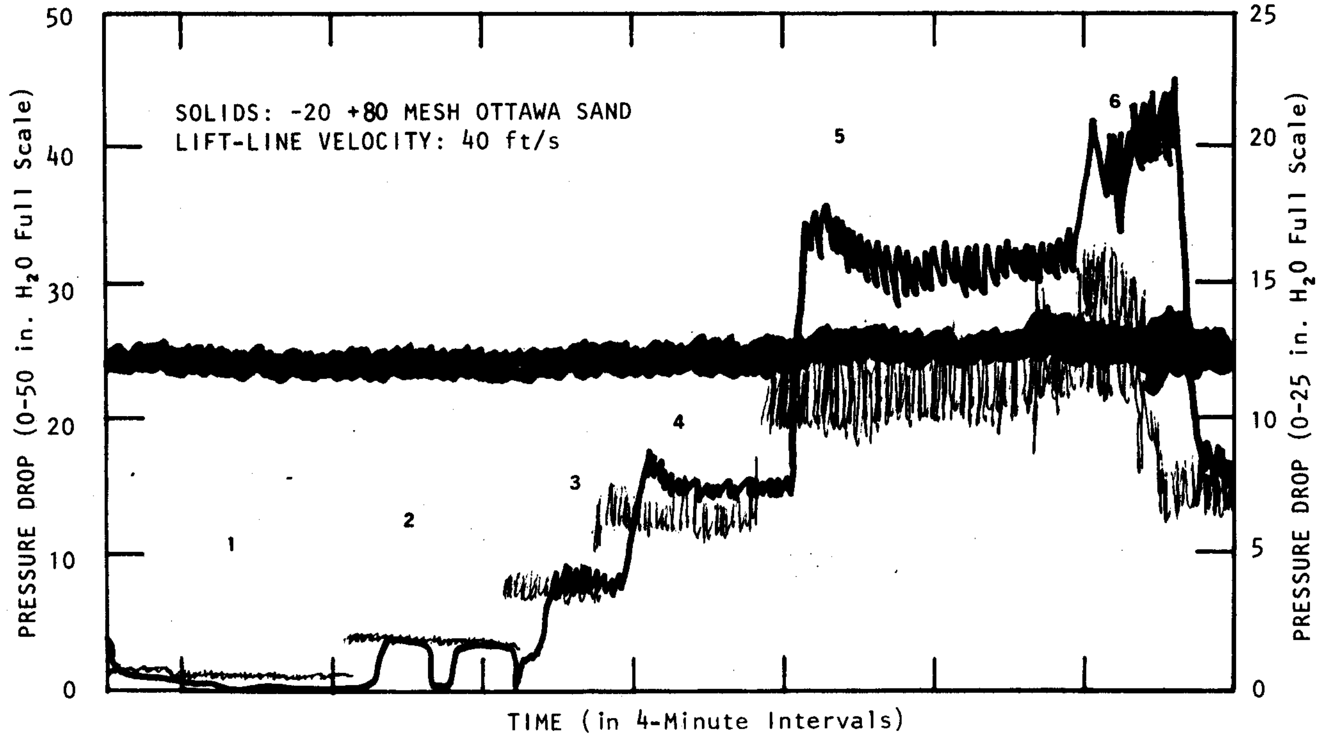
Figure 12. LIFT-LINE PRESSURE DROP FOR RUN HGD-3C



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min
1	0	0.1950
2	100	0.3399
3	425	0.5075
4	700	0.6495
5	1020	0.8409
6	1280	1.0032

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

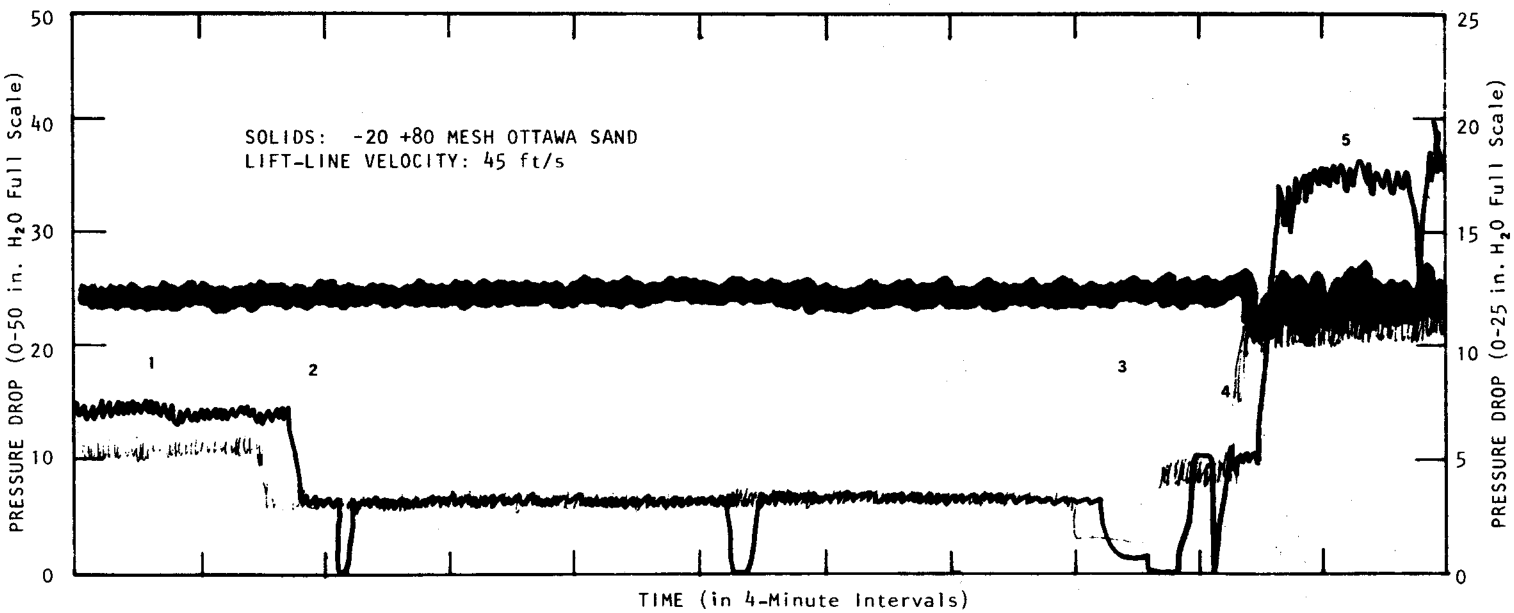
Figure 13. LIFT-LINE PRESSURE DROP FOR RUN HGD-3D



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min
1	0	0.3367
2	200	0.4290
3	1,500	0.6609
4	3,250	0.8890
5	7,600	1.3530
6	10,000	1.3530

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

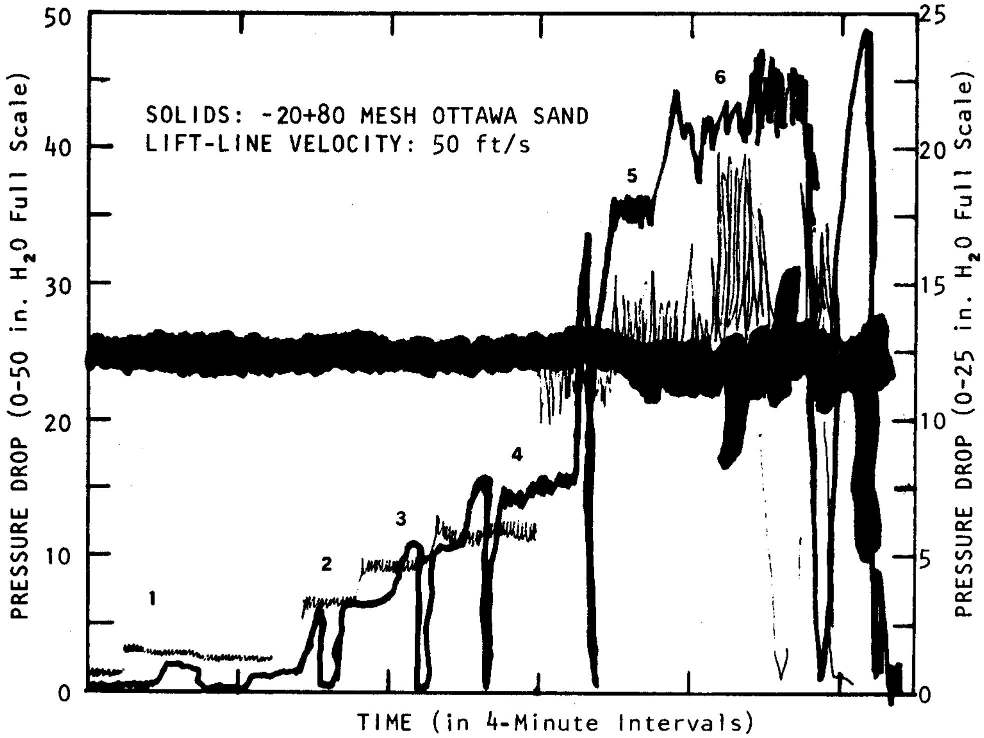
Figure 14. LIFT-LINE PRESSURE DROP FOR RUN HGD-4A



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min
1	3540	0.8687
2	1350	0.5615
3	350	0.3328
4	2265	0.6985
5	8450	1.1031

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

Figure 15. LIFT-LINE PRESSURE DROP FOR RUN HGD-4B



READING NO.	SOLIDS FLOW RATE, lb/hr	L-VALVE AERATION, ACF/min
1	245	0.3305
2	1000	0.5150
3	1325	0.5523
4	2340	0.6844
5	3690	0.8472
6	7550	1.0667

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

Figure 16. LIFT-LINE PRESSURE DROP FOR RUN HGD-4C

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

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

Other Test Configurations

Seven lift-line feeder configurations (Figures 17 through 23) are currently scheduled for testing. The lift pot and the L-valve (Figures 17 and 18, respectively) have already been tested. The J-valve (Figure 19), a nonmechanical valve similar to the L-valve, will be tested next. We will also investigate a reverse-seal leg (Figure 20) and a reverse-seal pot (Figure 21), and in addition, will test two modifications of the lift-pot design (Figures 22 and 23).

Task 9. Support Studies

Plant Effluent Processing

The effluent cleanup section operated during Test 68 and Test 69. We cleaned this section prior to Test 69 and will inspect it again after the test.

Test Methanation Systems and Catalysts

The IGT fixed-bed catalyst methanation section was put on standby for Test 69, but was not put on-line because of the early termination of this test. Chem Systems' liquid-phase methanation pilot unit is still being modified.

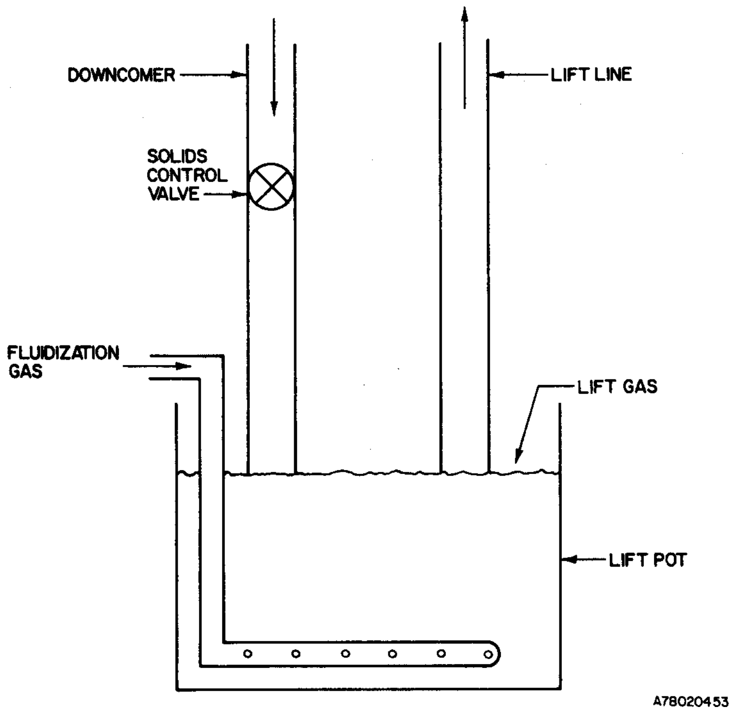


Figure 17. LIFT POT I

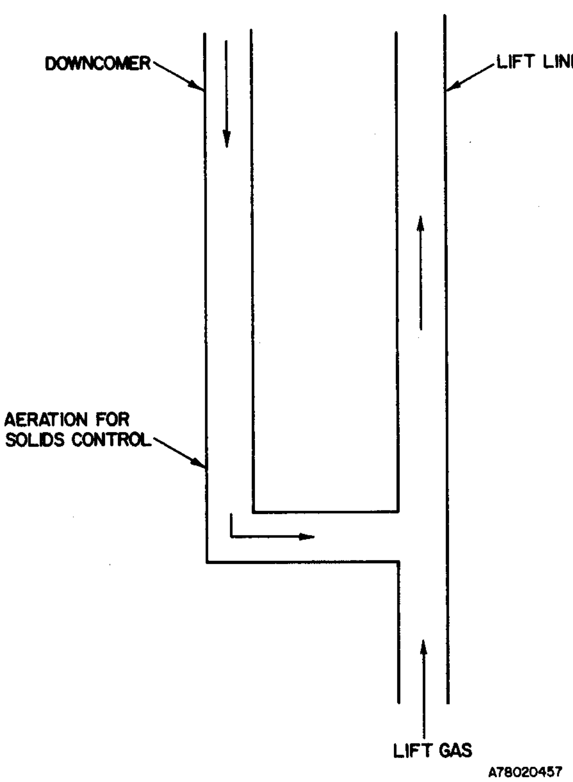


Figure 18. L-VALVE

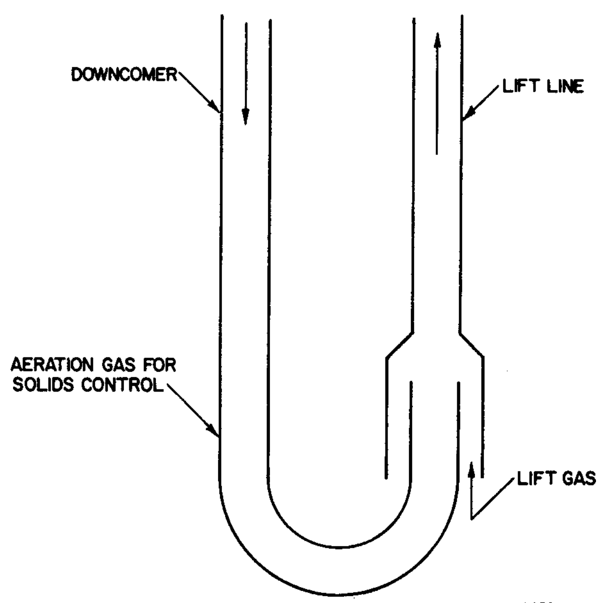


Figure 19. J-VALVE

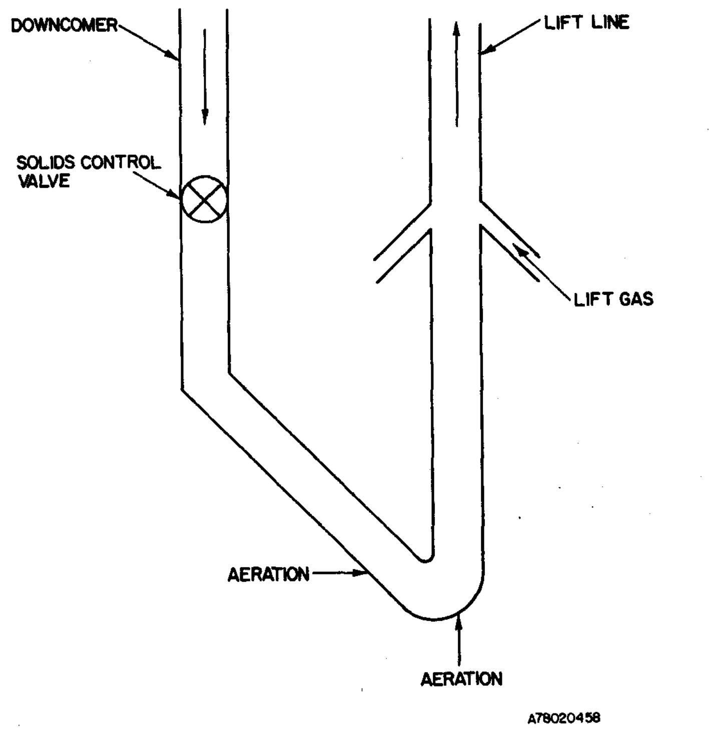


Figure 20. REVERSE SEAL LEG

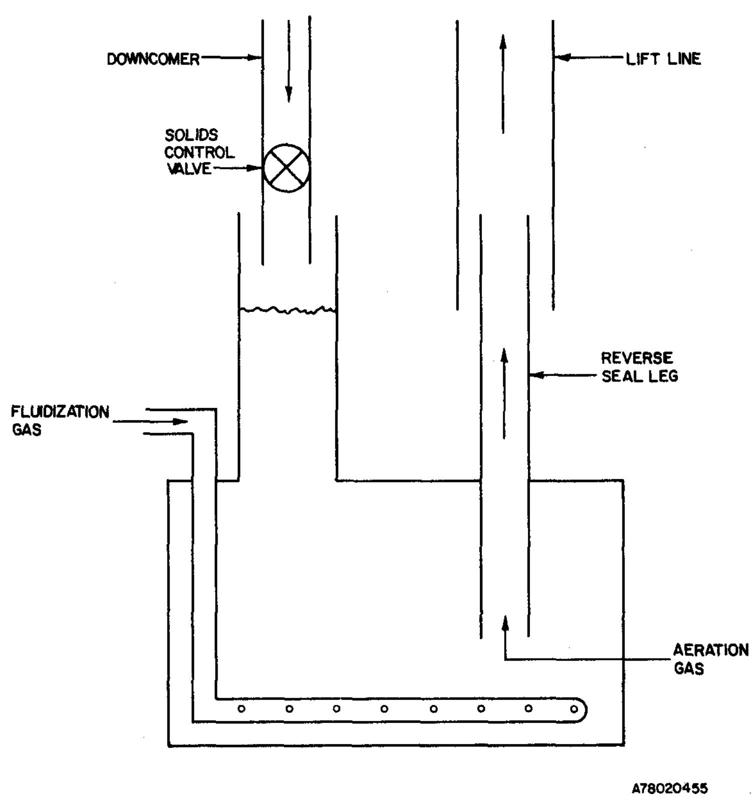


Figure 21. REVERSE-SEAL POT

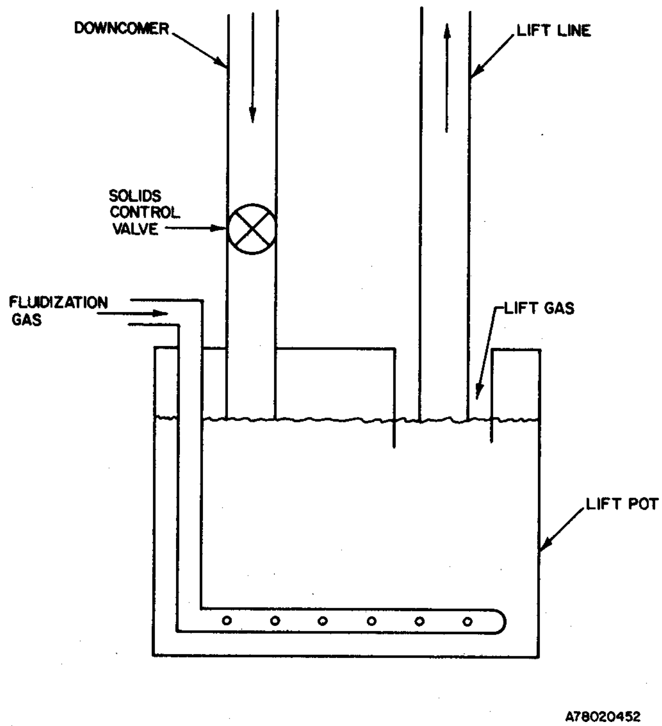


Figure 22. LIFT POT II

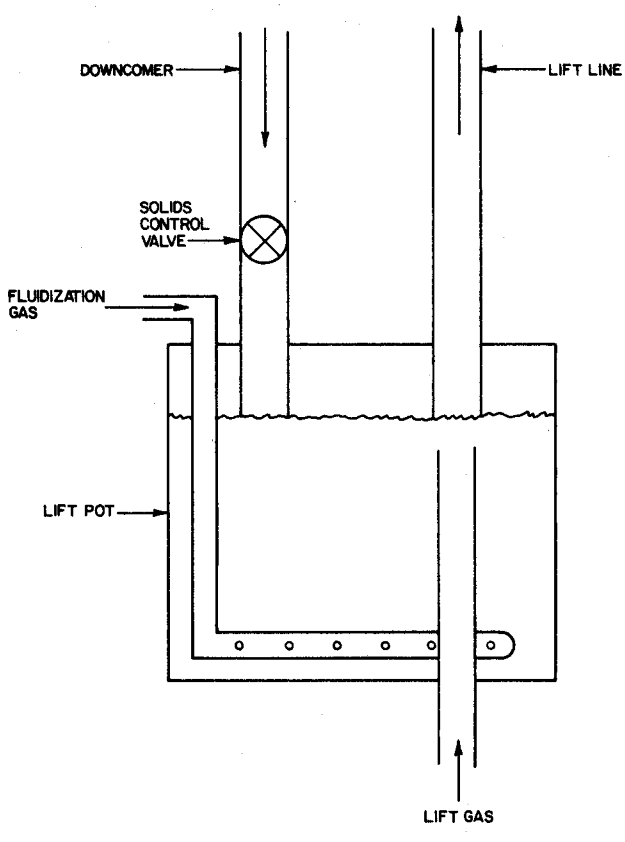


Figure 23. LIFT POT III

Investigate Hot-Oil Quench System

Summary

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

Objective

The existing oil-water separation system has had intermittent operational problems caused by the formation of an oil-solids-water layer in the quench system separation vessel (4.06-41). The three-phase layer may be formed by solids that get through the existing cyclone system during upset conditions and/or because the cyclone efficiency during steady operation allows a small amount of particulate matter (which builds up in the oil-water separation system) to pass through. We expect that the hot-water quench will add effective downstream capabilities for the removal of particulate matter that gets through the cyclone system during any mode of gasifier operation, thereby allowing smoother operation of the oil-water separation vessel during an entire gasification run. Our most important consideration was to design a scrubbing system that would maintain a product gas composition that would be essentially unchanged as it passed from the gasifier through the scrubber vessel.

Procedure

Water and toluene were studied as possible candidates for the scrubbing media. They were chosen because a) water is inherently the easiest and safest medium and b) our previous observed experience has shown that particulate matter is better removed or attracted to the toluene layer in our separation process.

The scrubbing vessel was designated as our existing prequench tower vessel (4.06-61). Product-gas quench and water/oil cooling and condensation will then be affected in the quench tower system or other system that is designed to meet the necessary heat-duty requirements of the scrubber contactor vessel off-gas.

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

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

Removal of the prequench vessel necessitated an evaluation of the possible modifications to the downstream system, which are required to effectively quench the gasifier product gas to the 1000°F level. We intended, again, to minimize equipment requirements. Two approaches were used: 1) modify the quench tower (4.06-01) as necessary for quenching the gas and 2) install an in-line cooler to reduce heat duty into the quench system and thereby minimize quench tower modifications.

To establish the quench system material balances, we used the University Computing Company's computer and available program, after making modifications to satisfy our process requirements. Once again, the computer used the Chao-Seider thermodynamic correlations to generate stream composition data.

Results

Scrubber Vessel

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

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

a. Toluene Advantages

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

b. Toluene Disadvantages

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

c. Water Advantages

- Available liquid stream.
- Noncombustible liquid.
- Solids/liquid separation method easily handled by cooling and filtration.
- Possible to operate at the dew point temperature with essentially unchanged gas composition.
- At temperatures higher than dew point, water partial pressure in product gas is higher than saturation pressure, therefore, minimal humidification problems occur.

d. Water Disadvantages

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

Downstream Quench Results

We estimate that the removal of approximately 16,000,000 Btu/hr is required of the quench system downstream of the scrubber contact vessel. Forty-seven plates and 300 gpm of 100°F water would be required if the quench tower was going to do quench duty alone. Without major modifications, the present system cannot handle these requirements.

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

Problem Areas

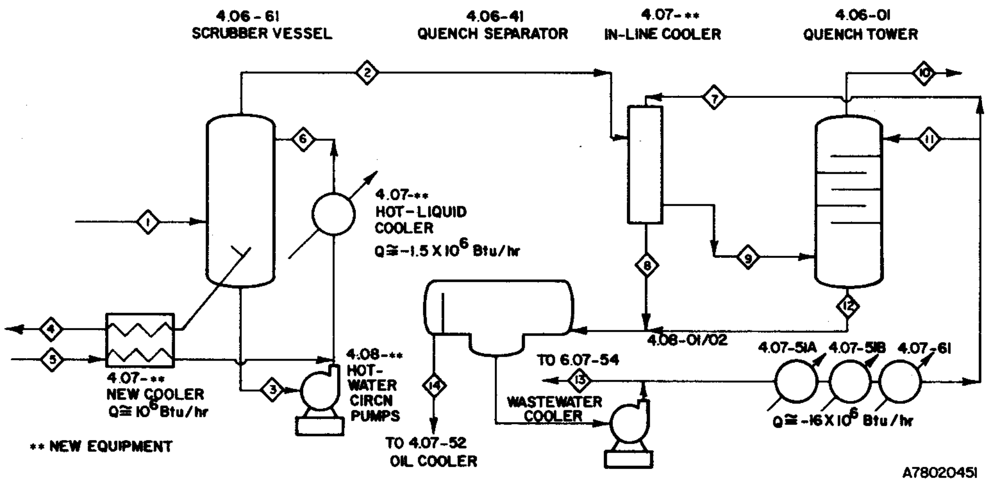
The exact configuration of the required in-line cooler is not known. Problems with flow regime can be expected, and we are studying effective ways to solve them. This cooler might possibly be purchased from an outside vendor. Water-scaling tendencies, if raw water is used, may present a problem in the contactor vessel.

Conclusions

We recommended that a hot-liquid scrubber be designed using water as the scrubbing medium. Figure 24 shows the conceptual process flow diagram for the bituminous case (90% carbon conversion) using water as the scrubbing medium. Minimal plant modifications are required. We estimate that the critical path to the project's completion will be the specification and procurement of the hot-liquid circulation pumps.

Engineering Services

Routine engineering services were conducted during January. In addition, a new steam-oxygen sparger was designed. The possibility of installing a double screen in the coal preparation section was evaluated. We found it to be feasible, and began constructing the double screen. Operating personnel are also preparing a reliability study of the HYGAS plant. Details of this study will be reported later.



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

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Figure 24. CONCEPTUAL HOT-WATER SCRUB PROCESS FLOW DIAGRAM

Relocation of solids transfer valve 339 from the high-temperature reactor to the steam-oxygen reactor is under way. A drawing of the new location and its linkages vs. the old location is presented in Figure 25. A scaled-down model of the valve was built, and its operation was tested and found to be satisfactory. We ordered materials for constructing the valve, and expect to start installing it soon.

Argonne National Laboratory's personnel collected operating data on the low-pressure slurry line for Test 69 on their two flow test meters. They planned to return and continue testing during Test 70.

IV. PROBLEMS

No critical problems were encountered during January.

V. PATENT STATUS

The work during January as reported herein is not considered patentable.

VI. WORK PLAN AND SCHEDULE

A work schedule for the HYGAS Process is presented in Figure 26.

Approved B. S. Lee
Bernard S. Lee
Executive Vice President

Signed Wilford G. Bair
Wilford G. Bair
Director

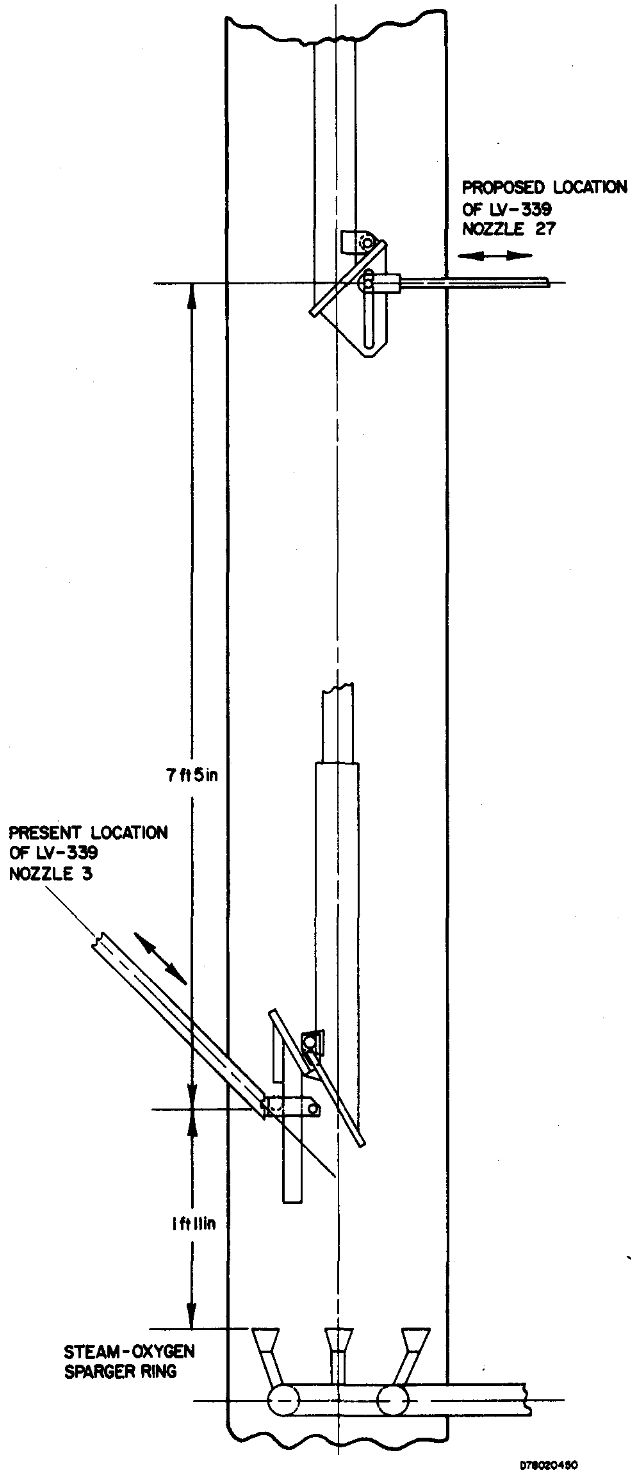

Figure 25. SOLIDS TRANSFER VALVE 339 — OLD VS. NEW LOCATIONS AND ITS LINKAGES

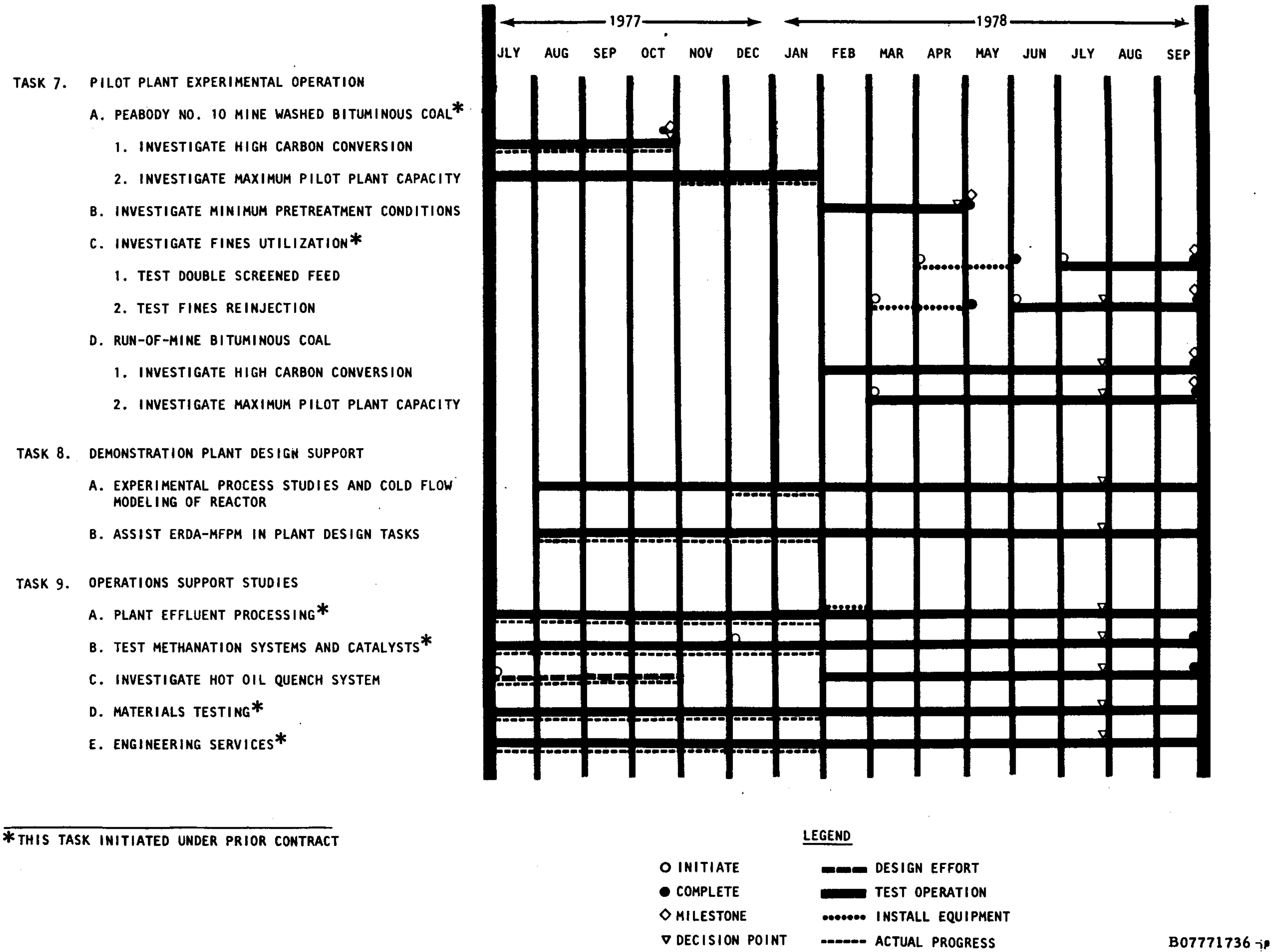

Figure 26. WORK SCHEDULE FOR HYGAS PROCESS FOR JANUARY 1978

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