

**INDUSTRIAL MEDIUM-Btu FUEL GAS
DEMONSTRATION-PLANT PROGRAM**

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MASTER

**TECHNICAL SUPPORT REPORT:
PILOT-PLANT AND
SUPPORT STUDIES**

Deliverable 46

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Submitted by
**MEMPHIS LIGHT, GAS AND WATER DIVISION
P.O. Box 430
Memphis, Tennessee 38101**

Prepared by
**Institute of Gas Technology
IIT Center, 3424 S. State Street
Chicago, Illinois 60616**

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MEMPHIS INDUSTRIAL FUEL GAS
DEMONSTRATION PLANT PROJECT

TECHNICAL SUPPORT REPORT - PART 3
(Deliverable #46)

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MEMPHIS LIGHT, GAS & WATER DIVISION
P.O. Box 430
MEMPHIS, TENNESSEE 38101

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INTRODUCTION

Program Summary

The United States Department of Energy (DOE) executed a contract with Memphis Light, Gas and Water Division (MLGW) which requires MLGW to perform process analysis, design, procurement, construction, testing, operation, and evaluation of a plant that will demonstrate the feasibility of converting high-sulfur bituminous coal to industrial fuel gas (IFG) with a heating value of 300+30 Btu per standard cubic foot (SCF). The demonstration plant is to be based on the U-GAS[®] Process, and its product gas is to be used in commercial applications in Memphis, Tennessee.

To perform this work, MLGW established an industrial team, which includes -

- Memphis Light, Gas and Water Division (MLGW), Memphis, Tennessee: The prime contractor and distributor of the industrial fuel gas
- Foster Wheeler Energy Corporation (FWEC), Livingston, New Jersey: The engineer-construction manager
- Institute of Gas Technology (IGT), Chicago, Illinois: The process developer
- Delta Refining Company (DRC), Memphis, Tennessee: To provide operating experience.

The contract specifies that the work is to be conducted in three phases. Phase I costs are financed entirely by DOE. The phases are -

- Phase I. Program Development and Conceptual Design
- Phase II. Demonstration-Plant Final Design, Procurement and Construction
- Phase III. Demonstration-Plant Operation

Task II of Phase I requires the preparation of a demonstration-plant process design based on the U-GAS Process. The demonstration-plant process design consists of -

- Establish the basis for designing the demonstration plant and pipeline distribution system
- Process engineering design of the demonstration plant including preparation of the process flow diagrams
- The Task IX Report includes the development of operating and process data for the design of the demonstration plant based on U-GAS technology.

Program Objectives

The specific objectives of Task IX of the demonstration-plant program were -

- To determine the requirement of pretreatment to prevent caking of the Western Kentucky coal in the gasifier
- To determine operating conditions suitable to produce a gas meeting the contract specifications of 300+30 Btu/SCF
- To determine the operating window for the ash-balanced operation by removal of ash as low-carbon agglomerates
- To determine suitable conditions to recycle and gasify coal fines entrained from the fluidized-bed gasifier
- To provide support studies for an understanding of ash agglomeration phenomena, agglomerate discharge, ash chemistry and to provide design data for the demonstration plant.

To make the overall program cost effective and attain the program objectives as quickly as possible, the program was divided into the following two major categories:

- I. Pilot Plant Operations
- II. Pilot Plant Technical Support

The objectives of these two categories were classified and defined as follows:

I. Objectives of the Pilot Plant Operations Studies

- A. Process Viability
- B. Mechanical Viability
- C. Coke Shakedown Tests
- D. Pretreatment Needs
- E. Ash Agglomeration
- F. Fines Reinjection
- G. Coal Operating Range
- H. Extended Coal Tests
- I. Pressure Operations
- J. Environmental Data

II. Objectives of Pilot Plant Technical Support Studies

- A. Cold-Flow Model
 - 1. New Venturi Configuration for the Pilot Plant
 - 2. Different Size Venturies
 - 3. Ash Agglomerate Discharge Mechanism
 - 4. Pressurized Model
- B. Bench-Scale Test Work and Ash Chemistry (Petrographic and Coal and Coke Chemistry Studies)
 - 1. Effects of Variables on Ash Agglomeration
 - 2. (Effect of Ash Chemistry on Ash Agglomeration)

C. Design Data for the Demonstration Plant

The pilot-plant test program and the technical support studies were conducted concurrently to complement each other.

SUMMARY, DISCUSSION, AND CONCLUSIONS

Pilot-Plant Operations

A total of 16 tests were conducted at the U-GAS pilot plant during the 15 months of operation. The pilot plant program was broadly classified into three categories, which were characterized by the feedstocks. These were -

- Run-of-Mine (ROM) Western Kentucky coal (8 tests)
- Coke (3 tests)
- Washed Western Kentucky Coal (5 tests)

Three tests were conducted with coke breeze to calibrate the unit after the necessary modifications were made.

Run-of-Mine Kentucky Coal

A total of eight tests (Runs 118 through 125) were performed with the ROM coal. In the initial tests (Runs 118 through 122), the objective was to determine the operating conditions necessary for stable gasification and agglomeration of the unwashed Western Kentucky coal. These tests revealed that a higher superficial velocity in the range of 4 ft/s is necessary to maintain stable fluidization conditions in the bed when operating at 1850° to 1900°F with this coal. This velocity was almost double the value used for the design of the gasifier and all the peripheral equipment. Operation at these conditions imposed a severe strain on the coal feed lockhopper system, the cyclones, and the downstream equipment resulting in severe mechanical problems. To overcome this limitation, the gasifier internal diameter was changed from 4 feet to 3 feet after the first five runs.

In the next series of tests (Runs 123 through 125) with ROM coal, successful stable gasification for about 170 hours was obtained. Both steam-air and steam-oxygen gas mixtures were used for gasification. Though long periods of steady-state operations were logged during some of these runs, continuous bed withdrawal was required to maintain constant ash levels in the bed. Sufficient ash could not be agglomerated and withdrawn from the fluidized bed to maintain an overall ash balance. Several attempts to increase the ash agglomeration rate by increasing bed temperature (up to 1950°F) and bed ash content (up to 70%) were not successful. Therefore, a program direction change was initiated that led to several modifications on the pilot-plant configuration. These changes and modifications are discussed under coke runs in the next subsection.

During the operations with ROM coal, a total of 137 tons of coal were processed in the U-GAS reactor over 258 hours of operation. Fluidized-bed temperatures from 1818° to 1950°F, reactor pressures from 23 to 31 psia, and superficial velocities from 3.1 to 4.5 ft/s were utilized. Heating value of gas up to 268 Btu/SCF and coal conversion efficiencies up to 81% were achieved during these tests. In all these tests, coal could be fed directly into the reactor without any need for pretreatment. The most notable of these tests were Run 118, in which sinter- and clinker-free ash agglomerates were produced during a 10-hour period of operation, and Run 124, which was the longest continuous test run conducted with this coal (168 hours).

Modifications and Coke Runs

After Run 126, the mechanical design of the sloping grid and the ash-agglomerating section were changed to provide more flexibility in introducing oxygen to the system and, at the same time, to remove ash agglomerates from the fluidized bed. Necessary modifications needed for 50 psig operation were also completed at this time. The modifications in the latter category included installation of a high-pressure, larger capacity feed screw, a surge hopper for continuous feed, and high-pressure relief valves. Provisions were made to blend coke and coal mixtures to avoid sudden transition in the feed to the reactor. In addition, a product-gas water scrubber was installed to provide information for the design of the demonstration plant and also to detect trace quantities of impurities in the product gas.

After completion of all the necessary modifications and configuration changes, a successful hot shakedown of the pilot-plant system was accomplished during Runs 127 and 128 with metallurgical coke breeze as feed material. Stable conditions for coke agglomeration were produced continuously at operating conditions of 1900° to 1950°F bed temperatures, 4 ft/s superficial velocity, 4-foot bed height, and 30% to 40% bed ash content. In addition to calibrating the unit, a major accomplishment of these shakedown runs was the obtaining of information that the ash agglomerates could be produced from the feedstock at relatively lower bed temperature if all the oxygen entered the reactor through the ash agglomeration zone. Also, removing all the oxygen from the grid area and introducing it through the center jet eliminated sinter and clinker formation in the entire reactor area.

Washed Western Kentucky Coal

The five tests (Runs 128 through 133) performed with washed Kentucky coal were highly successful. The changes in configuration and mode of operation completely eliminated the problem of ash sintering or clinkering, and, after five tests, the grid and the fluidized-bed area were free of any ash buildup. Washing the coal reduced the ash

content from about 19% to 12% while the size consist and the fines content of the coal were relatively unchanged. The feed size of washed coal was -1/4 X 0 inch. All of the tests in this category were of long duration with actual operating time on coal of 75 to 150 hours in each test and were planned to yield specific design information for the demonstration plant.

During these tests repeated stable gasification and ash agglomeration was achieved with both steam-air and steam-oxygen mixtures. Over 300 tons of coal were processed in the U-GAS reactor over the operating period of 424 hours. Reactor bed temperatures ranged from 1820° to 1885°F, pressures from 24 to 59 psia and superficial velocities from 3.2 to 4.3 ft/s. The raw product gas heating value reached up to 285 Btu/SCF. Coal fines from both the first and the second cyclones were recycled, yielding coal utilization efficiency of about 95%. The coal feed system operated without any problem with this highly caking coal and was optimized to reduce the level of tar and oils in the product gas to trace levels. The scrubber was continuously operated during the latter tests and showed no buildup of any tars or oils.

The last test in the program, Run 133, was performed at 45 psig, the design pressure of the pilot plant. All previous tests were at lower pressures of 5 to 10 psig. During the test, the system pressure was varied to provide information on gasification and ash agglomeration. At the elevated pressure, the system operated smoother, was easier to control, and fluidization was improved. There was no problem in forming or discharging ash agglomerates at the higher pressure although the coal feed increased up to about 1800 lb/hr. The bed operating temperature and the superficial velocity also did not require any changes. Excellent heat and material balance enclosures were obtained during these long duration tests and the results verified the computer model prediction of bed volumes, oxygen requirement, and gas composition.

The complete details about the individual runs are given in Appendix A. These runs produced the necessary data to confirm the ash agglomeration concept and provided sufficient data for the demonstration plant design including the environmental data.

Conclusions Based on the Pilot-Plant Tests

The conclusions from the pilot-plant tests are summarized for each objective in this category. Complete details of the pilot-plant operations and test data backing up the conclusions are presented in Appendix A.

Process Viability (U-GAS)

- Successful gasification tests were performed with
 - Metallurgical coke breeze
 - Western Kentucky coal.

- The process is capable of gasifying and producing ash agglomerates from raw coal.
- Up to 285 Btu/SCF gas was produced.
- Continuous ash-balanced ash agglomeration was achieved with Western Kentucky No. 9 coal.
- The gasifier internals were free of ash deposits during last 5 consecutive coal tests.
- Over three quarters of dust was reinjected into the system and gasified.
- The reactor operated between 20 psia and 60 psia in the temperature range of 1820° to 1900°F.
- The process produced design-specification gas.
- The steam and oxygen requirement matches the design basis.
- The gas yields match the design basis.

Mechanical Viability

- No serious erosion or corrosion of gasifier internals except the internal cyclone which is a pilot plant peculiarity.
- The process withstands large instantaneous upsets caused by
 - Temperature
 - Pressure
 - Superficial velocity.
- The process had relatively simple shutdowns and start-ups.
- The gasifier withstands process upsets without loss of ash agglomeration.

Coke Shakedown Tests

- Successful shakedown and agglomeration techniques were demonstrated with a 7-day continuous run.
- The new venturi and center jet configuration was demonstrated.
- Centralized oxygen feed reduced the temperature required for ash agglomeration and avoided the ash deposits.
- At least a 50°F range for producing ash agglomerates was established.

Pretreatment Needs

- Over 400 tons of caking coal were processed with continuous feed lasting for up to 7 days without pretreatment of coal.
- The FSI ranged between 5 to 7.5.
- Pretreatment was not essential for U-GAS Process.

Ash Agglomeration

- Repeated ash-balanced ash agglomeration was demonstrated.
- Required temperature, velocity, and reactor configuration were determined for high-quality ash-agglomerate discharge.
- The operating window was established:
 - Temperature: 1825° to 1900°F
 - Velocity: 3.25 to 4.5 ft/s
 - Bed Ash: 40% to 70%.

Fines Reinjection

- Over three quarters of overhead fines were reinjected.
- Second-stage fines recycle feasibility was demonstrated.
- Up to 95% coal conversion efficiency was obtained with fines rejection.

Coal Operating Range

- The following parameters were defined for stable gasifier operation with the Kentucky coal:
 - Temperature range: 1825° to 1900°F
 - Velocity Range: 3.25 to 4.5 ft/s
 - Bed ash content: 40% to 70%
 - Pressure range: 20 to 60 psia.
- The process allows flexible oxygen feed.
- There was 80% to 100% ash content in the agglomerate discharge.

Extended Coal Tests

- Run 124 with ROM coal gasified 84 tons of coal during 168 hours of operation with coal.
- Run 130 with washed coal gasified 88 tons of coal during 106 hours of operation with coal.
- Run 131 with washed coal gasified 70 tons of coal during 104 hours of operation with coal.
- Run 132 with washed coal gasified 47 tons of coal during 74 hours of operation with coal.
- Run 133 with washed coal gasified 104 tons of coal during 153 hours of operation with coal.

Pressure Operations

- The gasifier operated for 4.5 days at a pressure near 60 psia.
- Flexible and stable operation was indicated.
- The process has simple pressure control.
- Ash agglomeration formation and discharge were not affected by pressure in the range of operating conditions investigated and the amount of ash agglomerates produced.
- 90% + coal utilization was achieved under pressure operations.

Environmental Data

- Scrubber water samples, gas samples, and ash agglomerate samples were procured for the environmental data base.
- Environmental data was normalized with the coal feed for trace elements, minor constituents, and priority pollutants.

Pilot Plant Technical Support

Cold-Flow Model Tests

A total of 94 tests were conducted on the atmospheric half-sectional cold model. A pressure cold model was designed, built, and operated specifically for the Demonstration Plant Program. Four quantitative tests were conducted on the pressure cold model. The tests can be categorized as follows:

- Low-Pressure Cold Model Tests
 - Tests with alumina spheres (agglomerates) and acrylic molding powder bed
 1. Tests to determine the effect of obstructions and aeration rate in a jet above the throat of a 3-inch X 3-inch venturi (Runs 48 to 54)
 2. Tests in a 4-1/2 inch x 2-inch venturi
 - a. Discharge rate studies (Runs 55 to 59)
 - b. Center jet studies (Runs 70 to 82)
 3. Tests in a 3-inch X 2-inch venturi (Runs 61 to 64)
 4. Tests in a 2-inch X 2-inch venturi (Runs 65 to 68)
 - a. Discharge rate studies
 - b. Defluidization in the expander cone studies
 5. Tests on a 6-inch X 30-inch venturi

6. Tests on a 6-inch X 2-inch venturi (Runs 83 to 132)
 - a. Discharge rate studies (Concentration and venturi velocity)
 - b. Classification studies (Effect of configuration of piping below throat studied)
 - c. Pressure fluctuation studies
 - d. Discharge rate studies (Bed height, and superficial velocity studies)
- Coal, Coke, and Char Tests
 1. Tests with coal agglomerates, coke breeze feed bed (Run 133 to 140)
 2. Tests with coal agglomerates, char bed (Run 141)
 3. Agglomerate Separation
 4. Agglomerate char and coke elutriation velocities
- High-Pressure Cold Model Tests
 - Equipment and facility design and construction
 - Operating Procedure Manual
 - Tests with with coal agglomerates and char (Runs HP-1 through HP-4)

Low-Pressure Cold Model Tests

The low-pressure cold-model facility consists of a half-sectional vessel with a Plexiglas viewing window over a venturi, grid, and fluidized-bed section. Facilities for solids handling and removal and dust removal (cyclone) are provided. A blower and pneumatic instrumentation and controls enable separate flows to the grid, venturi, and central jet (if used) to be maintained. Initial tests were performed with acrylic molding powder substitute-bed material and alumina simulated agglomerates. The properties of these materials are shown in Table 1.

Table 1. PROPERTIES OF MATERIALS USED IN COLD-MODEL RUNS 48 TO 132

	<u>Acrylic (Plexiglas) Molding Powder</u>	<u>Alumina Spheres</u>
Mean Particle Diameter, μ	460	2520
Particle Density, lb/ft ³	72	157
Bulk Density, lb/ft ³	36	53

A brief synopsis of this experimental work and the conclusions derived from it follows. Details of the tests and data from the tests run are presented in Appendix B.

Tests With Alumina Spheres and Acrylic Molding Powder Bed

Tests to Determine the Effect of Obstructions and Aeration Rate in a Jet Above the Throat of a 3-inch X 3-inch Venturi (Runs 48 to 54)

Tests were performed to determine the effect of obstructions and jet aeration rate in the 3-inch x 3-inch venturi. Runs 48 through 52 were conducted to determine if an obstruction in the venturi throat, which was created by insertion horizontally of a 1/4-inch diameter copper tubing or aeration tap flow (located just above the venturi throat), would have any effect on the spheres discharge rate. Aeration rates of 0 to 24.0 SCF/hr were set, and discharge rate data determined. Higher aeration rates resulted in lower discharge rates. Obstruction tests (Runs 48 through 51) resulted in no effect of obstruction as measured by the percent of venturi throat area occluded at venturi velocities between 45 and 60 ft/s. Lowered discharge rates were observed with obstructions of up to 11% of the throat area at venturi velocities below 45 ft/s. Discharge rates were in the range of 0 to 0.5 lb/min in these tests.

Tests on a 4-1/2 inch X 2-inch Venturi (Runs 55 Through 82)

a. Discharge Rate Studies (Runs 55 Through 59)

Runs 55 through 59 were conducted to determine the discharge rate parameters of the 4-1/2-inch X 2-inch venturi. The standard conditions for further tests of the initial bed mass, superficial velocity, and initial agglomerate concentration were developed in Run 57. We found that the discharge rate decreased linearly with increasing venturi velocity, and classification fell off sharply at venturi velocities below 30 ft/s (discharge pipe velocities of approximately 8 ft/s). The discharge rate varied approximately linearly with the bed concentration.

b. Center Jet Studies (Runs 70 Through 82)

A series of 13 tests (Runs 70 through 82) were conducted to determine the effect of a center half-tube configuration in a 4-1/2 inch x 2-inch venturi on agglomerates discharge rates. The runs were conducted with two center tube sizes (1-inch and 1.5-inch ID) located at varying distances from the top of the venturi throat (0 inch and 14 inches). Some of the runs were conducted at varying throat velocities (18 to 76 ft/s) with a constant center tube velocity (0-243 ft/s); others were conducted with a constant throat velocity (43 to 47 ft/s) and varying tube velocity (60 to 225 ft/s). We found that the center tube configuration or its gas flow rate has a moderate effect on the discharge

rate. We were not able to correlate the magnitude of this effect with the configuration or velocities employed in the tests, however.

Tests on a 3-inch X 2-inch Venturi (Runs 61 Through 64)

A series of four test runs (Runs 61 through 64) were conducted on a 3-inch X 2-inch venturi configuration to further study the effect of varying venturi length and diameter on the rate of spheres discharge. These runs further confirmed that discharge rate is directly proportional to spheres concentration in the bed and inversely proportional to venturi velocity. We also found that a larger throat length has a detrimental effect on discharge rate by comparison with the 3-inch X 3-inch data.

Tests on a 2-inch X 2-inch Venturi (Runs 65 Through 68)

a. Discharge Rate Studies (Run 65)

Run 65 was conducted to determine the discharge behavior of a 2-inch x 2-inch venturi configuration and also to investigate the effect of throat diameter. Comparing the results of 4 1/2-inch X 2-inch, 3-inch X 2-inch, and 2-inch X 2-inch venturi tests, we concluded that increasing throat diameters increases discharge rate.

b. Defluidization in the Expander Cone Studies (Runs 66 Through 68)

Runs 66 through 68 demonstrate the condition of defluidization above the 2-inch X 2-inch venturi. A condition of defluidization of the bed occurs at a high concentration of spheres or at low venturi velocities. These three test runs established that there is a definite minimum velocity that must be exceeded at high spheres concentration in the bed to avoid a state of defluidization above the throat. We developed a linear relation in terms of spheres concentration in the bed and venturi velocity.

Tests on a 6-inch X 30-inch Venturi

Only one test run (Run 69) was conducted with the 6-inch X 30-inch venturi. The main objective of this run was to compare the results with earlier data on venturi discharge rate and also to study the effect of venturi length. The results further confirmed our earlier conclusion that increasing throat length results in decreased spheres discharge rate.

A total of 59 runs (Runs 83 through 141) were conducted using a 6-inch X 2-inch venturi configuration to study the effect of various parameters on the rate of spheres discharge. A synopsis of these studies follows.

Tests on a 6-inch X 2-inch Venturi (Runs 83 Through 132)a. Discharge Rate Studies (Concentration and Venturi Velocity)

Initial test runs (Runs 83 to 92) conducted with the 6-inch X 2-inch venturi indicated high discharge rates but poor classification. The initial configuration (Runs 83 through 87) had a 12-inch diameter half section below the throat that was modified by inserting an 8-inch diameter half section converging zone (Runs 88 through 91). As this modification could not restore the desired classification, various other alterations to the discharge system below the venturi section were attempted to achieve the objective.

As a first attempt towards classification, the 12-inch full-round section below the venturi section was replaced by an 8-inch full-round section to increase the gas velocity. The first test (Run 92) conducted with this new arrangement showed excellent classification.

b. Classification Studies (Effect of Configuration of Piping Below Throat)

The poor classification observed in Runs 83 through 91 in the discharge was believed to be caused by the low gas velocities (5-9ft/s) below the venturi. We felt we could entrain the acrylic back into the bed by increasing the gas velocities below the venturi. The full 12-inch Sched. 40 pipe (Runs 83 through 91) below the venturi was replaced by a full 8-inch Sched 40. pipe (Run 92 and 93). This resulted in improving the classification to acceptable levels, without reducing the spheres discharge. To improve the classification beyond what was achieved in Runs 92 and 93, we installed a long, 8-inch diameter half-section pipe for Runs 94 through 97. Although the classification was improved, the spheres discharge rate was reduced below the levels obtained in earlier runs at the same venturi velocity conditions. We hypothesized that the 8-inch diameter full-section pipe below the venturi may have produced some unsteadiness at the venturi throat, which helped us obtain more spheres discharge in Runs 92 and 93 than in Runs 94 through 97.

c. Pressure Fluctuation Studies

To create unsteadiness below the venturi throat, a ledge was installed and Runs 98 through 101 were conducted. The results showed even lower spheres discharge than the previous two configurations. Because pressure fluctuations were believed to be the cause of this result, pressure fluctuation studies were made, as described below.

Pressure recording instruments were employed in Run 102 through Run 115, to record the pressure traces at and below the venturi throat. Runs 102 through 115 were duplicates of previous runs made. Some attempts have been made to statistically relate the frequency observed on the pressure traces to variation of discharge rates at fixed values of bed concentration, but no correlation could be found. The particular values of venturi and superficial velocities employed, however, do appear to influence the relative behavior of the configurations with respect to sphere discharge rate. At this time, we decided to pick the most favorable case, an 8-inch half-section pipe below the venturi, and to investigate the effects of bed heights on the spheres discharge using this case.

d. Discharge Rate Studies, Bed Height, and Superficial Velocity Studies

Runs 116-132 were conducted at bed heights in the range of 6 to 48 inches above the grid. The spheres discharge rate results were correlated with the bed spheres concentration and the bed height. The discharge rate was directly proportional to the spheres concentration and inversely proportional to the bed height. Additional statistical analysis improved the correlation by incorporating the effect of gas velocities.

Coal, Coke, and Char Tests

Tests with Coal Agglomerates, Coke Breeze Feed Bed
(Runs 133 Through 140)

Runs 133-140 were conducted using coke breeze bed material and coal agglomerates taken from the AAG pilot plant runs. The discharge rates in these runs were considerably higher than achieved in previous acrylic/alumina runs at the same conditions. The agglomerate discharge rate was directly proportional to the agglomerates concentration in the bed. The classification was poor because of the comparable sizes of coke and agglomerate particles used. Removing selected cuts (- 10 + 14 mesh) from the coke breeze improved the classification.

Tests with Coal Agglomerates, Char Bed (Run 141)

Run 141 was conducted using coal agglomerates and char at a 17.2 ft/s venturi velocity. The char used was very representative of the size distribution because it was taken from the final bed of AAG pilot plant Runs 152 through 155. The agglomerate discharge rate appeared to be directly proportional to the agglomerates concentration in the bed, and classification was reasonable for this particular venturi velocity. The classification presumably could be improved by increasing the venturi velocity. Run 141 was conducted at the same venturi velocity as that of Run HP-4. (See Appendix B, "Pressure Cold Model.") One-half of the discharge materials from the Run HP-4 were

used directly in this run. The discharge rate in the atmospheric cold model (Run 141) was about 4 times higher than that of the pressurized (approximately 60 psia) model.

Agglomerate Separation

It was necessary in the course of these studies to develop an inexpensive but accurate procedure for separating agglomerates from char or coke breeze of overlapping sizes. This was required to determine the AAG operating parameters (bed ash agglomerate concentration and C-4 discharge agglomerate concentration) and similar parameters for the cold model (pressure and atmospheric) when operated with coal, char, or coke. A fluidization apparatus was employed and appropriate techniques were developed. These were checked by ash analysis and visual means and were adopted for use in these studies.

Agglomerate, Char and Coke Elutriation Velocities

This was conducted to further aid in understanding the mechanism of venturi discharge and classification, to obtain fundamental data on the material used in the tests, and to correlate this information with AAG Pilot Plant observations. A 2-inch ID Plexiglas column and metering equipment to deliver an air flow of up to 50 ft/s were set up. Measurements of the cumulative amount of a sample elutriated from the column as a function of velocity were made. We conducted such tests on agglomerates, char, coke breeze, and composite mixtures of these materials. We selected several samples from Run 150 of the AAG Pilot Plant for testing in this manner. These results were examined in the light of the venturi velocity which was maintained in the pilot plant at the time these paired samples of discharge and bed materials were obtained.

High-Pressure Cold Model

Equipment and Facility Design and Construction

A high-pressure cold-model test facility was designed and constructed for the purposes of obtaining information on the venturi and gasifier design of the demonstration plant. The cold-model vessel is 4 feet in diameter and is equipped with a 45 degree conical perforated plate-type grid and a 6-inch X 2-inch venturi with a 1.5-inch diameter Sched. 40 central jet. Complete flow, pressure, and temperature instrumentation and recording apparatus was provided.

Operating Procedure Manual

A manual for operators of the cold-model test facility was prepared. Procedures for start-up, run, data collection, and shutdown, and equipment specification, maintenance instructions, and safety procedures were included.

Tests with Coal Agglomerates and Char

We conducted four tests (Runs HP-1 through HP-4) and one equipment debugging run (HP-0) on the pressure cold model. Coal agglomerates and Western Kentucky coal char bed material were used in runs HP-1 through HP-4. Superficial velocities in the range of 1.8 to 2.9 ft/s, venturi velocities in the range of 15 to 39 ft/s at pressures in the range 10 to 45 psig were reached in the tests. Excellent classification of agglomerates from char in the venturi discharge was obtained. We obtained discharge rates of up to 15 lb/min (900 lbs/hr) in these tests. The higher pressure of operation resulted in lower mass discharge rates at a constant value of venturi velocity. Higher venturi velocity also resulted in a lower discharge rate. A critical velocity for classification appears to exist. The latter two findings are in accord with our findings in the atmospheric half-sectional cold model.

Conclusions Based on the Cold-Model Tests

New Venturi Configuration for the Pilot Plant

The observations on the quantity of material being discharged and the quality of solids movement in the expander core and fluidized bed above convinced us that the AAG pilot plant should convert from the 2-inch X 2-inch venturi then employed to a 4-1/2 inch X 2-inch venturi. This was done preceding Run 127. All subsequent AAG pilot plant runs employed the 4-1/2 inch x 2-inch venturi. Actual results in the pilot plant are discussed in Appendix A of this report.

Different Size Venturis

Tests with 2-inch X 2-inch, 3-inch X 2-inch, 4-1/2 inch X 2-inch and 6-inch X 2-inch venturis led to the conclusion that larger venturis with identical throat length will result in greater discharge rates. Tests with the 3-inch X 3-inch and 6-inch X 30-inch venturis, when compared with the results of the 6-inch X 2 inch and 3-inch X 2-inch venturis, resulted in the conclusion that longer throat length results in lesser discharge rates at otherwise identical conditions.

Ash Agglomerate Discharge Mechanism

Our observations on Runs 48 through 141 led us to propose a mechanism for ash agglomerate discharge. Agglomerate and bed material in the vicinity of the venturi throat is discharged into the pipe beneath the venturi in a pulse-like manner. The momentum and terminal velocity related characteristics of the agglomerates enable them to be discharged into the collecting hopper. Bed material is reentrained into the bed through the venturi if the venturi flow in the discharge pipe is maintained at a high enough rate. Thus, careful attention to the design of the piping will have to be made. Support work has identified the velocities required for separation of coal for agglomerates and bed material of particular size ranges.

An empirical correlation of discharge rate data incorporating the effects of bed height, bed ash agglomerate concentration, and velocities (for the grid and the venturi) was developed for the test results of Runs 116 through 132. The dependence of mass discharge rate on bed agglomerate concentration appeared to follow a similar power law dependence although the power appeared to be somewhat greater. The latter conclusion was based on somewhat limited data, however.

Pressurized Model

We obtained excellent classification of agglomerates from a fluidized bed of char in four runs (HP-1 through HP-4) in the pressurized cold model over a wide range of pressures (10 to 46 psi) and venturi velocities (15 to 39 ft/s). Although discharge rates as high as 15 lb/min were achieved, additional work for proper design of the demonstration-plant gasifier venturi is suggested because of the limited amount of information which was obtained on this equipment. This was due primarily to equipment delivery delays of an associated project whose equipment is integral with the equipment on this project and the mechanical failure of the compressor during Run HP-1.

Bench-Scale Test Work and Ash Chemistry Studies (Petrographic and Coal and Coke Chemistry Studies)

Bench-Scale Test Work

To understand the phenomenon of ash agglomeration, a bench-scale test setup was refurbished for the purposes of this program. The objective of the tests was to treat a sample of char, coke, or coal in a bed fluidized with nitrogen at a set maximum temperature for a predetermined time interval and determine the extent (if any) of agglomeration or sintering that occurred. A 2-inch stainless steel reactor was used in these tests. Preliminary work for this task involved making measurements of the minimum fluidization velocity of the materials used to ensure that adequate fluidization was maintained.

An agglomerating-sintering envelope for coke breeze (as a function of temperature and ash content) was developed at fixed fluidization velocity. Preliminary results for washed Western Kentucky coal char were obtained. Agglomerates were produced in the bench-scale reactor with the Western Kentucky coal char. Petrographic and microscopic examinations were performed to aid in the analysis of these results.

Ash Chemistry Studies

The chemistry of agglomeration of ash was investigated by several methods, including optical and electron microscopy, chemical analyses, and fusibility tests. Agglomeration of ash into rounded beads occurred with relative ease in runs on coke; the agglutinating agent is an iron-rich, relatively low-melting silicate in which other ash is embedded. Agglomeration of ash from coal was more difficult, because most of the iron was present in the form of sulfide and thus unavailable for reaction with aluminosilicates (clay minerals). The gas composition in the reactor bed is thermodynamically favorable for oxidation of the ferrous sulfide at regions near the oxygen inlet only. Oxidation of the ferrous sulfide and agglomeration of ash from coal were achieved by changing the mode of oxygen feed to the reactor.

Conclusions Based on the Bench-Scale Test Work and Ash Chemistry Studies

Effect of Variables on Ash Agglomeration

A relationship between fluidized-bed temperature and level of ash (implicitly also the composition of the ash) and agglomeration or sintering was determined. This was done extensively for coke with preliminary results of the same nature for Western Kentucky coal char. The ash composition of coke was shown not to differ appreciably from this coal's ash in any case. Other variables, such as particle size and fluidizing velocity, remain to be investigated. A detailed description of these analyses and the results are in Appendix B.

Effect of Ash Chemistry on Ash Agglomeration

The ferrous sulfide in the ash of the coal must be reduced to ferrous oxide if a low-melting ferrous aluminum silicate, the essential bonding matrix of the agglomerates, is to be formed. The gas composition in the bulk of the reactor bed is thermodynamically unfavorable for oxidation of ferrous sulfide except at regions near the oxygen inlets. A high oxygen concentration needs to be introduced somewhere in the reactor for agglomeration of the ash to take place from this standpoint. The section entitled "Ash Chemistry in the U-Gas Process" in Appendix B fully details the study of the chemistry of coal ash.

Design Data for the Demonstration Plant

A major portion of the requirement for the pilot plant operations and pilot plant technical support has been to obtain data for the preliminary design of the demonstration plant for the specific Western Kentucky feed coal and to provide additional data and technical information for incorporation into the final designing of the demonstration plant. From the pilot plant runs described in earlier sections it has been amply demonstrated that the U-GAS Process is a viable technology that incorporates simple control; and sufficient data for the preliminary design have been obtained. The pilot plant operation was supported by cold-flow model, bench-scale studies, and ash chemistry studies to arrive at the final data for the demonstration plant. The design data appears in detail in Appendix D.

The design data appearing in Appendix D are based on the following observations during the operations of pilot plant and support facilities:

- The support program firmly established process feasibility and provided a strong data base for completing the preliminary demonstration plant design.
- Four consecutive extended-period tests were conducted wherein good quality medium-Btu gas and high-ash-content (80 to 97 wt %) ash agglomerates were produced from Western Kentucky coal.
- A technique of feeding caking coals directly into the gasifier without pretreatment was perfected.
- Stable operability of the gasifier while recycling entrained coal fines back into the gasifier under continuous agglomerating conditions was demonstrated.
- Data related to environmental aspects of the U-GAS Process – particularly wastewater characteristics – which indicate presence of only trace quantities of tar and oils were provided.
- The pilot plant was operated for more than 100 hours at pressures of up to 60 psia (gasifier design pressure is 65 psia) proving applicability of ash-agglomeration technique and ash-agglomeration discharge mechanism at moderate elevated pressures.
- A broad operating window was established for the major operating variables of temperatures, superficial velocity, and bed-ash content.
- Coal-ash chemistry and bench-scale studies were conducted to further elucidate the phenomenon of ash agglomeration.
- Studies on large-scale cold-flow model of the bottom ash-agglomerating section of the U-GAS reactor were conducted. This is a critical area of the design effort because little or no scale-up information is available from past experience.

The design data includes the specification of coal, start-up material, gasifier material balances, gasifier-operating conditions, fines characteristics, ash-agglomerating characteristics, scrubber water analysis, saltation velocities, and the operating procedures.

Besides the experimental observation of the pilot plant, the scale-up of the pilot plant is supported by the mathematical and parametric analysis. The synopsis of the scale-up calculations also appears in Appendix D. These calculations are of a preliminary nature, and it is hoped that extensive work on this line will be done during the next phase of the development of the overall program.

Towards the end of the program, there was concern regarding the flowability of coal chars and ash agglomerates from the ash hopper in the demonstration plant. In the pilot plant, these materials are collected dry whereas in the demonstration plant they are removed as water slurry. Therefore, we conducted several tests to study the flowability and fluidization characteristics of coal char and ash particles in a water medium. Details of these experiments are given in Appendix D. The results of the experiments indicate that no difficulty is to be expected in discharging the ash agglomerates from a water-filled tank, as proposed in the demonstration plant.

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APPENDIX A. Pilot Plant Operations

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PILOT PLANT OPERATIONS

Pilot Plant Tests With Run-of-Mine Kentucky Coal

At the request of the U.S. Department of Energy (DOE), the AAG Test Program (Contract EX-76-C-01-2336) was interrupted on December 31, 1977, to conduct support work for this medium-Btu Industrial Fuel Gas Program (Contract EF-77-C-01-22582) supported by DOE through the Memphis Light, Gas and Water Division (MLGW) of the City of Memphis, Tennessee. Beginning January 1, 1978, all AAG pilot plant tests have been conducted to gather data for the design of a U-GAS demonstration plant using Western Kentucky No. 9 coal to be built at Memphis, Tennessee. The flow diagram of the U-GAS pilot plant is given in Figure A-1.

Physical and Chemical Properties of Kentucky No. 9 Coal

Table A-1 contains the proximate and ultimate analysis and some physical properties of run-of-mine (ROM) Kentucky No. 9 bituminous coal. Table A-2 contains the size distribution of the ROM coal used in the following tests.

Test Runs 118-125 - January-June 1978Test Run 118 - January 1978Summary

During the first test with Western Kentucky bituminous coal, the ash agglomerates were continuously produced during the 10-hour coal feed period.

Details

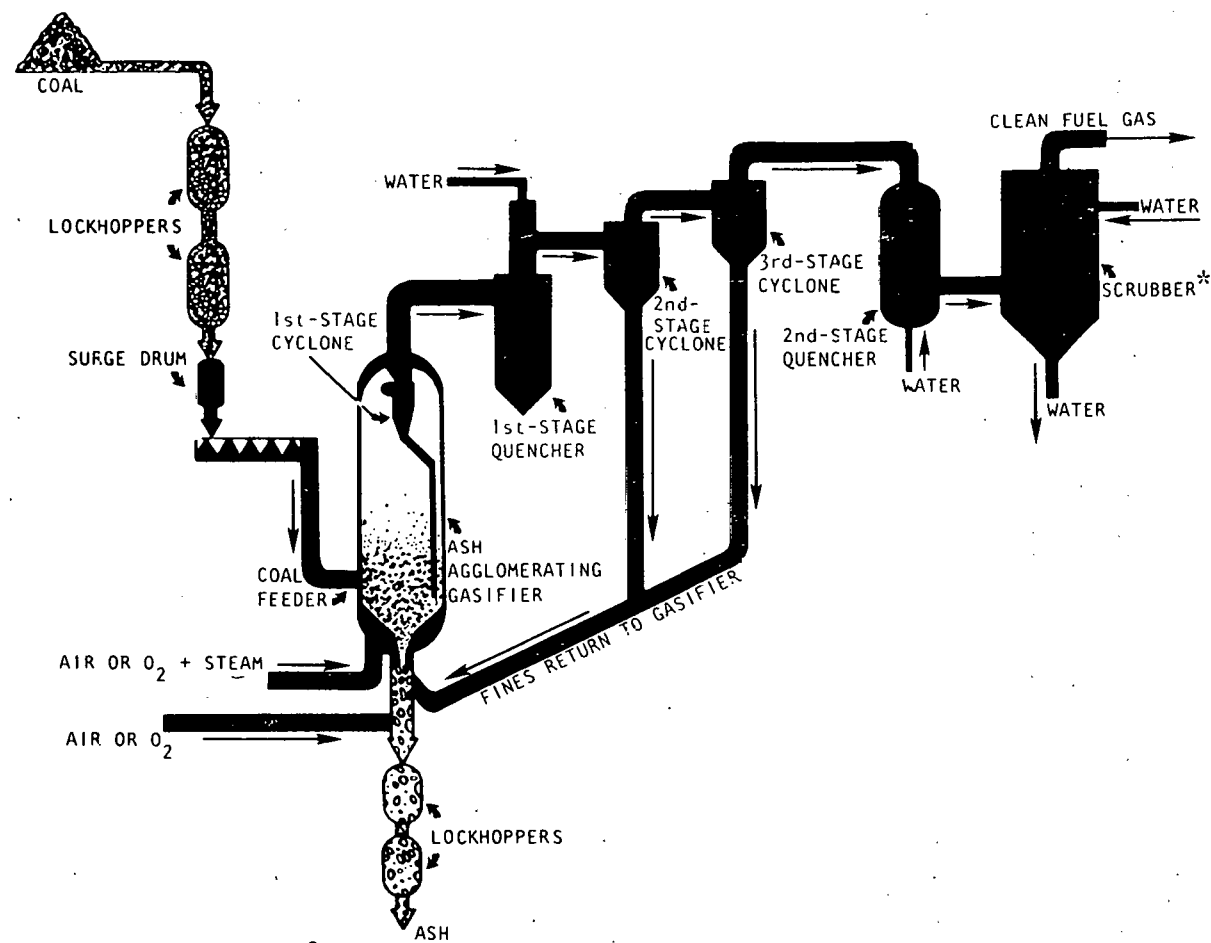
The test plan for Run 118 consisted of feeding and gasifying caking runs ROM Western Kentucky No. 9 bituminous coal at 4 ft/s superficial velocity and to produce and discharge ash agglomerate.

The total test run comprised 68 hours including 10 hours of continuous coal feed. Steam-air gasification produced an average operating temperature of 1846°F when coal was fed. No tar evolution was evident in the product gas or deposited on the cyclones. Ash agglomerates were produced during the coke feed (start-up) period just prior to introducing coal and continued during the entire 10-hour coal feed period.

Run 118 began on Saturday, January 21, 1978, and was brought up to the normal operating temperature (1850°F) around 1600 hours on January 22. After

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*Installed after August 1978.

Figure A-1. FLOW DIAGRAM FOR ASH AGGLOMERATING GASIFIER

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Table A-1. ANALYSIS OF WESTERN KENTUCKY ROM COAL

Proximate Analysis	<u>Wt. % (As Received)</u>
Moisture	1.6
Volatile Mater	34.4
Ash	18.9
Fixed Carbon	45.1
Ultimate Analysis	<u>Wt. % (Dry)</u>
Ash	19.18
Carbon	64.35
Hydrogen	4.41
Sulfur	4.60
Nitrogen	1.26
Oxygen	6.20
Bulk Density, lb/ft ³	60.8
Heating Value (dry basis) Btu/lb	11,570

Table A-2. SIZE DISTRIBUTION OF FEED MATERIAL (ROM Coal)

<u>Retained on USS Sieve Size</u>	<u>ROM Coal</u>
3	0.4
6	11.7
12	23.6
20	22.0
40	18.8
70	10.8
140	6.4
270	2.9
Pan	3.4

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some initial problems with the elevator screw feeder and a reactor screw feeder motor failure coke ash agglomeration was achieved and ROM coal was fed on January 23. The coke fluidized bed temperature of 1846°F was maintained with steam-air gasification when coal was fed. Ash agglomerates produced while feeding coke discharged continuously during the 10-hour coal feed period. The superficial velocity of 3.5 to 4 ft/s used, required a coal feed rate of ~1300 lb/hr. Test termination was required when the coal supply ran out at approximately 0400 hours on January 24.

An inspection of the reactor internals after cool down revealed a char deposit on the grid around the venturi, and ash deposits on the grid in the vicinity of the fines injection nozzle. An extreme case of erosion-corrosion severely damaged the internal cyclone. The damage was generally confined to the solids impinged areas of the barrel, upper sheet and outlet tube. As a result, a new cyclone design incorporating a refractory lining was expedited. Modifications were also made to increase the steam flow capacity to the reactor to facilitate steam-oxygen operation.

Test Run 119 - February 1978

Summary

In Run 119, through 168 hours of interrupted operation, ROM Kentucky coal was fed for only 2 hours. Solids feed system problems required modifications to the reactor screw feeder assembly.

Details

The test plan for Run 119 included operating with the ROM coal at 3 to 4 ft/s superficial velocity, reinjecting cyclone fines and gasifying in the steam-oxygen mode. When coal was fed, steam-air gasification produced an 1890°F bed temperature at 3.4 ft/s superficial velocity. Severe problems with the reactor screw feeder system forced a termination of the test run.

Run 119 was initiated on Monday, February 13, at 1400 hours. While operating with coke several feed system problems developed during the next several days. After rectifying those problems, the reactor was raised to the operating temperature of 1880°F at 0600 on Monday, February 20. At 0900 hours, coke feed was replaced by coal feed. At 1000 hours, second-stage cyclone fines, recycle was begun along with oxygen injection. The operation continued until 100 hours,

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at which time the reactor screw feeder jammed with solids. Due to the reoccurrence of this persistent problem, the test run was terminated.

Due to some evidence of sinter formation on the grid, the reactor's oxygen jet tube was extended to introduce the oxygen at the top of the grid directly above the venturi. The fines injection nozzle was extended horizontally, to 12 inches from the wall to minimize the possibility of concentrated fines contact with the grid surface. In other maintenance work, the air heater combustion blower blades were balanced to minimize a shutdown due to a bearing failure. The pitch (spacing) of the discharge end flighting of the reactor screw feeder was lengthened to improve material conveyability by the feeder. The reactor feeder discharge chute, was also altered to improve solids flow.

Test Run 120 - February-March 1978

Summary

During 107 hours of operation, coal was fed (very briefly) for about 1 hour only. A sudden drop in the reactor bed level and a quick sieve analysis of the second-stage cyclone fines, pointed to the non-functioning of the internal cyclone, possibly because of a plug or dipleg solids seal failure at the trickle valve. Because the fines carry over was excessive, the run was terminated to examine the internal cyclone.

Details

The test plan for Run 120 called for gasifying ROM Kentucky coal at an 1875° to 1900°F bed temperature and 4 ft/s superficial velocity with steam-air operation. Six oxygen-enriched steam-air gasifying medium was to be used to achieve coal ash agglomerates.

In Run 120, air-steam gasification produced an operating temperature of 1875°F with coke at a 4 ft/s superficial velocity. The internal cyclone performed inconsistently during the test. The balance of the test run was used to investigate the cause of the cyclone malfunction and to develop a solution to be applied during the subsequent tests.

Run 120 was initiated on Monday, February 27, at 1200 hours and the reactor was brought to an operating temperature (1850°F) on February 28 at 1400 hours. Shortly after 1700 hours, a fall in the reactor bed level was detected. At this time the internal cyclone pressure drop had fallen significantly (4P

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reduced by 1/3). Additionally, the cyclone dipleg pressure drop revealed that the dipleg was essentially empty. A sieve analysis of second-stage cyclone fines showed the presence of coarse particles; approximately 30% of the material was larger than 40 mesh (normally only 10% of the second-stage fines are larger than 100 mesh). Also the second-stage fines rate increased from 250 to 1000 lb/hr. These developments led to conclude that the internal cyclone was not functioning because of either a solids plug, excessive backgassing, or a broken dipleg. Hence the reactor was cooled down for examination.

The inspection did not indicate any mechanical damage to the cyclone, dipleg or trickle valve, hence the reactor was buttoned-up and the run was resumed. On Saturday, March 4, Run 120 was restarted and an operating temperature of 1838°F was achieved at 1400 hours on March 5. Operations with coke continued until 0900 hours on March 6 at 4 ft/s superficial velocity and steam-air gasification. Around 0925 hours the coal feed was initiated but at 1020 hours the coal feed was interrupted. Coke feed was resumed at 1230 hours on March 6, and one hour later the internal cyclone malfunctioned.

During the next 2 days, investigations continued with the assistance of personnel from Foster Wheeler Energy Corporation (FWEC) and Delta Refining Company (RFC) to determine the cause for the malfunctioning of the internal cyclone. At this time operations continued at 2.5 ft/s superficial velocity and 1850°F temperature with coke. As part of these investigations, the fluidized bed height was increased to 8 feet for 2-3 hours to determine the effect of immersing the trickle valve. The internal cyclone functioned under this condition, hence it was concluded that the trickle valve seal at the end of the dipleg was not effective. The test run was terminated at 0900 hours on March 9.

After the reactor cooled down, the internal cyclone was examined and found to be undamaged and to contain no solids accumulations. The trickle valve flapper was examined and it was determined that its seal was ineffective. To assure a positive seal, the trickle valve was removed and replaced with a reverse seal pot, wherein the dipleg is immersed in a fluidized bed of fines which is hung from the dipleg.

Test Run 121 - March 1978

Summary

ROM coal was fed for the last 4 of 23 hours of operation. The steam-air-

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oxygen temperature of 1910°F and a superficial velocity of 3.4 ft/s. Because of low ash discharge, the bed ash content rose from 30% to 67% within only 4 hours and resulted in a defluidization of the reactor bed.

Details

The test plan for Run 121 consisted of gasifying and producing a high ash discharge using ROM Kentucky coal while operating at a 1900° to 1910°F bed temperature and a 4 ft/s superficial velocity. Additional oxygen was to be used to achieve the desired bed temperature. In Run 121, the steam-air-oxygen mode of gasification produced a 1910°F bed temperature at 3.4 ft/s superficial velocity. Bed ash buildup caused defluidization, which eventually forced termination of the test.

Run 121 was initiated on Thursday, March 16, and the temperature reached 1905°F at 0800 hours on March 17. At this stage, a 3.4 ft/s superficial velocity was maintained using steam-air gasification medium. At 0835 hours coal feed replaced coke. A bed temperature of 1908° to 1910°F was maintained using an air-steam-oxygen mixture and a coal feed rate of 1200 lb/hr.

During this test the internal cyclone functioned well (second-stage fines rate 189 lb/hr). Total ash discharge at 1030 hours was considerably lower than the actual amount fed to the reactor. By 1200 hours the bed ash reached 67% and shortly thereafter the bed completely defluidized. Upon inspection, a deposit of black sinter was found covering the entire grid surface. To provide a means to operate at variable bed ash contents, without reaching critical levels, it was decided that an available alternate bed withdrawal system shall be activated to restrain the bed ash concentration to less than 50% by the removal of fluidized-bed solids.

Test Run 122 — March 1978

Summary

During 92 hours of test operation, ROM coal was fed continuously to the gasifier for 34 hours. The steam-air gasifying medium produced an operating temperature of 1800° to 1850°F. Bed ash content was maintained by a semi-continuous side withdrawal rate that averaged 250-300 lb/hr. Ash agglomerates were produced at a rate up to 100 lb/hr toward the end of the run. Eighteen hours after feeding coal, the internal cyclone stopped functioning but the test was continued. Later a failure in the automatic vent valve in the product

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gas line caused termination of the test.

Details

The test plan called for maintaining 40% to 45% bed ash content by achieving an overall ash balance through a continuous bed withdrawal while gasifying ROM Kentucky coal using air-steam medium at 1850°F and a 4 ft/s superficial velocity.

In Run 122, an operating temperature of 1800° to 1850°F was obtained when coal was fed using a steam-air gasifying medium. Steady-state gasification conditions were achieved, and high-ash solids were discharged at a rate of up to 100 lb/hr toward the end of the test. The bed ash content was maintained near 50% by a semicontinuous solids withdrawal from the bed at a rate that averaged 250 to 300 lb/hr. During the last 18 hours of operation, the internal cyclone malfunctioned, and a failure of an automatic vent or solids plug in the product gas line forced termination of the test.

Run 122 began on Sunday, March 26, and attained an operating temperature of 1830°F with coke at 1000 hours on March 28. While running with coke, the bed withdrawal system was calibrated. The discharged bed material was cooled down to 900°F with nitrogen. Coal feeding began at 0930 hours on March 29 and continued without interruption up to 0130 hours on March 30, whereupon increased second-stage cyclone fines and panel instrumentation indicated a plugged internal cyclone. The test run was continued without this internal recirculation of char fines until 1900 hours on March 30. A forced shutdown was necessary at this time because an automatic vent valve failed to open. The effective result was a complete stoppage of the product gas flow from the reactor until the rupture disc burst at 25 psig and terminated the test.

Upon cooling, a reactor inspection indicated a loosely attached deposit formed in the reactor. However, there were no indications of any high-ash deposits either on the grid or on the coal feed nozzle. Run 122 was followed by a one month refurbishing of the process system.

During the April 1978 turnaround the following changes were made to prepare the U-GAS pilot plant facility for 50 psig operation and improve the capacity compatibility of all components to the reactor.

- Support steel and refractory brick were installed to reduce the reactor diameter from 4 to 3 feet in the area of the fluidized bed.

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- A new internal cyclone was fabricated that included a refractory lining.
- A new bed withdrawal hopper (for 50 psi use) (C-7) was fabricated and installed.
- A revised second-stage fines reinjection system incorporating a reverse seal was fabricated and installed.
- The reactor screw feeder motor was changed to a two-speed unit.
- The new L-Valve type feed nozzle was installed on the reactor shell and piped to the screw feeder.
- The product gas scrubber foundation was poured and the scrubber was mounted.
- The butterfly valves needed to divert product gas from the scrubber to the incinerator or vent were installed. An electrical interlock between the two valves was also installed, such that one valve remains open at all times.
- Major instrument loop checks were completed.
- During this month, a FWEC team performed a Design Analysis, Risk Analysis and created a Configuration Control Procedure for modification of the U-GAS pilot plant equipment. All IGT equipment designs and installations were approved by the FWEC team. Also, a Safety Plan prepared by IGT was reviewed, revised, and approved by the MLGW Technical Subcommittee.

Test Run 123 - May 1978

Summary

In this first run with the new 3 foot diameter reactor, ROM Kentucky coal was fed for 38 of the 70 hours of operation. The gasification of mode was air-steam with a bed temperature of 1840°F and a superficial velocity of 4.3 ft/s. The bed ash content was maintained near 40% through use of the alternate bed discharge line. An instrumentation failure (erosion of an internally extended pressure tap) produced bed height measuring error. The reactor while being operated to maintain a 4 foot bed height was actually at 5 foot which completely immersed the internal cyclone dipleg. The turbulence of the fluid bed generated excessive cyclical stress causing the dipleg to break loose from the cyclone and fall into the bed. The test run was terminated when excessive fines carry over was detected.

Details

The test plan for Run 123 called for gasification with ash agglomeration of ROM Kentucky coal with steam and air at 1850°F and 4 ft/s superficial

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velocity. The internal cyclone was deactivated (sealpot aeration gas turned off) but the effect of the new expanded reactor freeboard on the fines carry over rate was to be determined. Air-steam gasification produced a 1840°F bed temperature at a superficial velocity of 4.28 ft/s. When the 40% bed ash level was reached, the alternate bed discharge line was activated because the instrument (pressure taps) failure immersed the cyclone dipleg and it fell onto the grid of the reactor. Upon evidence of high fines carryover and loss of bed level, the run was terminated.

Run 123 was initiated on Monday, May 8 at 1900 hours and a fluidized coke bed temperature of 1860°F was achieved at 1600 hours on May 9. Operations were switched to ROM coal feed at midnight on May 9. The bed ash level was allowed to reach 40%, and the bed withdrawal system was then activated. After a reactor feed loss on May 10 followed by a water leakage into the oil sump of the process compressor was corrected, coal feed was restored. At 2100 hours on May 10 the internal cyclone was activated. The use of the bed withdrawal line was continued to control the bed ash content between 40% to 50%.

Operations continued in this manner overnight from May 10 to 11 with an average 200 lb/hr of fines being removed from the system at a feed rate of 1038 lb/hr. At about 1040 hours on May 11, all pressure-drop readings for the internal cyclone fell to zero. Immediately, a drop in bed level and a simultaneous large increase in fines carryover to the external cyclone developed. An increase in feed rate and air flow was made and partially compensated for the drop in temperature to 1800°F.

During the balance of the test run, which was terminated at 2200 hours on May 11, a low (6 lb/hr) venturi discharge rate persisted. (It had been 20-25 lb/hr before 1200 hours on May 11.) The low discharge rate from the venturi gave evidence that something was blocking the venturi. At 2130 hours an examination of the bed view port revealed a darkened stagnant zone, and the unit was shut down.

After cooling a reactor examination revealed excessive erosion-corrosion of the internal cyclone outlet tube, several thermowells, and an extended pressure tap that essentially doubled the recorded bed height. The corroded thermowells were replaced and coated with an abrasion-resistant alumina. The dipleg (which was deformed) of the internal cyclone and the outlet tube were replaced. The diameter of the latter was decreased from 9 to 8 inches and

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was refractory-lined to resist erosion. To provide a smooth surface for the flow of bed solids along the gas distributor grid, the hex-head grid nozzles were replaced by 1/2-inch round bar stock having holes flush with the grid surface, directing the fluidizing gas vertically into the reactor. The use of an extended bed pressure tap was discontinued to measure the fluidized bed density.

Test Run 124 -- May-June 1978

Summary

Run 124 was by far the longest test run conducted with Western Kentucky No. 9 run-of-mine coal (305 hours of operation were logged). Coke ash agglomerates were produced at 1950°F using a 4 ft/s superficial velocity. Coal feed was introduced after lowering the operating temperature to 1875° to 1900°F while gasifying with oxygen-enriched steam-air at a superficial velocity of 4 ft/s. This gasifying mode also produced steady periods of coal operation at bed temperatures of 1820°F. Later, operations were conducted in the steam-oxygen mode at an average temperature of 1810° to 1815°F also with a superficial velocity of 4 ft/s. During operations with coal and steam-oxygen, the product gas heating value varied between 250 and 270 Btu/SCF. In both operating modes the bed ash content was maintained at 40% to 50% by means of the bed withdrawal system. The same procedure was used to operate at 50% to 60% ash content in the latter part of the steam-oxygen operating phase. When the coal silo ran empty and sufficient data had been obtained, a voluntary shutdown was initiated, and the test run was terminated. Upon reactor examination about 100 pounds of deposit was found partially covering the grid and the venturi throat. Sintered fines flakes were found lodged among the gas straightening vanes located below the neck of the venturi, so we decided to remove these vanes prior to Run 125.

Details

The test plan for Run 124 called for gasifying ROM Kentucky coal at 4 ft/s superficial velocity with steam-air, enriched air-steam, and the steam-oxygen mode of gasification. The bed ash content was to be maintained between 45% and 60% through using the alternate bed withdrawal line.

In Run 124 coal was fed for 168 hours, of which several periods of steady-state were achieved at a bed temperature near 1820°F. Initially the bed ash content was maintained at 40% to 50% and later at 50% to 60% by alternate

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withdrawal of bed material to achieve an overall ash balance. The product gas heating value varied between 250 and 270 Btu/SCF when operations were conducted in the steam-oxygen mode. A voluntary shutdown was initiated after 10 days of continuous operation.

Run 124 was initiated on Tuesday, May 23, at 2330 hours. After reactor preheating and maintaining 3 to 4 foot deep bed with coke, the start-up burner was ignited. After a slow bed temperature rise, a hot spot was observed on the reactor wall above the start-up burner nozzle. The reactor was shut down, drained of bed material, and opened for inspection. It was found that the start-up burner flame retention nozzle along with the reactor nozzle enclosing it had been partially destroyed. The burner flame had penetrated back through the light weight refractory and overheated the reactor wall. This damage was due to the use of an insufficient supply of cooling nitrogen within one annular space between the two nozzles. A solenoid valve with a direct read-out flow loop (orifice, transmitter, parallel indicator) was used to assure nitrogen flow at any time the burner was operational.

Run 124 was restarted on Saturday, May 27. The reactor was brought to an operating temperature of 1872°F around 1800 hours on May 28. The test run continued to 0600 hours on Monday, May 29, without any interruption. Early that morning, a failure in the drive mechanism of the reactor screw feeder occurred. The reactor was brought down to below char ignition temperature (1000°F) and idled. Tuesday, May 30, was spent locating parts, getting dimensions, and machining of a replacement part. The unit was repaired on Wednesday, May 31. It was restarted at approximately 1600 hours on May 31 and brought to a bed temperature of 1875°F at 2400 hours with air-steam operation. On Thursday, June 1, oxygen was added, and the unit temperature was raised to 1925°F around 0400 hours. The bed ash content was near 50%. At this point, the bed withdrawal system was activated to reduce the bed ash content to 30%. At 1800 hours on June 1 the bed temperature was raised to 1950°F by adding oxygen. Shortly after 2200 hours, feed problems developed because of a low plant air supply pressure. The feed was restored in approximately 3 hours. Significantly, ash agglomerates were produced from coke while operating at 1950° to 1975°F.

With the presence of ash agglomerates from coke and feed capability restored, operations were switched to coal feed at about 0230 hours on Friday,

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June 2. Test operations were now conducted at approximately 1875°F at a 4 ft/s superficial velocity and semicontinuous bed withdrawal to sustain a 40% to 50% bed ash content. Ash agglomerates first observed during coke feed at 2100 hours on June 1 continued to discharge during coal feed through 1100 hours on June 2. The feed rate was about 1000 lb/hr. Ash agglomerate discharge rate began to decline so the bed temperature was raised to about 1900°F on June 3 using oxygen-enriched steam-air. The bed ash content was maintained between 40% and 50% using the alternate discharge line, and the temperature was maintained for about 12 hours. The fines removal rate was about 160 lb/hr, and feed rate was 1275 lb/hr (0200 to 1600 hours on June 3). During this period some sinter formation was apparent because of the presence of sinter-like flakes discharging from the venturi. On June 4, operations continued at somewhat lower temperatures of 1870° to 1880°F (0200 to 1800 hours). The feed rate was reduced to 900 lb/hr, and the fines rate came down to 93 lb/hr, and the ash agglomerate discharge rate was zero. At 2000 hours on June 4 all venturi discharge ceased. An attempt was made to unplug the venturi, but this was not successful.

From 2000 hours on June 4 to 2000 hours on June 6 the bed temperature in the reactor was lowered to 1825° to 1850°F. By alternate bed discharge, the bed ash content was maintained at 40% to 50%, and the feed rate was approximately 800 lb/hr with the fines rate at 90 lb/hr.

Early on June 7, operations with coal were switched to steam-oxygen, and the test run was continued. Even though the venturi was blocked and ash agglomerates were not being produced, the test was providing very useful data on direct feeding of coal, fuel gas yields and composition, trace impurities in the product gas, and waste water composition. The bed ash content was maintained at 40% to 50% by alternate bed withdrawal, and the bed temperature was maintained at about 1820°F. The change in gasification mode produced an increase in feed rate to 1020 lb/hr, and operations continued up to 1200 hours. Shortly after 1200 hours, a feed interruption occurred because the lockhopper valves malfunctioned. The problem was corrected; coke feed was initiated, and the temperature was raised to 1827°F at about 1600 hours on June 7. Shortly after 1800 hours, coal feed replaced coke feed.

Gasification operations continued on June 8 at a temperature of approximately 1810°F with coal feed. The bed ash content was allowed to increase

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55% to 60%: The objective of increasing the ash level was to operate the unit at higher carbon conversions in the steam-oxygen mode. The feed rate was 1036 lb/hr, and the fines rate was approximately 124 lb/hr.

Shortly after 2200 hours on June 8 the internal cyclone was deactivated by turning off the aeration gas to the seal pot. The deactivation of the cyclone was evident by a decrease in the pressure drop across the cyclone accompanied by an increase in the fines carried over to the second-stage cyclone from 118 lb/hr to about 309 lb/hr. This operating objective was to ascertain the effectiveness of this internal cyclone. After 0200 hours on June 9 the internal cyclone was reactivated, and the recycle of external fines collected in the second-stage cyclone was started. The recycle of second-stage cyclone fines worked satisfactorily after some adjustment in the reverse seal gas rate.

At about 1200 hours on June 9 the coal supply was exhausted, and an orderly shutdown was initiated to terminate the test run at 1900 hours on June 9.

Examination of the reactor upon cooling revealed that some sintering had occurred, and, in some areas, clinkers had formed. Approximately 100 pounds of sintered clinker were removed from the venturi and grid areas. Complete examination of the internal cyclone showed no evidence of erosion-corrosion on any portions of the outer or inner shell, but showed some accumulation of ash particulate on the cyclone refractory lining itself, mostly on the scroll inlet and on a small area of the outlet tube. These materials were readily removed by light scraping.

The internal cyclone reverse seal pot contained only dust and a few of the scraped-down ash particles mentioned above. There was no damage to the grid, start-up burner, thermowells, pressure taps, or feed nozzles.

The oxygen injection pipe, which enters through the grid and terminates above the venturi, was severely eroded on its bottom side directly above the center of the venturi. Accordingly, this pipe was relocated such that it projects through the 45° conical grid directly up a distance of 4 feet normal to the grid surface and terminates at that point. An examination below the venturi neck revealed that some sintered fines discharging during the test run through the venturi had apparently accumulated among the gas straightening

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vanes located below the neck of the venturi. It was decided to remove these vanes prior to Run 125.

The sintered material, about 100 pounds, had permitted operation without any hindrance; this indicated that the deposit was not growing continuously during the test but was formed early in the test when operating at a relatively high temperature with coal. After lowering the temperature, the deposit growth was checked, and we were able to continue the test for several more days.

During Run 124, two malfunctions in the coal feed scale cycling system had occurred. Some gas (N_2) leakage at an automatic block valve flange had been detected and was rectified by improving the gasketing at that point. Upon recommendation from IGT's operations people, indicating lights are being furnished on the control panel illustrating the valve positions in this scale system so that the cause of any malfunction resulting in the loss of feed to the reactor can be more rapidly diagnosed.

Figure A-2 provides the time summary of Run 124, and Figures A-3 (a,b) and A-4 (a,b) provide a material and energy balance summary for two time periods from Run 124.

Test Run 125 — June 1978

Summary

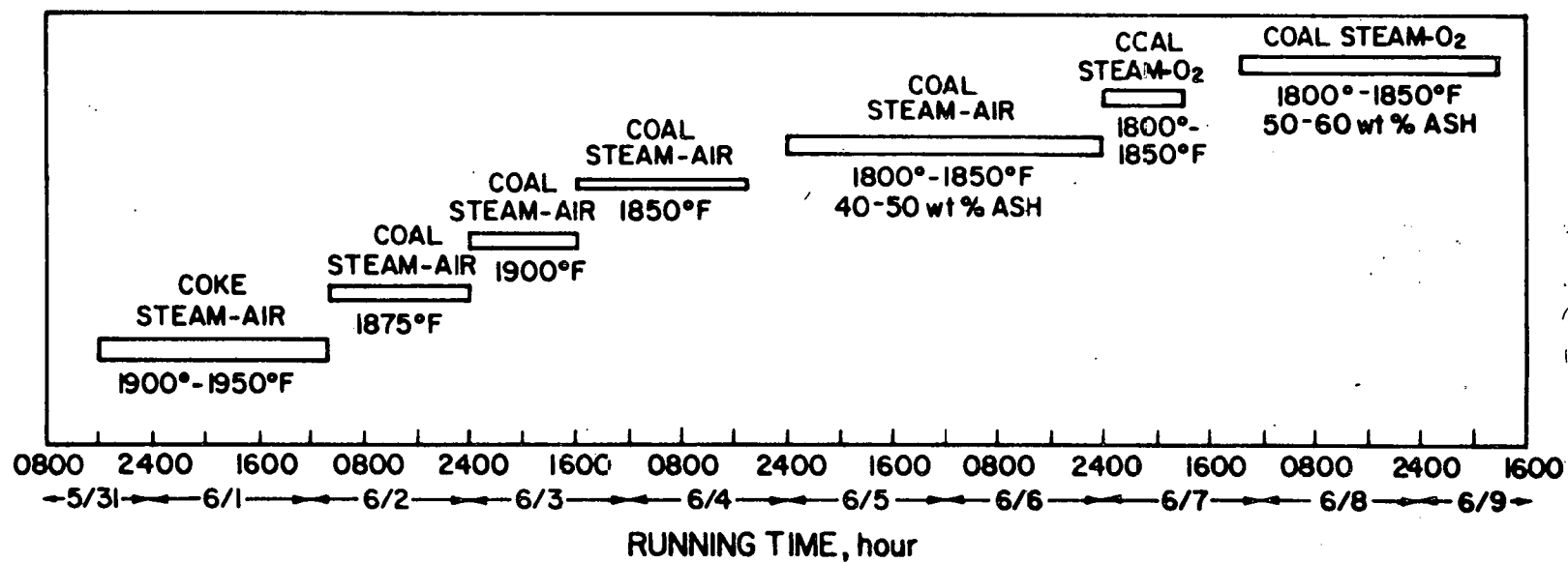
In Run 125, coke was fed to the gasifier for most of the 143 hours of operation, and coal feed was limited to about 1 hour. Using the steam-oxygen mode of gasification, ash agglomerates were produced from coke at a temperature of 1950°F and a superficial velocity of 4.3 ft/s. After gasifying and agglomerating coke ash continuously for 24 hours, coal was fed to the gasifier at the same operating conditions. However, the bed defluidized after only 1 hour of coal feed.

Details

The test plan for Run 125 consisted of determining the extent of ash agglomeration of ROM Kentucky coal at the same operating conditions that are required for coke ash agglomeration. The test plan requirement called for producing ash agglomerates from coke using the steam-oxygen gasification mode, a deactivated internal cyclone, and a 4 ft/s superficial velocity. Upon achieving ash agglomeration with coke for 24 hours we switched to coal and produced ash balanced operation (maintaining bed ash content below 60%). The

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Figure A-2. TIME SUMMARY OF RUN 124

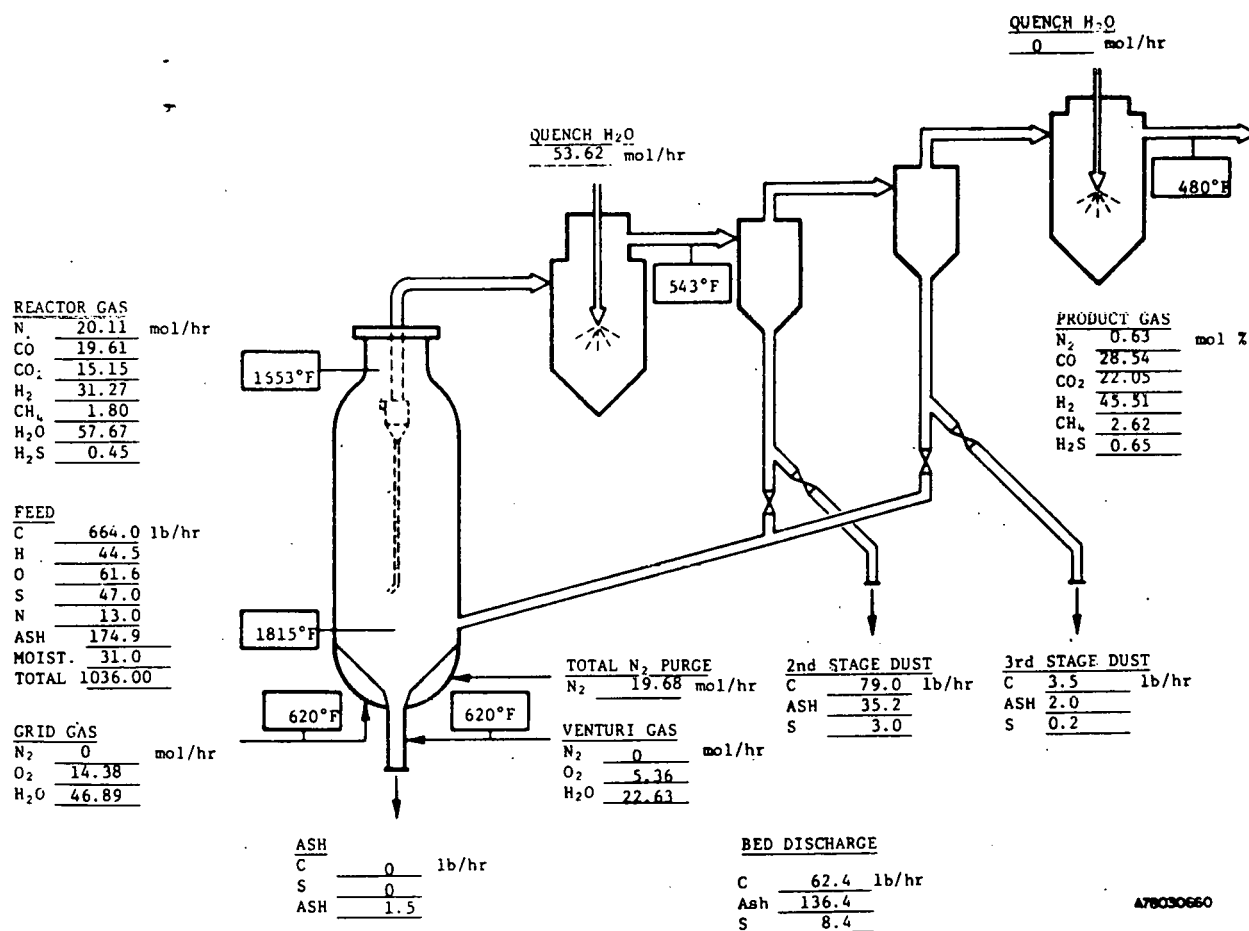


Figure A-3a. MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 124
(6/8/78; 0200-2200 Hours)

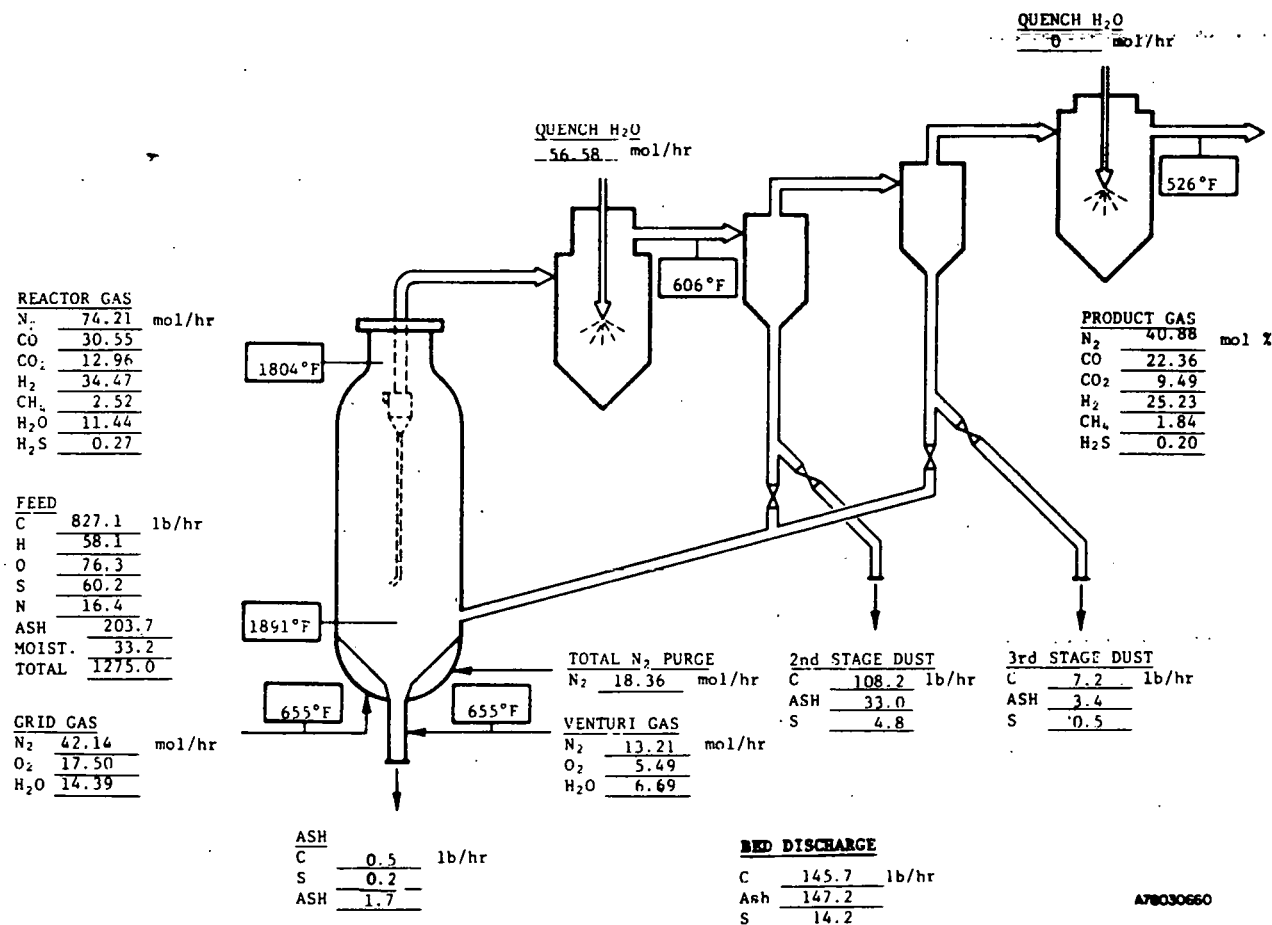
	ASH	CARBON	NITROGEN	OXYGEN	HYDROGEN	Btu/hr
Feed W. Ky #9 Coal	174.86	663.95	12.96	89.16	47.94	11,916,341
To Venturi	0	0	0	533.60	45.26	259,361
To Grid	0	0	0	1223.84	93.78	768,930
To Reactor	0	0	108.92	0	0	0
Purges	0	0	442.12	0	0	0
TOTAL INPUT, lb/hr	174.86	663.95	564.00	1846.6	186.98	12,944,632
Wall Losses	0	0	0	0	0	250,000
Venturi Discharge	1.52	0	0	0	0	978
Bed Discharge	136.36	62.37	0.37	0	0.40	1,067,421
2nd Stage Fines	35.16	78.99	0.49	0.12	0.58	1,247,598
3rd Stage Fines	1.99	3.54	0.02	0.02	0.03	56,327
Bed Inventory Change	-3.35	-3.03	-0.02	0	-0.02	-49,181
Reactor Gas	0	438.72	563.08	1721.28	185.98	10,045,862
TOTAL OUTPUT, lb/hr	171.68	580.59	563.94	1721.42	186.97	12,615,005
NET (OUTPUT-INPUT)	-3.18	-83.36	0	-125.18	0	-329,627
% Balance, Output/Input	98.18	87.44	100	93.20	100	97.45

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Figure A-3b. ASH AGGLOMERATING GASIFIER OPERATING CONDITIONS
 SUMMARY FOR RUN 124
 (6/8/78; 0200-2000 Hours)

Run No. <u>124</u> Date <u>6/8/78</u> Time Period <u>0200</u> to <u>2200</u>		
Venturi Diameter <u>3</u> inches Feed Material <u>W. Ky No. 9 Coal</u>		
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>		
1	Average Fluidized Bed Temperature, °F	1815
2	Average Reactor Pressure, psig	7.66
3	Venturi Pressure Drop, psi	0.35
4	Superficial Velocity, ft/s	5.28
5	Freeboard Velocity, ft/s	5.97
6	Venturi Velocity, ft/s	56.93
7	Grid Gas, mol/hr	61.27
8	Grid Gas Composition: %N ₂	0
9	%O ₂	23.47
10	%Steam	76.53
11	Venturi Gas, mol/hr	27.99
12	Venturi Gas Composition: %N ₂	0
13	%O ₂	19.15
14	%Steam	80.85
15	Fluidized Bed Density, lb/ft ³	28.88
16	Fluidized Bed Height, ft	4.89
17	Feed Composition, <u>1036.0</u> lb/hr, %C/%Ash	66.07/17.40
18	Average Fluidized Bed Composition, <u>6.61</u> lb/hr, %C/%Ash	45.83/50.73
19	Bed Discharge Composition, <u>207.9</u> lb/hr, %C/%Ash	30.00/65.59
20	Venturi Discharge Composition, <u>1.52</u> lb/hr, %C/%Ash	0/100
21	2nd Stage Fines Composition, <u>118.3</u> lb/hr, %C/%Ash	66.77/29.72
22	3rd Stage Fines Composition, <u>5.75</u> lb/hr, %C/%Ash	61.60/34.54
23	Carbon/Oxygen (Weight Basis)	1.05
24	Carbon/Steam (Weight Basis)	0.53

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	ASH	CARBON	NITROGEN	OXYGEN	HYDROGEN	Btu/hr
Feed W. Ky #9 Coal	203.66	827.07	16.39	105.84	61.80	14,902,821
To Venturi	0	0	369.88	282.72	13.38	168,409
To Grid	0	0	1179.92	790.24	28.78	428,209
To Reactor	0	0	107.24	0	0	0
Purges	0	0	406.84	0	0	0
TOTAL INPUT, lb/hr	203.66	827.07	2080.27	1178.80	103.96	15,499,439
Wall Losses	0	0	0	0	0	250,000
Venturi Discharge	1.74	0.45	0	0	0	11,196
Bed Discharge	147.15	145.73	1.49	0	0.87	2,400,230
2nd Stage Fines	33.03	108.19	0.87	0	0.62	1,710,932
3rd Stage Fines	3.43	7.15	0.05	0	0.05	116,060
Bed Inventory Change	-8.34	-7.56	-0.04	0	-0.04	-134,283
Reactor Gas	0	552.36	2077.88	1086.56	102.44	11,462,708
TOTAL OUTPUT, lb/hr	177.01	806.32	2080.25	1086.56	103.94	15,816,841
NET (OUTPUT-INPUT)	-26.65	-20.75	0	-92.24	0	317,402
% Balance, Output/Input	86.91	97.49	100	92.18	100	102.0

Figure A-4a. MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 124
(6/8/78; 0200-1600 Hours)

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Figure A-4b. ASH AGGLOMERATING GASIFIER OPERATING CONDITIONS
SUMMARY FOR RUN 124
(6/8/78; 0200-1600 Hours)

Run No. <u>124</u> Date <u>6/3/78</u> Time Period <u>0200 to 1600</u>		
Venturi Diameter <u>3</u> inches Feed Material <u>W. Ky. No. 9 Coal</u>		
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>		
1	Average Fluidized Bed Temperature, °F	1891
2	Average Reactor Pressure, psig	9.2
3	Venturi Pressure Drop, psi	0.56
4	Superficial Velocity, ft/s	4.88
5	Freeboard Velocity, ft/s	6.61
6	Venturi Velocity, ft/s	48.82
7	Grid Gas, mol/hr	74.0
8	Grid Gas Composition: %N ₂	56.92
9	%O ₂	23.64
10	%Steam	19.44
11	Venturi Gas, mol/hr	25.4
12	Venturi Gas Composition: %N ₂	52.03
13	%O ₂	21.62
14	%Steam	26.35
15	Fluidized Bed Density, lb/ft ³	27.25
16	Fluidized Bed Height, ft	4.60
17	Feed Composition, <u>1275.0</u> lb/hr, %C/%Ash	66.60/16.40
18	Average Fluidized Bed Composition, <u>-16.8</u> lb/hr, %C/%Ash	45.05/49.71
19	Bed Discharge Composition, <u>309.4</u> lb/hr, %C/%Ash	47.10/47.56
20	Venturi Discharge Composition, <u>2.4</u> lb/hr, %C/%Ash	18.80/72.56
21	2nd Stage Fines Composition, <u>147.5</u> lb/hr, %C/%Ash	73.35/22.39
22	3rd Stage Fines Composition, <u>11.2</u> lb/hr, %C/%Ash	63.85/30.61
23	Carbon/Oxygen (Weight Basis)	1.12
24	Carbon/Steam (Weight Basis)	2.18

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internal cyclone was to be activated if ash balanced operation was achieved without bed withdrawal.

Start-up was initiated on Monday, June 19, after accumulating a sufficient supply of coal over the previous weekend. While preheating with a coke fluidized bed two problems were noted. Oil and smoke were detected leaking from the bottom of the reactor. In addition, even though adjustments were made in the process gas flow to the venturi of the reactor, solids downflow could not be reduced, and particle size classification could not be achieved.

Investigation of the oil leakage problem revealed that the rental compressor had developed excessive piston ring wear. This compressor was shut down, and the alternate rental compressor was put on line.

The bed temperature rise was restrained while alterations were made to the venturi gas flow to achieve normal control of solids downflow and particle size. It was possible to reduce solids downflow to a minimum of 25 lb/hr, but there was an almost complete lack of particle size classification. It appeared that the revamped venturi (straightening vanes removed) would be incapable of ash agglomerate classification at the operating conditions. Accordingly, the unit was shut down, the bed removed, the venturi dismantled, the vanes reinstalled, and the unit reassembled all in 24 hours.

The run was restarted on Tuesday, June 20, and an operating temperature of 1853°F was achieved at approximately 1000 hours on June 21. The reinstalled vanes restored the venturi capability to size-classify bed material, although the velocities required were higher than those previously used. Operations were switched to the steam-oxygen gasification mode, and the temperature was raised to 1950°F by 1300 hours as per the test plan. Agglomerates began discharging at approximately 1500 hours, and temperatures were controlled at 1950° to 1960°F. The bed ash content rose to an average of 35%, and the bed sample composition included some agglomerates. Calculations made at 1900 hours produced a near-complete ash balance. Operations were being conducted with the internal cyclone deactivated. These conditions were maintained until 2400 hours on June 21, at which time a 1-hour feed interruption occurred. Steady operations were restored within an hour at 1950°F, and ash agglomerates were again discharged.

After another 10-hour steady-state period of ash agglomeration, coke feed was replaced by Western Kentucky coal at approximately 1200 hours on June 22.

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Shortly thereafter, a feed system electrical malfunction stopped feed to the reactor. The unit was cooled down and idled. During this period of idling (6 hours), problems with reheating of the reactor occurred. After several hours, it was diagnosed that water was leaking from the waste-heat boiler test loop. After draining considerable excess water from the downstream gas cooling and cleanup system, operations were restarted at 1100 hours on June 23. As temperatures of 1700° to 1800°F were reached, the charts on the panel board indicated a partial blockage of the venturi (1600 hours). Several attempts were made to achieve some solids discharge; finally, some large pieces of refractory were produced. Chart traces indicating normal operation were still not attained. Because refractory was still present in the venturi, temporary shutdown was initiated. The reactor was cooled down and opened, and a large quantity of refractory was removed from the venturi area.

The test was restarted on June 25 and was brought up to 1950°F using the steam-oxygen gasification mode. After steady-state ash balance conditions were achieved and ash agglomerates were produced with coke, the feed was switched to coal. This change was made on Monday, June 26, at 1700 hours.

As coal entered the reactor, the temperature began dropping at approximately 10°F/min, and significant increases in the oxygen supply were required to maintain the bed temperature above 1930°F. Simultaneously, the feed rate had to be increased to sustain bed level. Combined with reductions in the steam supply to hold the superficial velocity at about 4 ft/s, the oxygen concentrations reached 47% at the venturi and 32% at the grid. Within 10 minutes of reaching these concentrations and a temperature of 1940°F (at approximately 1800 hours) char recordings of the bed density, bed height, and venturi pressure drop began to give strong evidence of bed defluidization by means of a diminished magnitude in the trace fluctuations. A quick examination at the bed view port substantiated this conclusion: A darkening stagnant zone was seen. The reactor was therefore cooled, and approximately one-half of the bed was dropped out of the reactor through the side discharge line. The venturi was plugged, so the remainder of the material in the bed was cooled using nitrogen. Upon removal of residual char by vacuuming from the lower manway, approximately one 55-gallon drum of clinker was removed. The clinker was uniformly distributed in the bed, covering both the grid and the venturi area.

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After Run 125, some refractory repair work was done in the reactor free-board area and some parts of the internal cyclone.

Pilot Plant Tests with Coke Breeze

Test Runs 126-128 - July-October 1978

Test Run 126 - July 1978

Summary

Run 126 was also of short duration. The steam-air-oxygen mode of gasification produced coke ash agglomerates at temperatures near 1950°F. Agglomerate production was observed and recorded for a 3-hour period at a superficial velocity of 4.2 ft/s. A sudden defluidization and clinkering of the bed occurred while coke ash agglomerates were still being produced, forcing the termination of the test run. It appears that the use of oxygen jets projecting 4 inches above the grid surface may have created excessively high heat-release rates, which in turn produced clinkering.

Details

The test plan for Run 126 consisted of determining the extent of ash agglomeration achievable as a function of temperature and bed ash concentration for ROM Kentucky coal. Also, to produce ash agglomerates from coke at 1900° to 1950°F, with a deactivated internal cyclone, and 4 ft/s superficial velocity. Then introduce mixed feed at a lower temperature (1875°F) and finally operate with coal under steam-oxygen mode. Finally, activate the internal cyclone to determine its effects on the performance of the system.

Run 126 lasted for 79 hours during which no coal was fed to the reactor. Coke ash agglomerates were produced at 1950°F at 4.2 ft/s superficial velocity while operating under steam-air-oxygen mode. An apparent sudden defluidization of the bed while coke ash agglomerates were still being produced forced the termination of the test run.

Run 126 was begun on Saturday, July 8, and the newly installed refractory was dried out for about 2 days.

Coke feed was begun at approximately 1900 hours on Monday, July 10, and a temperature of 1900°F was reached at approximately 0500 hours on July 11. The unit was being operated in the steam-air mode with air only being directed through the venturi of the reactor and a mixture of air and steam being directed

through the grid to limit the oxygen concentration to approximately 15%. Oxygen was then added by jet injection during coke gasification to bring the temperature to 1950°F. Ash agglomeration using the jets during coal gasification would not be an operational mode change. By 0830 hours ash agglomerates were observed in the solids discharge from the venturi. By 0930 hours a rate of approximately 30 lb/hr was discharging with a feed rate of approximately 750 lb/hr of coke. Upon achieving the same ash discharge rate of a relatively high purity at 1030 hours, an ash balance was achieved. At this time the traces on the panel board charts of bed density and venturi pressure drop suddenly dropped, indicating that defluidization had occurred.

Efforts were made to loosen the venturi through the injection of steam, but they were not successful. The unit was brought down in temperature, and a total of 1000 pounds of bed material was removed through the side bed discharge.

This bed material contained many agglomerates and agglomerate clusters in addition to coke. Some of these clusters had fused together; some were just lightly attached and could be broken apart. The bed was examined and found to contain a considerable amount of residual agglomerates, char, and a large circular clinker that consisted primarily of ash agglomerates fused together with some char attached. The predominance of ash agglomerates was not unexpected because, during the test, the bed samples contained 39% ash before the unit was shut down. It was also observed that the bed samples contained many ash agglomerates. Some of the clusters of agglomerates were in the oxidized state; there was some evidence of clinker formation around the oxygen jet, and a heavy glassy clinker was found in the venturi. It appears that the oxygen jets, which had been used for the first time since Run 116, may have caused the clinkering and bed defluidization.

During the month of August, the AAG pilot plant was essentially prepared for 50 psig operation. All the modifications recommended in "AAG Pilot Plant - Design Analysis, Risk Analysis, and Configuration Control" were completed. Major changes included installation of the product gas scrubber, high-pressure reactor screw feeder, relief valves, and modifications on silo screw feeders and weigh-scale feed system.

The grid and venturi configurations were also changed. This latter change allows greater flexibility for the introduction of oxygen. The pure oxygen jet

(3/8" pipe) entering through a nozzle at the grid was replaced with a 1-1/2 inch pipe nozzle located at the center of the venturi. This center jet projects up to the bottom cone of the conical grid. To provide sufficient capacity for agglomerate removal, a new 4-1/2 inch diameter venturi was installed. A new gas flow loop (9405) was installed to supply process gas directly from the air heater to the central jet. This loop is in parallel with the 906 and 905 loops that lead, respectively, to the venturi (now annular space) and the grid. An additional loop (9303) supplies oxygen in parallel with the 9405 loop, so that flow from both loops enters the 1-1/2 inch diameter central jet.

A comprehensive "AAG Pilot Plant Training Manual" was prepared for use by engineers, technicians and others needing to know about the operation of the pilot plant.

This training manual covers the operation of the AAG Pilot Plant. All AAG reactor procedures and checklists have been included to facilitate understanding the process operation. Detailed descriptions and representations of the material (solids) handling systems, major instrumentation and control loops, and sampling procedures are included herein. Methods of data acquisition, reduction, analysis, storage, and reporting are briefly described. The configuration control and procedures employed to document process modifications are also specified. The operation and care of the solids handling equipment is mentioned so that maintenance problems may be reduced. The safety of the facility and the use of safe procedures is an important requirement of any pilot plant. The safety plan has been reviewed and updated to meet industrial standards. To attune all operating and staff personnel to process awareness and safe practices, the post-run meeting, scheduled to document verbal debriefing of the plant operators, has been stressed.

This training manual is intended for the instruction of new shift engineers and technicians. It can also serve as a "review" document for all experienced AAG operating personnel. The manual provides an accurate

description of the operation of the Ash Agglomeration Gasifier System including project report generation from the data produced. It is intended that utilization of the Configuration Control Procedures will continuously update this manual as changes are made in the AAG system.

Prior to Run 127, the following cold shakedown tests were conducted with the U-GAS Process equipment:

- The new start-up burner, suitable for pressure operation, was tested and operated up the heat input range required for bed ignition.
- The air heater was operated in the automatic control mode (for improved temperature control) after installing the new gas and air ratio regulators.
- Weigh screw feeders A and B were calibrated, and graphs were prepared to determine solids flow rate depending on the speed setting of the adjustable gear reducer. Both feeders can be adjusted from 500 to 3500 lb/hr, which will enable feed ratios of coal to coke to range from 12% to 100%.
- A speed change was made on the reactor screw feeder to yield adjustable feed rates from 450 to 2700 lb/hr.
- The reactor system was pressure-tested to 50 psig; small leaks were detected at the reactor screw feeder flanges and at two thermowells. All these connections were tightened to eliminate these leaks.
- Cold tests from the AAG system also included using process air at pressures up to 40 psig while operating the reactor screw feeder to obtain performance data for the newly installed 4-1/2 inch diameter venturi and 1-1/2 inch central jet pipe. Venturi annulus discharge rates were obtained at annulus velocities from 27 to 60 ft/s.
- To prevent possible venting of reactor gas in case of malfunctions within the scale system lockhopper electrical circuitry, several days of continuous solids feeding to the reactor were completed to test malfunction valve failures. Some circuitry modifications were required.

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Upon agreement at a meeting of the MLGW Technical Subcommittee an "Overall Program Plan" was generated for operation of the pilot plant for 6 months. This plan called for a 3 month technical review and altered the overall test program by substituting washed for ROM Kentucky coal in the following test plans. Initially IGT had to confirm through coke shakedown tests that this pilot plant was suitable for 50 psig operation and achieve coke gasification and ash agglomeration.

Test Run 127 — September 1978

Summary

A successful hot shakedown of the pilot plant system was accomplished during Run 127 with metallurgical coke as a feed material. Stable operating conditions were achieved at four set points (1 through 4). (See test plan, Table A-4.) Coke ash agglomerates were produced continuously during Set Point 4 operating conditions (1950°F temperature, 4 ft/s superficial velocity, 4-foot deep bed, and 35% bed ash content) using the steam-air-oxygen mode of gasification. The test run was terminated because elutriated solids eroded a hole in the second-stage cyclone. Operations at all four set points were carried out with a deactivated internal cyclone resulting in a high rate of fines being elutriated from the gasifier.

Details

The test plan for Run 127 consisted of conducting a hot shakedown of the U-GAS pilot plant using metallurgical coke as a start-up feed material. Also, to check the operability of the new process equipment consisting of the product gas scrubber, reactor screw feeder, 4-1/2 inch diameter venturi throat, 1-1/2 inch central jet nozzle located in the venturi throat, and the redesigned weigh-scale lockhopper feed system. Specifically, to carry out pilot plant operations under set point conditions 1 through 17, as described in Volume II of Overall Program Plan for Task IX — Technical Support, published in August 1978.

Run 127 encompassed 166 hours including 116 hours of metallurgical coke feed. A successful hot shakedown of the pilot plant system was accomplished, and operability of the process equipment was established. Stable operating conditions at four set points (1 through 4) were successfully achieved. At Set Point 4 conditions (1950°F bed temperature, 4 ft/s superficial velocity, and 35% bed ash content) coke ash agglomerates were produced continuously for

Table A-3. HOT SHAKEDOWN TEST WITH COKE OPERATING CONDITIONS
(Ash Agglomerating Gasifier)

Set Point	Period Hrs. $\pm$ 6	Bed Temp. $\pm$ 15 $^{\circ}$ F	Ash Conc. in Bed $\pm$ 5%	Superficial Velocity FPS $\pm$ 0.5	Venturi Annulus Vel. FPS $\pm$ 5	Bed Height Feet $\pm$ 0.5	Cyclone's Operating		
							Int.	II	III
1	12	1800	35	4	40*	4	No	No	No
2	12	1850	35	4	40	4	No	No	No
2A	12	1850**	35	4	40	4	No	No	No
3	12	1900	35	4	40	4	No	No	No
4	12	1950***	35	4	40	4	No	No	No
4B	12	1950****	35	4	40	4	Yes	No	No

OTHER OPERATING CONDITIONS REQUIRED

Mode = Steam-Air-Oxygen Process Gas

Process Gas to Grid = 15% oxygen conc. max.

Process Gas to Venturi Annulus = 21% Oxygen Conc. Max.

Process Gas to Central Jet = 50% Oxygen Conc. Max.

* Or as required to control venturi discharge classification capability.

** Minimum O₂ at grid under partial aggl. mode. May be repeated at 3A or 4A.

*** Assumed temperature for balanced ash agglomeration W/O bed withdrawal.
May require more set-points for ash agglomeration W/O bed withdrawal.

**** As required to sustain ash balance W/O bed withdrawal.

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20 hours using the steam-air-oxygen mode of gasification. With the deactivated internal cyclone, the second-stage cyclone shell was eroded at the inlet because of the large elutriation of fines. The resulting hole caused a product gas leak and for safety reasons, the run was terminated.

Run 127 was initiated on Wednesday, September 20, 1978. Stable operating conditions were accomplished at four set points. The operating conditions at each set point closely followed the overall test plan. The feedstock was Bethlehem fine coke breeze scalped to 1/4 X 0 inch. Air-steam was the gasification medium with oxygen added at the center jet when needed to attain the required set point temperatures. The effect of bed temperature on agglomerate formation was investigated at these set points under fixed conditions of superficial velocity, bed height, and bed ash concentration. At no time did the oxygen concentration at the grid, venturi, and center jet exceed the limitations of 15%, 21%, and 50%, respectively. Every attempt was made to hold the fixed parameters within the allowable deviations of ± 0.5 ft/s for the superficial velocity, ± 0.5 feet for the bed height, and $\pm 5\%$ for the bed ash. Bed samples were withdrawn and analyzed every hour for their percent ash content. Other samples (feed, venturi discharge, second- and third-stage cyclone fines, and withdrawn bed) were analyzed at least once in every 8-hour period. A computer check of the reactor operating conditions were obtained once every hour.

Set Point 1

The operating conditions of 1800°F at 4 ft/s superficial velocity were attained on September 23 at 1800 hours. The bed ash content was allowed to rise to achieve Set Point 1 (35%). Although bed height and superficial velocity met set point conditions, the venturi velocity had to be maintained around 60 ft/s to achieve selective solids classification. The beginning of Set Point 1 occurred at 1200 hours on September 23 when the bed ash content reached 33%. To maintain the conditions of Set Point 1 (1800°F, 4 ft/s superficial velocity, 4-foot bed height, and 35% bed ash content), a feed rate of about 470 lb/hr was required. A very occasional solids withdrawal from the bed through the side nozzle was required to control bed ash content. However, the rate of fines collected from the second-stage cyclone (the internal cyclone was deactivated) was about 215 lb/hr and contained approximately 20% ash. An overall ash balance for the system at the operating conditions of Set Point 1 indicated

that fines accumulation in the bed was minimal because of the high ash loss in the cyclones. Accordingly, bed ash concentration remained below 30% much of the time.

Set Point 2

Conditions of 1850°F, 4 ft/s superficial velocity, 4-foot bed height, and 35% bed ash content were attained at 0100 hours on September 24, 1978. The increase in bed temperature caused the rate of fines collected from the second-stage cyclone to increase to about 240 lb/hr, still containing about 20% ash. During the next 24 hours, the conditions of Set Point 2 were steadily maintained with the bed ash concentration averaging about 30%. No evidence of ash agglomerates was detected in the venturi discharge. Again, ash balances revealed that most of the ash that entered the reactor left with the overhead fines. The venturi annulus velocity remained at 50 to 60 ft/s.

Set Point 3

The conditions of Set Point 3 (1900°F temperature, 4 ft/s superficial velocity, 4-foot bed height, 35% ash in the bed) were achieved at 0200 hours on September 25, 1978. At this set point, ash levels of 35% were readily obtained during steady operation, yet only a trace of ash agglomerates was visible in the venturi discharge. These tiny agglomerates neither constituted any significant proportion of the discharge, nor did they grow in size. Thus, it was found undersirable to implement Set Point 3A, the diversion of all process oxygen entering the grid to the venturi area. A feed rate of 607 lb/hr was required to maintain a 4-foot bed height. The second-stage cyclone fines rate remained essentially the same as at Set Point 2. The overall ash balance was maintained by withdrawing approximately 50 lb/hr through the side discharge valve.

Set Point 4

Operating conditions at Set Point 4 (1950°F, 4 ft/s superficial velocity, 4-foot bed height, and 35% bed ash content) were achieved on September 26, 1978, at 0200 hours by adding extra oxygen through the center jet via the 9303 loop. The bed temperature was maintained at approximately 1945°F for about 4 hours, during which time the production of ash agglomerates was apparent from direct observation and from the rate of venturi discharge. However, the venturi discharge rate remained low for an overall ash balance. The feed rate at this

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stage reached about 900 lb/hr, and the second-stage cyclone fines rate rose to 320 lb/hr at 20% ash. An increase in bed temperature (more oxygen at the jet) of roughly 20°F resulted in a larger ash agglomerate discharge than was necessary in ash balanced operation. The feed rate had to be increased to 1200 lb/hr. Although the rate of fines remained on the high side (400 to 470 lb/hr), the ash content of these fines dropped to about 13%. Production of ash agglomerates continued under Set Point 4 conditions for about 20 hours (until 0400 hours on September 27) when it was discovered that product gas was leaking from the second-stage cyclone. The test run was immediately terminated at this point. Upon inspection, a hole was found in the cyclone.

The following conclusions can be drawn from the Set Points 1 through 4:

- Formation of ash agglomerates from coke feed at fixed conditions of superficial velocity, bed height, and bed ash concentration is a temperature-dependent phenomenon.
- A new 1-1/2 inch central jet nozzle was installed in the venturi throat for Run 127. Calculations were performed to determine the length of jet penetration. Merry's correlation shows a jet penetration length of 19 inches. This penetration length extends to the upper edge of the reactor grid. Accordingly, about 4 feet of bed still exists above the jet.

Test Run 128 - October 1978

Summary

The successful hot shakedown of the pilot plant system was completed during Run 128 with metallurgical coke as a feed material. Stable operation was achieved for six set point (3, 3A, 4AB, 5AB, 6AB, and 7AB). (See Table A-4.) Coke ash agglomerates were produced continuously from Set Point 3A (no O₂ at grid, 1900°F bed temperature, 4 ft/s superficial velocity, 30% to 40% bed ash content, and 4-foot bed height) throughout the remaining set points. The test run terminated because wall scale, jarred loose by thermal shock from a 14-hour interruption stemming from an air compressor failure, created a dead zone for a clinker to form, eventually causing bed defluidization.

Details

The test plan for Run 128 consisted of continuing the hot shakedown of the U-GAS pilot plant with metallurgical coke feed. Also, to operate the pilot plant at set point conditions 3 through 17 as summarized in Table II of Volume II of the Overall Program Plan for Taxk IX - Technical Support, published in August 1978.

Table A-4. CONTINUATION OF SHAKEDOWN TEST WITH OPERATIONS AT ASH
AGGLOMERATING CONDITIONS PRODUCED FROM TABLE A-3 COKE SHAKEDOWN DATA
(Ash Agglomerating Gasifier)

Set Point	Period Hrs. $\pm$ 6	Bed Temp. $\pm$ 15 $^{\circ}$ F $\pm$	Ash Conc. in Bed $\pm$ 5% $\pm$	Superficial Velocity FPS $\pm$ 0.5	Venturi Annulus Vel. FPS $\pm$ 5	Bed Height Feet $\pm$ 0.5	Cyclone's Operating		
							Int.	II	III
5	12	1950	35 *	4	30 **	4	No	No	No
6	12	1950	35	4	50 ***	4	No	No	No
7	12	1950	35	3	40 *	4	No	No	No
8	12	1950	35	4	40 *	2	No	No	No
9	12	1950	20	4	40 *	4	No	No	No
10	12	1950	45	4	40 *	4	No	No	No

OTHER OPERATING CONDITIONS REQUIRED

Mode - Steam, Air, Oxygen

Process Gas to Grid - 15% Oxygen Conc. Max.

Process Gas to Venturi Annulus - 21% Oxygen Conc. Max.

Process Gas to Central Jet 50% Oxygen Conc. Max.

+/- Assumed from Table I results.

++ Assumed.

* Optimum; as required to control venturi discharge classification capability.

** 10 FPS below optimum.

*** 10 FPS above optimum.

NOTE: Decision not to include internal cyclone will be reviewed based on results of 4B.

Table A-4, Cont. CONTINUATION OF SHAKEDOWN TEST WITH OPERATIONS AT ASH
AGGLOMERATING CONDITIONS PRODUCED FROM TABLE A-3 COKE SHAKEDOWN DATA
(Ash Agglomerating Gasifier)

Set Point	Period Hrs. $\pm$ 6	Bed Temp. $\pm$ 15°F	Ash Conc. in Bed $\pm$ 5%	Superficial Velocity FPS $\pm$ 0.5	Venturi Annulus Vel. FPS $\pm$ 5	Bed Height Feet $\pm$ 0.5	Cyclone's Operating		
							Int.	II	III
*11	12	1900**	20	4	40 +	4	No	No	No
12	12	1900**	30	4	40 +	4	No	No	No
13	12	1900**	40	4	40 +	4	No	No	No
*14	12	1900**	40	4	30 ++	4	No	No	No
*15	12	1900**	40	4	50 +++	4	No	No	No
*16	12	1900**	40	4	40 +	2	No	No	No
*17	12	1900**	40	3	40 +	4	No	No	No

OTHER OPERATING CONDITIONS REQUIRED

Mode - Steam, Air, Oxygen

Process Gas to Grid - 15% Oxygen Conc. Max.

Process Gas to Venturi Annulus 21% Oxygen Conc. Max.

Process Gas to Central Jet 50% Oxygen Conc. Max.

* If time permits.

** 50°F lower than set points 5 to 10.

+ Optimum; as required to control venturi discharge classification capability.

++ 10 FPS below optimum.

+++ 10 FPS above optimum.

The test Run 128 spanned 172 hours with 128 hours of metallurgical coke feed. Operating conditions of Set Point 3 were maintained for 14 hours. Set Point 3A (all oxygen entering at the grid flowing to the central jet) conditions were maintained for 24 hours. Ash agglomerates appeared in the venturi discharge. Set point 4AB (minimum oxygen at the grid, internal cyclone activated) conditions were maintained for 26 hours (1925°F). Set Point 5AB conditions (reduction in venturi flow by 5 mol/hr) were maintained for 6 hours (1925°F). Set point 6AB conditions (venturi flow increase by 5 mol/hr over Set Point 4AB) were maintained for 8 hours.

Operating conditions of Set Point 7AB were maintained for 8 hours with the bed temperature at 1900°F. The rental compressor failed, and the run halted because the standby compressor also would not work. Process air was restored after a 14-hour interruption. Set Point 7AB conditions were reestablished, but after 10 hours, no agglomerates had appeared in the discharge from the venturi. With the increase in bed temperature by 25°F ash agglomerates began discharging. The bed temperature was maintained for 4 hours at 1930°F. The reactor bed suddenly defluidized, ending the run.

Run 128 started on Monday, October 2, 1978. Stable operation was achieved for six set points. The operating conditions at each set point departed from the overall test plan in the following respects:

- a. Set Points 4AB, 5AB, and 6AB were achieved at a bed temperature of 1925°F, rather than the nominal set point of 1950°F. The test plan allows this flexibility because ash agglomerates were being produced and the bed ash content was being controlled totally by the venturi discharge rate.
- b. Set Point 7AB was similarly maintained at 1900°F, rather than the nominal value of 1950°F. Following the test run interruption the temperature in Set Point 7AB was increased from 1900° to 1925°F because ash agglomerates did not form at the lower temperature.
- c. The decrease and increase by 10 ft/s in gas velocity through the venturi annulus, called for in Set Points 5AB and 6AB, could not be achieved. Instead, a 5 mol/hr reduction and then a 5 mol/hr increase in flow through the venturi annulus, relative to Set Point 4AB, were used. A larger reduction than 5 mol/hr caused hot solids to discharge so rapidly that the venturi temperature exceeded 1000°F.
- d. The internal cyclone was activated following Set Point 3A and was kept operating to reduce dust production, which in Run 127 contributed to the second-stage cyclone erosion that shut down the run.

Otherwise, test parameters followed the test plan set points to within the specified range.

Set Point 3

The U-GAS gasifier attained the operating conditions of 1900°F at 4 ft/s superficial velocity on October 4 at 2200 hours. The bed ash content ranged generally within 30% to 40% as specified for Set Point 3. No agglomerates formed, and the alternate bed withdrawal was used to control the bed ash content. Coke feed rate was 585 lb/hr. Char fines discharged from the second-stage cyclone at 230 lb/hr. The rate of venturi discharge averaged 20 lb/hr.

Set Point 3A

The conditions of Set Point 3 were changed to those of 3A by switching all the oxygen (as air) entering the gasifier from the grid to the central jet. Operating conditions of Set Point 3A were attained on October 4 at 1600 hours. Although the bed temperature was still about 1900°F and the superficial velocity was close to 4 ft/s, the coke feed rate increased from 580 lb/hr, prevailing in Set Point 3, to 680 lb/hr. Char fines continued to discharge from the second-stage cyclone at 240 lb/hr; however, the rate of venturi discharge increased from 20 lb/hr at the start of Set Point 3A to 80 lb/hr, as ash agglomerates started to discharge. For a 6-hour period prior to the appearance of ash agglomerates, bed temperature was maintained at 1930°F to induce agglomerate formation. With the appearance of ash agglomerates, bed temperature was restored to its 1900°F set point value, and agglomerates continued to be discharged to the extent that the bed ash content was controlled solely by the ash agglomerate discharge from the venturi, and the alternate bed withdrawal ceased.

Set Point 4AB

Operating conditions of Set Point 4AB were attained on October 5 at 1700 hours. The bed temperature was maintained at 1925°F, lower than the nominal 1950°F value called for by the set point, but permitted under the test plan because ash agglomerates were being discharged and bed ash was controlled with our alternate bed withdrawal. With the internal cyclone activated, char fines removed by the second-stage cyclone decreased from 280 lb/hr to 190 lb/hr.

The coke feed rate ranged from about 550-600 lb/hr, and the venturi discharge rate varied from about 40-70 lb/hr, consisting almost entirely of ash agglomerates.

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Set Point 5AB

Gas velocity through the venturi annulus for Set Point 5AB was supposed to be 10 ft/s less than that in Set Point 4AB. However, when the venturi flow rate was reduced by only 5 mol/hr, the hot, reacting bed solids discharged so rapidly that the venturi heated up to over 1000°F. The IGT supervisors did not consider a higher venturi temperature safe, so the gas flow through the venturi annulus was reduced only 5 mol/hr, corresponding to 54 ft/s.

Operating conditions of Set Point 5AB were attained on October 6 at 2200 hours. The coke feed rate was 540 lb/hr. Char fines discharged from the second-stage cyclone at 190 lb/hr. The venturi discharge rate rose from 40 to 55 lb/hr, and the ash content in the venturi discharge decreased from 92% to 77.6% as more char from the bed came out.

Set Point 6AB

The gas velocity through the venturi annulus for Set Point 6AB was supposed to be 10 ft/s higher than that in Set Point 4AB. However, because gas flow was decreased 5 ft/s in Set Point 5AB, the IGT supervisors decided to hold the venturi gas flow increase to 5 mol/hr for Set Point 6AB. Operating conditions of Set Point 6AB were attained on October 7 at 0600 hours with gas flow to the venturi corresponding to 67 ft/s.

The coke feed rate averaged 520 lb/hr; char fines discharged from the second-stage cyclone at 200 lb/hr. The venturi discharge rate dropped from 55 lb/hr during the previous set point to 15 lb/hr, but then increased steadily during Set Point 6AB from 15 lb/hr to 30 lb/hr. The bed ash content also increased steadily from 18% to a peak of 31%, but declined thereafter.

Set Point 7AB

Operating conditions for Set Point 7AB were attained on October 7 at 1600 hours. The bed temperature was maintained between 1900° to 1915°F, and this temperature was sufficient to sustain ash agglomeration and not require alternate bed withdrawal. The superficial velocity was reduced to 3 ft/s, as called for in this set point.

The coke feed rate was 350 lb/hr, the lowest of any set point in this test run. With the internal cyclone continuing to operate, as in the previous set points, fines discharged from the second-stage cyclone at 140 lb/hr, also

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the lowest of any set point. Solids discharge rate from the venturi decreased during Set point 7AB from 70 lb/hr to 30 lb/hr as bed ash content declined from 19% to its lowest value during the run, 15%.

At 2430 hours the rental compressor failed, halting the test. Fluidization was sustained by nitrogen. Attempts to restart the compressor failed. The standby compressor was inoperable. Both compressors were repaired by 1400 hours on October 8 and the run resumed. The bed temperature had dropped to 400°F. Operating conditions for Set Point 7AB were reestablished on October 9 at 0600 hours.

Pieces of wall scale, some measuring 3/8-inch thick X 4 inches long, were present in the venturi discharge. No agglomerates formed, and the bed ash content increased steadily. To induce agglomerate formation, the bed temperature was increased to 1925°F at 1400 hours. Small agglomerates appeared in the venturi discharge. Despite a gas velocity through the venturi annulus of 50 ft/s, which was lower than in the first part of Set Point 7AB, discharge from the venturi did not exceed 20 lb/hr until, abruptly, at 2000 hours it increased to 40 lb/hr. At 2300 hours on October 9, all discharge from the venturi ceased, and bed density chart traces indicated partial defluidization.

The reactor was cooled and opened for inspection. A piece of clinker surrounded by wall scales was found in the venturi, lying at the juncture of the venturi and the grid, partly blocking the central jet. The reactor grid had no attached ash accumulation.

Some ash scale had accumulated on the refractory wall surface within the fluidized bed area. Thick deposits (up to 2 inches) of rough-surfaced ash scale had accumulated on all internal cyclone refractory-lined surfaces.

Explanation of Shutdown

The sudden cooling of the reactor following loss of the air compressor during Set Point 7AB loosened the wall scale, which subsequently fell off when operating conditions were restored. Some of this loosened wall scale wedged in the venturi expander, around the central jet, creating a defluidized zone and clinker formation.

Conclusions

1. Use of functioning internal cyclone reduced char fines carryover to external cyclone by at least 50 lb/hr.

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2. Ash agglomerates appear to form at a lower bed temperature when all the oxygen supplied enters the gasifier at the center. (This conclusion comes from comparing results of Run 127 with those of Run 128.)
3. Shakedown with coke was completed.

A low process air pressure switch (PSL 901) was installed to cut off the process air and oxygen if low supply pressure occurs and simultaneously to replace it with nitrogen or steam to sustain bed fluidization in the reactor.

Test Plan Change

At an MLGW Technical Subcommittee meeting held on October 19, 1978, Table II of Volume II of the August 1978 Program was revised due to results of Runs 127 and 128. The revised test plan appears in Table A-5.

Pilot Plant Tests With Washed Kentucky Coal

Physical and Chemical Properties of Washed Coal

Table A-6 contains the proximate and ultimate analysis and some physical properties of the washed Western Kentucky bituminous coal. Table A-7 contains the size distribution of washed coal used during the following tests.

Test Runs 129-133 -- October 1978-February 1979

Test Run 129 -- October 1978

Summary

Run 129 was a successful test of the pilot plant system using washed Kentucky No. 9 coal as a feed material. Stable operation was achieved for two set points (1 and 2). During Set Point 2, coal was blended as an increasingly large fraction of the feed, and, shortly after using all coal for the gasifier feed, the scrubber became plugged. The run was terminated voluntarily because prolonged venting of product gas was considered a safety hazard.

Details

The test plan for Run 129 consisted of achieving an ash-balanced operation with the formation of ash agglomerates, using washed Kentucky coal. The pilot plant was to be operated at Set Points 1 through 14 (Table A-5), according to Table III revised of the Overall Program Plan for Task IX, Volume II.

Test Run 129 spanned 66 hours, of which August 1978 metallurgical coke

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Table A-5. MODIFIED TEST PLAN FOR U-GAS TESTS

Set Pt.	Period ± 2 hrs	Bed Temp. ± 150°F	Ash Conc. ± 5	Sup. Vel. ± .5	Vent. Vel. ± 5	Bed Ht. ± .5	Cyclones	
							Int.	Ext. (Recycle)
1	6	1825	15	4	50*	4	Yes	No
2	6	1825	30	4	50	4	No	No
3**	6	1850	30	4	50	4	No	No
4	6	1850	30	4	50***	4	No	No
5**	6	1875	30	4	50	4	No	No
6	6	1875	30	4	50***	4	No	No
7**	6	1900	30	4	50	4	No	No
8	6	1900	30	4	50***	4	No	No
9	6	1850	30	4	50	3	No	No
10	6	1850	30	3	50	4	No	No
11	6	1850	45	4	50	4	No	No
12	6	1900	45	4	50	4	No	No
13	24	1900****	30	4	50	4	Yes	No
14	48	1900****	30	4	50	4	Yes	Yes

OTHER OPERATING CONDITIONS

Mode: Steam, Air Oxygen.

Process Gas to Grid: Steam only. If oxygen uses 15% max. conc.

Process Gas to Venturi Annulus: 21% oxygen conc. max.

Process Gas to Central Jet: 60% oxygen conc. max.

Set Point Period will be increased to 24 hours for ash balanced operation achieved.

- * Or optimum as required to control venturi discharge clarification.
- ** Introduce oxygen at grid with maximum concentration of 15%, only if agglomerate presence confirmed at this set point.
- *** Decrease venturi gas flow rate by about 20%.
- **** Or at conditions where no bed withdrawal is necessary.

Table A-6. ANALYSIS OF WESTERN KENTUCKY WASHED COAL

Proximate Analysis	Wt % (As Received)
Moisture	3.1
Volatile Matter	35.8
Ash	12.0
Fixed Carbon	49.1
Ultimate Analysis	Wt % (Dry)
Ash	12.10
Carbon	72.18
Hydrogen	4.51
Sulfur	3.10
Nitrogen	1.34
Oxygen	6.77
Bulk Density, lb/ft ³	50.4
Heating Value (Dry Basis), Btu/lb	12,498

Table A-7. SIZE DISTRIBUTION OF FEED MATERIAL (Washed Coal)

<u>Retained on USS Sieve Size</u>	<u>Washed Coal</u>
3	0.3
6	14.8
12	25.8
20	21.5
40	18.9
70	11.4
140	4.7
270	1.7
Pan	0.9

was fed for 32 hours, a coke and washed coal mixture for 8 hours, and washed Kentucky coal for 8 hours. Set Point 1 conditions were maintained for 12 hours. Set Point 2 conditions were maintained for 14 hours. During the test, the product gas scrubber trays (perforated plates) plugged with a black coal-tar having naphthalene odor. The gas was diverted around the scrubber and the trap cleaned, reassembled, and put back on line. The trays again plugged; coal feed was discontinued, and coke feed was started. The run was voluntarily ended due to atmospheric pollution caused by venting, to the atmosphere, product gas that contained some char dust.

Run 129 started on Sunday, October 21, 1978. Stable operation of the gasifier was achieved at the first two set points of Table III test plan. Operating conditions were close to those called for in the overall test plan.

Set Point 1

The U-GAS gasifier attained the operating conditions of 1825°F at 4 ft/s superficial velocity with coke feed. The internal cyclone was activated, as called for in Set Point 1. Prior to its activation, char fines discharged from the second-stage cyclone at about 200 lb/hr. Following the activation of the internal cyclone, the second-stage cyclone fines discharge rate dropped to 126 lb/hr.

All operating conditions for Set Point 1 were attained on October 23 at 1400 hours. The coke feed rate was 310 lb/hr, and solids discharge from the venturi varied from 7 to 12 lb/hr; however, no ash agglomerates formed. The ash concentration in the bed was kept at 15% through alternate bed withdrawal, which varied from 30 to 50 lb/hr.

Set Point 2

Operating conditions for Set Point 2 were attained on October 24 at 1800 hours. Following the completion of Set Point 1, the internal cyclone was deactivated, as called for in the test plan, and a mixture of coal and coke, initially 20% coal but increasing to 80% coal, was fed to the gasifier. At 1000 hours on October 25, only coal was fed. Two and one-half hours after using 100% coal feed, the pressure at the inlet to the scrubber increased suddenly from 0.1 to 1.0 psig; reactor gas issued from the scrubber water drain; and scrubber water was carrying over to the incinerator, downstream.

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The reactor gas was vented to the atmosphere and the scrubber was inspected. The two scrubber decks were covered with a black, sticky substance that smelled like naphthalene. The scrubber decks were cleaned and the scrubber was put back on line. Reactor gas to the scrubber was quenched to 400°F, lower than the usual 480°F value, and closer to the 180°F temperature prevailing in the scrubber. Despite this, within 1-1/2 hours the plugging reoccurred.

Coke feed was restored replacing the coal feed to the gasifier and the scrubber, once again, was inspected. The tar-like substance had again plugged the 1/10-inch diameter holes of the impingement baffle plates (decks) of the scrubber. Continuing to vent toxic gases and particulates to the air would have constituted pollution and a safety hazard. IGT supervisors decided to end the run rather than operate with an intermittently plugging scrubber.

The reactor was cooled, the bed was drained and the manways were opened. A piece of firebrick from the lining at the lower manway had fallen and lodged in the venturi expander beside the central jet pipe. The central jet pipe had melted on one side. There was no buildup of ash on the grid or on the refractory walls of the reactor.

The first quencher was examined. No deposits were found, but the inner shell was deformed and cracked in several places upstream of the atomized quench water nozzle.

Conclusions

1. Use of the internal cyclone reduced coke char fines loss by over 70 lb/hr.
2. Product gas contains an organic material that condenses and solidifies between 180°F (the scrubber temperature) and 480°F (the quench temperature), when the washed Kentucky coal was fed to the gasifier.

Following Run 129, several repairs and changes were made to the pilot plant:

- The first-stage quencher shell was repaired where high-temperature stress had deformed it, and the spray nozzle lance was lengthened to provide a better spray pattern in the quencher.
- The central jet pipe was replaced with a new pipe coated with alumina for abrasion resistance.
- A by-pass line was installed around the scrubber. The lower deck of the scrubber was replaced with a perforated plate with 1/2-inch openings, and the upper deck was cleaned and replaced. Four additional spray nozzles were installed.

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Test Run 130 — November 1978

Summary

This test plan was a continuation of Run 129 using the revised Table III and starting with Set Point 3. The U-GAS® Pilot Plant operated at four set points during Run 130 with washed Kentucky coal as feed. Ash agglomerates were produced at each of the set point operating conditions, and ash-balanced operation was achieved at each set point except Set Point 3. The first three set points (3, 4, and 6) used steam-air-oxygen fluidization, and the last set point (6A) used steam-oxygen.

The gasifier operated at Set Point 6A (1875°F bed temperature, 4 ft/s superficial velocity, 4-foot bed height) for 22 hours and produced a gas with a higher heating value of about 280 Btu/SCF.

The test run was intentionally terminated following unsuccessful efforts to activate the internal cyclone. Inspection of the reactor following Run 130 revealed that the grid, venturi, and reactor walls were free of ash accumulation, as was the feed nozzle, but the internal cyclone experienced a buildup of ash on its refractory-lined inlet surfaces. The outlet from the internal cyclone experienced severe erosion and corrosion, and the dipleg had fallen.

Run 130 was a very successful pilot plant run. Almost 88 tons of coal were gasified over a period of 106 hours. Coal ash agglomerates were produced under ash-balanced conditions for a period of 72 hours. The venturi ash discharge contained spherical ash agglomerates, about 10 mesh (0.075 in.) in diameter, and the average ash content of the discharged solids exceeded 70%.

Details

The test plan for Run 130 consisted of achieving an ash-balanced operation with the formation of ash agglomerates using washed Kentucky coal. Also to operate the pilot plant at Set Points 3 through 14 according to Table III in the addendum to the overall program plan for Task IX, Volume II, issued on October 19, 1978 (same as Table A-5).

The test run spanned 178 hours with metallurgical coke fed into the gasifier for 42 hours and 88 tons of washed Kentucky coal fed for 106 hours, and ash-balanced operation was maintained for 72 hours.

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The operating conditions of Set Point 3 were maintained for 10 hours. Small ash agglomerates began appearing in the discharge from the venturi annulus in the last hour of this set point.

The operating conditions of Set Point 4 were maintained for 32 hours. Ash agglomerates continued to be discharged. After the gasifier was operated at Set Point 4 for 18 hours, the side bed discharge was discontinued.

Operation at Set Point 5 was not carried out. Instead, oxygen was introduced at the grid, and the operating values of Set Point 4 were maintained for an additional 6 hours.

The operating conditions of Set Point 6 were maintained for a total of 16 hours. Ash agglomerates continued to be discharged. After 8 hours of operation at Set Point 6, a sudden loss of steam supply caused the bed temperature to drop below 1100°F. The steam supply was restored within 1-1/2 hours, and coke and coal were fed into the gasifier for 6 hours until operating conditions at Set Point 6 were restored. These operating conditions were maintained for an additional 8 hours. Ash agglomerates began to be discharged 2 hours after Set Point 6 operating conditions were restored.

The operating conditions of Set Point 6A (Set Point 13 without the internal cyclone activated) were maintained for 22 hours. During this period gasification was by steam and oxygen only; no compressed air was used. Attempts to activate the internal cyclone failed, and ash agglomerates continued to be discharged from the venturi annulus.

Coke feed was restored and fed for an additional 24 hours after siderite (iron ore) became mixed with the coal supply and fed into the gasifier.

The run was voluntarily terminated after more attempts to activate the internal cyclone failed. The post-run inspection showed that the dipleg to the internal cyclone had fallen into the bed, and the seal pot at the end of the dipleg had a hole at the bottom. The grid and venturi area were free of any ash deposits. The internal cyclone experienced a heavy accumulation of ash at its refractory-lined inlet.

Run 130 started on Wednesday, November 1, 1978. Stable operation was achieved for three set points with steam-air-oxygen gasification and for one set point with steam-oxygen gasification. The operating conditions at each set point complied with the overall Test Plan. (Table III Revised)

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Set Point 3

The U-GAS gasifier first attained the operating condition of an 1850°F bed temperature, and a 4 ft/s superficial velocity with coke feed. The internal cyclone was not activated, and the second- and third-stage cyclone fines were collected and removed. A mixture of coal and coke, was then fed to the gasifier, initially 20% coal but ultimately increasing to 100% coal at 1000 hours on November 3.

At 1400 hours the product gas scrubber, which had been operating satisfactorily, became plugged at the upper infringement deck, as in the previous run. The product gas was diverted by bypass the scrubber.

The coal feed rate averaged 1175 lb/hr. The second-stage fines removal rate was 355 lb/hr. The side bed discharge was used at the end of Set Point 3 to control the bed ash content. The gas velocity through the venturi annulus was about 59 ft/s, which allowed about 17 lb/hr of solids to discharge. Coinciding with the appearance of ash agglomerates, an increase in total solids discharge occurred at about 2300 hours. The ash agglomerates were rounded, averaging about 10 mesh (0.075 inch) in diameter.

Set Point 4

The conditions of Set Point 4 (1850°F, 4 ft/s superficial velocity, 4-foot bed height, and a reduced velocity through the venturi annulus) were achieved at 0200 hours on November 4.

The gas velocity through the venturi annulus was at first reduced, by decreasing the molar flow to achieve 52 ft/s, but a venturi temperature rise raised it to 60 ft/s. Solids discharge was about 25 lb/hr and did contain ash agglomerates. The ash content of the venturi discharge exceeded 60%. At 1400 hours on November 4, the venturi annulus gas velocity was again reduced to about 51 ft/s, which increased the venturi solids discharge rate. Additional char now discharged raised the venturi line temperature from 800°F to more than 1000°F. The steam concentration of the venturi gas was raised to lower the temperature. The solids now discharged from the venturi at an average 50 lb/hr with a bed ash content between 40% and 50%.

The alternate bed discharge was discontinued after 1800 hours on November 4. An increase in coal feed rate from 1625 to 1830 lb/hr was required due to an increase in second-stage fines discharge at 495 lb/hr. Ash-balanced operation

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was maintained from this point for a period of 72 hours.

Set Point 5 called for increasing the bed temperature and introducing oxygen at the grid. Oxygen, comprising 4% of the grid gas flow, was introduced at the grid at 1030 hours on November 5, but the other operating values of Set Point 4 were maintained for an additional 6 hours to determine if the discharge of venturi solids was affected. The rate of solids discharge from the venturi remained at about 60 lb/hr including many ash agglomerates. After sufficient data obtained operations was moved to the next set point. (6)

Set Point 6

The operating conditions of Set Point 6 (1875°F bed temperature, 4 ft/s superficial velocity, 4-foot bed height, oxygen flowing to the grid) were attained at 2000 hours on November 5.

The gas velocity through the venturi annulus dropped to 46 ft/s from 51 ft/s with the addition of more steam, and the solids discharge from the venturi increased to 90 lb/hr. The bed ash content began falling now from 53% ultimately reaching 26%.

The coal feed rate now averaged 1800 lb/hr as the second-stage fines discharge rate averaged 457 lb/hr. The side bed discharge was activated for 2 hours before the bed ash analyses confirmed a decreasing bed ash content. The alternate discharge was maintained for 4 more hours during a later steam supply interruption as a precaution, then discontinued for the remainder of the test run.

At 0500 hours on November 6, an interruption in the steam supply occurred. An automatic steam-failure valve-interlock system shut off the flow of oxygen to the gasifier, and nitrogen flow cooled the bed to 1040°F. The steam supply was restored at 0600 hours, and operations were resumed with coke feed to re-establish operating temperatures and avoid the formation and condensation of tars from coal. At 1200 hours, Set Point 6 operating conditions had been restored and ash agglomerates appeared in the solids discharge from the venturi 3 hours later.

Set Point 6A

At 1900 hours on November 6, it was decided to switch to steam-oxygen fluidization at the operating conditions of Set Point 6, as called for in the

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10/19/78 addendum to the test program herein called 6A.

A coal feed rate increase to 1965 lb/hr was needed as the production of second-stage fines increased to 547 lb/hr. The ash content of the solids discharged from the venturi exceeded 70% and contained many ash agglomerates as the rate varied from 60 to 79 lb/hr. The bed ash content averaged 50% at this time.

Beginning at 1000 hours on November 7, attempts were made to activate the internal cyclone, as listed for Set Point 13. However, despite an increase in aeration at the seal pot and pulse-blasting of the dipleg with nitrogen and water, the rate of fines carry over to the second-stage cyclone did not diminish.

At 1115 hours on November 7 an interruption in the city water supply to the quenchers occurred, and the coal feed rate and bed temperature were lowered, which resulted in a loss of bed height. Within 1/4 hour the water supply and operating set point conditions were restored. Ash agglomerates continued to discharge throughout the interruption.

The steam-oxygen gasification was producing product gas having a heating value of 280 Btu/SCF. Figure A5 shows material and energy balance summary for Set Point 6A conditions in Run 130.

At 1800 hours a hot spot was observed on the incinerator steel stack. The scrubber, cleaned of tar, was reactivated to lower the gas temperature to the incinerator. The incinerator did operate 100° cooler, but the scrubber became plugged within 4 hours, and once again had to be bypassed. The hot spot remained but grew no larger and was monitored at two hour intervals.

Between midnight and 0200 hours on November 8, the bed density was observed to be steadily increasing. Bed ash content results on samples taken at 2200 hours, revealed 100% ash. Coke was substituted for coal feed at 0200 hours upon receiving task analysis report. The coal feed, it was found, was contaminated with sidrite (iron ore) which inadvertently had been piled on top of the U-GAS washed Kentucky coal supply during clean-up operations underway at the steam-iron facility.

Any attempts at continuing the test were concentrated on coke feed flushing of the gasifier bed of iron. With the bed temperature at 1500°F during

this coke flushing period an attempt was made to activate the internal cyclone by filling the seal pot through raising the bed level; however, the collection efficiency of the internal cyclone did not improve. Despite a 12 hour effort.

It was decided then to terminate the run at 0200 hours on November 9, to inspect the reactor for iron contamination and to determine why the internal cyclone was not functioning.

The reactor was cooled and opened for inspection. The internal cyclone dipleg had fallen and rested on the grid. The outlet pipe of the internal cyclone was severely eroded and corroded from inside and out. The bottom cover of the internal cyclone reverse seal pot was deformed and eroded; it also had a hole in it.

The grid and the venturi were completely free of ash accumulation, as were the feed nozzle and reactor walls, but the refractory-lined portions of the internal cyclone experienced a heavy buildup of ash. There was no evidence of iron deposits.

Although the alumina coating on the central jet pipe had worn off in spots, there was no damage to the pipe itself. However, six of the thermowells monitoring reactor temperature were so eroded and corroded as to require replacement.

In the incinerator, the refractory was missing at the location of the hot spot and was assumed to have fallen off.

The product gas scrubber decks and bottom cone had accumulated a tar-char solids composite that interfered with water discharge and gas flow.

Achievements

1. Run 130 resulted in ash-balanced operation for 72 hours with washed Kentucky coal feed and the production of ash agglomerates from coal, at bed temperatures between 1850° and 1875°F, for a period of 86 hours.
2. Even though 88 tons of coal were gasified over a period of 106 hours and the system experienced two major upsets, the gasifier internals – including the grid, venturi, and feed nozzle – remained free of ash deposits.
3. Product gas having a higher heating value of 280 Btu/SCF was produced with steam-oxygen gasification of the coal.

Following Run 130, several repairs and changes were made to the pilot plant:

- The dipleg to the internal cyclone was shortened by 2 feet and reinstalled. The seal pot was repaired, shortened, and put back on the dipleg. The outlet pipe and refractory of the internal cyclone were repaired.
- The corroded thermowells were replaced and lined with alumina.
- The worn central jet pipe was recoated with boron oxide.
- The product gas scrubber decks and bottom cone were cleaned of solids. The scrubber was converted to a spray tower by adding 3 nozzles and removing the decks.
- Following the discovery of a leak, the cone to the second-stage cyclone was replaced.
- A crack in the C-4 hopper shell near the ash nozzle was repaired and reinforced following the discovery of a leak. The C-3 hopper ash nozzle was also stiffened.

Test Run 131 — December 1978

Summary

During an MLCW Technical Subcommittee meeting, the 11/15/78 revision to Table A-5 was prepared containing four set points (Table A-8). In Run 131 the U-GAS pilot plant operated successfully at the conditions of Set Point 1 with washed Kentucky coal as the results of Run 130 were duplicated with ash-balanced operation under ash agglomerating conditions for 3 days. In all, 69.3 tons of coal were gasified during 4-1/2 days of coal feed. The ash agglomerate discharge varied between 75% and 85% ash by weight.

During a 35 hour period Set Point 1 operating conditions under the internal cyclone returned most of the reactor fines to the bed, and the conversion of coal reached 86%.

The return of second-stage fines to the bed, under Set Point 3, conditions produced 3 hours of steady state operation and a 92% coal conversion efficiency. A loss of second-stage fines return caused a brief rise in the reactor bed temperature to 1950°F before the oxygen feed could be reduced and the temperature lowered back to set point conditions. Ash agglomerates 1/8 to 1/4 inch in diameter, the largest yet made with coal

Table A-8. U-GAS PILOT PLANT COAL TEST RUNS
(Revised 11/15/78)

Set Pt.	Period, hr (+2)	Bed Temperature, °F ^c (+15)	Ash Concentration, % (+5)	Superficial Velocity, ft/s (+0.5)	Venturi Velocity, ft/s ^e (+5)	Bed Height, ft (+0.5)	Cyclones	
							Int.	Ext. (Recycle)
1	6	1850	40	4	50	4	Yes	No
2	6	1850	50 ^d	4	50	6	Yes	No
3 ^a	48	1850	50 ^d	4	50	4 ^f	Yes	Yes
4 ^b	6	1850	40	3	50	4 ^f	Yes	Yes

OTHER OPERATING CONDITIONS

Mode: Steam, Oxygen

Process Gas to Grid: Steam only. If oxygen, max. conc. of 15%

Process Gas to Venturi Annulus: 30% oxygen conc. max.

Process Gas to Central Jet: 75% oxygen conc. max.

^aIf external cyclone not successful go back to Set Point 2 conditions.

^bAfter 5 days of coal testing.

^cAs required for agglomeration.

^dMaximum.

^eOr optimum as required to control venturi discharge.

^fSet point at 6-foot depth if Set Point 2 is successful.

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feed, discharged from the venturi following this bed temperature upset.

The internal cyclone failed to function after the recycling of second-stage cyclone fines ended. During attempts to re-achieve Set Point 2 operating conditions, the dust carryover exceeded 1100 lb/hr and wore a hole in the second-stage cyclone, terminating the run.

A post-run inspection revealed that the reactor grid area was free of ash deposits and undamaged; however, the refractory-lined inlet scroll section of the internal cyclone was nearly plugged with ash deposit.

Details

The test plan for Run 131 presented in Table A-8 was Table III in the addendum to the overall program plan for Task IX, Volume II, issued on October 19, 1978, and now revised again on November 15.

The test run spanned 196 hours with metallurgical coke fed into the gasifier for 70 hours and 69.3 tons of washed Kentucky coal fed for 104 hours. Ash agglomerates were produced from coal throughout the run, except during two loss-of-feed process upsets. Ash-balanced operation was maintained for 68 hours.

The operating conditions for Set Point 1 were maintained for 35 hours and followed a feed interruption that led to the activation of the internal cyclone. Prior to the feed interruption, the internal cyclone resisted extensive attempts at activation. Ash-balanced operation was maintained at Set Point 1 for 23 hours.

Operation at Set Point 2 was postponed. Instead, the second-stage fines were recycled into the reactor, corresponding with Set Point 3.

The operating conditions for Set Point 3 were maintained for about 3 hours. Loss of the second-stage fines recycle led to a brief bed temperature excursion that resulted in the production of 1/8 to 1/4-inch-diameter ash agglomerates. The internal cyclone ceased functioning at this time.

The operating conditions for Set Point 2 could not be established despite efforts to get the internal cyclone to function again. A 6-foot bed height was maintained for 6 hours until a hole formed in the second-stage cyclone, terminating the run,

The post-run inspection revealed a near-plugging of the refractory-lined inlet section of the internal cyclone from a deposit comprised of 98% ash. The seal pot was unobstructed, the grid and venturi area were free of any ash deposits, and the thermowells and central jet pipe were undamaged.

Set Point 1

The U-GAS gasifier attained stable conditions at an 1850°F bed temperature, 4 ft/s superficial velocity and a 4-foot bed height with coke upon start-up. The carryover of fines to the second-stage cyclone varied between 196 and 228 lb/hr with coke feeding at the rate of 450 to 600 lb/hr. From 0100 hours on December 5 to 1900 hours on December 6, the seal pot aeration was adjusted in order to activate the internal cyclone and reduce the carryover of fines, as called for in Set Point 1. However, pressure drop measurements along the cyclone dipleg indicated that the seal pot aeration gas was flowing up the dipleg, rather than out of the seal pot, such that the collected fines were not being returned to the reactor bed.

At 1200 hours on December 6, the superficial velocity was reduced to 3 ft/s, but the operation of the internal cyclone did not improve. The set point superficial velocity of 4 ft/s was restored with the introduction of coal feed.

A mixture of coal and coke, initially 20% coal but ultimately increasing to 100% coal, was fed to the gasifier starting at 2115 hours on December 6. By 0135 hours December 7, 100% coal feed was attained. Ash agglomerates began appearing in the discharge of solids from the venturi at 1030 hours on December 7. The side bed withdrawal proved unnecessary and was discounted at 1700 hours. Ash-balanced operation continued with an increase in the venturi solids discharge rate to 40 and then to 60 lb/hr. The coal feed rate averaged 1500 lb/hr, and the carryover of fines removed by the second-stage cyclone was 450 lb/hr.

The higher heating value of the nitrogen and water-free reactor gas was 278 Btu/SCF, and the coal conversion was 72%.

At 2300 hours on December 7, a feed interruption occurred that required a checkout of the scale system. Oxygen flow to the reactor was stopped. The bed cooled to 750°F and was drained. The feed problem was traced to the reactor screw feeder; the chain driving it was slipping. A link was removed

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from the chain, and coke feed was restored at 0100 hours on December 8.

At 1025 hours on December 8, coal feed was returned, initially as a 20% coal mixture with coke, and by 1545 hours 100% coal feed was attained. At this time the internal cyclone began functioning, as indicated by reduced carryover of fines removed by the second-stage cyclone. The coal feed rate now averaged 1039 lb/hr, with 144 Btu/hr of fines removed by the second-stage cyclone. Ash agglomerates again began appearing in the solids discharge from the venturi. These agglomerates were smaller than those present on December 7 and averaging (-10 mesh), resembling those of Run 130. Simultaneously an 80 lb/hr of venturi solids discharge averaged 86% by weight ash as bed ash content rose to 70%.

At 1500 hours on December 9, the bed temperature was raised to 1880°F by increasing the oxygen input concentration to promote the formation of larger, denser ash agglomerates, and the venture throat velocity was reduced from 72 to 67 ft/s to increase the ash withdrawal rate. The reactor bed ash content then decreased to from 50% to 60%. During the ash-balanced operation, the rate of venturi solids discharge averaged 130 lbs/hr with a coal feed rate of 1176 lb/hr, and the fines carryover removed by the second-stage cyclone averaged 113 lb/hr.

The higher heating value of the nitrogen and water-free reactor gas was about 260 Btu/SCF, and the coal conversion was 86%. Based on the coal conversion, 64% of the fines was returned to the bed and gasified to extinction. Figure A-6 shows material and energy balance summary for Run 131. (Set Point 1).

At 0140 hours on December 10, the rate of fines carryover from the internal cyclone rose to 288 lb/hr, dropping the fluidized bed height by 0.5 feet. During the next 12 hours, numerous seal pot aeration rate changes were made, but no reduction in fines carryover was achieved.

In the afternoon of December 10, it was decided to bypass Set Point 2 (increasing the bed height to 6 feet), to proceed instead with Set Point 3 at a 4-foot bed height by recycling the second-stage cyclone fines, and to try Set Point 2 conditions later.

Set Point 3

At 1700 hours on December 10, the temperature of the reactor gas from the first-stage quencher was increased to 1000°F, thereby increasing the temperature

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of the fines returned to the bed from the second-stage cyclone. The bed temperature varied initially, but stabilized within 15 to 20 minutes following the addition of oxygen, as the second-stage cyclone fines were returned to the bed. The quench water flow rate, the bed density, and the fines recycle rate remained steady for about 3 hours. The higher heating value of the nitrogen and water-free reactor gas was 271 Btu/SCF, and the coal conversion was about 92%. At 2030 hours, the recycling of fines abruptly ceased as nitrogen, used to aerate the second-stage cyclone seal pot, backgassed up the diplet of the second-stage cyclone, the result of a decrease in solids density in the dipleg of the cyclone.

With the loss of the 800°F fines entering the bed, the reactor bed temperature rose rapidly to 1950°F before the operators could reduce the oxygen feed and bring the bed temperature back to set point conditions. In response to this temperature upset, the bed density decreased from about 36 to 30 lb/ft³, and the fluidized bed height dropped. The venturi discharge rate, determined periodically through the discharge of batches through a lockhopper arrangement, increased temporarily by nearly 100% to 320 lb/hr just after the temperature excursion. The venturi discharge was 96% ash by weight and consisted almost entirely of 1/8 to 1/4-inch diameter ash agglomerates, the largest yet produced from operation with this coal. The bed ash content simultaneously dropped from 60% to less than 40%.

With the ending of fines recycling, the fines carryover removed by the second-stage cyclone was found to have increased to 551 lb/hr, indicating that the internal cyclone no longer functioned. Over the next 12 hours, various changes in the nitrogen aeration to the internal cyclone diplet and seal pot were made but did not reduce the fines carryover from the internal cyclone.

Set Point 2

It was decided to proceed with Set Point 2 despite the relatively large fines carryover and, if successful, to return to Set Point 3.

After 3 hours of increased coal feed, the bed height had been raised to 5 feet when the reactor screw feeder jammed at 1400 hours on December 11, temporarily interrupting the coal feed and preventing the attainment of a 6-foot bed height. The reactor was idled while the feed failure was examined.

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It was found that a roll pin used to secure the handle on the feed sample ball valve had sheared, enabling the valve to remain open when a feed sample was taken using a lockhopper method. The failed valve enabled the reactor to vent through the reactor feed screw and out the feed sample line. The effect was that the coal discharging from the reactor feed screw was prevented from entering the feed chute, thereby plugging the screw discharge pipe.

By 1700 hours on December 11 the feed chute was cleared of coal, but the reactor bed temperature had dropped to 600°F. The operating conditions of 1850°F and a 6-foot bed height sor Set Point 3 were achieved at 0345 hours on December 12, after a period of reheating and start-up with coke feed, followed by blending increasing amounts of coal with the coke feed until 100% coal feed was attained.

Because the internal cyclone was not functioning, the fines carryover removed by the second-stage cyclone was very high (1134 lb/hr). The coal feed rate needed to maintain the bed height was also high (2370 lb/hr), because of the high loss of fines.

At 0820 hours on December 12, the second-stage cyclone developed a product gas and dust leak, forcing the termination of Run 131.

The reactor required about 36 hours to cool so that the internal cyclone could be lifted and inspected. The refractory-lined inlet scroll section was almost completely plugged with a deposit comprised of 98% ash. The seal pot and dipleg were free of any accumulation and were unobstructed.

The reactor grid area, including thermowells, the central jet pipe, and the fines injection and feed nozzles, were free of any deposits.

During Run 131, the product gas scrubber operated for 43 hours during coal feed, the longest of any coal test, but, eventually, the water drain became plugged with tar despite numerous efforts to clear the drain by poking a rod through the obstruction. Twenty-seven pounds of tar were removed from the scrubber following Run 131, representing about 0.02% of the weight of coal fed to the gasifier while the scrubber was operational.

Following Run 131, several repairs and changes were made to the pilot plant:

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- The refractory lining was removed from the internal cyclone. The seal pot on the dipleg of the internal cyclone was also removed, and the dipleg was extended to discharge fines within the fluidized bed.
- A section approximately 14 inches X 20 inches was replaced on the barrel portion of the second-stage cyclone opposite the inlet. Ultrasonic tests indicated severe wear at this location from the heavy fines loading of Run 131.
- Solenoid valves and piping for the pressure and differential pressure switches were installed for the second- and third-stage cyclone fines removal at the C-5 and C-6 lockhoppers. Two cylinder-operated ball valves were installed as part of this lockhopper arrangement for removing these fines at 50 psi operation.
- Differential pressure switches were installed to ensure the safe removal of ash from the C-4 ash lockhopper and to facilitate safe pressurization at the weigh scale surge hopper of the coal feed system, also for 50 psi operation.
- The coal feed capacity per cycle was raised from 150 to 300 pounds and time delays were shortened to allow the coal screw feeder to achieve rates of 3200 lb/hr.
- Piping was altered to allow coal feed at the lower reactor feed nozzle, 6 inches above the grid. Coal previously was fed through a nozzle 2-1/2 feet above the grid.
- The grid fluidization nozzles with 1/8-inch holes were replaced with nozzles having a 3/32-inch hole to permit a greater flow for operation at 50 psi.
- The incinerator firebrick nozzle assembly was removed, and the annular space where the off-gases from the reactor enter was cleaned. The nozzle was repaired and replaced. A new shroud was installed around the nozzle, and new firebrick was emplaced.
- Reinforcing pads were installed on the C-3 and C-4 ash lockhoppers, below the gasifier. The hoppers were hydrotested at 125 psig and 70°F to ensure pressure-tightness at 50 psig and 800°F.

Test Run 132 - January-February 1979

Summary

At an MLW Technical Subcommittee meeting held on 12/21/78 Table III was again revised for use in Run 132 as Table A-9 herein. In Run 132, the U-GAS pilot plant operated at Set Point 1 of the 12/21/78 Table III test plan. However, operation at other set point conditions were not completed because of a malfunctioning internal cyclone. A total of 47 tons of coal

Table A-9. U-GAS PILOT PLANT COAL TEST RUNS^a (Revised 12/21/78)

Set Point	Length of Period, hr (+2)	Bed Temperature, ^b °F(+15)	Ash Concentration, % (+15)	Superficial Velocity, ft/s (+1)	Venturi Velocity, ^c ft/s(+5)	Bed Height, in. (+0.5)	Cyclones		Pressure, psig (+10)
							Internal	External (Recycle)	
1	24	1875	50	4	50	4	Yes	No	15
2	8	1875	50	4	50	4	Yes	Yes	15
3	8	1875	50	4	50	4	Yes	No	45
4	72 ^{d,e}	1875	50	4	50	4	Yes	Yes	45

Other Operating Conditions

Mode: Steam, Oxygen

Process Gas to Grid: Steam only, if maximum oxygen concentration of 15%

Process Gas to Venturi Annulus: 30% oxygen concentration (maximum)

Process Gas to Central Jet: 75% oxygen concentration (maximum)

^a Because of new conditions imposed by pressure operation, it is not possible to predict the exact time required to achieve the above set points. Several other sets of information have been identified that are important for the design of the demonstration plant. As time and funding permits, the following will be investigated in the pilot plant:

- Effect of bed heights
- Effect of superficial velocity
- Use of unwashed Kentucky No. 9 coal
- Turndown and load followings
- Range of operating variables

^b As required for agglomeration.

^c Or optimum as required to control venturi discharge.

^d If not feasible, then at Set Point 3.

^e ± 8 hours.

was gasified during 74 hours of ash-balanced operation. Run 132 was interrupted by a feed problem; the internal cyclone and reactor were checked at that time and found free of any ash accumulation. After Run 132 was restarted, the bed height was raised to 6 feet in an effort to improve the internal cyclone performance, but it still allowed excessive fines carryover. Other measures also failed to reduce the fines carryover, which continued to increase with time. A high pressure drop caused the bed to evacuate through the diplet of the internal cyclone, thereby terminating the run.

A post-run inspection revealed a deposit of devolatilized tar and char covering the reactor walls; this was a result of coal being fed to the empty reactor. The internal cyclone had accumulated ash, completely filling the barrel and scroll inlet. This accumulation began on the refractory lining, eroded a hole in the barrel, and created the high pressure drop that evacuated the bed. The product gas scrubber operated throughout the run without interruption, indicating that little or no tar was present in the raw gas.

Details

The test plan for Run 132 is outlined in Table A-6. This Table A-9 originally appeared as Table 3 in the "Addendum" to Overall Program Plan for Task IX, Volume II, October 1978, and revised on 10/19, and 11/15 and now presented as the 12/21/78 revision.

The test run spanned 174 hours with 47.0 tons of washed Kentucky coal fed for 58 hours. Ash agglomerates and 270 to 280 Btu/SCF product gas were produced from coal throughout the run except during a loss-of-feed operational upset. Ash-balanced operation was maintained for 68 hours over two separate periods of 32 and 36 hours.

The operating conditions for Set Point 1 were maintained for 10 hours; ash-balanced operation was sustained throughout Set Point 1. The internal cyclone operated efficiently for about 14 hours from the start of Set Point 1, but then allowed an increasingly greater fines carryover to the second-stage cyclone.

A coal-feed jam in the reactor screw feeder enabled the bed temperature to rise far enough to cause the oxygen feed to shut off automatically and the unit to begin cooling. After the jam was cleared, coke would not flow from the storage hopper of the silo because of moisture condensation in the hopper.

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By the time the coke feed was restored the reactor had cooled to 500°F.

A higher pressure drop than normal across the venturi led to a decision to empty the cooled bed and inspect the reactor. The venturi was found to be unobstructed. A friable, 2-1/2 inch-diameter conglomerate of smaller ash agglomerates was found adhering to a thermocouple that projected through the grid. The rest of the reactor was free of ash accumulation. The internal cyclone had a thin ash deposit at its barrel section, but no erosion was visible.

The run was restarted, and several different approaches were used to improve the efficiency of the internal cyclone before proceeding with Set Point 2. However, the internal cyclone allowed a steadily increasing fines carryover to the second-stage cyclone. Increasing and lowering the bed height between 4 and 6 feed did not reduce the carryover of fines.

Ash had been accumulating within the internal cyclone and created a sufficiently high back pressure at the freeboard to allow gas bypassing, which resulted in the evacuation of the reactor bed through the dipleg and termination of the run.

The post-run inspection revealed that the brick wall and part of the reactor bed area were covered with char and carbonaceous matter. The internal cyclone shell below the inlet.

The first-stage quencher's inner shell had deformed, but the second- and third-stage cyclones and second-stage quencher were free of surface wear.

Run 132 began on Friday, January 26, 1979. Reactor gas samples were taken every 2 hours using a helium tracer for the first time; a measured amount of helium was added to the reactor through the grid 10 minutes before the gas sample bomb was taken. The purpose of using the tracer was to improve the closure of heat and material balances through a direct measurement of reactor gas composition. The standard method for obtaining reactor gas composition at the U-GAS pilot plant used a measured nitrogen input related to measured reactor dry gas composition.

Set Point 1

The gasifier was raised to a bed temperature of 1875°F at a superficial velocity of 4 ft/s and a bed height of 4 feet with coke feed. Next, a

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mixture of coal and coke, which was initially 20% coal but increased to 100% coal, was fed to the gasifier. By 0600 hours on January 28, the gasifier feed consisted entirely of coal. The bed ash content had increased to about 50% by 1400 hours on January 28 when ash agglomerates began to be discharged from the venturi, increasing the ash content of the discharged from the venturi, increasing the ash content of the discharged material. Fines carryover to the second-stage cyclone was about 9% of the coal feed, or 77 lb/hr, indicating satisfactory operation of the internal cyclone. All the operating conditions of Set Point 1 were thereby attained by 1400 hours on January 28.

Set Point 1 operating conditions were maintained over the next 14 hours except for a period of 4 hours near midnight when the bed height dropped below the set point limit of 3.5 feet, reaching as low as 3 feet, before raised back to 4 feet. During this period the bed temperature also increased slightly beyond the upper set point limit of 1890^oF. At 0400 hours the higher heating value of the nitrogen- and water-free product gas was 279 Btu/SCF.

At 0600 hours on January 29, the carryover of fines to the second- and third-stage cycloned had increased to more than 17% of the coal feed, indicating unsatisfactory operation of the internal cyclone. The venturi discharge material consisted of ash agglomerates mixed with char, having an average ash content between 90% and 95%. The reactor bed ash content was around 68%.

At 1400 hours on January 29, the weigh scale system malfunctioned, resulting in a premature coal feed dump from the weigh hopper into the surge hopper. While this was being emptied, the reactor feed screw became jammed with coal, interrupting the coal limit, which shut off the oxygen supply. The bed began to cool. An increased pressure drop across the venturi and an increased bed density were observed, indicating the possibility of an obstruction at the venturi annulus. After the coal was emptied from the feed system surge hopper, the feed was switched to coke to increase the bed temperature. However, moisture had condensed from the walls of the Silo B hopper, and coke would not flow. By the time coke feed was restored the bed had cooled to 500^oF.

It was decided to empty the bed and inspect the venturi for ash accumulation. No accumulation was found on the grid or in the vicinity of the venturi; however, an ash conglomerate had formed surrounding the grid thermocouple (No. 7), consisting of smaller ash agglomerates fused together.

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Although firmly attached, the 3-1/2 inch-diameter chunk proved to be friable and was easily removed.

The reactor was too warm to enter, but the venturi was probed from the lower manway and found to be free of any obstruction. The internal cyclone was checked by dropping a weight that emerged from the bottom of the dipleg. A mirror was used to inspect the inside. A thin layer of ash had accumulated on the barrel, but the small amount found meant that the barrel and scroll inlet to the internal cyclone were probably clean. The internal cyclone dipleg bore no trace of erosion.

Prior to restarting Run 132 on February 1, the reactor was sealed and pressure-checked. The pressure check revealed a severe leak at a stainless steel expansion joint in the product gas line. The bellows had corroded, and the expansion joint had to be replaced.

Run 132 was restarted with autoignition of the coke bed taking place at 1030 hours on February 3. However, condensed moisture in the Silo B hopper again prevented coke from feeding to the gasifier. The wet coke was removed and replaced with dry coke from the main silo. The bed was brought to operating temperature, and the usual practice of blending coal with coke was followed so that, by 2315 hours on February 3, 100% coal feed was attained.

A decision was made to operate a 6-foot bed depth to improve the efficiency of the internal cyclone by avoiding back-gassing of reactor gas through the dipleg. The fines carryover averaged 16% of the coal feed. Ash agglomerates began to appear in the venturi char discharge at 0530 on February 4. Oxygen was added at the grid for 12 hours, comprising up to 5% of the grid flow, starting at 0400 on February 4. This was discontinued after some conglomerates of fused ash agglomerates were found in the venturi discharge. At 0600 hours the higher heating value of the nitrogen- and water-free product gas was 275 Btu/SCF.

At 1700 hours on February 4, the 6-foot bed height was reduced to 4 feet. At 1600 hours before the change, fines carryover amounted to 243 lb/hr, 17.3% of the coal feed rate. Following the drop in bed height at 1800 hours, the carryover was 390 lb/hr, 31.6% of the coal feed rate. The fines carryover gradually increased, and- by about 1200 hours on February 5, it was decided to terminate the run.

A high reactor freeboard back pressure was observed at this time, which was found to be the result of an 8-pai pressure drop through the internal cyclone. At 1300 hours (during shutdown) the bed density and height recordings fell to zero. The bed and some coal feed had evacuated through the internal cyclone dipleg to the downstream cyclones and quenchers. Five hundred pounds of solids were recovered from the second-stage cyclone solids collection hopper.

The empty reactor took 3 days to cool, because of reignition of the congealed carbonaceous coal residue. An inspection of the reactor was conducted. A large portion of the reactor brick wall (the bed area) was covered with carbonaceous material and char. Particular dense accumulations of this loosely adhering char were found at the feed nozzles, start-up burner, thermowells, and manway openings. Some char had also accumulated on the outside of the internal cyclone. The carbonaceous material came from the raw coal, which was fed to the reactor briefly while it was empty.

The inlet and barrel sections of the internal cyclone were packed full of solidified ash and char. A hole had been eroded from the inside through the cyclone shell below the inlet opening.

It appears that ash had accumulated on the refractory remaining at the barrel section and outlet tube of the internal cyclone. This narrowed the gas passage and created high velocities, which were responsible for the eroded hole and high back pressure that caused the bed to evacuate through the dipleg.

The first-stage quencher was deformed at its inner shell. The second-stage cyclone was ultrasonically checked for wear; no significant wear was found. The third-stage cyclone, second-stage quencher, and the product gas scrubber were free of any tar deposits and surface wear.

- The hole in the barrel portion of the internal cyclone was patched, and the refractory lining remaining at the barrel and outlet tube was removed.
- The warped lower 2 feet of the inner shell of the first-stage quencher was replaced.
- The internal cyclone dipleg exit pipe was modified to provide a 30-degree offset to prevent backgassing of bubbles from the fluidized bed.
- Three damaged thermowells in the upper bed and lower freeboard were replaced.

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- The L-valve feed nozzle was extended 3 inches into the reactor to minimize char deposits on the reactor wall in the vicinity of this feed port.
- A system for cooling the product gas with nitrogen or steam injection prior to the internal cyclone was installed.

Conclusion

It appears that lowering the coal feed injection point reduced the formation of condensible tars to trace levels, as indicated by the uninterrupted operation of the product gas scrubber and clean scrubber effluent.

Test Run 133 — February 1979

Summary

Run 133 accomplished the major objectives specified by the Technical Subcommittee prior to the run as well as the objectives of Task 9, Phase 1, of the overall program. For the first time, the U-GAS pilot plant operated continuously with coal feed for 138 hours at a reactor pressure of 42 psig (almost 60 psia). In all, 102 tons of coal were gasified. Ash-balanced operation was achieved throughout the run with the production of ash agglomerates and 250 to 260 Btu/SCF product gas. The run had five nearly continuous steady-state periods of operation, the longest of which lasted 3 days. The run was voluntarily terminated.

During Run 133, attempts to return the second-stage cyclone fines to the bed were not successful due to backgassing up the second-stage cyclone diplet. From information gained during and after Run 133, however, the dipleg height was increased by 3 feet and future runs should not experience the backgassing problem.

The post-run inspection revealed no significant accumulation of ash on the reactor walls, coal feed nozzle, or grid. However, the internal cyclone inlet was clogged with ash particulates and an ash deposit extended from the scroll inlet to the inside barrel. The product-gas scrubber, which operated satisfactorily during Run 133, experienced a minor accumulation of char fines.

Details

The test plan for Run 133 consisted of the following:

- Achieve ash-balanced operation of the U-GAS pilot plant with the formation of ash agglomerates, using washed Kentucky No. 9 coal as feed.

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- Operate the plant according to the conditions set forth in the Technical Subcommittee Meeting of February 13, 1979:

1. Bed depth: 4 feet
2. Bed temperature: 1850^oF
3. Superficial velocity: 4 ft/s
4. Internal cyclone: activated.

The technical Subcommittee promulgated the following test schedule:

1. Heat-up elapsed time, approximately 24 hours.
2. Start-up with coke, 6 to 8 hours.
3. Change from coal/coke blend to coal, 2 hours.
4. Ash agglomerate at atmospheric pressure, 6 hours.
5. Increase pressure at a rate of 5 psi/hr to a maximum of 60 psia (45 psig), 8 hours.
6. Run to the end of the coal supply with concurrent drying, 3 to 4 days.

During Run 133, the U-GAS pilot plant operated as follows:

- The test run spanned 193 hours with 102 tons of washed Kentucky coal fed for 138 hours. Ash agglomerates and 250 to 260 Btu/SCF product gas were produced from coal throughout 128 hours of continuous operation. Ash-balanced operation was maintained throughout the coal feed period.
- Heatup of the initial coke bed took 30 hours.
- Coke was fed to the gasifier for one hour, then coal was fed (101.6 tons) without being blended with coke.
- After ash agglomerates began to discharge, the reactor was operated at 13 psig for 4 hours.
- Over a period of 4 hours, the operating pressure was increased to 38 psig, at a bed temperature between 1830^o to 1850^oF and 3.5 ft/s superficial velocity.
- A feed problem developed at the higher operating pressure which eventually required a lower reactor pressure to remedy. The reactor pressure was then increased from 13 to 40 psig over a period of 12 hours, including 6 hours of operation at 28 psig.
- Stable operation at a 40 to 43 psig operating pressure was maintained for a period of 58 hours.

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- An electrical problem with the ash discharge lockhopper system required lowering the bed temperature for about 4 hours.
- Stable operation was maintained an additional 14 hours.
- The reactor pressure was reduced to 30 psig and stable operation maintained for 8 hours. During this period, several attempts were made to return the second-stage cyclone fines to the reactor; these were not successful.
- The reactor pressure was increased to 42 psig, and stable operation maintained for 10 hours.
- A feed interruption required cooling the reactor bed. The feed was restored and the run was voluntarily terminated, to facilitate an inspection.
- The post-run inspection revealed no accumulation of ash on the reactor walls or grid area. The internal cyclone inlet was filled with an ash accumulation. A heavy ash accumulation also existed in the scroll section and extended into the barrel and down to the cone. The internal cyclone outlet pipe had an ash accumulation on the inside.

The product gas scrubber experienced a minor accumulation of fines on the conical bottom.

Run 133 began Tuesday, February 20, 1979. The start-up burners were lit at 2000 hours on February 20, 1979, to heat the fluidized coke bed to the point of self-sustained reaction. At 0100 hours on February 21, 1979, the spark plug and flame-sensing element in the start-up burner malfunctioned, and repairs delayed the resumption of heating the bed until 1600 hours on February 21, 1979. Self-sustained reaction in the fluidized coke bed occurred 9 hour later, at 0100 hours on February 22, 1979.

The coke feed flowed slowly because of the addition of moisture condensing from the walls of the hopper in which it was stored. It was decided to feed coal immediately rather than wait until the wet coke was displaced with dry coke. The coal feed was not blended first with coke, as had been the pilot-plant practice, so that at 0200 hours on February 22, 1979, the feed consisted entirely of coal.

Ash agglomerates began to appear in the discharge of bed solids from the venturi at 1400 hours on February 22, 1979. The pilot plant operated an additional 4 hours at near-atmospheric pressure (14 psig). The coal feed rate averaged 1294 pounds/hour during this period, and the fines carryover past

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the internal cyclone averaged 131 pounds/hour. The bed solids discharge from the venturi averaged 185 pounds/hour, consisting almost entirely of ash agglomerates. A small (31 pound/hour) side withdrawal of bed solids was used starting at 1600 hours as a precaution to prevent the bed ash content from rising above 70%. The bed ash content stayed between 50% to 60%, and, at 2400 hours, the side withdrawal was discontinued.

Beginning at 1815 hours on February 22, 1979, the reactor pressure was raised in increments of 5 psi until it reached 40 psig at 2200 hours. Ash agglomerates continued to form and be discharged.

At about 0200 hours, and again at 0600 hours, 1400 hours, and 2000 hours on February 23, 1979, short feed interruptions occurred. These interruptions were caused by slow pressurization of the C-2 surge hopper. The pressure regulator on the nitrogen line to the hopper was removed, solving the problem. The reactor pressure had been reduced to 14 psig to allow coal to feed without interruption, and was then increased 5 psi per hour until it reached 28 psig at 0200 hours.

Uneven coal feed due to backgassing at the L-valve coal feed line was remedied by adjusting the aeration to the line while operating the reactor at 28 psig from 0200 through 0600 hours on February 24. A steady-state period of operation ensued.

During the steady-state period (1000 through 2200 hours on February 24, 1979, the reactor operated at 42 psig with a bed temperature of 1879°F. The coal feed rate averaged 1780 pounds/hour. The bed solids discharge rate from the venturi averaged 162 pounds/hour, consisting almost entirely of ash agglomerates. Nearly 96% of the discharge was ash. The fines carryover past the internal cyclone averaged 211 pounds/hour or about 11.9% of the coal feed. During previous near-atmospheric pressure test runs, a carryover of fines averaging about 12% of the coal feed has indicated satisfactory operation of the internal cyclone. The higher heating value of the water- and nitrogen-free product gas ranged from 254 to 263 Btu/SCF during this period.

At 2400 hours on February 24, 1979, it was decided to lower the bed temperature to 1850°F. A steady-state period of operation ensued from 0200 through 1000 hours on February 25, 1979. During this period, the reactor operated at 41 psig with a bed temperature of 1854°F. The coal feed rate

averaged 1420 pounds/hour. The bed solids discharge rate from the venturi averaged 62 pounds/hour, consisting almost entirely of ash agglomerates. This discharge was 95% ash. The fines carryover past the internal cyclone averaged 219 pounds/hour, or about 15.4% of the coal feed. The higher heating value of the water- and nitrogen-free product gas ranged from 247 to 255 Btu/SCF during this period.

After 1000 hours on February 25, 1979, the flow of steam and oxygen to the venturi was reduced, resulting in an increased rate of bed solids discharge from the venturi. A steady-state period of operation ensued, from 1200 hours on February 25, to 2000 hours on February 26, 1979. During this period, the reactor operated at 43 psig with a bed temperature of 1852°F. The coal feed rate averaged 1550 pounds/hour. The bed solids discharge rate from the venturi averaged 132 pounds/hour, consisting almost entirely of ash agglomerates varying in size from 20 mesh to 1/4 inch. This discharge was 92% ash. The fines carryover past the internal cyclone averaged 225 pounds/hour, or about 14.5% of the coal feed. The higher heating value of the water- and nitrogen-free product gas was 248 to 261 Btu/SCF during this period.

At 2200 hours on February 26, 1979, a short circuit developed in the electrical system of the ash discharge lockhopper. After this was repaired, another steady-state period of operation ensued, from 2400 hours on February 26 through 0800 hours on February 27, 1979. During this period, the reactor operated at 41.3 psig with a bed temperature of 1853°F. The coal feed rate averaged 1730 pounds/hour. The bed solids discharge rate from the venturi averaged 104 pounds/hour, consisting mostly of ash agglomerates. This discharge was 88% ash. The fines carryover past the internal cyclone averaged 246 pounds/hour, or about 14.2% of the coal feed. The higher heating value of the water- and nitrogen-free product gas was 250 Btu/SCF during this period.

At 1500 hours on February 27, 1979, an attempt was made to return the second-stage cyclone fines to the bed at a reduced pressure of 30 psig. Six attempts were made, but a seal of solids in the dipleg could not be maintained, preventing the fines from returning to the bed. The reactor pressure was even reduced to 14 psig, but the backgassing problem continued.

At 2030 hours on February 27, 1979, the attempts to return fines to the bed ceased and the reactor pressure was increased to 43 psig. A steady-state

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period of operation ensued, from 2200 hours on February 27 through 0800 hours on February 28, 1979. During this period, the reactor operated at 42 psig with a bed temperature of 1847°F. The coal feed rate averaged 1580 pounds/hour. The bed solids discharge rate from the venturi averaged 118 pounds/hour, consisting mostly of ash agglomerates. This discharge was 87% ash. The reactor bed was 40% ash, lower by 10% to 15% than during the previous steady-state periods. The fines carryover past the internal cyclone averaged 245 pounds/hour, or about 15.5% of the coal feed. The higher heating value of the nitrogen- and water-free product gas was 251 Btu/SCF during this period.

At 1000 hours on February 28, 1979, backgassing at the L-valve coal feed line caused coal to jam the reactor screw feeder. Coal feed was restored by 1100 hours, but 20 minutes later, the problem recurred. A vent in the coal feed surge hopper was opening prematurely. The reactor was idled while the circuitry was repaired. By 1400 hours, the problem was solved, but the bed temperature had dropped to 800°F.

At 1800 hours on February 28, 1979, it was decided to terminate Run 133 to allow an inspection of the reactor without its having been restarted. Figures A-7 and A-8 show material and energy balance summary for Run 133.

The reactor cooled and was opened on March 2. There was no accumulation of ash on the reactor walls, at the coal feed nozzle, nor on the grid, although a small buildup occurred at the dead zone created by a thermowell protruding from the grid. The internal cyclone was opened for inspection. The inlet was filled with ash particulates, some of which had fused into a mass adhering to the inside metallic walls of the cyclone. The accumulation extended from the scroll section into the barrel section, where it was as thick as 4 inches. The outlet pipe from the internal cyclone also had a 1-inch deposit of ash on the inside.

The product gas scrubber had operated satisfactorily during the run and was free of tar buildup. A thin layer of fines had accumulated at the conical bottom.

Fluidization tests of the second-stage cyclone fines showed them to have a fluidized density of 18.5 lb/ft³. Based on the recorded total pressure drop across the second-stage cyclone, a 17-foot height of fluidized fines in the second-stage cyclone dipleg should have prevented any backgassing of purge

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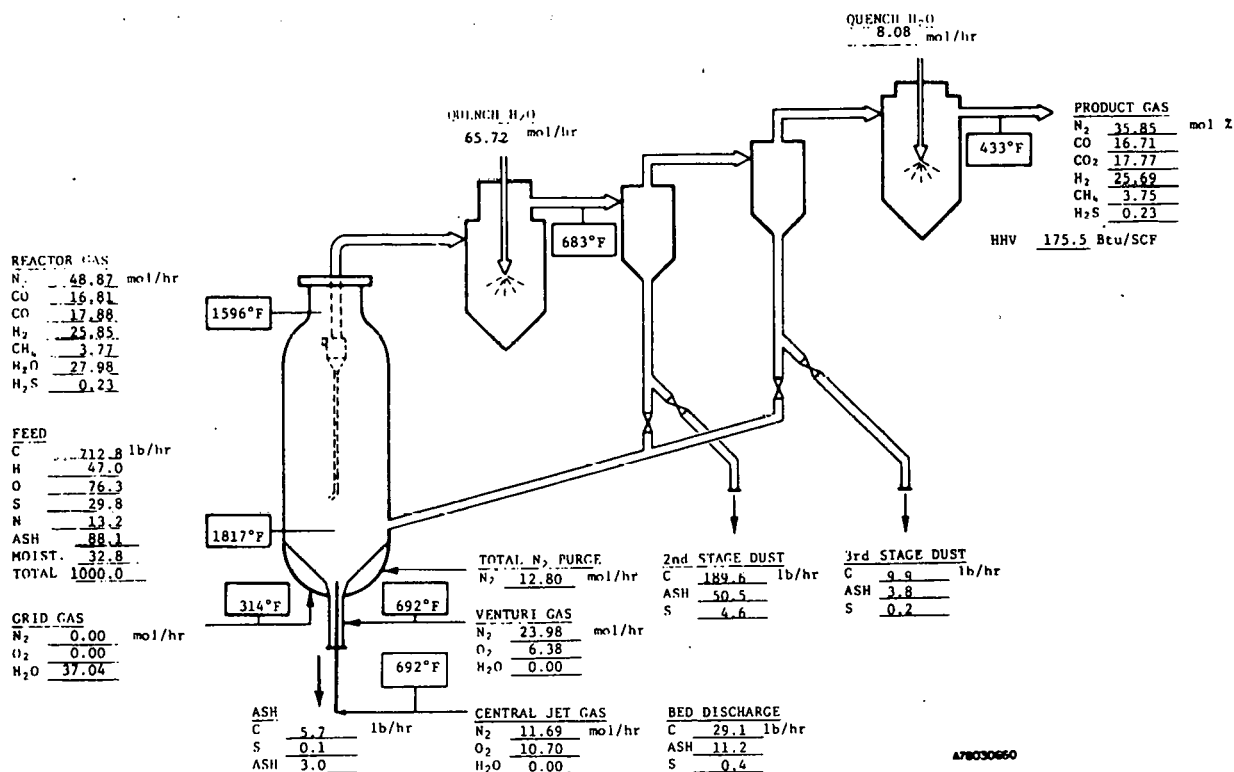
nitrogen or reactor gas; the actual dipleg height was 19 feet. An examination of the second-stage cyclone dipleg pressure drop traces during the shutdown revealed that a single nitrogen purge accounted for most of the variation indicative of backgassing. When this flow was reduced, the backgassing ceased.

Following Run 133, the second-stage cyclone dipleg height was increased by 3 feet to 22 feet. This height should ensure that the second-stage cyclone fines will be successfully returned to the bed in future tests.

The successful completion of Run 133 accomplished the major objectives of Task 9, Phase 1, of the MLGW Contract 8803. All U-GAS pilot plant operations under this task concluded on March 20, 1979.

Figures A-5 (a,b) through A-20 (a,b) are some of the mass and energy balances for steady-state periods from Runs 129 through 133. These figures also include operational variables such as cyclones in use, reactor and venturi diameters, and gasification mode.

Figure A-5a. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 129



AGG MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 129

DATE 10/24-10/24/78

TIME PERIOD 1400-2000

BAIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	712.8	38.1	13.2	29.3	50.7	105.4	1000.0	12205097.
TO VENTURI			671.9		0.0	204.0	875.9	131496.
TO GRID			0.0		74.7	532.6	667.3	520709.
TO CENTRAL JET			327.5		0.0	342.3	669.9	64102.
PURGES			358.6				358.6	3478.
TOTAL INPUT	712.8	38.1	1371.1	29.3	125.3	1244.4	3571.6	12860779.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTRI DISCHARGE	5.7	3.0	0.0	0.1	0.0	0.0	9.0	38335.
BED DISCHARGE	29.1	11.2	0.2	0.4	0.0	0.0	41.0	443082.
2ND STAGE FINES	189.6	50.5	1.7	4.6	1.0	0.5	243.0	2958170.
3RD STAGE FINES	9.9	3.8	0.1	0.2	0.1	0.0	14.0	150770.
BED INV. CHANGE	-3.0	-1.1	-0.0	-0.0	-0.0	0.0	-4.2	-2431.
REACTOR GAS	461.6		1369.1	7.3	124.2	1288.3	3251.0	3025294.
TOTAL OUTPUT	692.9	67.5	1371.1	12.6	125.3	1289.3	3558.8	12913270.
NET (OUT-IN)	-19.9	-20.6	0.0	-17.2	0.0	44.9	-12.3	52491.
%BALANCE OUT/IN	97.2	76.6	100.0	42.2	100.0	103.6	99.6	100.

Figure A-5b. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 129

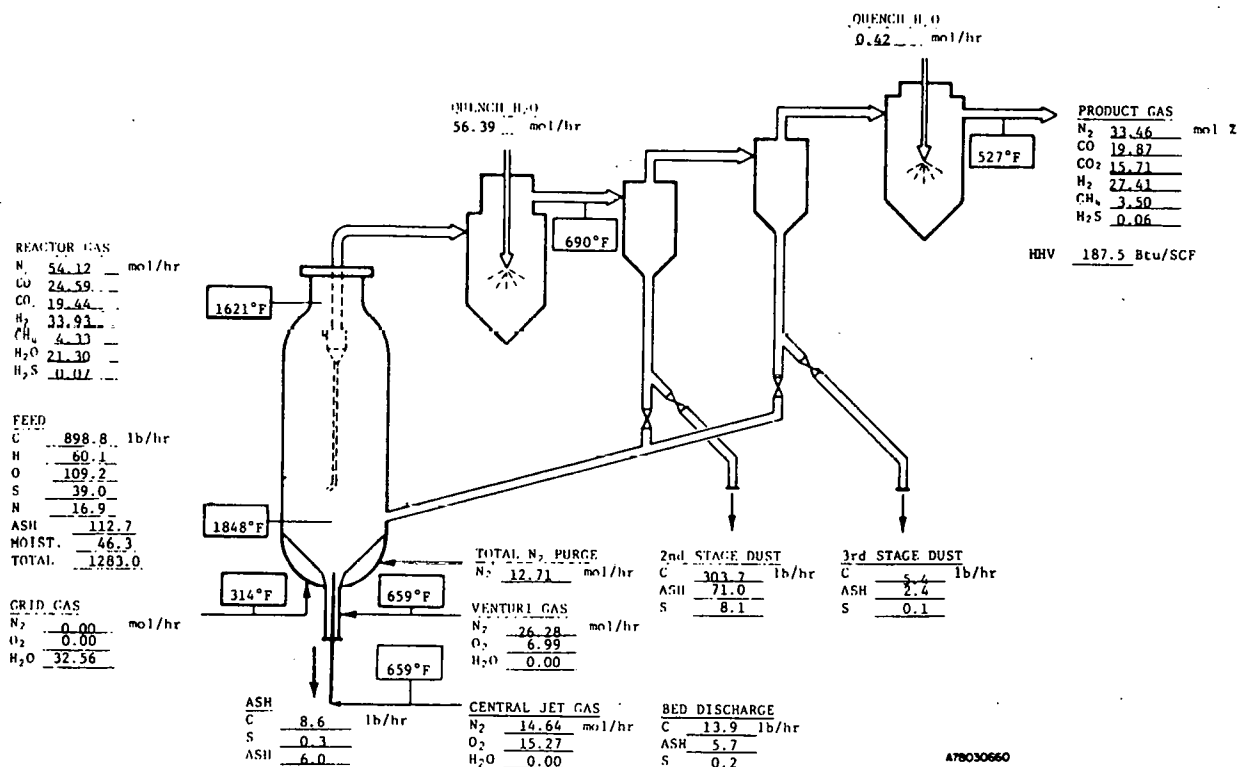
AAG OPERATING CONDITIONS SUMMARY

Run No. 129 Date 10/24/78 Time Period 1400 to 2000
Venturi Diameter 4.5 inches Feed Material Washed W.Ky. No.9 Coal
Cyclone Fines Collector: 2, Internal 0, External 2

1	Average Fluidized Bed Temperature, °F	1817
2	Average Reactor Pressure, psig	12.40
3	Venturi Pressure Drop, psi	0.52
4	Superficial Velocity, ft/s	3.2
5	Freeboard Velocity, ft/s	2.7
6	Venturi Velocity, ft/s	40.7
7	Grid Gas, mol/hr	37.04
8	Grid Gas Composition: %N ₂	0.0
9	%O ₂	0.0
10	%Steam	100.0
11	Venturi Gas, mol/hr	30.36
12	Venturi Gas Composition: %N ₂	79.0
13	%O ₂	21.0
14	%Steam	0.0
15	Central Jet Gas, mol/hr	22.39
16	Central Jet Gas Composition: %N ₂	52.22
17	%O ₂	47.78
18	%Steam	0.0
19	Fluidized Bed Density, lb/ft ³	31.90
20	Fluidized Bed Height, ft	4.32
21	Feed Composition, <u>1000</u> lb/hr, %C/%Ash	73.7 / 9.11
22	Average Fluidized Bed Composition, <u>-4.2</u> lb/hr, %C/%Ash	72.02 / 26.42
23	Bed Discharge Composition, <u>41</u> lb/hr, %C/%Ash	71.05 / 27.30
24	Venturi Discharge Composition, <u>9</u> lb/hr, %C/%Ash	63.84 / 33.85
25	2nd Stage Fines Composition, <u>248</u> lb/hr, %C/%Ash	76.46 / 20.38
26	3rd Stage Fines Composition, <u>14</u> lb/hr, %C/%Ash	70.40 / 27.42
27	Carbon/Oxygen (Weight Basis)	1.3
28	Carbon/Steam (Weight Basis)	1.1

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.63	12.7	19.9	18.3	19.0	15.0	9.33	3.13	1.95							
Fluidized Bed	0.15	10.9	35.8	21.1	20.8	8.76	1.59	0.47	0.41							
Discharged Bed	0.07	10.2	37.5	21.7	22.8	6.68	0.80	0.10	0.17							
Ash	4.84	51.8	31.4	5.88	4.53	0.69	0.26	0.17	0.35							
2nd Stage Fines						7.62	29.5	-	-	17.8	8.29	4.02	19.6	9.10	3.40	0.70
3rd Stage Fines						-	-	-	-	-	2.60	11.7	33.1	20.9	22.0	9.70

Figure A-6a. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130



ASH MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 130 DATE 11/ 3-11/ 3/78 TIME PERIOD 1000-2000

..BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

♦ INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
♦ FEED	898.8	112.7	16.9	39.0	65.3	150.3	1283.0	15530274.
♦ TO VENTURI			736.3		0.0	223.6	959.8	136163.
♦ TO GRID			0.0		65.6	521.0	586.6	475672.
♦ TO CENTRAL JET			410.1		0.0	488.7	898.7	75683.
♦ PURGES			356.0				356.0	2923.
♦ TOTAL INPUT	898.8	112.7	1519.3	39.0	130.9	1383.5	4084.2	16145032.
♦ WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
♦ VENTRI DISCHARGE	8.6	6.0	0.1	0.3	0.0	0.1	15.0	133606.
♦ BED DISCHARGE	13.9	5.7	0.1	0.2	0.1	0.0	20.0	212741.
♦ 2ND STAGE FINES	303.7	71.0	3.0	8.1	1.9	0.3	388.0	4705349.
♦ 3RD STAGE FINES	5.4	2.4	0.0	0.1	0.0	0.0	8.0	83342.
♦ BED INV. CHANGE	-4.8	-2.6	-0.0	-0.1	-0.0	-0.0	-7.5	-4288.
♦ REACTOR GAS	580.2		1516.2	2.2	128.9	1356.2	3583.7	11219744.
♦ TOTAL OUTPUT	907.1	82.6	1519.3	10.8	130.9	1356.5	4007.2	16600494.
♦ NET (OUT-IN)	8.3	-30.1	0.0	-28.2	-0.0	-27.1	-77.1	455462.
♦ %BALANCE OUT/IN	100.9	73.3	100.0	27.7	100.0	98.0	98.1	103.

Figure A-6b. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130

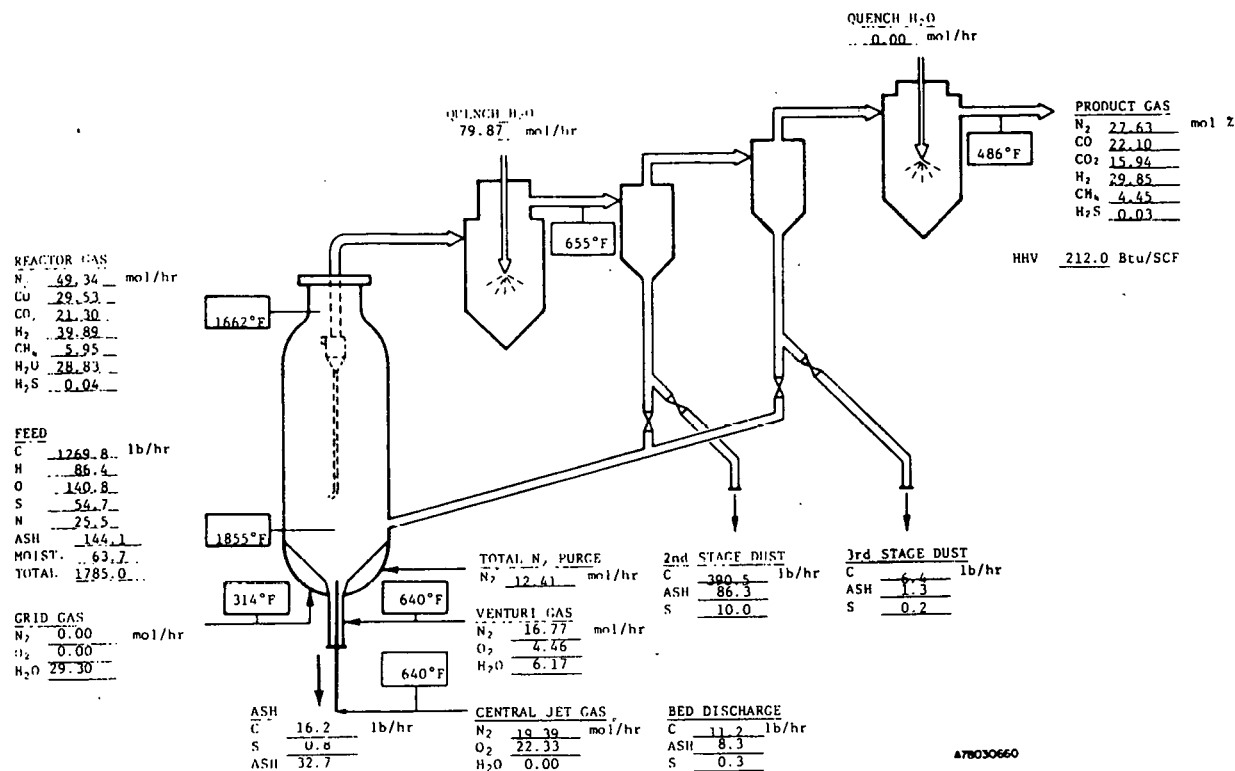
AAG OPERATING CONDITIONS SUMMARY

Run No. 130 Date 11/3/78 Time Period 1000 to 2000
Venturi Diameter 4.5 inches Feed Material Washed W.Ky.No.9 Coal
Cyclone Fines Collector: 2, Internal 0, External 2

1	Average Fluidized Bed Temperature, °F	1848
2	Average Reactor Pressure, psig	9.10
3	Venturi Pressure Drop, psi	0.50
4	Superficial Velocity, ft/s	3.9
5	Freeboard Velocity, ft/s	3.5
6	Venturi Velocity, ft/s	59.3
7	Grid Gas, mol/hr	32.56
8	Grid Gas Composition: %N ₂	0.0
9	%O ₂	0.0
10	%Steam	100.0
11	Venturi Gas, mol/hr	33.27
12	Venturi Gas Composition: %N ₂	79.00
13	%O ₂	21.00
14	%Steam	0.00
15	Central Jet Gas, mol/hr	29.91
16	Central Jet Gas Composition: %N ₂	48.94
17	%O ₂	51.06
18	%Steam	0.00
19	Fluidized Bed Density, lb/ft ³	29.01
20	Fluidized Bed Height, ft	4.21
21	Feed Composition, <u>1283</u> lb/hr, %C/%Ash	72.68 / 9.11
22	Average Fluidized Bed Composition, <u>-7.5</u> lb/hr, %C/%Ash	63.32 / 34.32
23	Bed Discharge Composition, <u>20</u> lb/hr, %C/%Ash	69.41 / 28.74
24	Venturi Discharge Composition, <u>15</u> lb/hr, %C/%Ash	57.16 / 40.10
25	2nd Stage Fines Composition, <u>388</u> lb/hr, %C/%Ash	78.28 / 18.30
26	3rd Stage Fines Composition, <u>8</u> lb/hr, %C/%Ash	67.66 / 29.79
27	Carbon/Oxygen (Weight Basis)	1.3
28	Carbon/Steam (Weight Basis)	1.5

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.23	8.01	18.6	20.1	20.3	16.2	9.78	4.21	2.44							
Fluidized Bed	0.13	6.22	29.3	27.8	28.7	6.62	0.80	0.17	0.23							
Discharged Bed	0.12	12.1	35.0	21.8	22.4	6.71	1.59	0.15	0.16							
Ash	1.02	30.6	39.4	13.4	13.4	0.89	0.45	0.29	0.51							
2nd Stage Fines																
3rd Stage Fines																

Figure A-7a.. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130



GAS MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 130

DATE 11/ 4-11/ 5/78

TIME PERIOD 1600-0600

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
• FEED	1269.8	144.1	25.5	54.7	93.5	197.4	1785.0	21985852.
• TO VENTURI			469.8		12.4	241.4	723.6	301114.
• TO GRID			0.0		59.1	468.8	527.9	428047.
• TO CENTRAL JET			543.3		0.0	714.4	1257.7	95258.
• PURGES			347.6				347.6	2508.
• TOTAL INPUT	1269.8	144.1	1386.2	54.7	165.0	1622.0	4641.8	22717521.
• WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
• VNTRI DISCHARGE	16.2	32.7	0.3	0.8	0.1	0.0	50.0	263124.
• BED DISCHARGE	11.2	8.3	0.1	0.3	0.0	0.0	20.0	176303.
• 2ND STAGE FINES	390.5	86.3	3.7	10.0	2.2	3.3	496.0	6061362.
• 3RD STAGE FINES	6.4	1.3	0.0	0.2	0.1	0.0	8.0	99831.
• BED INV. CHANGE	-1.8	-1.3	-0.0	-0.0	-0.0	-0.0	-3.1	-1749.
• REACTOR GAS	681.4		1382.1	1.4	162.6	1615.5	3843.1	13613428.
• TOTAL OUTPUT	1104.0	127.2	1386.2	12.6	165.0	1618.9	4414.0	20462299.
• NET (OUT-IN)	-165.8	-16.9	0.0	-42.1	0.0	-3.1	-227.9	-2255222.
• %BALANCE OUT/IN	86.9	88.3	100.0	23.1	100.0	99.8	95.1	90.

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Figure A-7b. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130

AAG OPERATING CONDITIONS SUMMARY

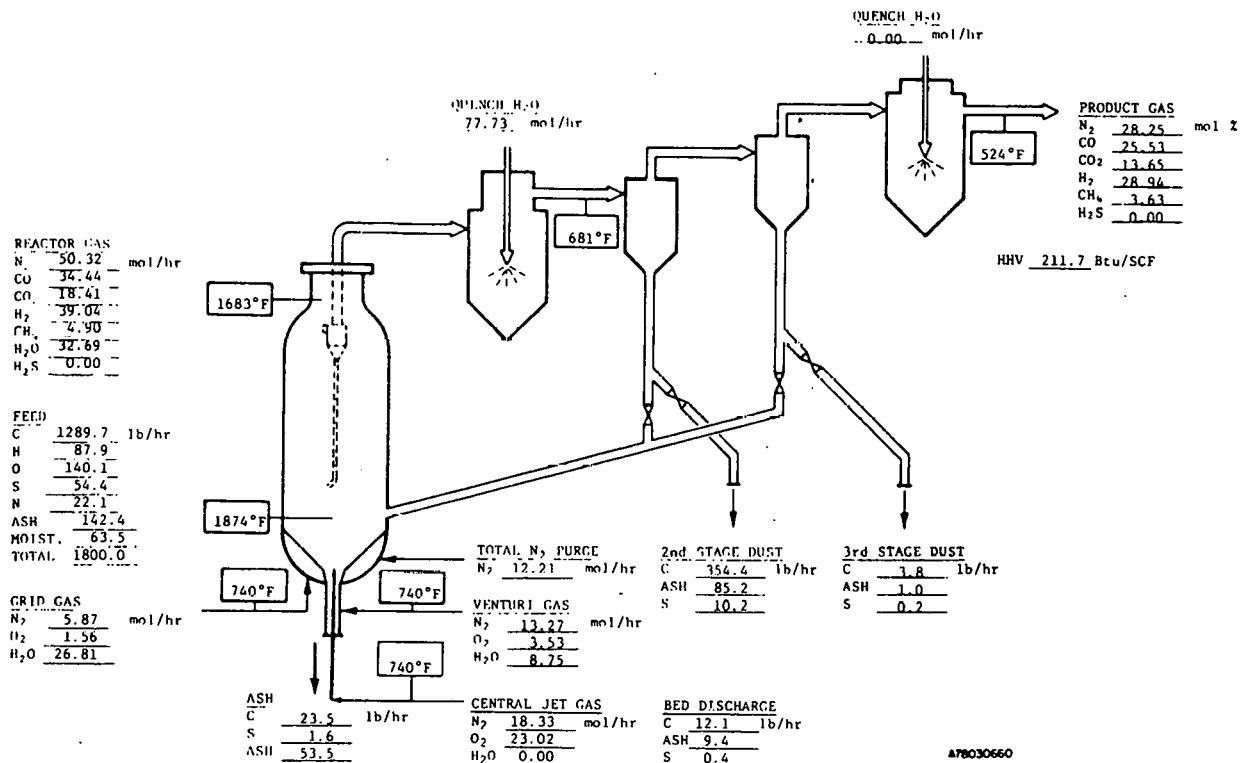
11/4/78-

Run No. 130 Date 11/5/78 Time Period 1600 to 0600Venturi Diameter 4.5 inches Feed Material Washed W.Ky.No.9 CoalCyclone Fines Collector: 2, Internal 0, External 2

1	Average Fluidized Bed Temperature, °F	1855
2	Average Reactor Pressure, psig	8.57
3	Venturi Pressure Drop, psi	0.42
4	Superficial Velocity, ft/s	4.1
5	Freeboard Velocity, ft/s	4.0
6	Venturi Velocity, ft/s	50.8
7	Grid Gas, mol/hr	29.3
8	Grid Gas Composition: %N ₂	0.0
9	%O ₂	0.0
10	%Steam	100.0
11	Venturi Gas, mol/hr	27.4
12	Venturi Gas Composition: %N ₂	61.21
13	%O ₂	16.27
14	%Steam	22.52
15	Central Jet Gas, mol/hr	41.72
16	Central Jet Gas Composition: %N ₂	46.49
17	%O ₂	53.51
18	%Steam	0.00
19	Fluidized Bed Density, lb/ft ³	24.93
20	Fluidized Bed Height, ft	4.15
21	Feed Composition, <u>1785</u> lb/hr, %C/%Ash	73.77 / 8.37
22	Average Fluidized Bed Composition, <u>-3.1</u> lb/hr, %C/%Ash	55.81 / 41.77
23	Bed Discharge Composition, <u>20</u> lb/hr, %C/%Ash	56.18 / 41.39
24	Venturi Discharge Composition, <u>50</u> lb/hr, %C/%Ash	32.41 / 65.39
25	2nd Stage Fines Composition, <u>496</u> lb/hr, %C/%Ash	78.73 / 17.39
26	3rd Stage Fines Composition, <u>8</u> lb/hr, %C/%Ash	79.41 / 16.25
27	Carbon/Oxygen (Weight Basis)	1.3
28	Carbon/Steam (Weight Basis)	2.0

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.24	13.7	24.1	19.2	20.3	12.7	6.12	2.34	1.34							
Fluidized Bed	0.00	3.88	27.5	34.6	25.6	7.01	0.84	0.25	0.29							
Discharged Bed	0.03	5.47	33.9	27.6	28.2	4.00	0.46	0.12	0.18							
Ash	0.19	9.83	30.8	28.2	28.5	1.65	0.43	0.22	0.24							
2nd Stage Fines																
3rd Stage Fines																

Figure A-8a. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130



ASB MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 130

DATE 11/ 5-11/ 6/78

TIME PERIOD 2000-0400

BAIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

♦ INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
♦ FEED	1289.7	142.4	22.1	54.4	95.0	196.6	1800.0	22372551.
♦ TO VENTURI			371.8		17.6	252.9	642.3	396611.
♦ TO GRID			164.4		54.0	478.9	697.4	426441.
♦ TO CENTRAL JET			513.4		0.0	736.7	1250.1	106856.
♦ PURGES			342.0				342.0	4169.
♦ TOTAL INPUT	1289.7	142.4	1413.7	54.4	166.7	1665.0	4731.8	23189772.
♦ WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
♦ VNTRI DISCHARGE	23.5	53.5	0.4	1.6	0.1	0.0	79.0	389718.
♦ BED DISCHARGE	12.1	9.4	0.1	0.4	0.0	0.0	22.0	190059.
♦ 2ND STAGE FINES	354.4	85.2	3.6	10.2	2.1	0.4	456.0	5487572.
♦ 3RD STAGE FINES	3.8	1.0	0.0	0.2	0.0	0.0	5.0	59195.
♦ BED INV. CHANGE	-6.0	-7.0	-0.1	-0.3	-0.0	-0.0	-13.4	-7220.
♦ REACTOR GAS	692.9		1409.7	0.0	164.4	1663.1	3930.1	13841290.
♦ TOTAL OUTPUT	1080.7	142.0	1413.7	12.1	166.7	1663.5	4478.7	20210614.
♦ NET (OUT-IN)	-208.9	-0.4	0.0	-42.3	0.0	-1.5	-253.1	-2979157.
♦ %BALANCE OUT/IN	83.8	99.7	100.0	22.2	100.0	99.9	94.7	87.

Figure A-8b. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130

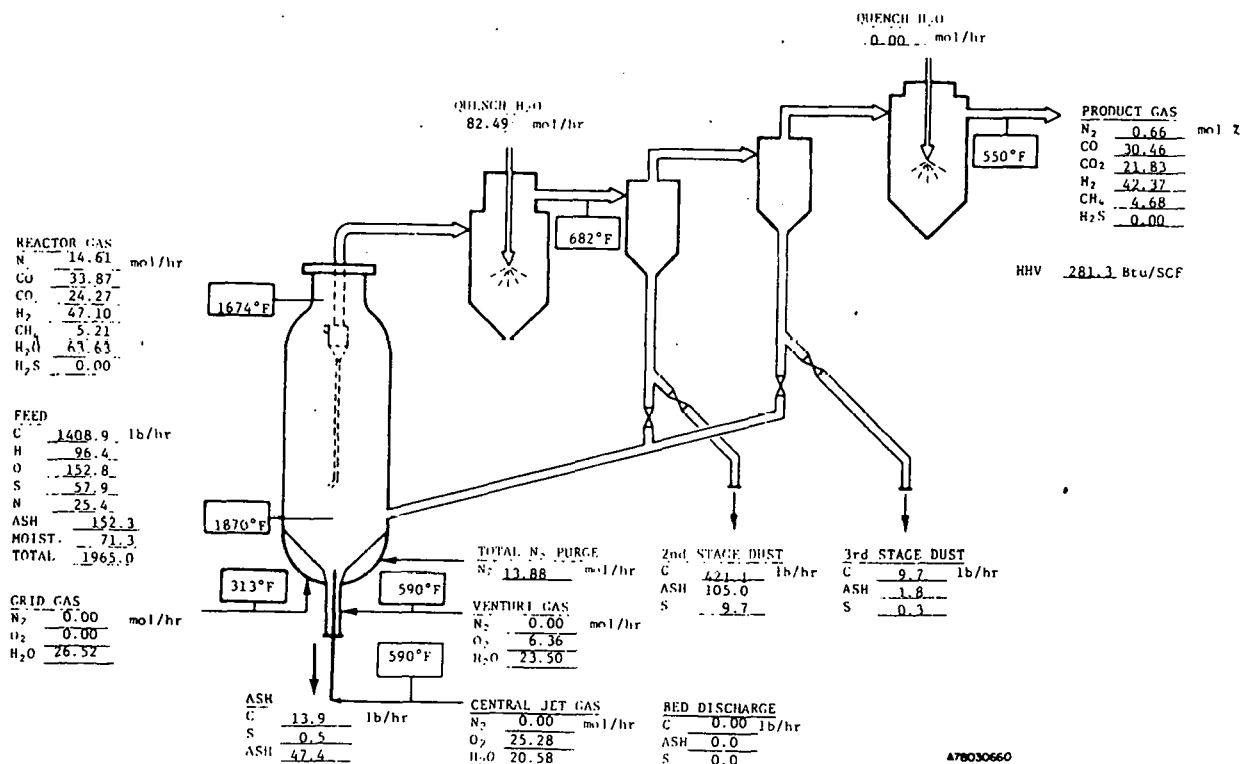
AAG OPERATING CONDITIONS SUMMARY

Run No. 130 Date 11/5/78-
11/6/78 Time Period 2000 to 0400
Venturi Diameter 4.5 inches Feed Material Washed W.Ky. No.9 Coal
Cyclone Fines Collector: 2, Internal 0, External 2

1	Average Fluidized Bed Temperature, °F	1874
2	Average Reactor Pressure, psig	9.50
3	Venturi Pressure Drop, psi	0.36
4	Superficial Velocity, ft/s	4.1
5	Freeboard Velocity, ft/s	4.0
6	Venturi Velocity, ft/s	45.9
7	Grid Gas, mol/hr	34.24
8	Grid Gas Composition: %N ₂	17.14
9	%O ₂	4.56
10	%Steam	78.30
11	Venturi Gas, mol/hr	25.55
12	Venturi Gas Composition: %N ₂	51.95
13	%O ₂	13.81
14	%Steam	34.25
15	Central Jet Gas, mol/hr	41.35
16	Central Jet Gas Composition: %N ₂	44.32
17	%O ₂	55.68
18	%Steam	0.00
19	Fluidized Bed Density, lb/ft ³	24.46
20	Fluidized Bed Height, ft	4.17
21	Feed Composition, <u>1800</u> lb/hr, %C/%Ash	74.27 / 8.20
22	Average Fluidized Bed Composition, <u>-13.41</u> lb/hr, %C/%Ash	45.09 / 52.22
23	Bed Discharge Composition, <u>22</u> lb/hr, %C/%Ash	54.84 / 42.61
24	Venturi Discharge Composition, <u>79</u> lb/hr, %C/%Ash	29.78 / 67.66
25	2nd Stage Fines Composition, <u>456</u> lb/hr, %C/%Ash	77.73 / 18.68
26	3rd Stage Fines Composition, <u>5</u> lb/hr, %C/%Ash	75.42 / 20.00
27	Carbon/Oxygen (Weight Basis)	1.2
28	Carbon/Steam (Weight Basis)	2.0

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.23	17.7	30.9	20.1	17.1	8.68	3.54	1.06	0.66							
Fluidized Bed	0.00	3.93	26.0	30.6	30.5	5.55	1.82	0.78	0.81							
Discharged Bed	0.03	5.47	33.9	27.6	28.2	4.00	0.46	0.12	0.18							
Ash	0.06	7.98	27.5	28.7	34.0	0.73	0.36	0.25	0.36							
2nd Stage Fines																
3rd Stage Fines																

Figure A-9a. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130



ASH MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 130

DATE 11/ 6-11/ 7/78

TIME PERIOD 2200-0800

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	1408.9	152.3	25.4	57.9	104.4	216.2	1965.0	24432137.
TO VENTURI			0.0		47.4	579.5	626.9	700339.
TO GRID			0.0		53.5	424.3	477.8	384496.
TO CENTRAL JET			0.0		41.5	1138.3	1179.8	499384.
PURGES			388.8				388.8	1741.
TOTAL INPUT	1408.9	152.3	414.2	57.9	246.7	2358.2	4638.2	25518713.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTRI DISCHARGE	13.9	47.4	0.2	0.5	0.1	0.0	62.0	233972.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	421.1	105.0	4.6	9.7	2.3	4.3	547.0	6571281.
3RD STAGE FINES	9.7	1.8	0.1	0.3	0.1	0.1	12.0	152468.
BED INV. CHANGE	-1.2	-1.2	-0.0	-0.0	-0.0	-0.0	-2.4	-1315.
REACTOR GAS	760.1		409.3	0.0	244.2	2336.5	3750.2	15672880.
TOTAL OUTPUT	1203.6	153.0	414.2	10.4	246.7	2340.9	4368.8	22879286.
NET (OUT-IN)	-205.3	0.7	0.0	-47.6	-0.0	-17.3	-269.5	-2639428.
%BALANCE OUT/IN	85.4	100.5	100.0	17.9	100.0	99.3	94.2	90.

Figure A-9b. ASH ACCLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130

AAG OPERATING CONDITIONS SUMMARY

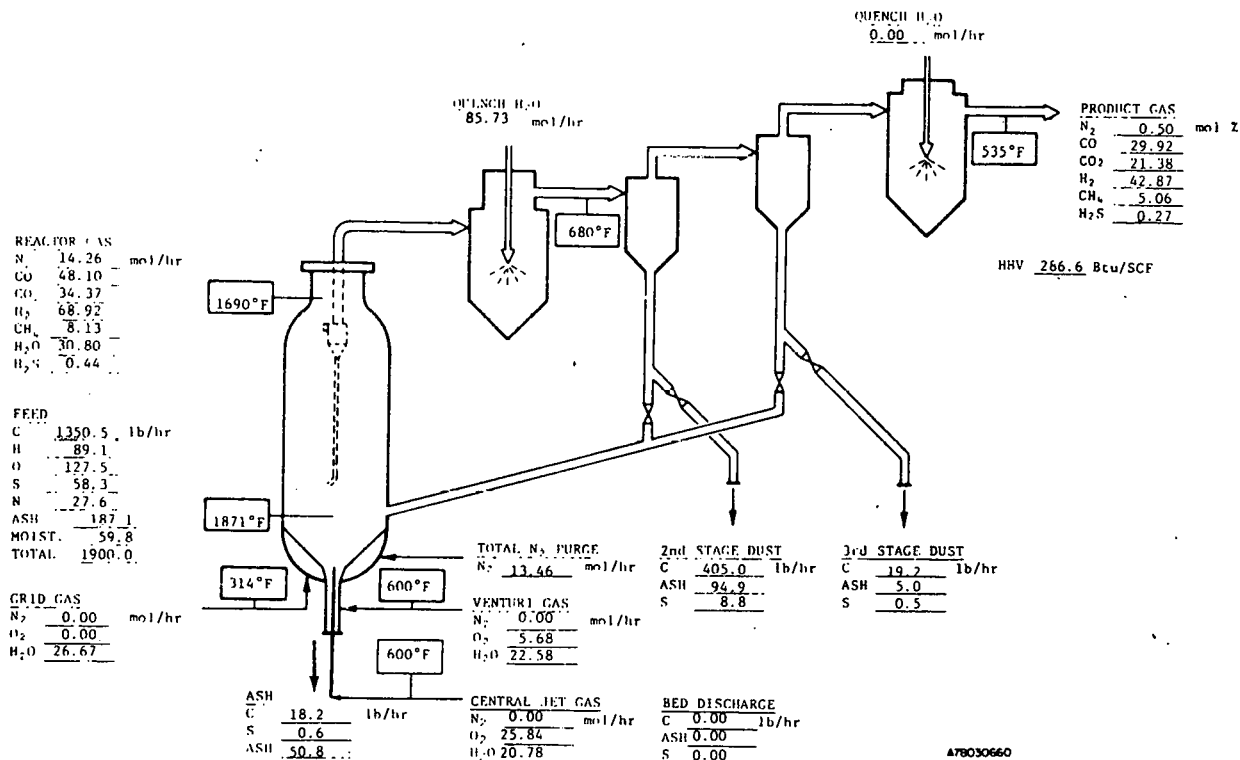
11/6/78-

Run No. 130 Date 11/7/78 Time Period 2200 to 0800
 Venturi Diameter 4.5 inches Feed Material Washed W.Ky.No.9 Coal
 Cyclone Fines Collector: 2, Internal 0, External 2

1	Average Fluidized Bed Temperature, °F	1870
2	Average Reactor Pressure, psig	8.85
3	Venturi Pressure Drop, psi	0.45
4	Superficial Velocity, ft/s	4.3
5	Freeboard Velocity, ft/s	4.3
6	Venturi Velocity, ft/s	57.2
7	Grid Gas, mol/hr	26.52
8	Grid Gas Composition: %N ₂	0.0
9	%O ₂	0.0
10	%Steam	100.0
11	Venturi Gas, mol/hr	29.86
12	Venturi Gas Composition: %N ₂	0.0
13	%O ₂	21.29
14	%Steam	78.71
15	Central Jet Gas, mol/hr	45.86
16	Central Jet Gas Composition: %N ₂	0.0
17	%O ₂	55.13
18	%Steam	44.87
19	Fluidized Bed Density, lb/ft ³	26.36
20	Fluidized Bed Height, ft	3.90
21	Feed Composition, <u>1965</u> lb/hr, %C/%Ash	74.40 / 8.04
22	Average Fluidized Bed Composition, <u>-2.4</u> lb/hr, %C/%Ash	48.44 / 49.73
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	0.0 / 0.0
24	Venturi Discharge Composition, <u>62</u> lb/hr, %C/%Ash	22.39 / 76.42
25	2nd Stage Fines Composition, <u>547</u> lb/hr, %C/%Ash	76.99 / 19.19
26	3rd Stage Fines Composition, <u>12</u> lb/hr, %C/%Ash	80.47 / 14.94
27	Carbon/Oxygen (Weight Basis)	1.0
28	Carbon/Steam (Weight Basis)	1.1

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.20	15.5	29.4	21.9	21.4	7.63	2.33	0.79	0.81							
Fluidized Bed	0.05	3.36	25.9	30.9	31.3	6.27	1.31	0.44	0.47							
Discharged Bed	-	-	-	-	-	-	-	-	-							
Ash	0.04	3.16	13.8	40.0	40.7	1.49	0.49	0.13	0.18							
2nd Stage Fines																
3rd Stage Fines																

Figure A-10a. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130



ASR MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 130

DATE 11/7-11/7/78

TIME PERIOD 1400-2000

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	1350.5	187.1	27.6	58.3	95.8	180.7	1900.0	23253976.
TO VENTURI			0.0		45.5	543.0	588.5	697453.
TO GRID			0.0		53.8	426.7	480.5	339625.
TO CENTRAL JET			0.0		41.9	1159.3	1201.2	510704.
PURGES			377.1				377.1	3095.
TOTAL INPUT	1350.5	187.1	404.7	58.3	237.0	2309.7	4547.2	24344148.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTRI DISCHARGE	18.2	50.3	0.2	0.6	0.1	0.0	70.0	303164.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	405.0	94.9	4.9	8.8	2.0	1.3	517.0	6255815.
3RD STAGE FINES	19.2	5.0	0.2	0.5	0.1	0.0	25.0	298431.
BED INV. CHANGE	-4.3	-3.4	-0.0	-0.1	-0.0	-0.0	-7.8	-4378.
REACTOR GAS	1087.3		399.4	14.0	234.7	2362.4	4097.8	20887068.
TOTAL OUTPUT	1525.5	147.4	404.7	23.9	237.0	2363.7	4702.0	27990099.
NET (OUT-IN)	175.0	-39.8	0.0	-34.5	0.0	54.0	154.8	3645951.
MBALANCE OUT/IN	113.0	78.7	100.0	40.9	100.0	102.3	103.4	115.

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Figure A-10b. ASH AGGLOMERATING GASIFIER
OPERATING CONDITIONS SUMMARY FOR RUN 130

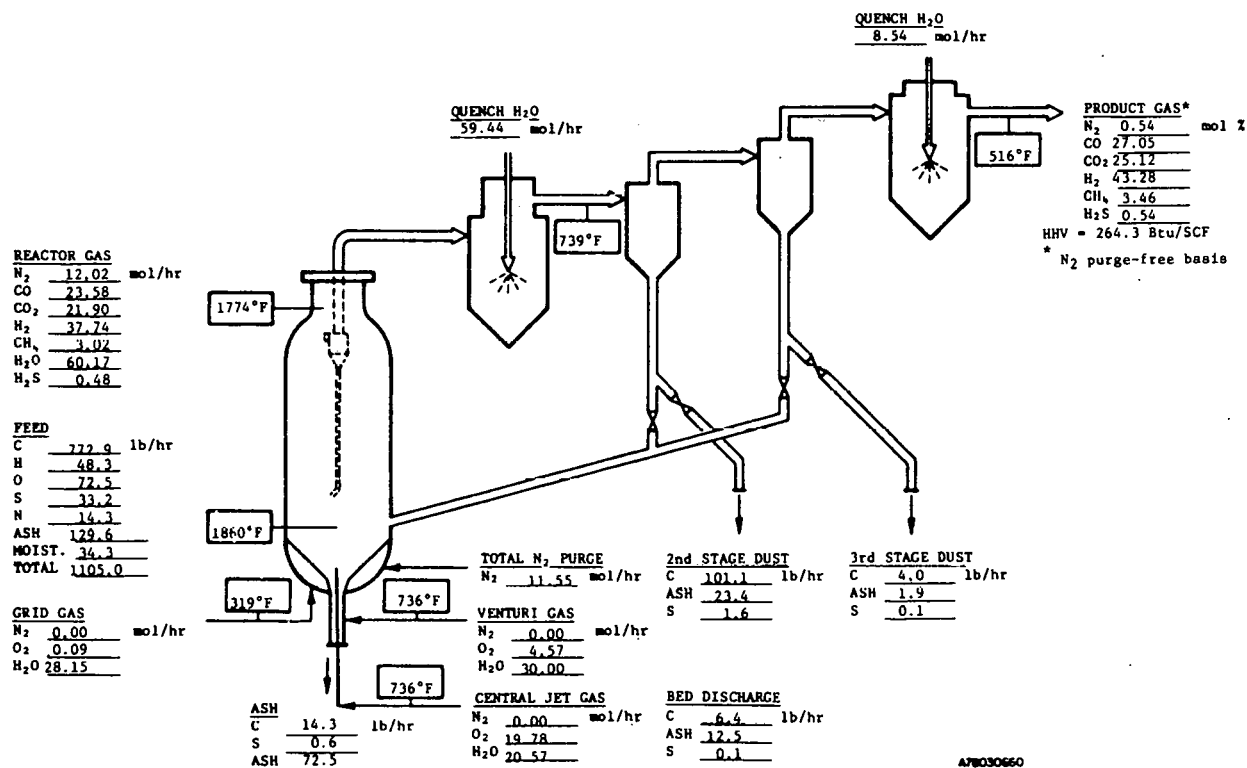
AAG OPERATING CONDITIONS SUMMARY

Run No. 130 Date 11/7/78 Time Period 1400 to 2000
Venturi Diameter 4.5 inches Feed Material Washed W.Ky.No.9 Coal
Cyclone Fines Collector: 2, Internal 0, External 2

1	Average Fluidized Bed Temperature, °F	1871
2	Average Reactor Pressure, psig	9.35
3	Venturi Pressure Drop, psi	0.41
4	Superficial Velocity, ft/s	4.2
5	Freeboard Velocity, ft/s	4.6
6	Venturi Velocity, ft/s	51.0
7	Grid Gas, mol/hr	26.67
8	Grid Gas Composition: %N ₂	0.00
9	%O ₂	0.00
10	%Steam	100.00
11	Venturi Gas, mol/hr	28.26
12	Venturi Gas Composition: %N ₂	0.00
13	%O ₂	20.09
14	%Steam	79.91
15	Central Jet Gas, mol/hr	46.62
16	Central Jet Gas Composition: %N ₂	0.00
17	%O ₂	55.42
18	%Steam	44.58
19	Fluidized Bed Density, lb/ft ³	24.93
20	Fluidized Bed Height, ft	4.05
21	Feed Composition, <u>1900</u> lb/hr, %C/%Ash	73.39/10.17
22	Average Fluidized Bed Composition, <u>-7.8</u> lb/hr, %C/%Ash	54.91/43.11
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	0.0 / 0.0
24	Venturi Discharge Composition, <u>70</u> lb/hr, %C/%Ash	26.07/72.57
25	2nd Stage Fines Composition, <u>517</u> lb/hr, %C/%Ash	78.34/18.36
26	3rd Stage Fines Composition, <u>25</u> lb/hr, %C/%Ash	76.74/19.95
27	Carbon/Oxygen (Weight Basis)	1.0
28	Carbon/Steam (Weight Basis)	1.1

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.14	9.40	25.0	23.0	23.4	10.4	5.67	2.36	0.63							
Fluidized Bed	0.09	4.82	30.4	28.9	29.8	4.68	0.75	0.19	0.28							
Discharged Bed	-	-	-	-	-	-	-	-	-							
Ash	0.25	9.59	26.9	31.3	30.6	0.58	0.33	0.20	0.30							
2nd Stage Fines																
3rd Stage Fines																

Figure A-11a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 131



U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 131

DATE 12/8-12/10/78

TIME PERIOD 2000-0200

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHERWISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	772.9	129.6	14.3	33.2	52.1	102.9	1105.0	13382171.
TO VENTURI			0.0		60.5	626.2	686.7	701378.
TO GRID			0.0		56.7	453.4	510.1	349520.
TO CENTRAL JET			0.0		41.5	962.0	1003.5	527287.
PURGES			323.6				323.6	-724.
TOTAL INPUT	772.9	129.6	337.9	33.2	210.8	2144.5	3628.8	14959631.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VNTRI DISCHARGE	14.3	72.5	0.3	0.6	0.0	0.0	86.0	257801.
BED DISCHARGE	6.4	12.5	0.1	0.1	0.0	0.0	19.0	105784.
2ND STAGE FINES	101.1	23.4	0.8	1.6	0.3	0.7	128.0	1546489.
3RD STAGE FINES	4.0	1.9	0.0	0.1	0.0	0.0	6.0	62118.
BED INV. CHANGE	-17.7	4.7	-0.1	-0.3	-0.0	-1.1	-15.3	-9538.
REACTOR GAS	582.0		336.8	15.2	210.5	2040.8	3185.4	12243937.
TOTAL OUTPUT	690.1	115.0	337.9	17.4	210.8	2040.4	3411.6	14456590.
NET (OUT-IN)	-82.8	-14.5	0.0	-15.8	0.0	-104.1	-217.2	-503041.
%BALANCE OUT/IN	89.3	88.8	100.0	52.3	100.0	95.1	94.0	97.

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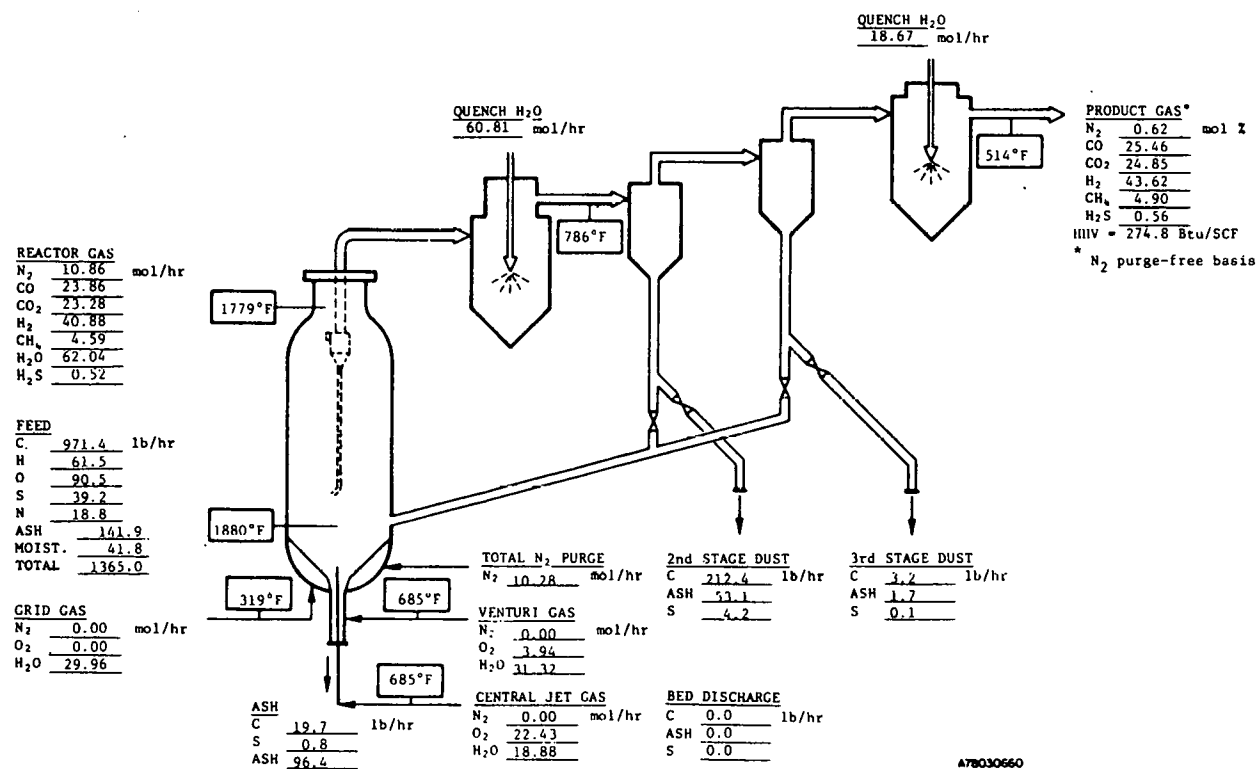
Figure A-11b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 131

AAG OPERATING CONDITIONS SUMMARY															
Run No. <u>131</u> Date <u>12/8-12/10</u> ^{/78} Time Period <u>2000</u> to <u>0200</u>															
Venturi Diameter <u>4.5</u> inches Feed Material <u>Washed W. Ky. #9 coal</u>															
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>															
1	Average Fluidized Bed Temperature, °F										1860				
2	Average Reactor Pressure, psig										9.4				
3	Venturi Pressure Drop, psi										0.69				
4	Superficial Velocity, ft/s 3 ft reactor diameter										4.2				
5	Freeboard Velocity, ft/s 4 ft reactor diameter										3.4				
6	Venturi Velocity, ft/s										67.8				
7	Grid Gas, mol/hr										28.24				
8	Grid Gas Composition: %N ₂										0.00				
9	%O ₂										0.34				
10	%Steam										99.66				
11	Venturi Gas, mol/hr										34.57				
12	Venturi Gas Composition: %N ₂										0.00				
13	%O ₂										13.21				
14	%Steam										86.79				
15	Central Jet Gas, mol/hr										40.35				
16	Central Jet Gas Composition: %N ₂										0.00				
17	%O ₂										49.02				
18	%Steam										50.98				
19	Fluidized Bed Density, lb/ft ³										38.23				
20	Fluidized Bed Height, ft										4.53				
21	Feed Composition, <u>1105</u> lb/hr, %C/%Ash										72.18/12.10				
22	Average Fluidized Bed Composition, <u>15.3</u> lb/hr, %C/%Ash										32.74/67.06				
23	Bed Discharge Composition, <u>19</u> lb/hr, %C/%Ash										33.45/65.93				
24	Venturi Discharge Composition, <u>86</u> lb/hr, %C/%Ash										16.57/84.30				
25	2nd Stage Fines Composition, <u>128</u> lb/hr, %C/%Ash										79.02/18.32				
26	3rd Stage Fines Composition, <u>6</u> lb/hr, %C/%Ash										66.08/31.34				
27	Carbon/Oxygen (Weight Basis)										0.99				
28	Carbon/Steam (Weight Basis)										0.55				

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.35	14.80	25.76	21.53	18.86	11.35	4.73	1.73	0.89							
Fluidized Bed	0.05	2.51	11.29	29.16	38.42	11.82	4.60	1.54	0.61							
Discharged Bed	0.06	1.00	5.20	12.90	44.09	21.89	9.60	4.17	1.09							
Ash	0.25	7.57	17.68	35.56	31.43	5.18	1.39	0.64	0.30							
2nd Stage Fines					0.03	0.72	8.52			9.98	5.08	8.54	34.47	21.27	10.37	1.02
3rd Stage Fines					-	-	-			-	-	3.67	5.25	8.83	49.67	32.58

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Figure A-12a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 131



U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 131

DATE 12/10-12/10/78

TIME PERIOD 0600-1400

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	971.4	141.9	18.8	39.2	66.2	127.6	1365.0	16835751.
TO VENTURI			0.0		63.1	627.2	690.3	711958.
TO GRID			0.0		60.4	479.4	539.8	368829.
TO CENTRAL JET			0.0		38.1	1019.9	1057.9	471479.
PURGES			288.0				288.0	-215.
TOTAL INPUT	971.4	141.9	306.8	39.2	227.8	2254.0	3941.0	18437803.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VNTRI DISCHARGE	19.7	96.4	0.3	0.8	0.0	0.0	116.0	362594.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	212.4	53.1	2.2	4.2	0.7	1.3	274.0	3306946.
3RD STAGE FINES	3.2	1.7	0.0	0.1	0.0	0.0	5.0	49850.
BED INV. CHANGE	9.4	-7.2	0.0	0.1	-0.0	0.0	2.9	2935.
REACTOR GAS	620.8		304.2	16.7	227.1	2119.3	3288.1	13438804.
TOTAL OUTPUT	865.5	144.0	306.8	21.8	227.8	2120.7	3686.5	17411129.
NET (OUT-IN)	-105.9	2.1	0.0	-17.3	0.0	-133.4	-254.5	-1026674.
%BALANCE OUT/IN	89.1	101.5	100.0	55.7	100.0	94.1	93.5	94.

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Figure A-12b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 131

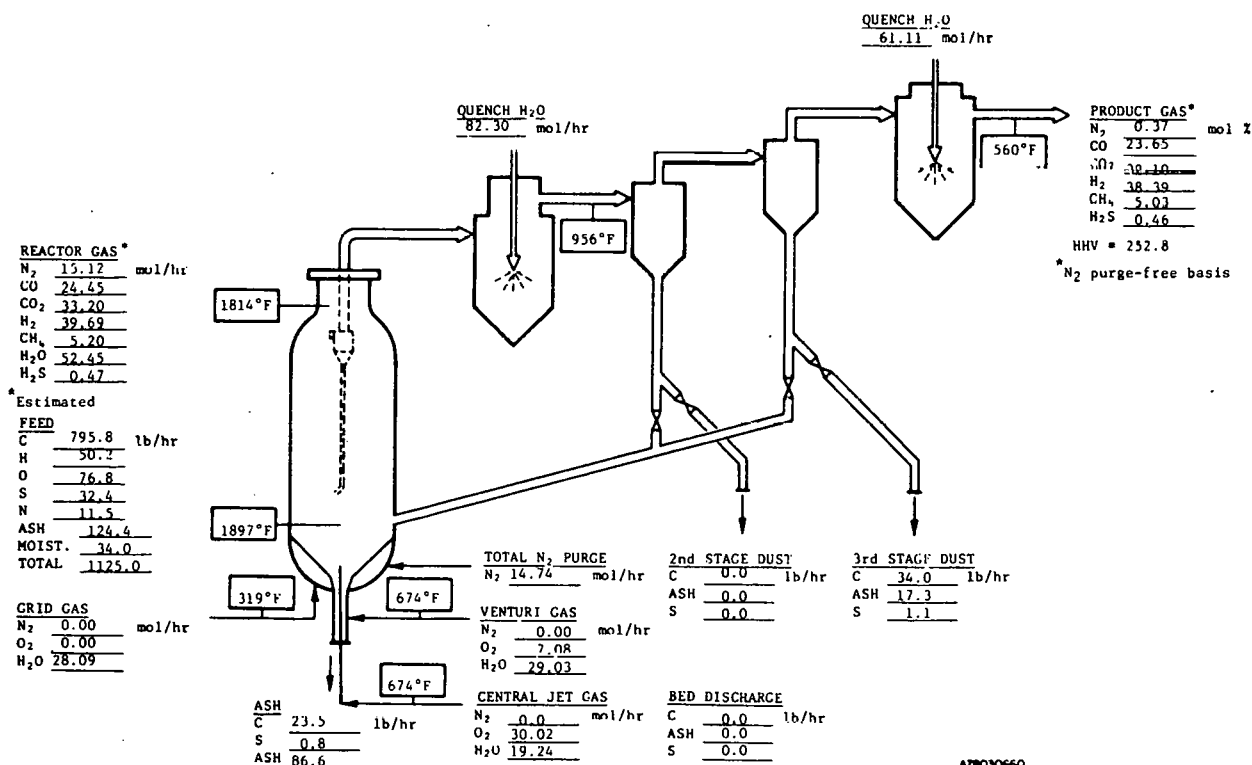
AAG OPERATING CONDITIONS SUMMARY

Run No. 131 Date 12/10/78 Time Period 0600 to 1400
 Venturi Diameter 4.5 inches Feed Material Washed W. Ky. #9 coal
 Cyclone Fines Collector: 2, Internal 0, External 2

1	Average Fluidized Bed Temperature, °F	1880
2	Average Reactor Pressure, psig	9.4
3	Venturi Pressure Drop, psi	0.59
4	Superficial Velocity, ft/s <u>3 ft reactor diameter</u>	4.3
5	Freeboard Velocity, ft/s <u>4 ft reactor diameter</u>	3.6
6	Venturi Velocity, ft/s	65.2
7	Grid Gas, mol/hr	29.96
8	Grid Gas Composition: %N ₂	0.00
9	%O ₂	0.00
10	%Steam	100.00
11	Venturi Gas, mol/hr	35.26
12	Venturi Gas Composition: %N ₂	0.00
13	%O ₂	11.17
14	%Steam	88.83
15	Central Jet Gas, mol/hr	41.31
16	Central Jet Gas Composition: %N ₂	0.00
17	%O ₂	54.30
18	%Steam	45.70
19	Fluidized Bed Density, lb/ft ³	33.02
20	Fluidized Bed Height, ft	4.22
21	Feed Composition, <u>1365</u> lb/hr, %C/%Ash	73.41/10.72
22	Average Fluidized Bed Composition, <u>2.9</u> lb/hr, %C/%Ash	35.08/65.22
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	0/0
24	Venturi Discharge Composition, <u>116</u> lb/hr, %C/%Ash	16.99/83.08
25	2nd Stage Fines Composition, <u>274</u> lb/hr, %C/%Ash	77.52/19.38
26	3rd Stage Fines Composition, <u>5</u> lb/hr, %C/%Ash	63.47/
27	Carbon/Oxygen (Weight Basis)	1.15
28	Carbon/Steam (Weight Basis)	0.67

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.31	17.14	32.96	26.11	15.89	5.49	1.39	0.46	0.25							
Fluidized Bed	0.02	3.17	15.21	45.83	26.59	6.57	1.67	0.04	0.54							
Discharged Bed	—	—	—	—	—	—	—	—	—							
Ash	0.15	5.15	15.16	58.16	18.56	1.61	0.63	0.31	0.27							
2nd Stage Fines					0.10	4.40	27.25			10.8	7.15	9.25	22.13	12.10	6.05	0.73
3rd Stage Fines					—	—	—			—	—	3.00	2.50	10.50	45.50	38.50

Figure A-13a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 131



U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY

RUN NO. 131

DATE 12/10-12/10/78

TIME PERIOD 1980-2040

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

*****	C	ASH	N	S	H	O	TOTAL	*****	BTU/HR
* INPUT									
* FEED	795.8	124.4	11.5	32.4	54.0	107.0	1125.0	*	13814559.
* TO VENTURI			0.0		58.5	591.1	749.6	*	658321.
* TO GRID			0.0		56.6	449.4	506.0	*	330924.
* TO CENTRAL JET			0.0		38.8	1268.6	1307.4	*	504523.
* PURGES			412.8				412.8	*	1335.
* TOTAL INPUT	795.8	124.4	424.2	32.4	207.9	2516.0	4100.7	*	15319661.
* WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	*	250000.
* VNTRI DISCHARGE	23.5	86.6	0.3	0.8	0.0	0.0	109.0	*	399971.
* BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	*	0.
* 2ND STAGE FINES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	*	0.
* 3RD STAGE FINES	34.0	17.3	0.3	1.1	0.2	0.0	53.0	*	542791.
* BED INV. CHANGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	*	0.
* REACTOR GAS	754.2		423.6	15.1	207.7	2292.7	3693.3	*	13576776.
* TOTAL OUTPUT	811.7	103.9	424.2	17.0	207.9	2292.8	3857.5	*	14769538.
* NET (OUT-IN)	15.9	-20.5	0.0	-15.4	0.0	-223.3	-243.2	*	-550122.
* BALANCE OUT/IN	102.0	83.5	100.0	52.6	100.0	91.1	94.1	*	96.

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Figure A-13b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 131

AAG OPERATING CONDITIONS SUMMARY

Run No. 131 Date 12-10-78 Time Period 1900 to 2040
 Venturi Diameter 4.5 inches Feed Material Washed W. Ky. #9 coal
 Cyclone Fines Collector: 1, Internal 0, External 1

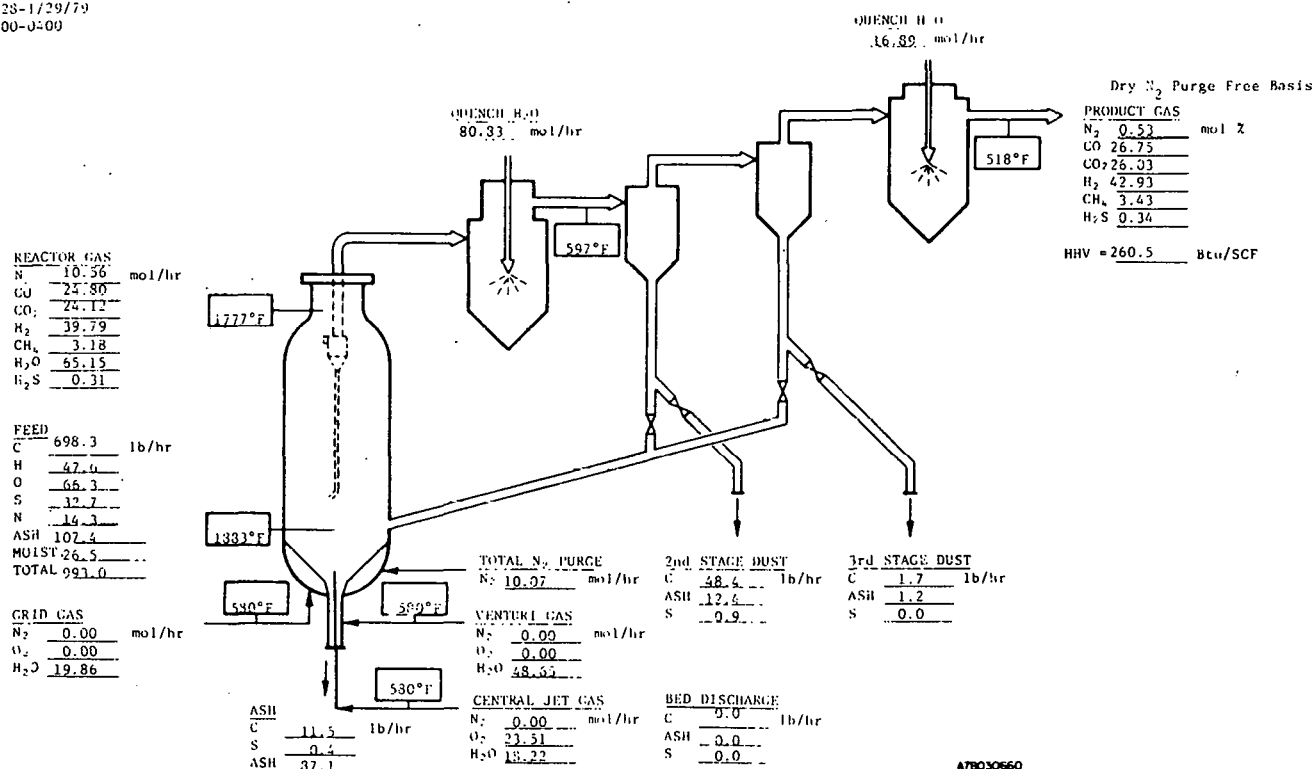
1	Average Fluidized Bed Temperature, °F	1897
2	Average Reactor Pressure, psig	11.8
3	Venturi Pressure Drop, psi	0.62
4	Superficial Velocity, ft/s <u>3 ft reactor diameter</u>	4.2
5	Freeboard Velocity, ft/s <u>4 ft reactor diameter</u>	3.4
6	Venturi Velocity, ft/s	61.1
7	Grid Gas, mol/hr	28.09
8	Grid Gas Composition: %N ₂	0.00
9	%O ₂	0.00
10	%Steam	100.00
11	Venturi Gas, mol/hr	36.11
12	Venturi Gas Composition: %N ₂	0.00
13	%O ₂	19.61
14	%Steam	80.39
15	Central Jet Gas, mol/hr	49.27
16	Central Jet Gas Composition: %N ₂	0.00
17	%O ₂	60.94
18	%Steam	39.06
19	Fluidized Bed Density, lb/ft ³	34.56
20	Fluidized Bed Height, ft	3.84
21	Feed Composition, <u>1125</u> lb/hr, %C/%Ash	72.94/11.40
22	Average Fluidized Bed Composition, <u>0</u> lb/hr, %C/%Ash	--
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	--
24	Venturi Discharge Composition, <u>109</u> lb/hr, %C/%Ash	21.55/79.42
25	2nd Stage Fines Composition, <u>0</u> lb/hr, %C/%Ash	--
26	3rd Stage Fines Composition, <u>53</u> lb/hr, %C/%Ash	64.20/32.64
27	Carbon/Oxygen (Weight Basis)	0.7
28	Carbon/Steam (Weight Basis)	0.6

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.62	21.52	38.45	24.04	11.08	3.14	.67	.25	.23							
Fluidized Bed	0.03	2.78	15.36	44.28	19.45	9.59	5.65	1.86	1.00							
Discharged Bed	-	-	-	-	-	-	-	-	-							
Ash	0.18	10.48	23.56	55.90	8.91	0.57	0.16	0.08	0.16							
2nd Stage Fines																
3rd Stage Fines																

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Figure A-14a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 132

Run 132
1/28-1/29/79
1600-0400



RUN NO. 132

DATE 1/28-1/29/79

TIME PERIOD 1600-0400

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHERWISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	698.3	107.4	14.3	32.7	50.5	89.8	993.0	12524704.
TO VENTURI			0.0		38.1	778.4	876.5	11198339.
TO GRID			0.0		40.0	317.8	357.8	246800.
TO CENTRAL JET			0.0		36.7	1043.8	1080.5	417782.
PURGES			282.1				282.1	-281.
TOTAL INPUT	698.3	107.4	296.4	32.7	225.4	2229.8	3589.9	14308637.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTRI DISCHARGE	11.5	37.1	0.1	0.4	0.0	0.0	38.0	162580.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	48.4	12.4	0.3	0.9	0.2	0.0	62.0	745034.
3RD STAGE FINES	1.7	1.2	0.0	0.0	0.0	0.0	3.0	27112.
BED INV. CHANGE	1.1	-2.6	0.1	0.4	0.1	0.0	-1.1	-437.
REACTOR GAS	625.2		295.8	10.4	225.0	2211.0	3367.4	12910701.
TOTAL OUTPUT	687.8	98.2	296.4	12.1	225.4	2211.0	3530.8	14095050.
NET (OUT-IN)	-10.5	-9.2	0.0	-20.6	0.0	-13.8	-59.1	-213583.
%BALANCE OUT-IN	99.5	91.4	100.0	36.9	100.0	99.2	98.4	99.

Figure A-14b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 132

AAG OPERATING CONDITIONS SUMMARY

Run No. 132 Date 1/28-1/29/79 Time Period 1600 to 0400
 Venturi Diameter 4.5 inches Feed Material Washed W. Ky. #9 Coal
 Cyclone Fines Collector: 3, Internal 1, External 2

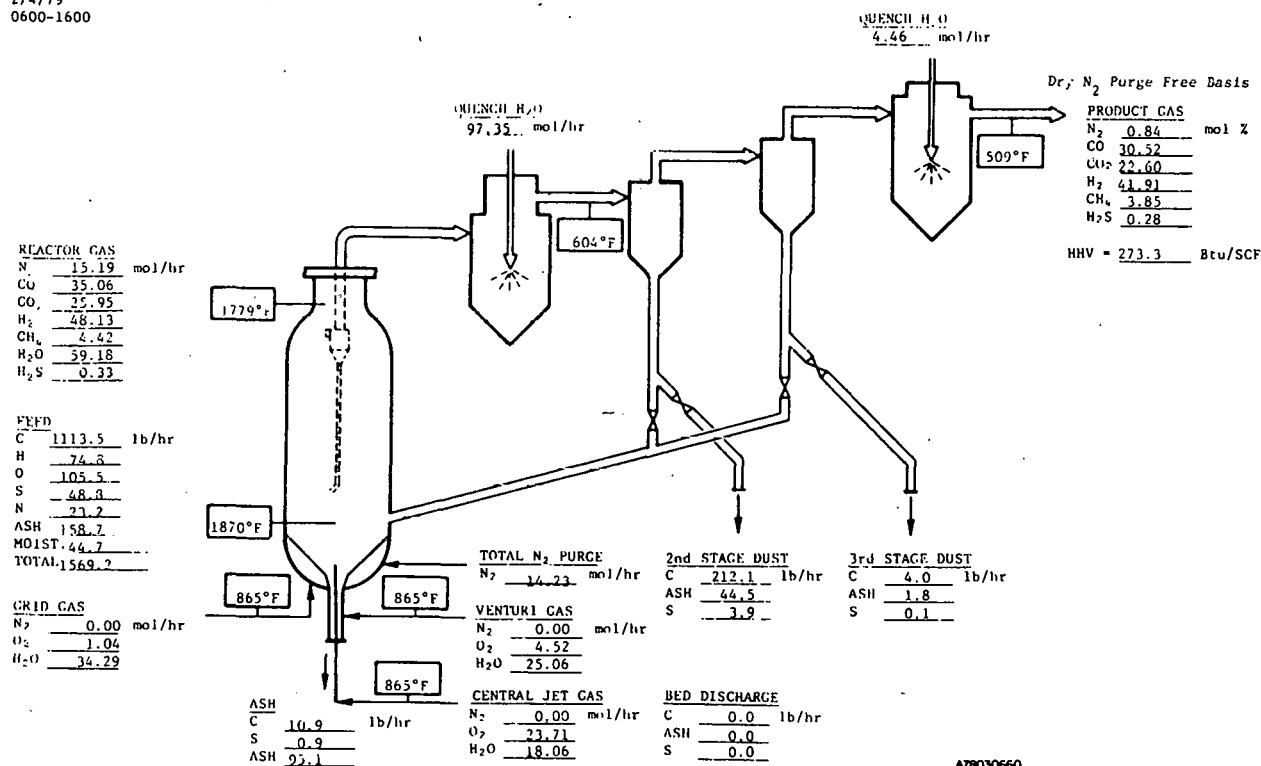
1	Average Fluidized Bed Temperature, °F	1883
2	Average Reactor Pressure, psig	12.5
3	Venturi Pressure Drop, psi	0.67
4	Superficial Velocity, ft/s	4.0
5	Freeboard Velocity, ft/s	3.4
6	Venturi Velocity, ft/s	78.3
7	Grid Gas, mol/hr	19.86
8	Grid Gas Composition: %N ₂	0.00
9	%O ₂	0.00
10	%Steam	100.00
11	Venturi Gas, mol/hr	48.65
12	Venturi Gas Composition: %N ₂	0.00
13	%O ₂	0.00
14	%Steam	100.00
15	Central Jet Gas, mol/hr	41.73
16	Central Jet Gas Composition: %N ₂	0.00
17	%O ₂	56.34
18	%Steam	43.66
19	Fluidized Bed Density, lb/ft ³	36.62
20	Fluidized Bed Height, ft	3.61
21	Feed Composition, <u>993</u> lb/hr, %C/%Ash	72.25/11.11
22	Average Fluidized Bed Composition, <u>11</u> lb/hr, %C/%Ash	28.30/71.82
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	0.0/0.0
24	Venturi Discharge Composition, <u>98</u> lb/hr, %C/%Ash	11.73/88.92
25	2nd Stage Fines Composition, <u>62</u> lb/hr, %C/%Ash	78.00/19.92
26	3rd Stage Fines Composition, <u>3</u> lb/hr, %C/%Ash	57.10/41.16
27	Coal/Oxygen (Weight Basis)	1.32
28	Coal/Steam (Weight Basis)	0.64

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.0	17.9	28.6	24.6	17.9	7.9	2.3	0.5	0.4							
Fluidized Bed	0.0	2.2	15.8	34.9	33.3	7.5	4.4	1.5	0.4							
Discharged Bed																
Ash	0.0	2.3	12.9	40.6	37.8	3.4	-	-	3.0							
2nd Stage Fines					0.0	0.0	0.0			25.7	3.5	9.2	26.2	19.5	13.8	2.1
3rd Stage Fines					0.0	0.0	0.0			71.0	8.9	0.5	1.2	1.6	7.4	9.4

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Figure A-15a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 132

Run 132
2/4/79
0600-1600



RUN NO. 132

DATE 2/4 - 2/4/79

TIME PERIOD 0600-1200

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHERWISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	1113.5	158.7	23.2	48.8	79.8	145.2	1569.0	19932385.
TO VENTURI			0.0		50.5	545.5	596.1	620430.
TO GRID			0.0		69.1	531.3	600.4	553929.
TO CENTRAL JET			0.0		36.4	1047.6	1084.0	443360.
PURGES			393.6				393.6	-1091.
TOTAL INPUT	1113.5	158.7	421.8	48.8	235.9	2320.2	4293.3	21544073.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTURI DISCHARGE	10.9	95.1	0.2	0.9	0.1	0.0	106.0	153038.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	212.1	44.5	1.7	3.9	1.0	0.3	263.0	3261545.
3RD STAGE FINES	4.0	1.8	0.0	0.1	0.0	0.0	6.0	62506.
BED INV. CHANGE	-5.2	-0.3	0.1	0.3	-0.0	0.0	-5.0	-3587.
REACTOR GAS	785.2		427.6	10.6	234.3	2338.2	3796.5	15847145.
TOTAL OUTPUT	1007.1	141.1	427.6	15.3	235.9	2338.4	4165.9	19570646.
NET (OUT-IN)	-106.4	-17.6	5.9	-33.0	0.0	13.3	-132.9	-1973427.
%BALANCE OUT-IN	90.4	88.9	101.4	32.4	100.0	100.3	96.9	91.

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Figure A-15b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 132

AAG OPERATING CONDITIONS SUMMARY

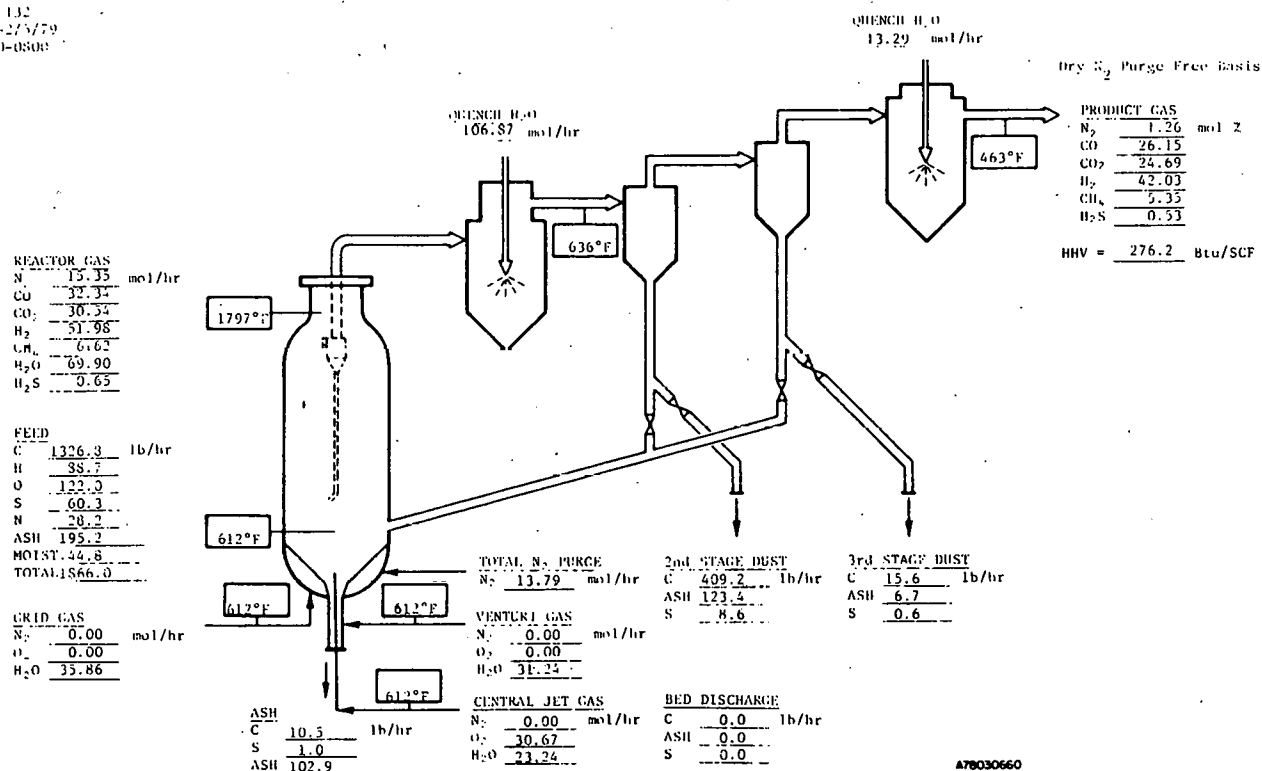
Run No. 132 Date 2/4/79 Time Period 0600 to 1600
 Venturi Diameter 4.5 inches Feed Material Washed W. Ky. #9 Coal
 Cyclone Fines Collector: 2, Internal 0, External 2

1	Average Fluidized Bed Temperature, °F	1870
2	Average Reactor Pressure, psig	13.1
3	Venturi Pressure Drop, psi	0.53
4	Superficial Velocity, ft/s	3.8
5	Freeboard Velocity, ft/s	3.9
6	Venturi Velocity, ft/s	42.1
7	Grid Gas, mol/hr	35.33
8	Grid Gas Composition: %N ₂	0.00
9	%O ₂	2.93
10	%Steam	97.07
11	Venturi Gas, mol/hr	29.58
12	Venturi Gas Composition: %N ₂	0.00
13	%O ₂	15.27
14	%Steam	84.73
15	Central Jet Gas, mol/hr	41.77
16	Central Jet Gas Composition: %N ₂	0.00
17	%O ₂	56.75
18	%Steam	43.25
19	Fluidized Bed Density, lb/ft ³	28.24
20	Fluidized Bed Height, ft	5.68
21	Feed Composition, <u>1569</u> lb/hr, %C/%Ash	73.05/10.41
22	Average Fluidized Bed Composition, <u>-5.0</u> lb/hr, %C/%Ash	42.15/56.87
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	0.0/0.0
24	Venturi Discharge Composition, <u>106</u> lb/hr, %C/%Ash	10.30/89.68
25	2nd Stage Fines Composition, <u>263</u> lb/hr, %C/%Ash	80.65/16.92
26	3rd Stage Fines Composition, <u>6</u> lb/hr, %C/%Ash	67.00/30.65
27	Coal/Oxygen Weight Basis)	1.68
28	Coal/Steam (Weight Basis)	1.12

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.0	19.0	32.6	26.1	15.5	5.1	1.1	0.3	0.3							
Fluidized Bed	0.0	3.0	23.7	46.0	13.8	8.1	4.0	0.9	0.5							
Discharged Bed																
Ash	0.0	10.1	37.2	49.9	1.8	0.3	-	-	0.7							
2nd Stage Fines					4.2	4.7	27.4			11.9	2.5	9.4	23.5	10.9	4.8	0.7
3rd Stage Fines					-	-	-			0.4	0.8	10.9	12.3	15.4	36.0	23.7

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Figure A-16a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 132

Run 132
2/4-2/5/79
2200-0300

RUN NO. 132

DATE 2/4-2/5/79

TIME PERIOD 2200-0300

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	1326.3	195.2	28.2	60.3	93.7	161.3	1866.0	23663060.
TO VENTURI			0.0		63.0	499.3	562.3	729493.
TO GRID			0.0		72.3	573.3	645.6	554654.
TO CENTRAL JET			0.0		46.9	1353.2	1400.1	536269.
PURGES			336.3				336.3	-1922.
TOTAL INPUT	1326.3	195.2	414.5	60.3	275.3	2533.6	4361.3	25431553.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTURI DISCHARGE	10.5	102.9	0.2	1.0	0.1	0.0	113.0	146266.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	409.2	123.4	4.0	8.6	2.0	0.0	546.0	6330392.
3RD STAGE FINES	15.6	6.7	0.1	0.6	0.1	0.0	23.0	246275.
BED INV. CHANGE	4.8	-5.3	0.0	0.0	-0.0	0.0	-1.0	515.
REACTOR GAS	334.0		434.3	21.1	273.7	2612.9	4176.0	17435090.
TOTAL OUTPUT	1274.1	227.2	434.3	31.2	275.3	2612.9	4355.5	24408538.
NET (OUT-IN)	-52.7	31.9	19.3	-29.1	0.0	24.3	-5.8	-1073015.
%BALANCE OUT-IN	96.0	116.4	104.6	51.8	100.0	100.9	99.9	96.

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Figure A-16b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 132

AAG OPERATING CONDITIONS SUMMARY

Run No. 132 Date 2/4-2/5/79 Time Period 2200 to 0800
 Venturi Diameter 4.5 inches Feed Material Washed W.Ky. #9 Coal
 Cyclone Fines Collector: 2, Internal 0, External 2

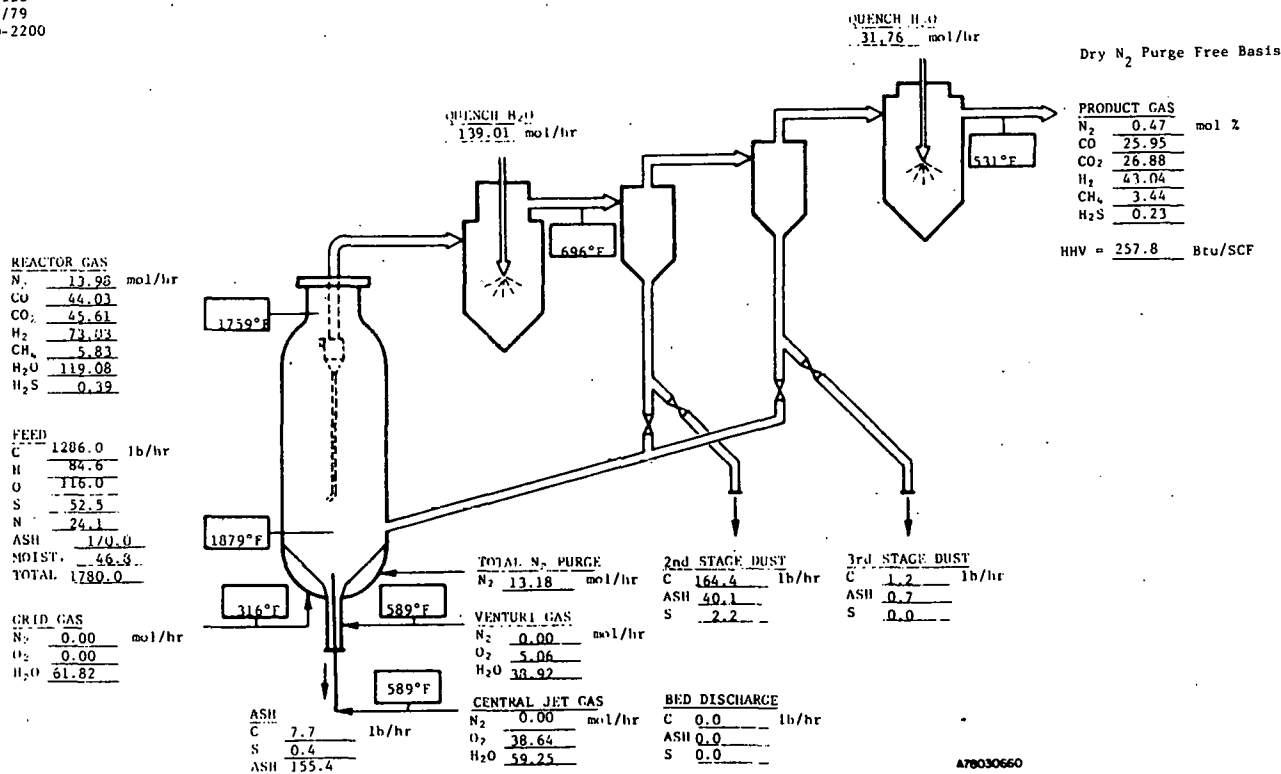
1	Average Fluidized Bed Temperature, °F	1856
2	Average Reactor Pressure, psig	16.6
3	Venturi Pressure Drop, psi	0.48
4	Superficial Velocity, ft/s	3.8
5	Freeboard Velocity, ft/s	3.9
6	Venturi Velocity, ft/s	34.4
7	Grid Gas, mol/hr	35.86
8	Grid Gas Composition: %N ₂	0.00
9	%O ₂	0.00
10	%Steam	100.00
11	Venturi Gas, mol/hr	31.24
12	Venturi Gas Composition: %N ₂	0.00
13	%O ₂	0.00
14	%Steam	100.00
15	Central Jet Gas, mol/hr	53.91
16	Central Jet Gas Composition: %N ₂	0.00
17	%O ₂	56.89
18	%Steam	43.11
19	Fluidized Bed Density, lb/ft ³	25.20
20	Fluidized Bed Height, ft	5.05
21	Feed Composition, <u>1866</u> lb/hr, %C/%Ash	72.85/10.72
22	Average Fluidized Bed Composition, <u>-1.0</u> lb/hr, %C/%Ash	51.55/47.30
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	0.0/0.0
24	Venturi Discharge Composition, <u>113</u> lb/hr, %C/%Ash	9.30/91.03
25	2nd Stage Fines Composition, <u>546</u> lb/hr, %C/%Ash	74.95/22.60
26	3rd Stage Fines Composition, <u>23</u> lb/hr, %C/%Ash	67.80/28.98
27	Coal/Oxygen (Weight Basis)	1.90
28	Coal/Steam (Weight Basis)	1.15

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8	40	20	10	5
Feed	0.0	20.1	30.6	23.8	16.1	6.3	1.9	0.6	0.6							
Fluidized Bed	0.0	4.4	26.3	40.3	17.7	7.9	2.3	0.8	0.3							
Discharged Bed																
Ash	0.0	7.3	24.7	61.3	6.0	0.4	-	-	0.3							
2nd Stage Fines					9.9	8.9	19.6			7.4	3.0	11.8	23.5	10.8	4.3	0.8
3rd Stage Fines					-	0.8	2.3			2.1	5.9	4.9	15.7	15.9	28.9	23.5

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Figure A-17a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 133

Run 133
2/24/79
1000-2200



RUN NO. 133

DATE 2/24-2/24/79

TIME PERIOD 1000-2200

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHERWISE NOTED

INPUT	C	ASH	N	S	O	TOTAL	BTU/HR
FEED	1286.0	170.0	24.1	52.5	39.3	157.5	1780.0
TO VENTURI			0.0		73.5	73.5	318835.
TO GRID			0.0		124.6	983.1	1034745.
TO CENTRAL JET			0.0		119.5	2184.4	1395177.
PURGES			369.2			369.2	2496.
TOTAL INPUT	1286.0	170.0	393.3	52.5	412.4	4115.7	6429.9
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTRI DISCHARGE	7.7	155.4	0.2	0.4	0.1	0.0	158900.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	164.4	40.1	1.5	2.2	0.7	0.2	209.0
3RD STAGE FINES	1.2	0.7	0.0	0.0	0.0	0.0	2.0
BED INV. CHANGE	7.4	-37.2	0.1	-0.0	0.0	0.0	-29.2
REACTOR GAS	1145.6		391.5	12.4	411.6	4069.0	6030.1
TOTAL OUTPUT	1326.3	159.1	393.3	15.0	412.4	4069.2	6375.1
NET (OUT-IN)	40.3	-11.0	0.0	-37.6	0.0	-46.5	-54.3
%BALANCE OUT/IN	103.1	93.6	100.0	28.5	100.0	98.9	99.1

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Figure A-17b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 133

AAG OPERATING CONDITIONS SUMMARY

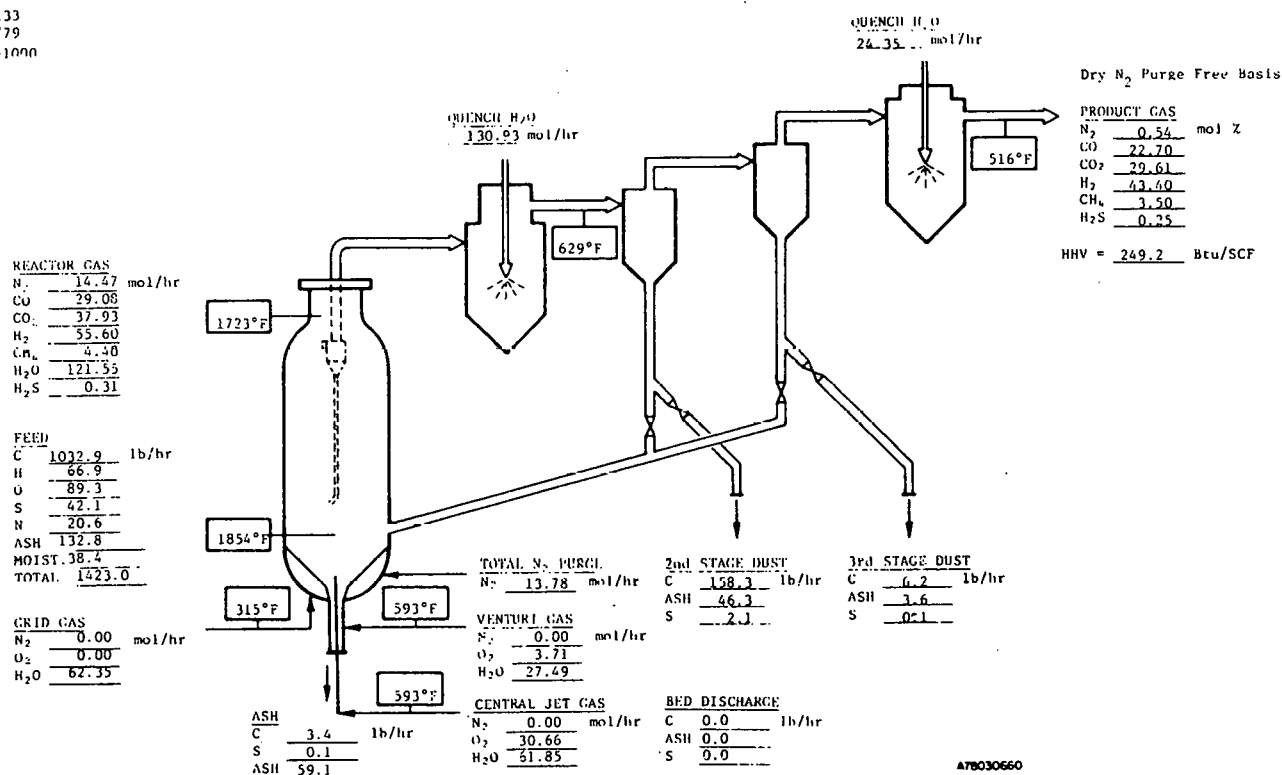
Run No. 133 Date 2/24/79 Time Period 1000 to 2200
 Venturi Diameter 4.5 inches Feed Material Washed W.Ky. #9 Coal
 Cyclone Fines Collector: 3, Internal 1, External 2

1	Average Fluidized Bed Temperature, °F	1879
2	Average Reactor Pressure, psig	41.6
3	Venturi Pressure Drop, psi	0.58
4	Superficial Velocity, ft/s	3.6
5	Freeboard Velocity, ft/s	3.0
6	Venturi Velocity, ft/s	29.4
7	Grid Gas, mol/hr	61.82
8	Grid Gas Composition: %N ₂	0.00
9	%O ₂	0.00
10	%Steam	100.00
11	Venturi Gas, mol/hr	43.98
12	Venturi Gas Composition: %N ₂	0.00
13	%O ₂	11.51
14	%Steam	88.49
15	Central Jet Gas, mol/hr	97.89
16	Central Jet Gas Composition: %N ₂	0.00
17	%O ₂	39.47
18	%Steam	60.53
19	Fluidized Bed Density, lb/ft ³	31.27
20	Fluidized Bed Height, ft	4.26
21	Feed Composition, <u>1780</u> lb/hr, %C/%Ash	74.20/9.81
22	Average Fluidized Bed Composition, <u>-29.2</u> lb/hr, %C/%Ash	36.23/63.60
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	0.00/0.00
24	Venturi Discharge Composition, <u>162</u> lb/hr, %C/%Ash	4.75/95.93
25	2nd Stage Fines Composition, <u>209</u> lb/hr, %C/%Ash	78.67/19.19
26	3rd Stage Fines Composition, <u>2</u> lb/hr, %C/%Ash	62.10/35.57
27	Coal/Oxygen (Weight Basis)	1.27
28	Coal/Steam (Weight Basis)	0.62

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.0	16.1	31.9	26.4	14.6	6.1	2.4	1.4	1.1							
Fluidized Bed	0.0	1.7	14.1	44.8	23.2	3.7	5.9	1.2	0.4							
Discharged Bed																
Ash	0.0	1.9	21.4	70.7	4.6	0.7			0.7							
2nd Stage Fines					0.0	0.6	13.5			15.5	5.5	13.8	28.8	14.9	6.8	0.6
3rd Stage Fines					1.1	0.0	0.3			1.0	3.4	9.5	15.0	14.9	32.9	21.9

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Figure A-18a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 133

Run 133
2/25/79
0200-1000

RUN NO. 133

DATE 2/25-2/26/79

TIME PERIOD 0200-1000

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	1032.9	132.8	20.6	42.1	71.3	123.4	1423.0	18345672.
TO VENTURI			0.0		55.4	558.6	614.0	650384.
TO GRID			0.0		125.7	997.6	1123.3	1066491.
TO CENTRAL JET			0.0		124.7	1970.6	2095.3	1460652.
PURGES			336.0				336.0	-319.
TOTAL INPUT	1032.9	132.8	406.6	42.1	377.0	3650.2	5641.6	21523080.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTURI DISCHARGE	3.4	59.1	0.1	0.1	0.2	0.0	62.0	53538.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	158.3	46.3	1.2	2.1	0.9	0.1	209.0	2437453.
3RD STAGE FINES	6.2	3.6	0.1	0.1	0.0	0.0	10.0	96978.
BED INV. CHANGE	-7.3	36.2	-0.0	-0.1	-0.0	0.0	28.4	11269.
REACTOR GAS	357.9		405.3	10.1	375.3	3623.3	5272.9	18667999.
TOTAL OUTPUT	1013.0	145.1	406.6	12.4	377.0	3623.9	5583.1	21522238.
NET (OUT-IN)	-14.9	12.3	0.0	-29.7	0.0	-26.3	-53.5	-342.
%BALANCE OUT/IN	98.6	109.3	100.0	29.3	100.0	99.3	99.0	100.

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Figure A-18b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 133

AAG OPERATING CONDITIONS SUMMARY

Run No. 133 Date 2/25/79 Time Period 0200 to 1000Venturi Diameter 4.5 inches Feed Material Washed W. Ky. #9 CoalCyclone Fines Collector: 3, Internal 1, External 2

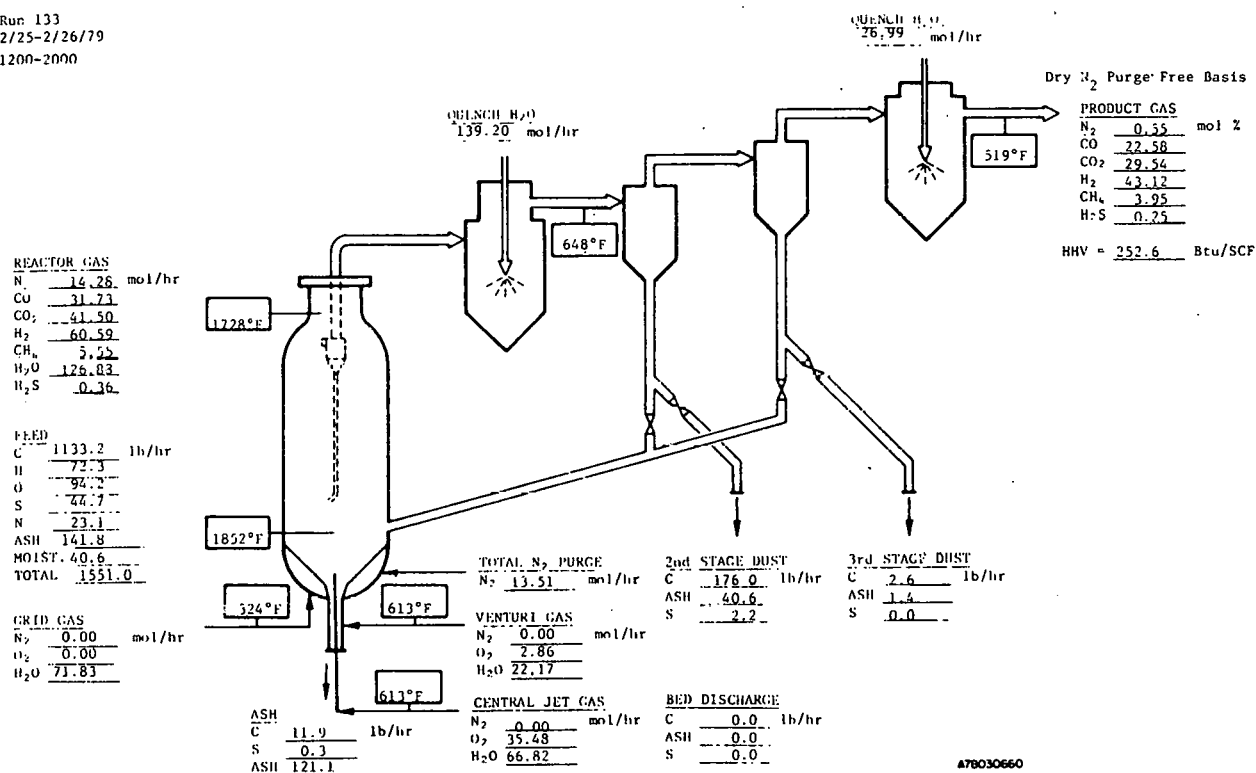
1	Average Fluidized Bed Temperature, °F	1854
2	Average Reactor Pressure, psig	41.3
3	Venturi Pressure Drop, psi	0.68
4	Superficial Velocity, ft/s	3.2
5	Freeboard Velocity, ft/s	2.6
6	Venturi Velocity, ft/s	17.3
7	Grid Gas, mol/hr	62.35
8	Grid Gas Composition: %N ₂	0.00
9	%O ₂	0.00
10	%Steam	100.00
11	Venturi Gas, mol/hr	31.20
12	Venturi Gas Composition: %N ₂	0.00
13	%O ₂	11.89
14	%Steam	83.11
15	Central Jet Gas, mol/hr	92.51
16	Central Jet Gas Composition: %N ₂	0.00
17	%O ₂	33.14
18	%Steam	66.86
19	Fluidized Bed Density, lb/ft ³	35.23
20	Fluidized Bed Height, ft	4.32
21	Feed Composition, <u>1423</u> lb/hr, %C/%Ash	74.60/9.59
22	Average Fluidized Bed Composition, <u>28.4</u> lb/hr, %C/%Ash	37.00/62.76
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash	0.00/0.00
24	Venturi Discharge Composition, <u>62</u> lb/hr, %C/%Ash	5.45/95.27
25	2nd Stage Fines Composition, <u>209</u> lb/hr, %C/%Ash	75.75/22.15
26	3rd Stage Fines Composition, <u>10</u> lb/hr, %C/%Ash	61.90/36.13
27	Coal/Oxygen (Weight Basis)	1.29
28	Coal/Steam (Weight Basis)	0.52

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.0	19.4	31.5	26.9	13.9	5.3	2.0	0.7	0.3							
Fluidized Bed	0.0	1.8	13.7	40.8	25.3	10.0	6.7	1.4	0.3							
Discharged Bed																
Ash	0.0	3.7	22.9	58.6	14.7	0.1	-	-	0.0							
2nd Stage Fines					0.0	0.6	10.1			13.4	4.2	12.1	30.4	14.6	11.1	3.5
3rd Stage Fines					0.0	1.3	6.9			8.3	8.2	8.2	22.4	15.4	18.6	10.7

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Figure A-19a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 133

Run 133
2/25-2/26/79
1200-2000



RUN NO. 133

DATE 2/25- 2/26/79

TIME PERIOD 1200-2000

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHERWISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	1133.2	141.8	23.1	44.7	77.8	130.3	1551.0	20050030.
TO VENTURI			0.0		44.7	446.2	490.9	523543.
TO GRID			0.0		144.3	1149.3	1294.1	316224.
TO CENTRAL JET			0.0		134.7	2204.4	2339.1	1592132.
PURGES			378.5				378.5	3183.
TOTAL INPUT	1133.2	141.8	401.6	44.7	402.0	3930.2	6053.5	23090167.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VENTRI DISCHARGE	11.9	121.1	0.1	0.3	0.1	0.0	132.0	204040.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	176.0	40.6	1.3	2.2	0.9	0.1	221.0	2687339.
3RD STAGE FINES	2.6	1.4	0.0	0.0	0.0	0.0	4.0	39877.
BED INV. CHANGE	-0.4	-5.9	0.0	-0.3	-0.0	0.0	-6.3	-2920.
REACTOR GAS	945.3		400.1	11.5	401.0	3365.5	5623.4	20399758.
TOTAL OUTPUT	1135.3	157.1	401.6	13.1	402.0	3365.6	5974.7	23578594.
NET (OUT-IN)	2.1	15.3	0.0	-31.6	-0.0	-64.6	-78.9	483426.
%BALANCE OUT/IN	100.2	110.3	100.0	29.3	100.0	93.4	93.7	102.

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Figure A-19b. U-GAS PIIOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 133

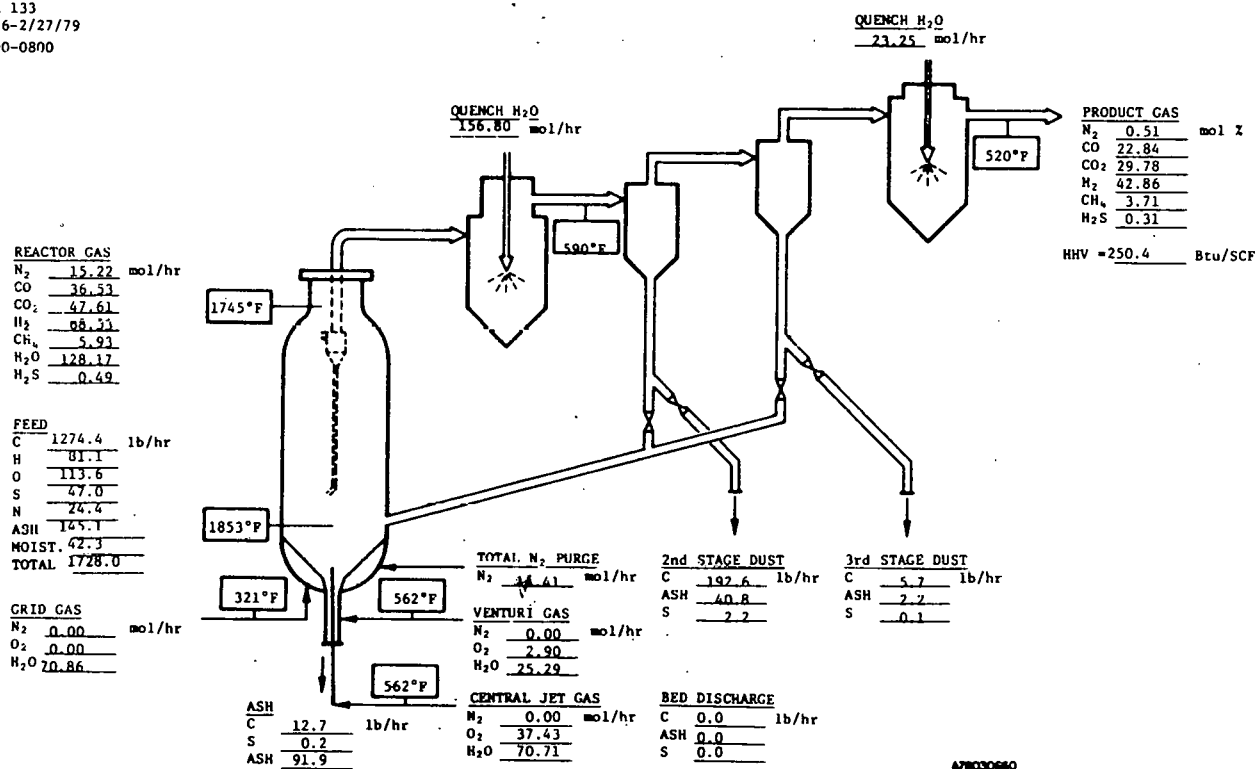
AAG OPERATING CONDITIONS SUMMARY																
Run No. <u>133</u> Date <u>2/25-2/26/79</u> Time Period <u>1200</u> to <u>2000</u>																
Venturi Diameter <u>4.5</u> inches Feed Material <u>Washed W.Ky.#9 Coal</u>																
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>																
1	Average Fluidized Bed Temperature, °F										1852					
2	Average Reactor Pressure, psig										42.8					
3	Venturi Pressure Drop, psi										0.65					
4	Superficial Velocity, ft/s										3.4					
5	Freeboard Velocity, ft/s										2.7					
6	Venturi Velocity, ft/s										15.4					
7	Grid Gas, mol/hr										71.83					
8	Grid Gas Composition: %N ₂										0.00					
9	%O ₂										0.00					
10	%Steam										100.00					
11	Venturi Gas, mol/hr										25.03					
12	Venturi Gas Composition: %N ₂										0.00					
13	%O ₂										11.42					
14	%Steam										88.58					
15	Central Jet Gas, mol/hr										102.30					
16	Central Jet Gas Composition: %N ₂										0.00					
17	%O ₂										34.68					
18	%Steam										65.32					
19	Fluidized Bed Density, lb/ft ³										33.20					
20	Fluidized Bed Height, ft										4.19					
21	Feed Composition, <u>1551</u> lb/hr, %C/%Ash										75.03/9.39					
22	Average Fluidized Bed Composition, <u>-6.3</u> lb/hr, %C/%Ash										39.20/60.49					
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash										0.0/0.0					
24	Venturi Discharge Composition, <u>132</u> lb/hr, %C/%Ash										9.0/91.72					
25	2nd Stage Fines Composition, <u>221</u> lb/hr, %C/%Ash										79.62/18.37					
26	3rd Stage Fines Composition, <u>4</u> lb/hr, %C/%Ash										64.06/33.80					
27	Coal/Oxygen (Weight Basis)										1.26					
28	Coal/Steam (Weight Basis)										0.54					

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.0	18.6	36.2	28.2	11.1	3.8	1.5	0.4	0.2							
Fluidized Bed	0.0	2.5	12.3	41.0	26.9	9.6	6.0	1.4	0.3							
Discharged Bed																
Ash	0.0	2.5	8.6	59.4	27.0	2.1	-	-	0.4							
2nd Stage Fines					0.0	1.4	16.8			16.4	4.6	14.4	27.9	12.5	5.3	0.7
3rd Stage Fines					0.0	0.0	2.8			4.6	1.1	7.3	22.9	18.6	27.4	15.3

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Figure A-20a. U-GAS PILOT PLANT MATERIAL AND ENERGY BALANCE SUMMARY FOR RUN 133

Run 133
2/26-2/27/79
2400-0800



RUN NO. 133

DATE 2/26- 2/27/79

TIME PERIOD 2400-0800

BASIS= 1 HR. ALL UNITS IN LBS UNLESS OTHER WISE NOTED

INPUT	C	ASH	N	S	H	O	TOTAL	BTU/HR
FEED	1274.4	145.1	24.4	47.0	35.3	151.2	1728.0	22569355.
TO VENTURI			0.0		51.0	437.4	548.4	589116.
TO GRID			0.0		142.9	1133.3	1276.6	939329.
TO CENTRAL JET			0.0		142.6	2329.1	2471.6	1651149.
PURGES			403.7				403.7	4383.
TOTAL INPUT	1274.4	145.1	428.1	47.0	422.2	4111.5	6428.4	25753333.
WALL LOSSES	0.0	0.0	0.0	0.0	0.0	0.0	0.0	250000.
VNTRI DISCHARGE	12.7	91.9	0.2	0.2	0.1	0.0	104.0	202721.
BED DISCHARGE	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.
2ND STAGE FINES	192.6	40.8	1.4	2.2	0.7	1.2	239.0	2942965.
3RD STAGE FINES	5.7	2.2	0.1	0.1	0.0	0.0	3.0	37223.
BED INV. CHANGE	4.9	-2.2	0.0	0.1	-0.0	0.0	3.2	2226.
REACTOR GAS	1030.3		426.4	16.4	421.4	4153.7	6103.8	22500966.
TOTAL OUTPUT	1296.8	132.7	428.1	19.0	422.2	4153.9	6453.7	25936102.
NET (OUT-IN)	22.4	-12.4	0.0	-23.0	-0.0	43.4	30.3	232769.
%BALANCE OUT/IN	101.8	91.4	100.0	40.4	100.0	101.2	100.5	101.

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Figure A-20b. U-GAS PILOT PLANT OPERATING CONDITIONS SUMMARY FOR RUN 133.

AAG OPERATING CONDITIONS SUMMARY																
Run No. <u>133</u> Date <u>2/26-2/27/79</u> Time Period <u>2400</u> to <u>0800</u>																
Venturi Diameter <u>4.5</u> inches Feed Material <u>Washed W.Ky. #9 Coal</u>																
Cyclone Fines Collector: <u>3</u> , Internal <u>1</u> , External <u>2</u>																
1	Average Fluidized Bed Temperature, °F										1853					
2	Average Reactor Pressure, psig										41.3					
3	Venturi Pressure Drop, psi										0.67					
4	Superficial Velocity, ft/s										3.6					
5	Freeboard Velocity, ft/s										3.0					
6	Venturi Velocity, ft/s										17.0					
7	Grid Gas, mol/hr										70.86					
8	Grid Gas Composition: %N ₂										0.00					
9	%O ₂										0.00					
10	%Steam										100.00					
11	Venturi Gas, mol/hr										28.19					
12	Venturi Gas Composition: %N ₂										0.00					
13	%O ₂										10.29					
14	%Steam										89.71					
15	Central Jet Gas, mol/hr										108.14					
16	Central Jet Gas Composition: %N ₂										0.00					
17	%O ₂										34.61					
18	%Steam										65.39					
19	Fluidized Bed Density, lb/ft ³										33.31					
20	Fluidized Bed Height, ft										4.12					
21	Feed Composition, <u>1728</u> lb/hr, %C/%Ash										75.60/8.61					
22	Average Fluidized Bed Composition, <u>3.2</u> lb/hr, %C/%Ash										47.20/52.13					
23	Bed Discharge Composition, <u>0</u> lb/hr, %C/%Ash										0.0/0.0					
24	Venturi Discharge Composition, <u>104</u> lb/hr, %C/%Ash										12.25/88.36					
25	2nd Stage Fines Composition, <u>239</u> lb/hr, %C/%Ash										80.60/17.08					
26	3rd Stage Fines Composition, <u>8</u> lb/hr, %C/%Ash										70.65/27.35					
27	Coal/Oxygen (Weight Basis)										1.34					
28	Coal/Steam (Weight Basis)										0.57					

U.S. Sieve Size, wt %	3	6	12	20	40	70	140	270	PAN	200	230	50.8 μ	40 μ	20 μ	10 μ	5 μ
Feed	0.0	10.0	27.3	36.6	13.4	7.8	3.7	0.9	0.3							
Fluidized Bed	0.0	12.8	27.9	33.8	12.4	7.2	4.5	1.0	0.4							
Discharged Bed																
Ash	0.0	3.9	27.0	60.5	7.9	0.3	-	-	0.4							
2nd Stage Fines					0.0	2.9	26.4			17.8	1.9	11.7	23.8	10.7	4.0	0.8
3rd Stage Fines					0.0	0.0	2.8			3.1	0.0	7.0	29.7	25.6	24.2	7.6

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APPENDIX B. Cold-Flow Model Tests

B-1

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LOW-PRESSURE COLD-FLOW MODEL TESTS

Introduction

The low-pressure cold-flow model test facility was used to conduct a research program on the behavior of the venturi discharge device and on other aspects of the ash-agglomerating gasifier. The overall objective was to determine the variables which affect the discharge of agglomerates from the gasifier in order that the latter can be properly designed to discharge the required amount of ash. At the inception of this work, virtually nothing was known about the behavior of the venturi device. In our research, we investigated several different venturi configurations of varying diameter, throat length, and discharge-pipe configuration. We investigated the role of agglomerate concentration, bed height (initial mass), agglomerate size and density (by the use of alumina spheres simulated agglomerates, and coal agglomerates), venturi velocity, and bed superficial velocity. We determined the influence of many of these variables on agglomerate discharge behavior.

The tests in all cases were performed in the equipment shown in its initial configuration in Figure B-1. Various changes were made from time to time, details of which are presented in the individual investigations which follow. The equipment consisted essentially of a semicircular 4-ft diameter fluidized bed with a Plexiglas® viewing face. Separately removable venturi, grid, and fluidized bed sections were provided. The venturi section and the related piping were the only sections which had to be modified in the course of the subsequent investigations. A centrifugal blower provided the source of fluidizing gas. A maximum discharge pressure of 7 psig could be attained, and this was adequate to overcome the pressure drop in the lines, orifices, control valves, and the equipment itself. Separate flows to venturi, grid, and subsequently central jet were provided. A feed hopper located above the vessel was provided and was fed by pneumatic means. Subsequent feeding from the hopper into the vessel was by gravity. Agglomerates could be fed along with bed material from the hopper or separately via a lock hopper. A cyclone was provided to collect fines fed or produced in the system, but no attempt at returning or recirculating these fines was made.

Although run procedures were improved and developed throughout these investigations, a typical procedure was as follows.

A charge of bed material powder was pneumatically fed to the feed hopper prior to the run. A high-velocity gas jet, sufficient to prevent any material from falling into the collecting hoppers, was set along with a sufficient grid flow to prevent weepage of

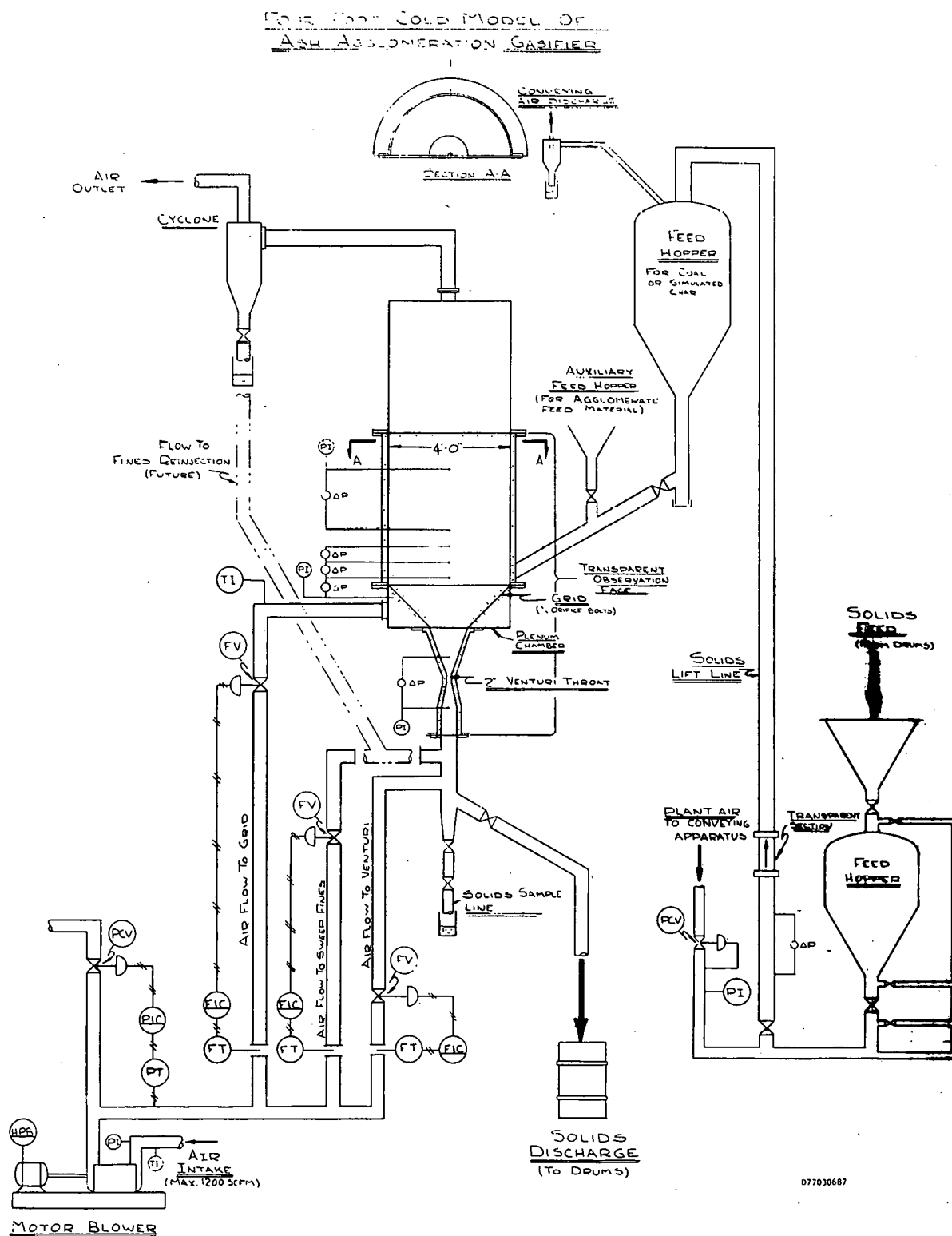


Figure B-1. FOUR-FOOT COLD MODEL OF ASH-AGGLOMERATING GASIFIER

material into the grid plenum chamber. The valve on the line from the feed hopper into the bed was then opened, the solids flowed by gravity into the vessel, and a fluidized bed was established. Agglomerates (alumina pellets) were then fed through a separate charge hopper until the desired amount was added to the bed. The gas flow rates were then set to the conditions desired for the run, and we then commenced collection of agglomerates falling through the venturi.

The first sample, collected when the velocities were being adjusted, was not included in the subsequent data analysis. Two to three minute intervals were employed for each sample collected, of which six to eight were collected in each run. The discharge samples collected were weighed, and agglomerates and char were separated by screening - the amounts of each being separately determined. The final bed material collected at the end of a run was also weighed and separated into bed material and agglomerates. This information was then used to calculate the average concentration of agglomerates $X_{b(avg)}$ prevailing in the bed during the time each of the samples was collected.

The materials used in tests 48-132 were acrylic molding powder bed material and alumina spheres. In tests 133-141 coke bed materials or final bed from runs 131 or 150 was used with coal agglomerates (-10+20 mesh) produced in various pilot-plant tests. Properties of the former materials are shown in Table B-1.

Table B-1. PROPERTIES OF MATERIALS USED IN COLD MODEL TESTS 48-132

	Acrylic (Plexiglas) Molding Powder	Alumina Spheres*
Mean Particle Diameter,	450	2520
Particle Density, lb/ft ³	72	157
Bulk Density, lb/ft ³	36	53

* Catalyst and Chemicals Inc. CS 222 catalyst support.

Although we enhanced our knowledge of venturi behavior immensely in the course of this work, a great deal still needs to be done. The value of this work is primarily qualitative in nature because of the semicircular design of the equipment. A full circular model capable of operation at up to 200 psig has been constructed and initially operated in a parallel effort to this work. (See Appendix B.) Results of these studies are expected to more accurately represent the conditions that will prevail in the demonstration plant. The work done here, however, has defined the scope of such work, enabling more efficient use of the equipment and manpower in that effort.

The balance of this report on cold model studies is divided into individual studies as itemized below.

Tests to Determine the Effects of Obstructions and Aeration Rate in a Jet Above the Throat of a 3 X 3 Inch Venturi

Summary

Seven tests were conducted on the half-section, 4-foot diameter ash-agglomerator cold-flow model. This work was accompanied by modification of the throat and expander cone of the 3 x 3 in. venturi to determine the effect of obstructions and aeration on the rate of discharge of spheres from the fluidized bed. The test results indicate that these modifications do not significantly affect the discharge rate of spheres from the bed.

Objective

Observations and results of previous test runs on the cold model established that there are various operating parameters influencing the rate of discharge of spheres from the bed. Important factors are venturi velocity, average sphere size, particle density of bed material and spheres, bed height, and mass fraction of spheres in the bed. The objective of the tests conducted was to determine the influence of some of these variables on venturi discharge rate when the venturi was aerated and when obstructions were implanted in the venturi. Additional effects of parameters related to these two practices were also examined.

Equipment Setup

Figure B-2 shows a view of the venturi section used in this test work. The obstruction consisted of one extendable 1/4-in. diameter copper tubing. Aeration was produced by nitrogen introduced through a rotameter located at a point just above the venturi throat.

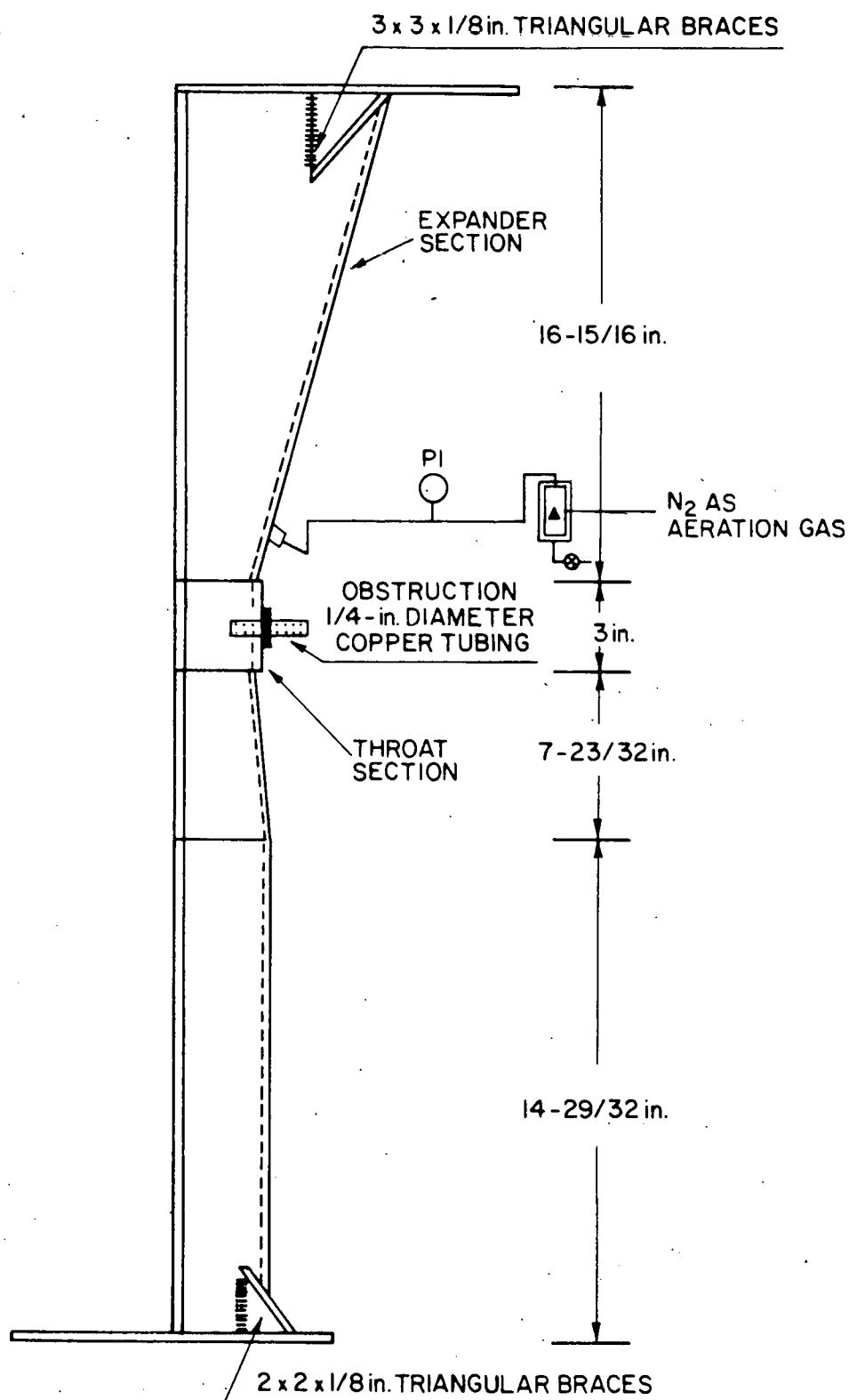
Operation and Results

Runs 48 and 49

Runs 48 and 49 were conducted as trials to determine the optimum locations for aeration and the obstruction to generate significant data.

Run 50

This run was conducted using the obstruction location shown in Figure B-2 and at varying venturi velocities and weight fractions of spheres in the bed. The results show extremely low discharge at the velocities used.



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Figure B-2. AAG COLD MODEL: VIEW OF THE
3 x 3 in. VENTURI THROAT

Run 51

This test was conducted using four different obstructional areas and six different venturi velocities. All other operating parameters were similar to those of Run 50. Figure B-3 shows that total discharge through the venturi varied with venturi velocities but the percentage of sphere yield fell sharply below a velocity of 32.1 ft/s. An obstruction at the venturi has little effect upon solids discharge rate. This figure also shows that net sphere yield is relatively constant at venturi velocities down to 30.6 ft/s.

Run 52

Run 52 demonstrated the effect of added aeration on venturi discharge rate. Figure B-4 shows that discharge rate was inhibited by aeration at lower venturi velocities. As usual, venturi discharge varied inversely with venturi velocity, with the yield of spheres diminishing rapidly at velocities below 32.1 ft/s.

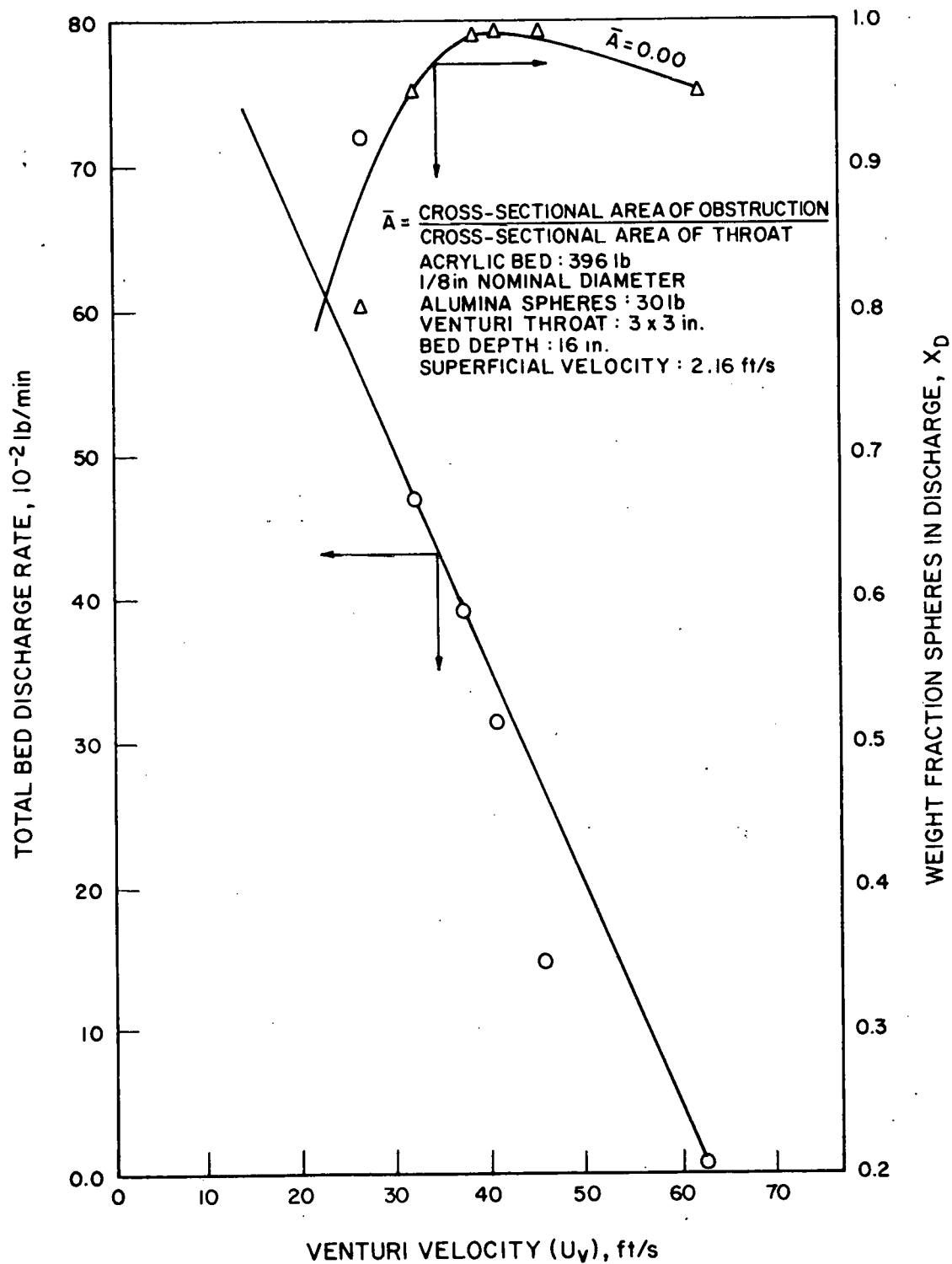
Runs 53 and 54

These two runs were conducted to determine the effect of superficial velocity on sphere discharge rate at a constant venturi velocity. Two venturi velocities were selected, both near the point at which sphere yield in the discharge decreases. (See Figure B-5.) Visually, the improved (greater) discharge rate appeared to be the result of more intensive solids mixing at the higher superficial velocities.

Conclusions

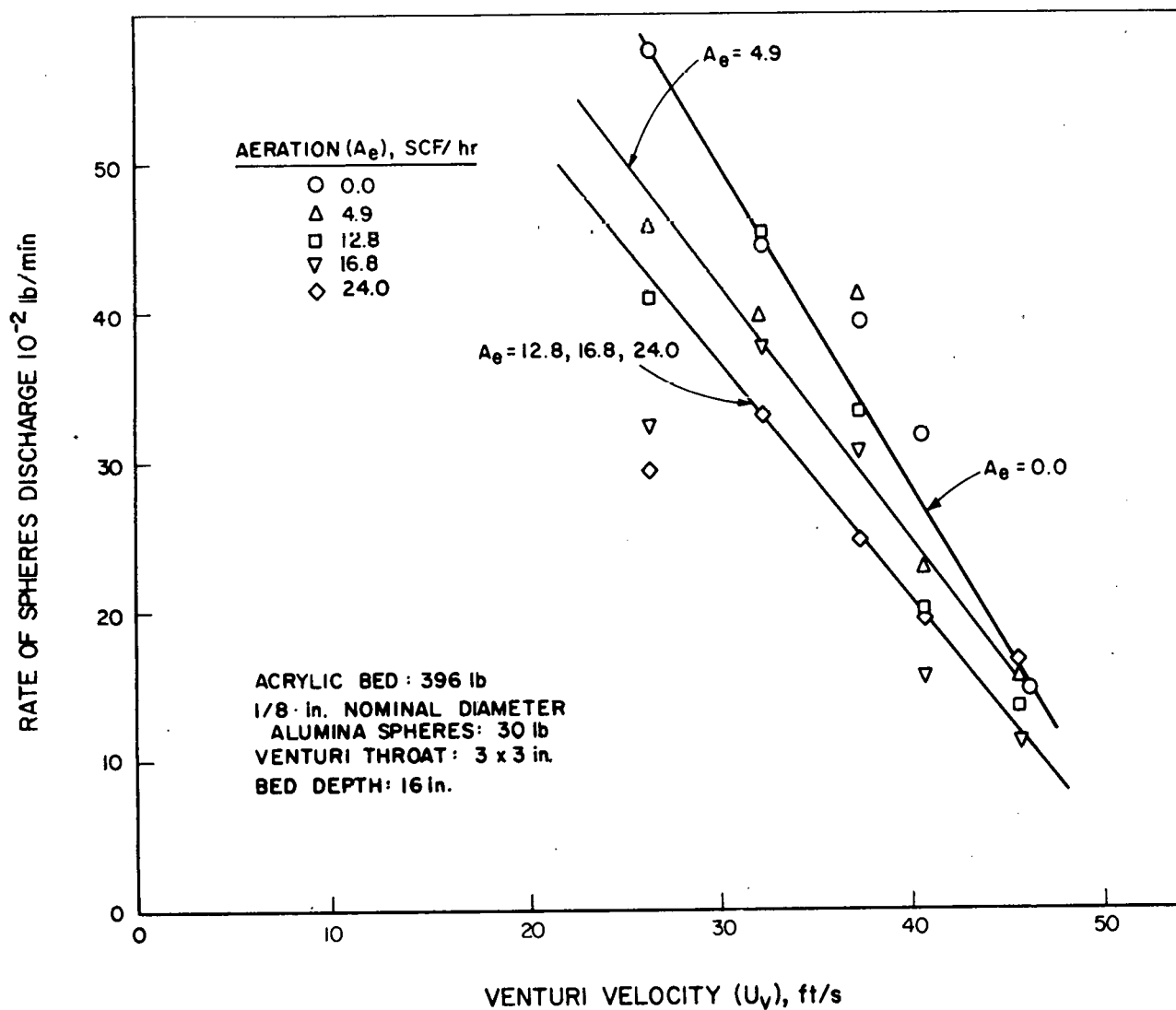
The conclusions of Runs 48 through 54 can be summarized as follows:

- The obstruction of up to 10% of the cross-sectional area of the venturi throat affected the discharge rate by a maximum of approximately 25%, but the effect diminished as venturi velocity was increased and disappeared at venturi velocities above 46.0 ft/s. (See Figure B-6.)
- There appears to be a relationship between venturi velocity and sphere concentration in the discharge and total discharge rate corresponding to other fixed operating parameters. As the venturi velocity is lowered, the discharge rate increases; however, this is accompanied by a lower concentration of spheres in the discharge. Increasing the velocity above the point at which the classification of spheres is virtually complete (approximately 95% spheres in the discharge) has a negligible effect on sphere concentration in the discharge. Lowering the venturi velocity below this "critical" velocity produces a sharp decrease in sphere concentration in the venturi. The effect of venturi velocity on the rate of sphere discharge is, however, linear over the range of venturi velocity. The effect of aeration was also slight, and diminished at aeration rates above 13 SCF/hr. In both cases discharge rates were diminished. Therefore, these types of modifications and experiments were not continued.



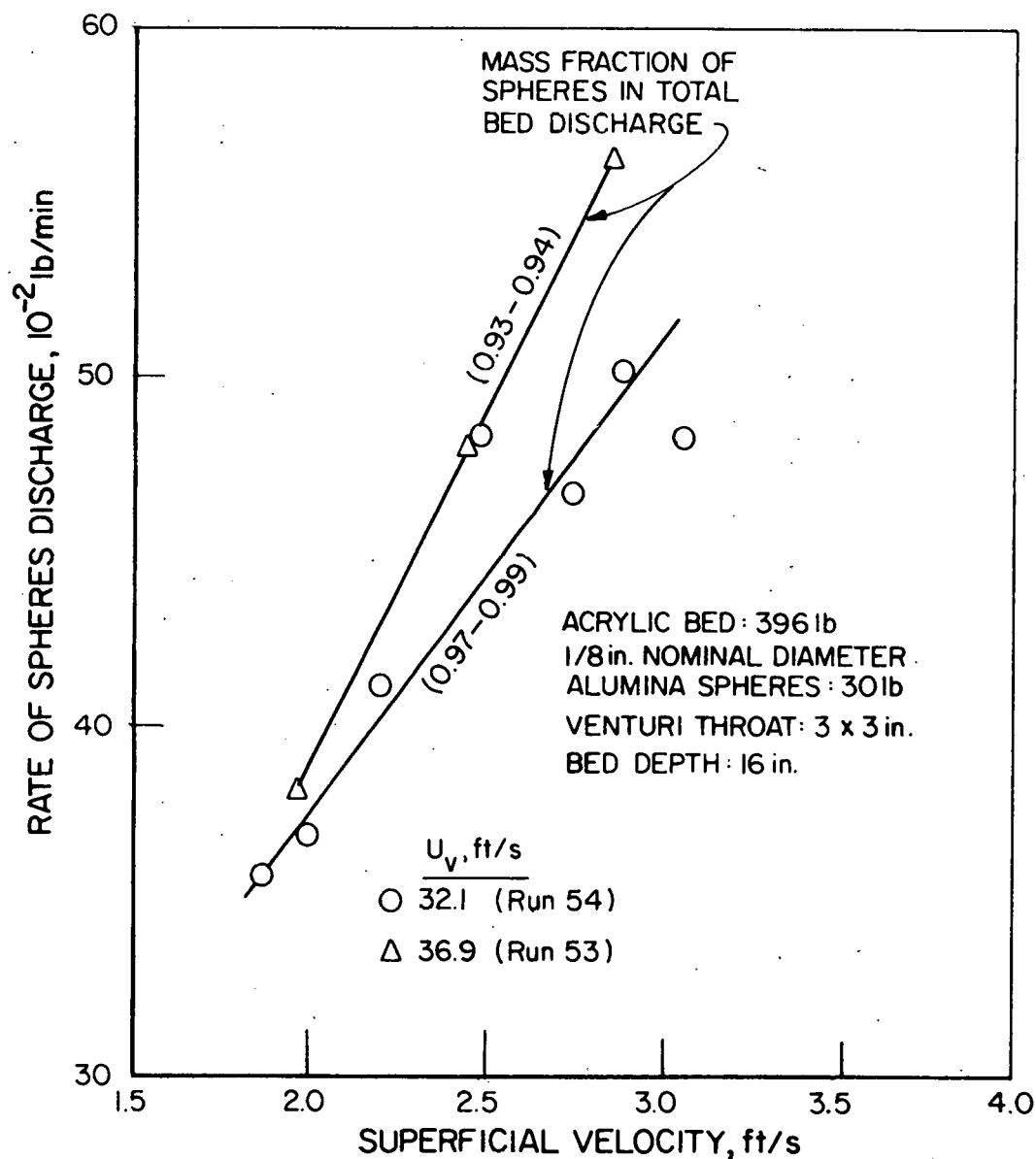
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Figure B-3. AAG COLD MODEL: VENTURI VELOCITY VERSUS RATE OF TOTAL BED DISCHARGE AND WEIGHT FRACTION OF SPHERES IN DISCHARGE (Run 51)



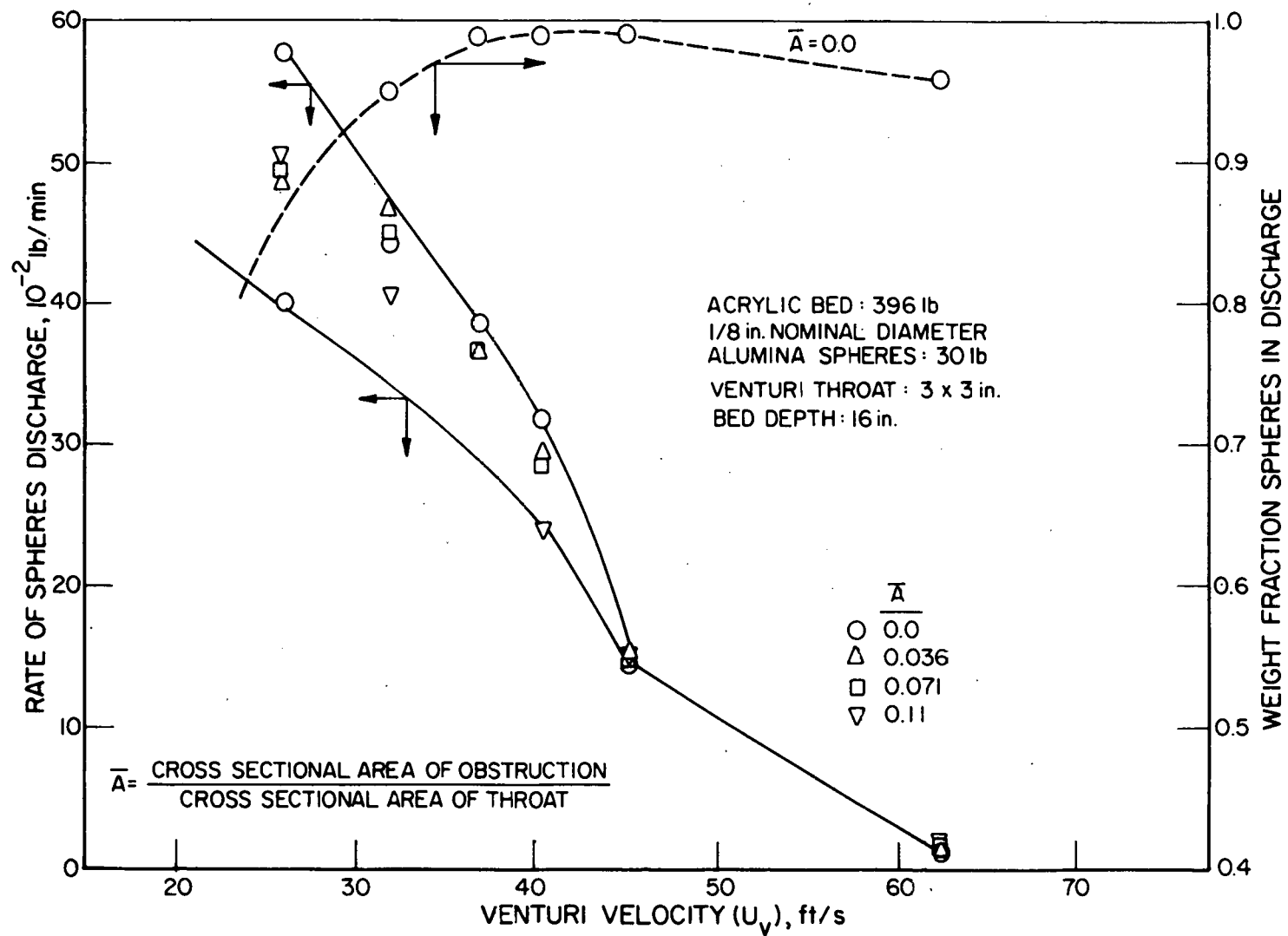
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Figure B-4. AAG COLD MODEL: VENTURI VELOCITY VERSUS
 RATE OF SPHERE DISCHARGE (Run 52)



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Figure B-5. AAG COLD MODEL: RATE OF SPHERE DISCHARGE
VERSUS SUPERFICIAL VELOCITY (Runs 53 and 54)



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Figure B-6. AAG COLD MODEL: VENTURI VELOCITY VERSUS RATE OF SPHERE DISCHARGE AND WEIGHT FRACTION OF SPHERES IN DISCHARGE (Run 51)

Tests on a 4-1/2 X 2 in. Venturia. Discharge Rate StudiesSummary

Five tests were conducted on the half-section 4-1/2 x 2 in. venturi throat. The effect of sphere mass concentration in the bed on the sphere discharge rate was examined. The effect of venturi velocity at a fixed sphere concentration at a superficial velocity of 2.16 ft/s and at two bed heights was determined, and a basis for comparison with future runs was selected (Run 57).

Objective

The main objectives of these test runs were --

- To determine the sphere discharge rate at varying mass fractions of spheres in the bed
- To determine the sphere discharge rate at varying venturi velocities
- To determine the sphere discharge rate at varying superficial velocities while keeping venturi velocities constant.

Equipment Setup and Modification

There was no major change in equipment setup. The 3 x 3 in. venturi throat was replaced by a half-section 4-1/2 x 2 in. throat. (See Figure B-7.) We also replaced the 18 x 3 in. Plexiglas bed discharge collecting tube by a 1-ft³ cylindrical conical steel vessel. This enabled us to operate the unit at much lower venturi velocities at which the discharge rate is higher and large amounts of material are collected in the sampling.

Operation and ResultsRun 55

This run demonstrated the effect of three different mass fractions (0.02, 0.07, and 0.09) in the bed on spheres discharge rate. As established earlier, the discharge rate increased linearly with an increased mass fraction of spheres in the bed. (See Figure B-8.)

Run 56

This test was conducted to determine the effect of varying venturi velocities on sphere discharge rate in relation to a fixed superficial velocity, bed height, and mass fraction of spheres in the bed. Results indicate that sphere discharge rate varied inversely with venturi velocity and directly with an increased mass fraction of spheres in the bed. (See Figure B-9.)

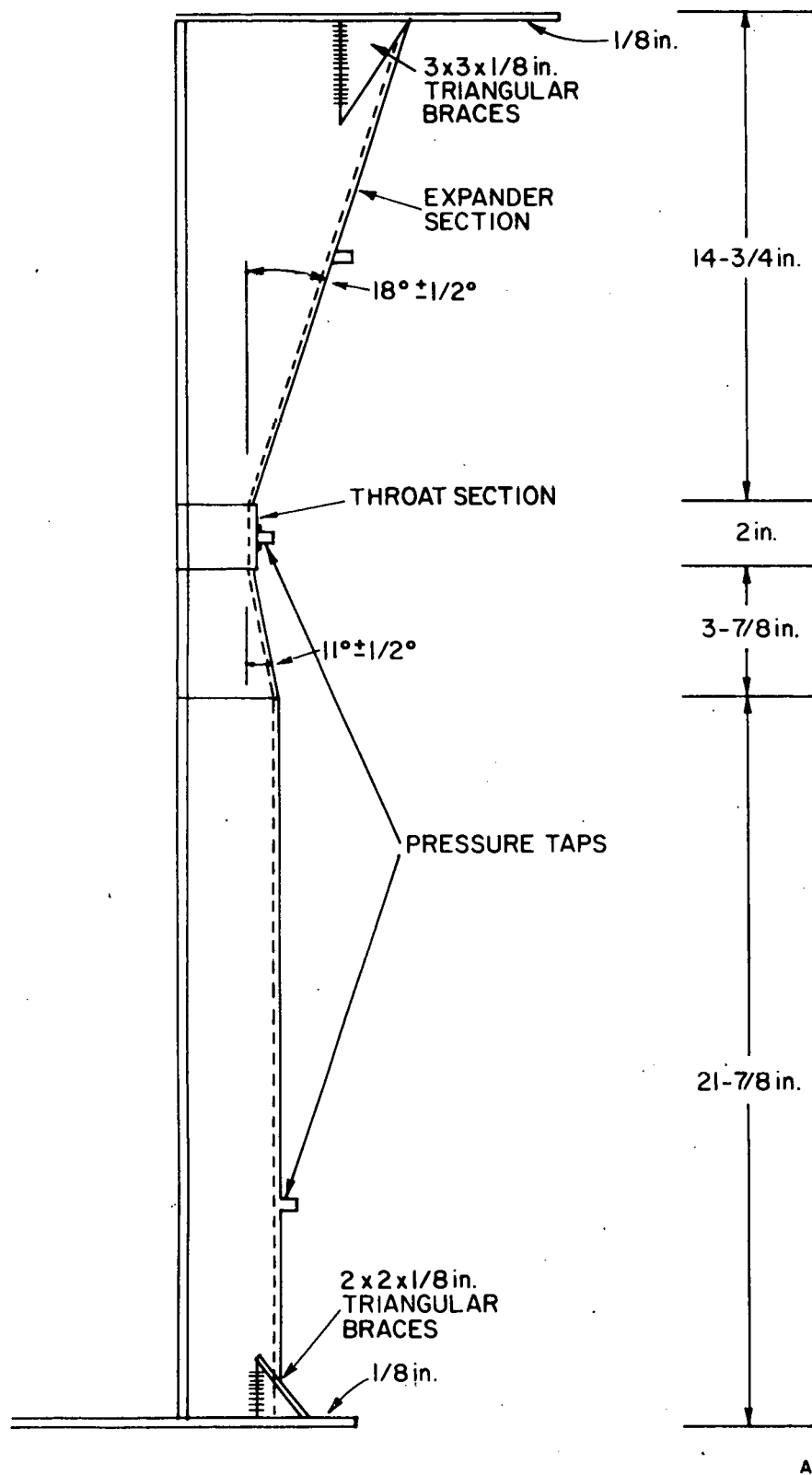
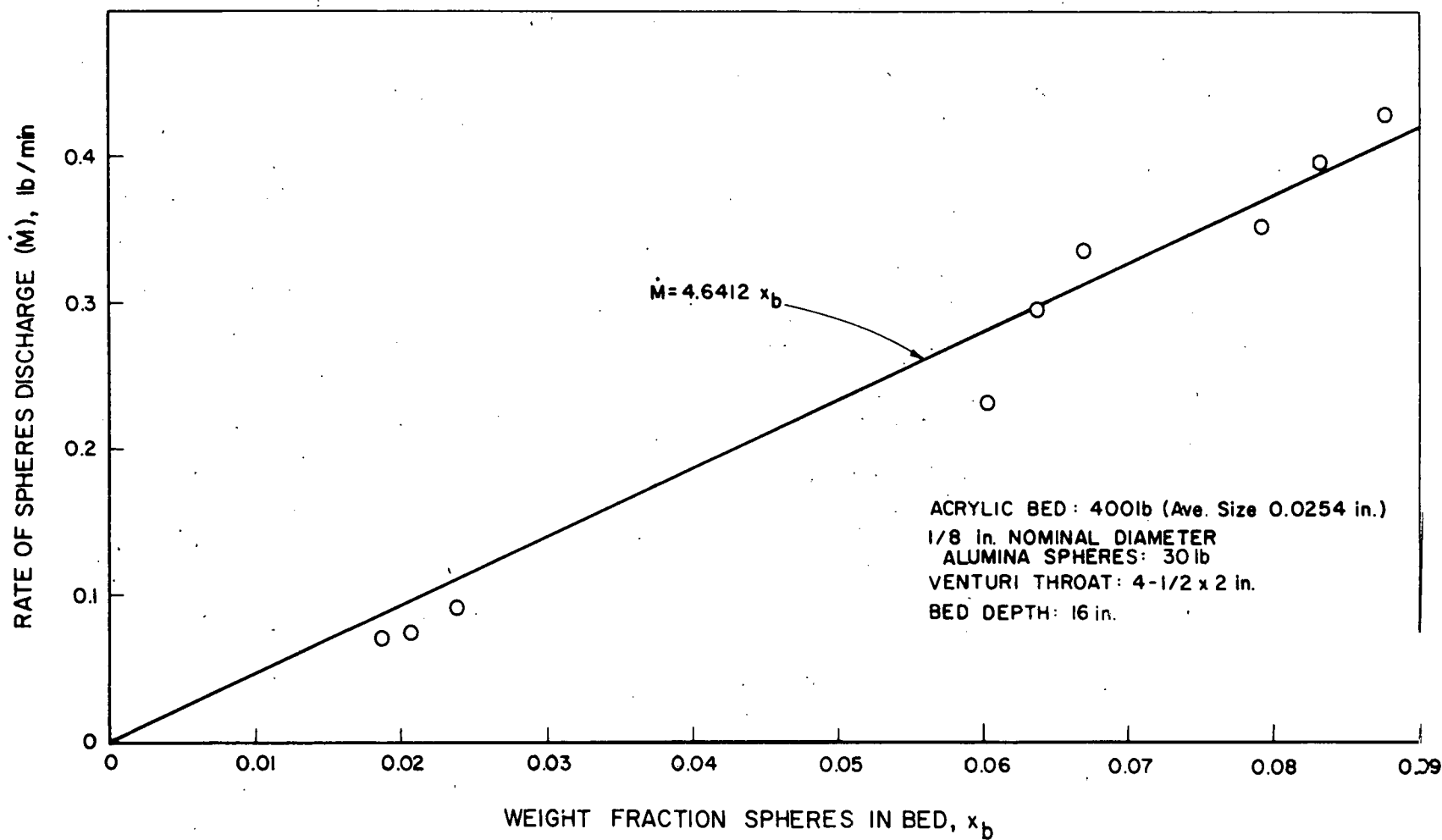


Figure B-7. AAG COLD MODEL: VIEW OF THE
4-1/2 x 2 in. VENTURI THROAT



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Figure B-8. AAG COLD MODEL: WEIGHT FRACTION OF SPHERES IN BED
 VERSUS RATE OF SPHERE DISCHARGE (Run 55)

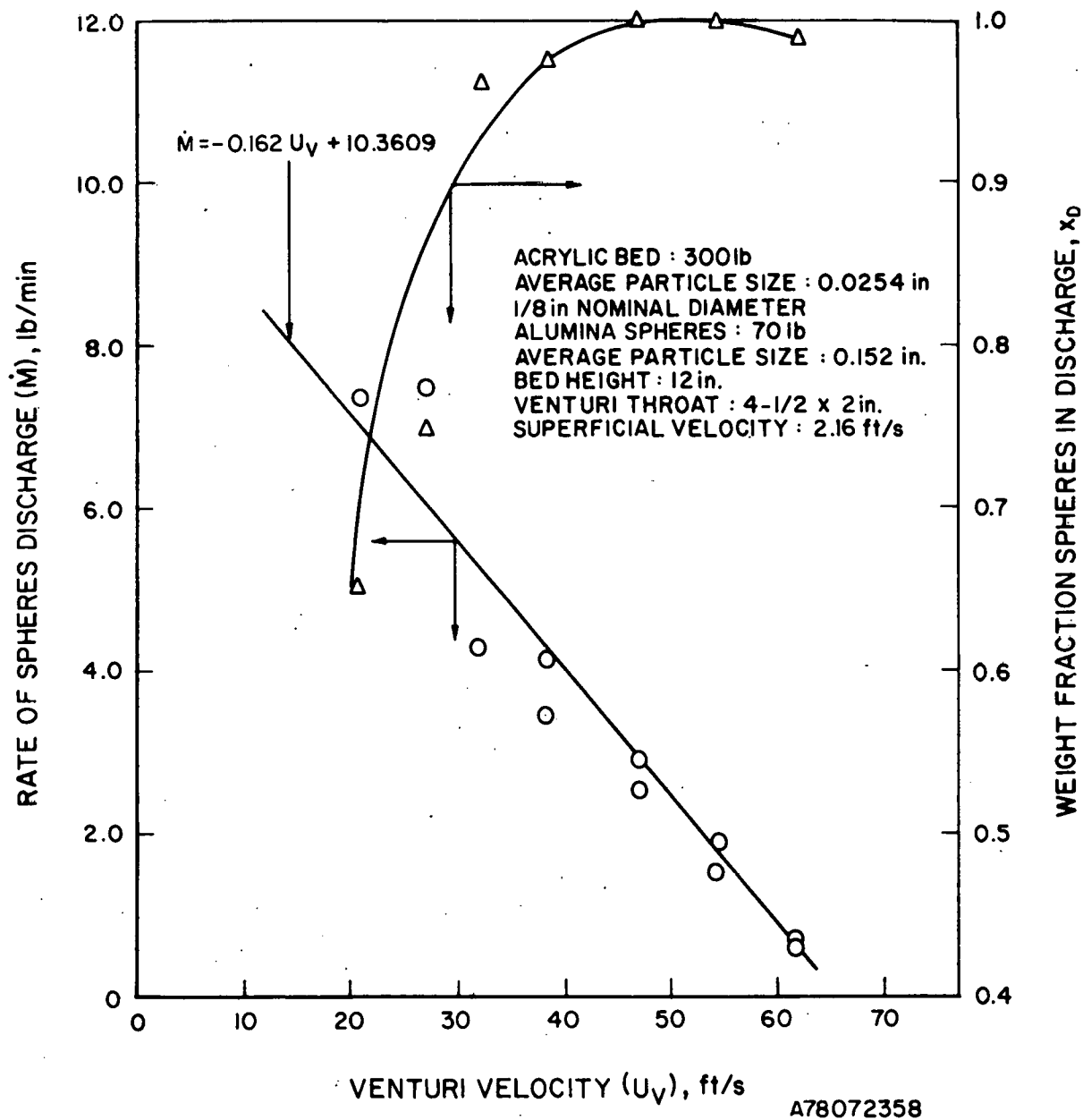


Figure B-9. AAG COLD MODEL: VENTURI VELOCITY VERSUS RATE OF SPHERE DISCHARGE AND WEIGHT FRACTION OF SPHERES IN DISCHARGE (Run 56)

Run 57

This run was a repetition of Run 56 except that the bed height was increased from 12 to 16 inches. This condition will be the basis of comparison for all our future runs. We also made a film showing the fluidization pattern in the expander at different operating venturi velocities.

Runs 58 and 59

These tests demonstrated the effect of varying superficial velocities on sphere discharge rate.

Conclusions

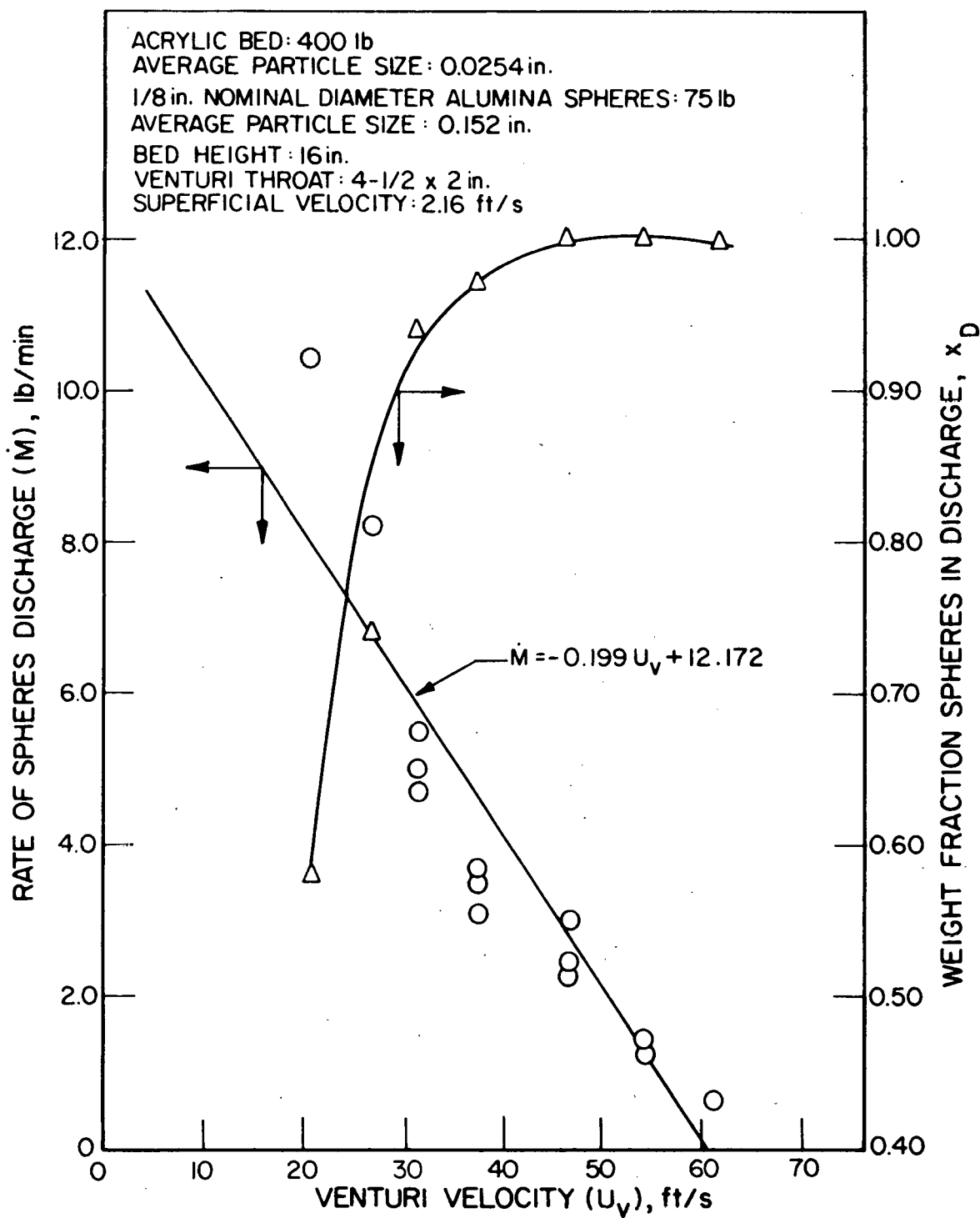
- The discharge rate of spheres was found to be linearly proportional to the mass concentration of spheres in the bed up to 9 weight percent mass concentration. The coefficient of proportionality between the discharge rate in lb/min and the concentration in lb spheres/lb total bed was 4.64 for the conditions of Run 55 (venturi velocity of 61.8 ft/s).
- The effects of venturi velocity on the mass concentration of spheres in the discharge and on mass discharge was determined and is shown for Runs 56 and 57 in Figures B-9 and B-10. The qualitative behavior is as described in the "Conclusions" section for the 3 x 3 in. venturi. Run 57 will form the basis of cross-venturi comparisons. The regression fit for the discharge rate (lb/min) versus venturi velocity (ft/s) for this run was found to be $\dot{M} = 12.17 - 0.199 U_v$.
- The results of Runs 56 and 57 indicate that bed height does have an effect on the sphere discharge rate. This effect appears very slight; however, a direct determination of the effect of bed height is not possible from these tests because the sphere concentrations in the bed differed by approximately 3%. This was investigated in more detail subsequently (see p.B-58) in the 6 X 2 in. venturi.

b. Center Jet StudiesSummary

A series of 13 tests (Runs 70 through 82) were conducted on the half-section cold-flow model with a venturi configuration of 4-1/2 X 2 inches with a center half-tube of 1 and 1-1/2 inch ID. The tube in these tests terminated at the top of the 4-1/2 inch throat, and 14 inches above the latter point. The inclusion of the tube at the center or gas flow through the tube produces moderate effects on the rate of sphere discharge.

Objective

The objective was to determine the effect of a center half-tube on discharge rate from the 4-1/2 X 2 inch configuration. We were concerned with the effect of locating the terminus of the tube at varying distances from the venturi throat and with varying flows in the annular venturi region and the tube.



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Figure B-10. AAG COLD MODEL: VENTURI VELOCITY VERSUS
 RATE OF SPHERE DISCHARGE AND WEIGHT FRACTION OF
 SPHERES IN DISCHARGE (Run 57)

Operation and Results

Runs 70-73

These test runs were conducted using 1-inch ID half-section tube (terminating at the top of the venturi throat). Both the venturi and tube velocities were varied. The rate of sphere discharge, as compared with the rate achieved without the tube (Run 57), was slightly changed.

Runs 74-75

These tests were conducted at the same conditions as those of Runs 70-73, except that the 1-inch ID half-section tube terminated 14 inches above the top of the venturi. No major change in the rate of sphere discharge, as compared with the rate achieved without the tube and with the tube (Runs 70-73), was observed.

Runs 76-77

These runs were conducted with a 1-1/2 in. ID half-section tube terminating at the top of the venturi throat.

Run 78

This test run was conducted by allowing the 1-1/2 inch ID half-section tube (terminating at the top of the venturi throat) to fill with spheres and with no gas flow in the tube while varying flows in the annular venturi region. No major change in the rate of sphere discharge as compared with the rate achieved without the tube (Run 57) and with the tube (Runs 76 and 77) was observed.

Run 79

All the test conditions were kept constant in this run, and discharge was collected every 3 minutes, thus enabling us to observe the effect of the mass fraction of spheres in the bed on the rate of sphere discharge as the run proceeded. The data for this run are shown in Table B-2 and sphere discharge rate versus initial sphere concentration in the bed is plotted in Figure B-11. The relationship is curvilinear, in contrast to previously observed results where a linear relationship was found. The data here extend to much higher concentrations than have been used in previous tests (~16% maximum); this may account for the difference between this and earlier observations.

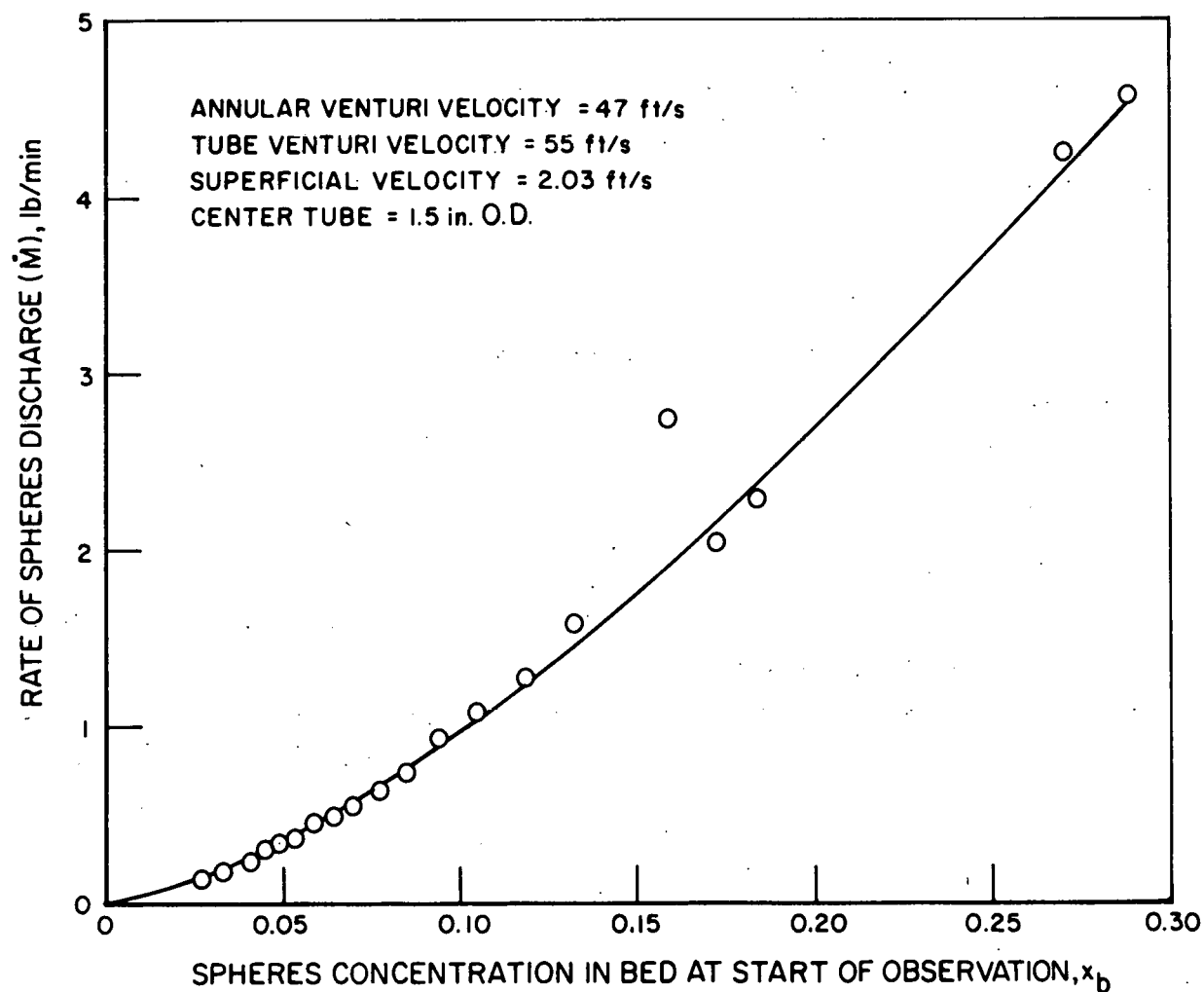
Table B-2. SUMMARY OF COLD-MODEL RUN CONDITIONS (Runs 70 through 82)

Run No.*	Venturi** Configuration	Tube ID	Venturi Velocity (U_v)	Tube Velocity (U_t)	Weight Fraction Spheres in Bed, x_b
	in.		ft/s		
70	4.5 x 2/0	1.0	31.2-67.1	243.0	0.080-0.097
71	4.5 x 2/0	1.0	34.7-67.4	99.0	0.085-0.099
72	4.5 x 2/0	1.0	43.4	60.7-225.0	0.110-0.130
73	4.5 x 2/0	1.0	26.9-69.2	74.4	0.117-0.154
74	4.5 x 2/14	1.0	31.7-69.5	75.7	0.116-0.149
75	4.5 x 2/14	1.0	43.7	62.5-227.0	0.112-0.131
76	4.5 x 2/0	1.5	0-75.3	60.5	0.119-0.158
77	4.5 x 2/0	1.5	47.2	56.0-105.0	0.126-0.143
78	4.5 x 2/0	1.5	0-75.45	0.00	0.099-0.154
79	4.5 x 2/0	1.5	47.00	55.16	0.022-0.289
80	4.5 x 2/14	1.5	47.5-47.8	61.2-121.6	0.138-0.151
81	4.5 x 2/14	1.5	17.04-76.72	61.3-61.8	0.119-0.153
82	4.5 x 2/14	1.5	18.77-75.58	0.00	0.119-0.157

* The following parameters were common to all tests:

Sphere Diameter = 0.152 in.
Average Particle Diameter = 0.0254 in.
Sphere Density = 157 lb/ft³
Bed Particle Density = 72.0 lb/ft³
Bed Bulk Density = 36.0 lb/ft³
Superficial Velocity = 2.0-2.3 ft/s

** Last number (after /) indicates height of center tube above top of venturi throat.



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Figure B-11. RATE OF SPHERE DISCHARGE VERSUS SPHERES CONCENTRATION IN BED AT START OF OBSERVATION (Run 79)

Run 80

In this test run, tube velocity was varied, and corresponding discharge of the bed material was noted at a constant venturi velocity. No major change in discharge rate was observed.

Run 81

This test was conducted to determine the effect of venturi velocity on the rate of total bed discharge and to compare the results with those of Run 78. No major difference was observed in the rate of total discharge using two different lengths of center tube.

Run 82

This run was the same as Run 78 except that the center tube was extended 14 inches above the throat. No major change in the rate of total discharge was observed.

Conclusions

The test results of Runs 70 through 82, and Run 57, confirmed that the location of a center tube of 1 or 1-1/2 inch ID in relation to the 4-1/2 inch venturi configuration has only a moderate effect on the discharge rate, as is true of the gas flow rate in the center tube. We were not able to correlate the tube location or tube flow rate with the observed deviations from the base case results (Run 57). The discharge rate, however, is dependent on venturi velocity and sphere concentration in the bed.

Tests on a 3 X 2 in. VenturiSummary

We conducted four runs on the half-section cold-flow model. These runs enabled a comparison of results from 3 X 2 venturis with those obtained previously from 4-1/2 X 2 in. venturis and 3 x 3 in. venturis. Additional information on the effect of sphere concentration on discharge rate was obtained.

Objective

The objective of Runs 61 through 64 was as follows:

- To determine discharge rates at varying venturi velocities on 3 X 2 in. venturis

Equipment Setup

There was no major change in equipment setup. We modified the 3 X 3 in. throat to 3 X 2 in. by removing a section 3-in. in diameter X 1-in. long, and we compensated for this loss by adding a section 3-in. in diameter X 1-in. long to the lower, straight section of the venturi.

Operation and Results

Test conditions for Runs 61 through 64 are shown in Table B-3. The standard 16-in. bed height and 3 X 2 in. units were used. The tests were run varying either venturi velocity (Run 62), superficial velocity (Runs 63 and 64), or sphere weight fraction (Run 61), holding other parameters at the indicated constant or nearly constant values. Discharge rates and the weight fractions of spheres in the discharge were measured in all the tests and added to the data file. The effect of venturi throat length on discharge rate is illustrated in Figure B-12. The solid lines are the discharge rate versus venturi velocity curves for 3 X 2 and 3 X 3 in. (Runs 48-54) venturis, but the concentrations of spheres for the two lines were different at 0.16 and 0.07, respectively. The effect of concentration would tend to exaggerate the difference between the lines. It was observed in Run 61 that the effect of concentration on M is one of direct proportionality, specifically -

$$M = 1.2 X_b \quad (B-1)$$

This latter correlation was developed for the condition of Run 61 ($U_v=62.1$ ft/s) and is shown along with 4-1/2 X 2 in. venturi data in Figure B-13. If the coefficient of proportionality (1.2) does not change with venturi velocity, then the 3 X 2 in. venturi data can be extrapolated to a sphere concentration of 0.07 by simple multiplication by the ratio of 0.07 to 0.16 by virtue of the above relationship. This is shown as the dashed line in Figure B-12. In any case one actual data point (62.1 ft/s) can be used. With this treatment of the data we see that the discharge rates for a 3 X 2 in. venturi are considerably higher than those for the 3 X 3 in. venturi. The effect of venturi throat

Table B-3. SUMMARY OF COLD-MODEL RUN CONDITIONS (Runs 61-64)

Run No.*	Venturi Configuration	Venturi Velocity (U_V) ft/s	Superficial Velocity (U_S), ft/s	Total Mass Bed (M), lb	Weight Fraction Spheres in Bed, X_b
61	3x2	62.1	2.09	400	0.024-0.091
62	3x2	32-67	2.11	400	0.13-0.16
63	3x2	38.4	1.8-3.6	400	0.14-0.15
64	3x2	61.2	1.9-3.7	400	0.15

* The following parameters were common to all tests:

Sphere Diameter = 0.152 in.
 Average Particle Diameter = 0.0254 in.
 Settled Bed Height (No Spheres) = 16 in.
 Sphere Density = 157 lb/ft³
 Bed Particle Density = 72.0 lb/ft³
 Bed Bulk Density = 36.0 lb/ft³

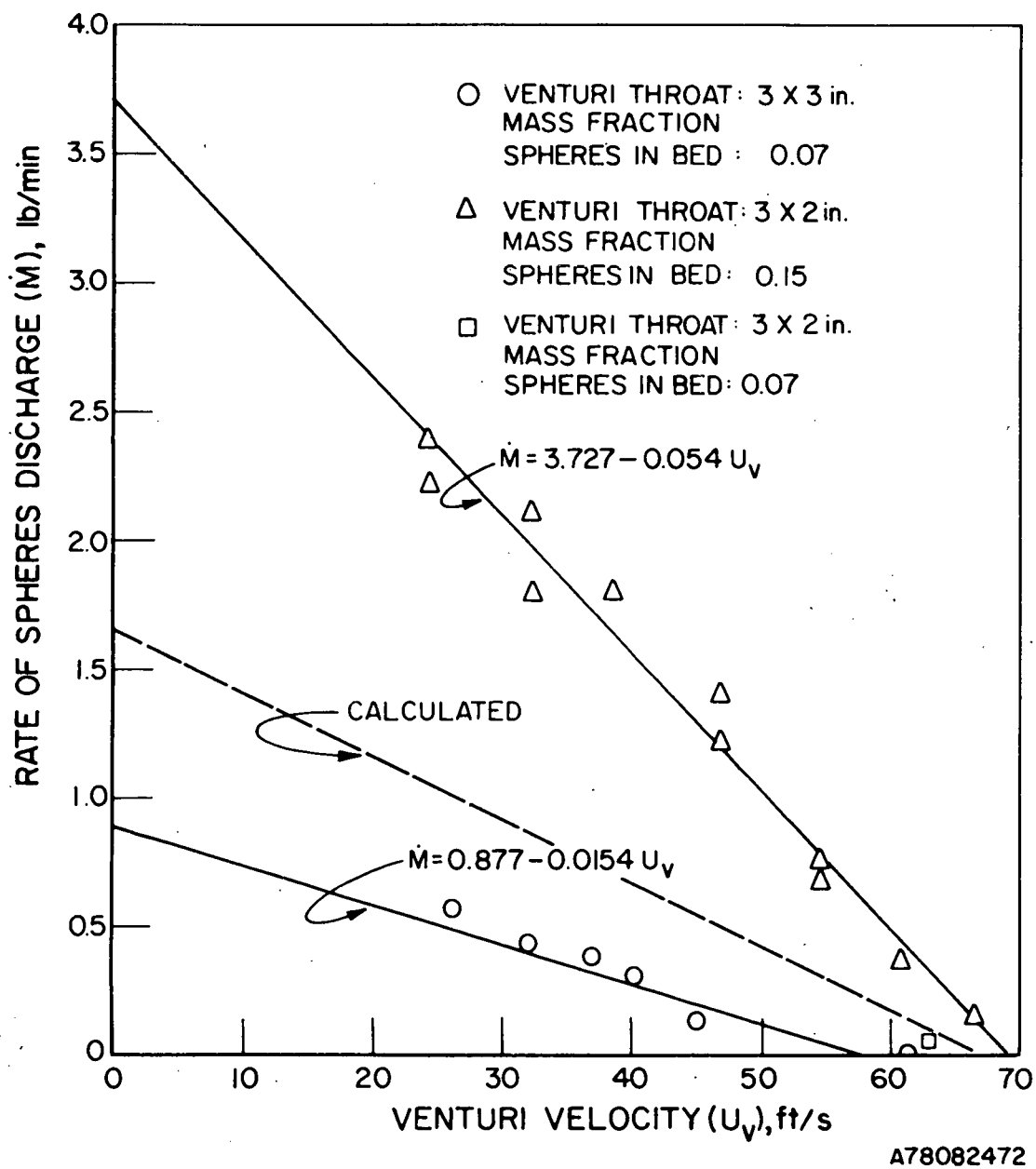


Figure B-12. VENTURI VELOCITY VERSUS RATE OF SPHERE DISCHARGE

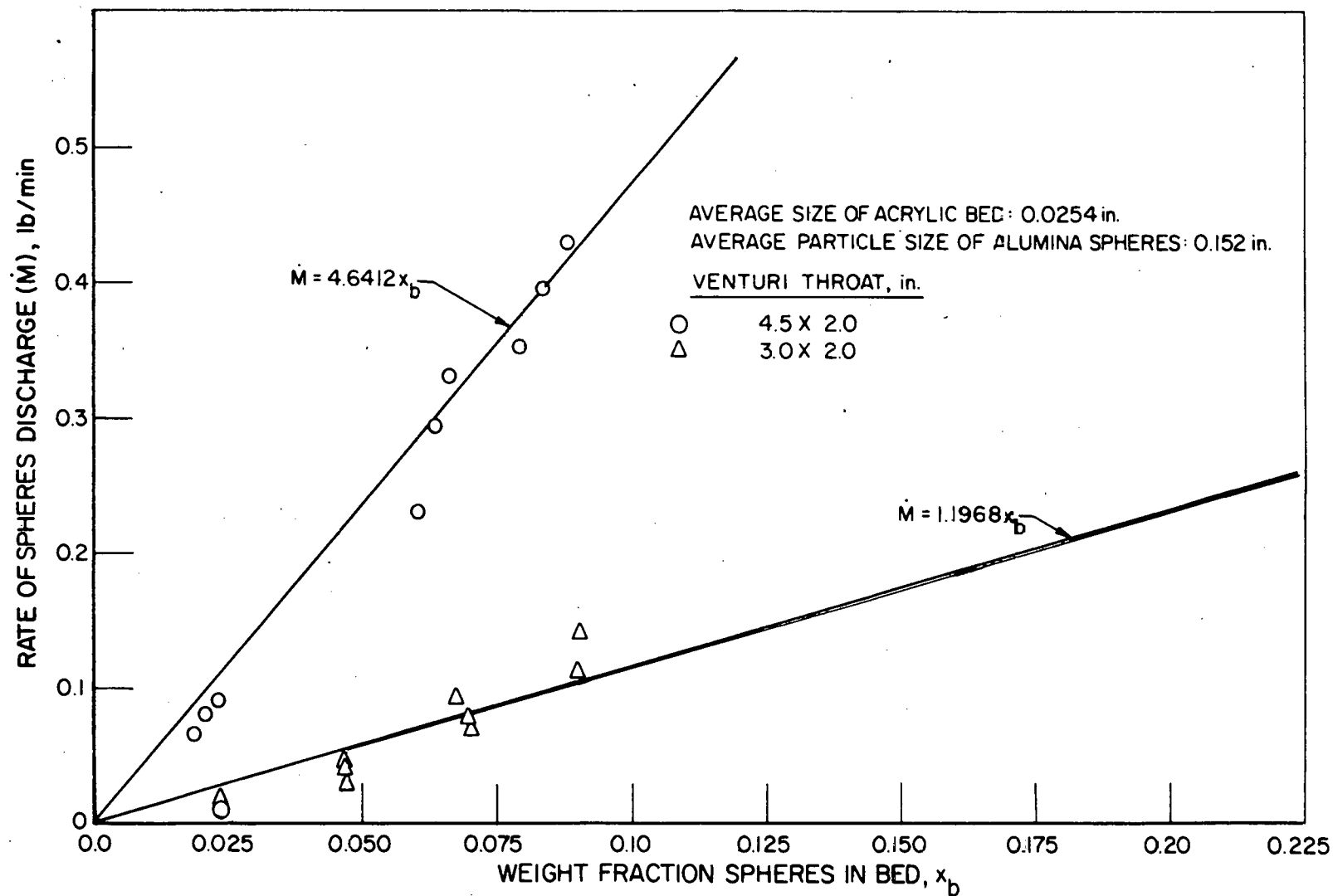


Figure B-13. WEIGHT FRACTION OF SPHERES IN THE BED VERSUS RATE OF SPHERE DISCHARGE FOR THE AAG COLD MODEL

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length seems to be to lessen discharge, all other conditions remaining the same. The effects of varying superficial velocity have been recorded in Runs 63 and 64.

Conclusions

Larger venturis with the same throat length yield larger discharge. The effect of venturi throat length seems to be such that at larger throat lengths the discharge rate is decreased.

Tests on a 2 X 2 in. Venturi

a. Discharge Rate Studies

Summary

We conducted four runs on the half-section cold-flow model. These runs enabled a comparison of results from 2 X 2 in. venturis with those obtained previously from 4-1/2 X 2 and 3 X 2 in. venturis. Additional information on the effect of sphere concentration on discharge rate was obtained. A condition of defluidization of the bed occurs at a high concentration of spheres or at low venturi velocities. We determined the maximum concentration of spheres to avoid defluidization above the 2 X 2 in. venturi at venturi velocities of 27 to 77 ft/s.

Objective

The objectives of Runs 65 through 68 were as follows:

- To determine discharge rates at varying venturi velocities on 2 X 2 in. venturis
- To determine the discharge rates at varying mass fractions of spheres in the bed on 2 X 2 in. venturis.

Equipment Setup

Runs on the 2 X 2 in. throat used the same venturi throat that had been used in Runs 1 through 30.

Operation and Results

Test conditions for Runs 65 through 68 are shown in Table B-4. The standard 16-in. bed height and 2 X 2 in. units were used. The tests were run varying either venturi velocity (Run 65), or sphere weight fraction (Runs 67 and 68), holding other parameters at the indicated constant or nearly constant values. Discharge rates and the weight fractions of spheres in the discharge were measured in all the tests and added to the data file.

Table B-4. SUMMARY OF COLD-MODEL RUN CONDITIONS (Runs 65-68)

<u>Run No.*</u>	<u>Venturi Configuration</u>	<u>Venturi Velocity (U_v), ft/s</u>	<u>Superficial Velocity (U_s), ft/s</u>	<u>Weight Fraction Spheres in Bed, X_b</u>
65	2x2	38-108	2.07	0.15-0.16
66	2x2	61.7	2.07	0.02-0.25
67	2x2	38.8	2.05	0.08-0.23
68	2x2	106	2.12	0.16-0.33

* The following parameters were common to all tests:

Sphere Diameter = 0.152 in.
 Average Particle Diameter = 0.0254 in.
 Settled Bed Height (No Spheres) = 16 in.
 Sphere Density = 157 lb/ft³
 Bed Particle Density = 72.0 lb/ft³
 Bed Bulk Density = 36.0 lb/ft³
 Initial Mass of Bed = 400 lb

Figure B-14 compares the venturi velocity versus rate of sphere discharge in the 2 X 2 in. venturi (Runs 65-68) versus those of 3 X 2 in. (Runs 61-64) and 4-1/2 X 2 in. (Runs 55-59) venturis. The sphere discharge rate for a 2 X 2 in. venturi is always less than the larger diameter venturis at ostensibly identical conditions.

Conclusions

Discharge rates for a 2 X 2 in. venturi are less than those of a 3 X 2 in. and 4-1/2 in. venturi. As found in previous studies, here also the discharge rates are linearly proportional to the venturi velocity. Also, a smaller venturi with the same throat length yields smaller discharge.

b. Defluidization Studies

Summary

Runs 65-68, used in the discharge rate study, were used to study defluidization of the bed. These runs used a 2 X 2 in. venturi. A condition of defluidization of the bed occurs at a high concentration of spheres or at low venturi velocities. We determined the maximum concentration of spheres to avoid defluidization above the 2 X 2 in. venturi at venturi velocities of 27 to 77 ft/s.

Objective

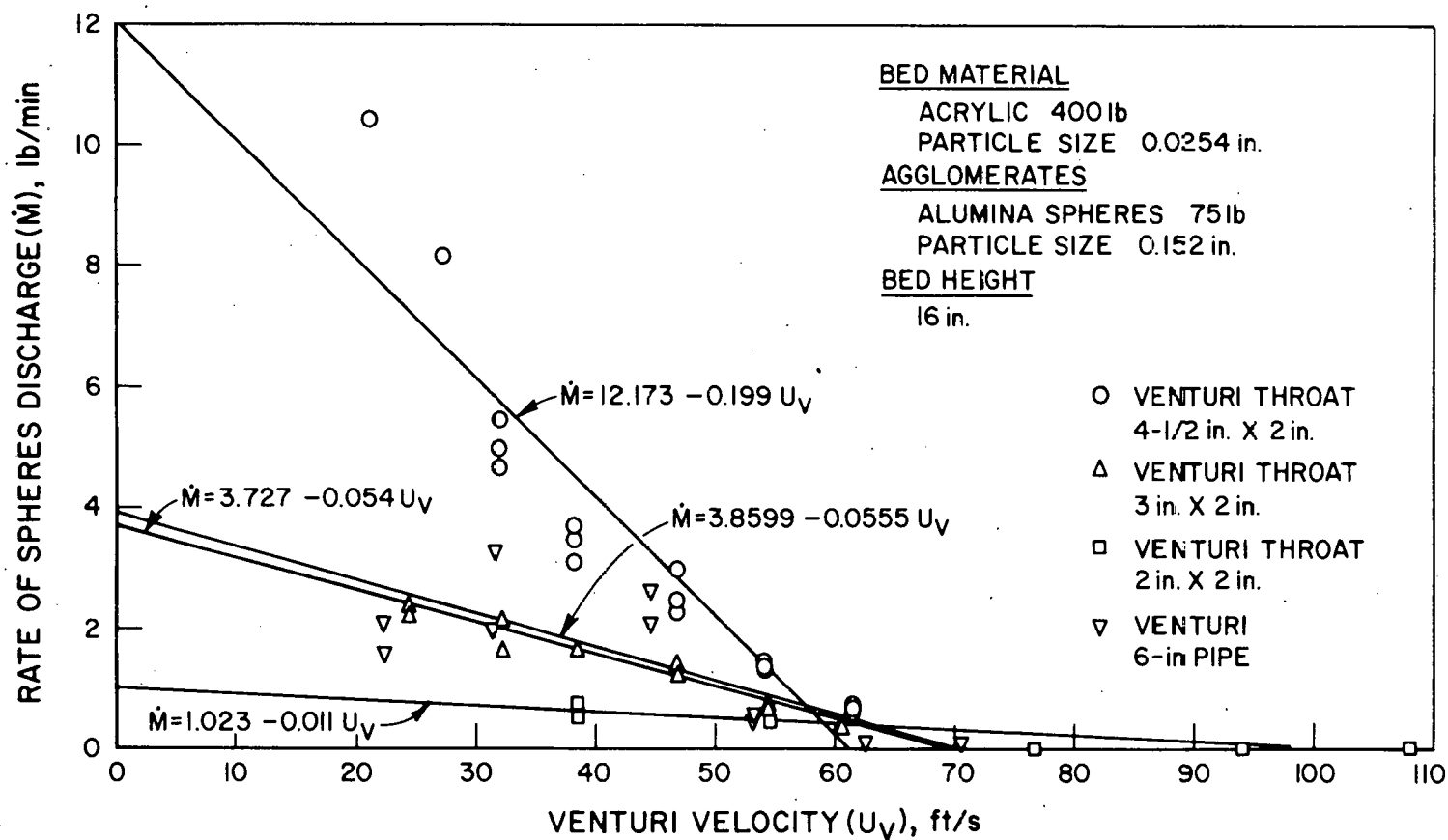
The objective of Runs 65 through 68 was, in addition to discharge rate study, to determine the defluidization concentration at varying venturi velocities in a 2 X 2 in. venturi.

Equipment Setup

Same as equipment setup in discharge rate study.

Operation and Results

In Runs 65-68, we observed defluidization of the bed above the venturi and noted the venturi velocity and sphere concentration at which this occurred (for the 2 X 2 in. venturi). In this type of defluidization spheres segregate out of the bed and bridge over the venturi. Eventually all of the spheres in the bed segregate out above the bridged section. It is possible to maintain fluidization of the bed while this is occurring. The defluidization results obtained in several runs are shown in Figure B-15. A linear relation of venturi velocity and defluidization sphere concentration is apparent. The regression fit of the data gives the linear equation -



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Figure B-14. AAG COLD MODEL: VENTURI VELOCITY VERSUS RATE OF SPHERE DISCHARGE

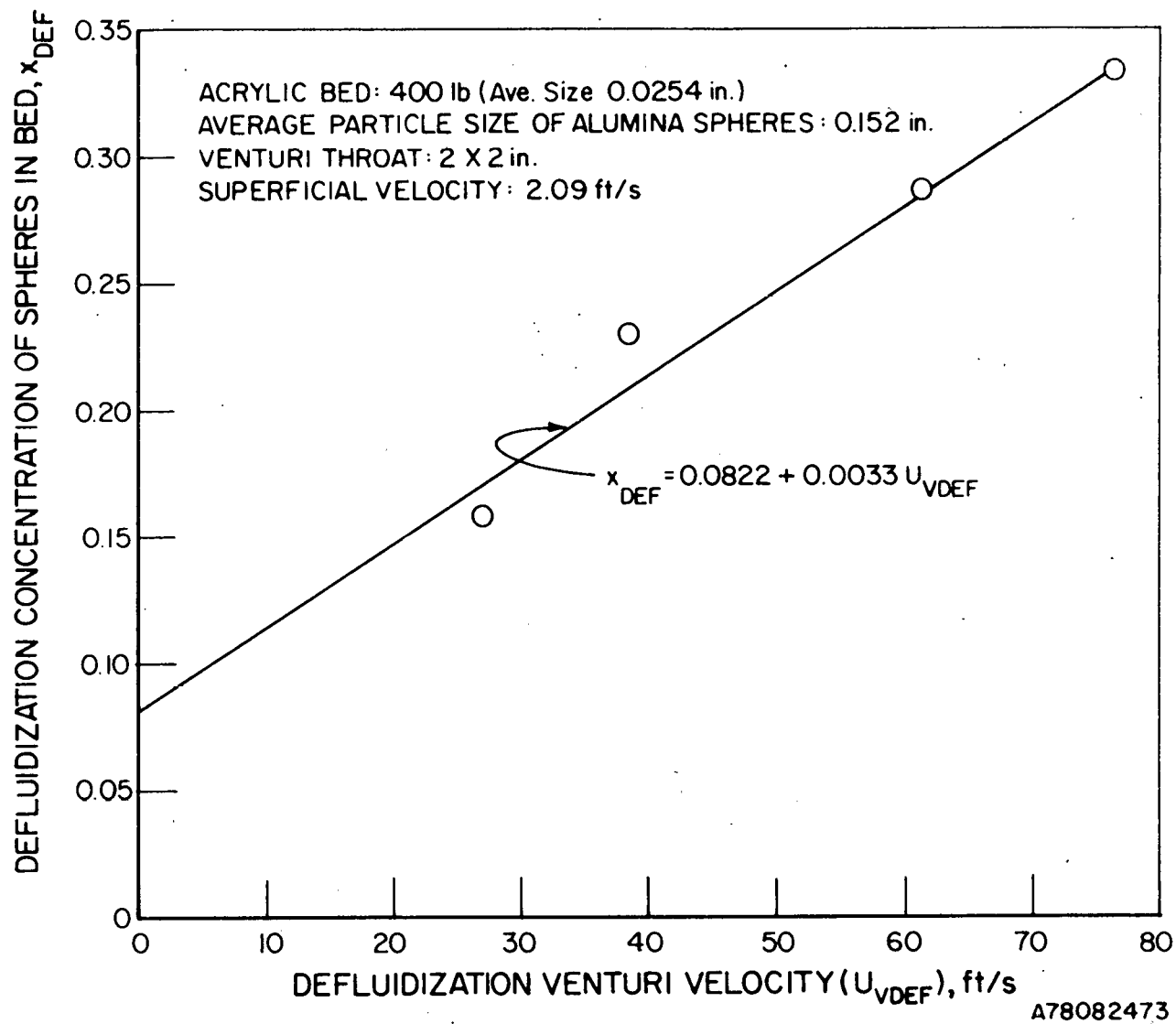


Figure B-15. DEFLUIDIZATION CONCENTRATION VERSUS VENTURI VELOCITY
 FOR THE AAG COLD MODEL

$$x_{DEF} = 0.082 + 0.0033U_v \quad (B-2)$$

where -

x_{DEF} = sphere concentration at defluidization, (lb spheres)/(lb spheres + lb bed material)

U_v = venturi velocity at defluidization, ft/s

There is a definite minimum velocity that must be exceeded at high sphere concentrations to avoid defluidization of spheres above the venturi. The relationship of defluidization velocity to sphere concentration appears to be linear.

Tests With a 6 X 30 in. Configuration

Summary

Run 69 was performed using a 6 X 30 in. configuration for the lower section of the cold model. A rate of sphere discharge lower than that of the 4-1/2 X 2 in. venturi and greater than that of the 2 X 2 or 3 X 3 in. venturi was obtained. The discharge rate in the 3 X 2 in. venturi was approximately equivalent to that obtained in the present configuration.

Objective

The objective was to perform the standard tests on the 6 X 30 in. configuration and to compare these results with earlier data.

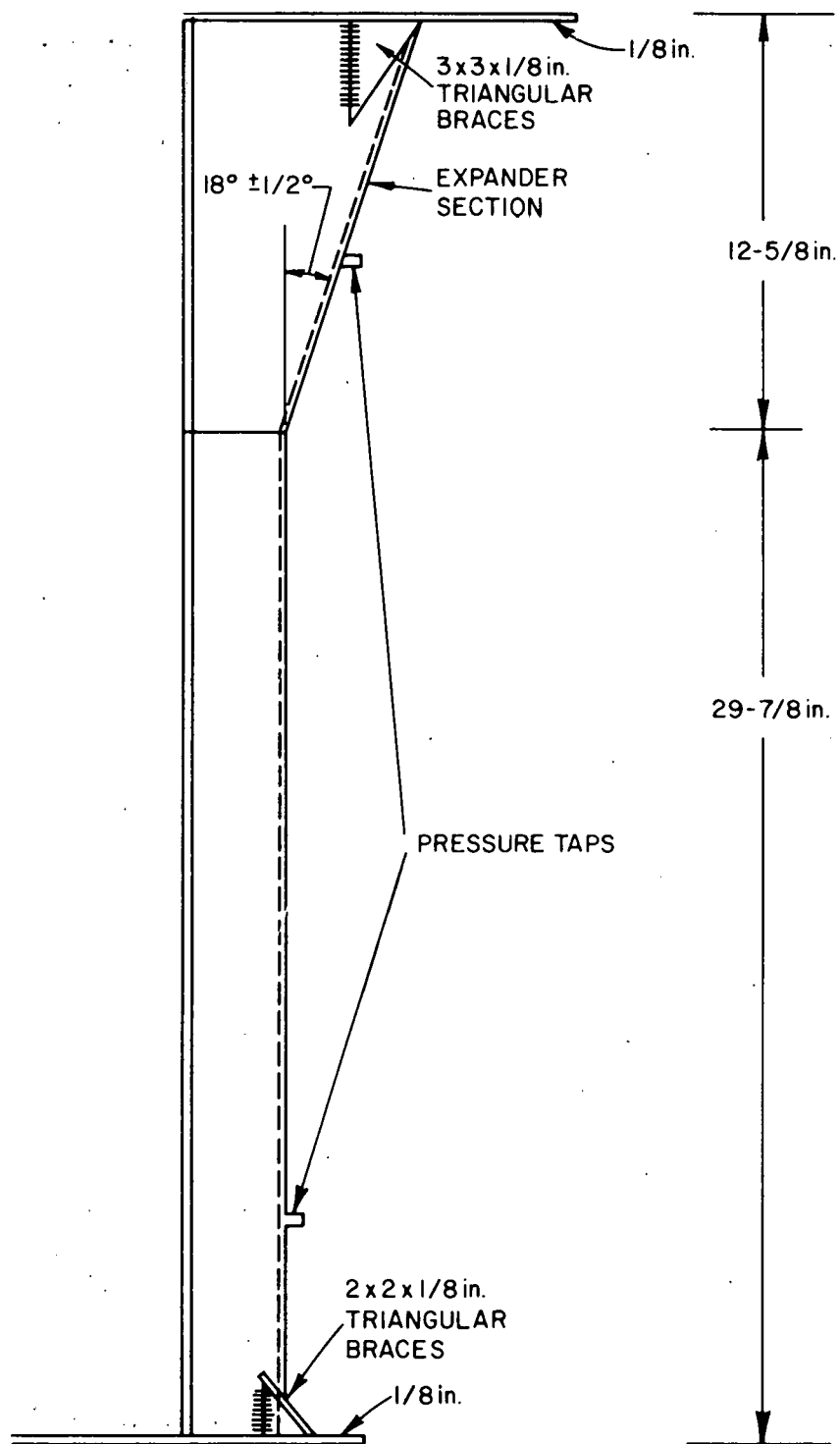
Equipment Setup and Modification

A 6-inch throat half-diameter by 30-inch-long (nominal) lower section was installed. This section is shown in Figure B-16. No other modifications were required.

Operation and Results

Run 69 was the only run utilizing this configuration. The conditions of this run were a venturi velocity of from 22 to 70 ft/s at bed agglomerate concentration of from 12% to 16%. The sphere discharge rate versus venturi velocity behavior was determined and is shown in Figure B-14. Also shown in this figure are the results for 4-1/2 X 2, 3 X 2, and 2 X 2 in. configurations. All conditions other than venturi velocity were ostensibly identical in all these runs. The figure shows that the sphere discharge rate in the 6 X 30 in. configuration is lower than that of the 4-1/2 X 2 in. configuration and higher than that of the 2 X 2 in. configuration at any fixed value of venturi velocity. The sphere discharge-rate behavior is quite similar to that of the 3 X 2 in. configuration.

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Figure B-16. AAG COLD MODEL: VIEW OF THE
6 x 30 INCH VENTURI SECTION

Conclusions

The 6 X 30 in. configuration performs less satisfactorily than the 4-1/2 X 2 in. configuration. Increasing the length of the throat region has a detrimental effect on discharge rate. This has already been determined with the 3 X 3 and 3 X 2 in. configurations; the former had a decidedly lower discharge rate at the same venturi velocity and bed spheres concentration.

Tests on a 6 X 2 in. Venturi

a. Discharge Rate Studies

Summary

We conducted five tests (Runs 83 through 87) in the half-section model using a 6 X 2 in. venturi configuration. These runs resulted in higher discharges and, unfortunately, higher concentration of bed material in discharge.

Objective

The objective of these tests was to investigate the behavior of the 6 X 2 in. based venturi configuration.

Operation and Results

These runs are summarized in Table B-5. The use of the 6 X 2 in. venturi required greater venturi flows for the same venturi velocity than the smaller 4.5 X 2 in. used previously. At a value of superficial velocity of approximately 2 ft/s, as used in all tests following Run 57, the velocity through the grid bolt was inadequate to prevent some weeping of bed material into the plenum chamber below the grid. Thus, higher superficial velocities in some of the tests were required. Runs 83 through 87 resulted in discharge rates substantially higher than those observed in the 4.5 X 2 in. venturi configuration at comparable velocities. These higher rates were accompanied by larger amounts of bed material. At velocities as high as approximately 70 ft/s (Run 85), the weight fraction of spheres in the discharge ranged from 0.56 to 0.88. In contrast, runs with the 4.5 X 2 in. venturi at comparable velocities achieved concentrations of approximately 99%.

The poor classification was thought to be because of inadequate (too low) gas velocities below the venturi throat. The velocities in Runs 83 through 87 were in the range of 5.6 to 8.7 ft/s in the 12-in. Schedule 40 pipe below the venturi. Because this velocity is close to the bed material's terminal velocity, we felt we could re-entrain this material by using higher velocities in the pipe. This initiated the classification study, which is the subject of the next section.

Table B-5. SUMMARY OF COLD-FLOW MODEL RUN CONDITIONS (Runs 83 through 87)

<u>Run No.*</u>	<u>Venturi** Configuration</u>	<u>Venturi Velocity</u>	<u>Superficial Velocity</u>	<u>Weight Fraction Spheres in Bed, x_b (avg)</u>	<u>Weight Fraction Spheres in Discharge, x_d</u>
83	6 x 2 - 12/12	44.3	2.07	0.131 - 0.219	0.227 - 0.500
84	6 x 2 - 12/12	61.0	2.43	0.038 - 0.253	0.220 - 0.754
85	6 x 2 - 12/12	68.7	2.51	0.104 - 0.234	0.565 - 0.883
86	6 x 2 - 12/12	44.8	2.11	0.045 - 0.231	0.034 - 0.569
87	6 x 2 - 12/12	59.2	3.28	0.015 - 0.229	0.077 - 0.569

* The following parameters were common to all tests:

Sphere Diameter = 0.152 in.

Average Particle Diameter = 0.0254 in.

Sphere Density = 157 lb/ft³

Bed Particle Density = 72.0 lb/ft³

Bed Bulk Density = 36.0 lb/ft³

** The number after the hyphen indicates diameter of half-section below throat (in.). The number after the slash indicates the pipe diameter below venturi.

Conclusions

Initial tests of the 6 X 2 inch venturi configuration resulted in higher concentration of bed material in the discharge. Because this would imply loss of carbonaceous material and concomitant reduction in efficiency in the actual U-GAS® Process reactor, we modified the system to improve this aspect of performance. We determined that inadequate velocities were being maintained in the discharge line below the venturi.

For reasonable values of bed sphere concentration $X_{b(avg)}$, i.e., above 8%, the sphere discharge rate $\dot{M}$ appears to follow a relationship of the form -

$$\dot{M} = \beta X_{b(avg)}^n \quad (B-2)$$

where n is the slope of the log-log plot of $\dot{M}$ versus $X_{b(avg)}$ and varies from 1.4 to 2.

b. Classification Studies

Summary

A series of tests (Runs 88 through 101) were conducted on the half-section cold-flow model with a 6 X 2in. venturi. For Runs 88-91, a converging zone consisting of a half-section 21-3/16 inches long and 8 inches in diameter (Figure B-17) and a 7 ft 9 inch long, 12 in. fully round section (Schedule 40) was below the converging zone. In Runs 92-93, a 7 ft 9 in. long, 8 in. fully round section (schedule 40) replaced the 12 in. pipe. In Runs 94-97, a 7 ft 9 in. long, 8 in. half-round section replaced the 8 in. fully round pipe. In Runs 98-101 a ledge in the shape of a segment of a circle was installed below the venturi where the transition from half to fully round took place prior to Run 94. All these configurations were used to attempt to improve the classification at lower venturi velocities. The sphere discharge rate was lowered as result of these changes, and the sensitivity of the venturi velocity to sphere-discharge rate was lessened as well.

Objective

The objective of these runs was to investigate the classification of spheres in the total bed discharge.

Equipment Setup

Runs 88-91 were conducted by modifying the venturi of Runs 83-87 and reducing the diameter of the converging half-section below the throat to 8 in. (Figure B-17). In Runs 92-93, a 7 ft 9 in. long, 8 in. (schedule 40) fully round section replaced the 12 in. (schedule 40) fully round section with the same length below the converging zone. In Runs 94 through 97, a 7 ft 9 in. long, 8-in. diameter pipe was replaced by a half-section

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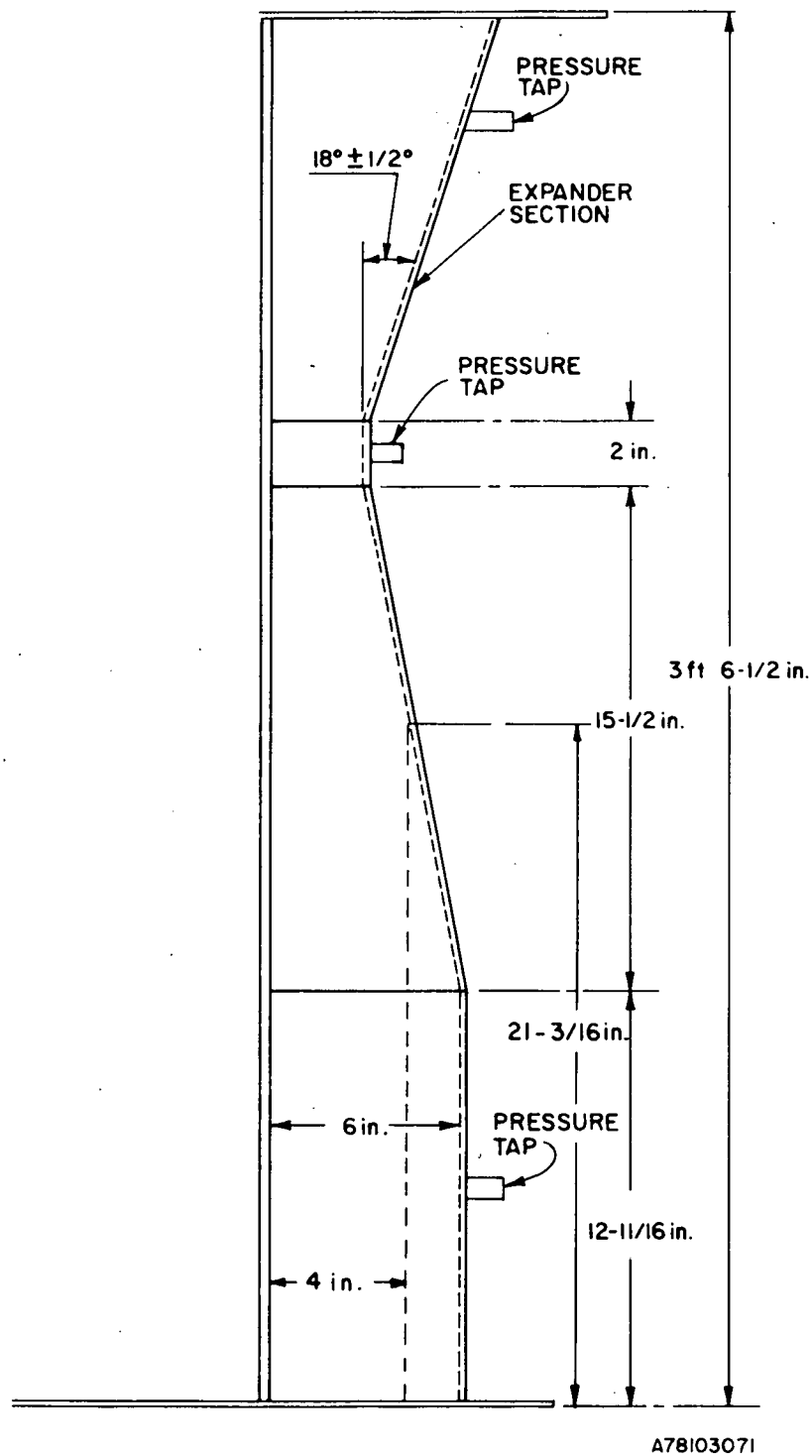


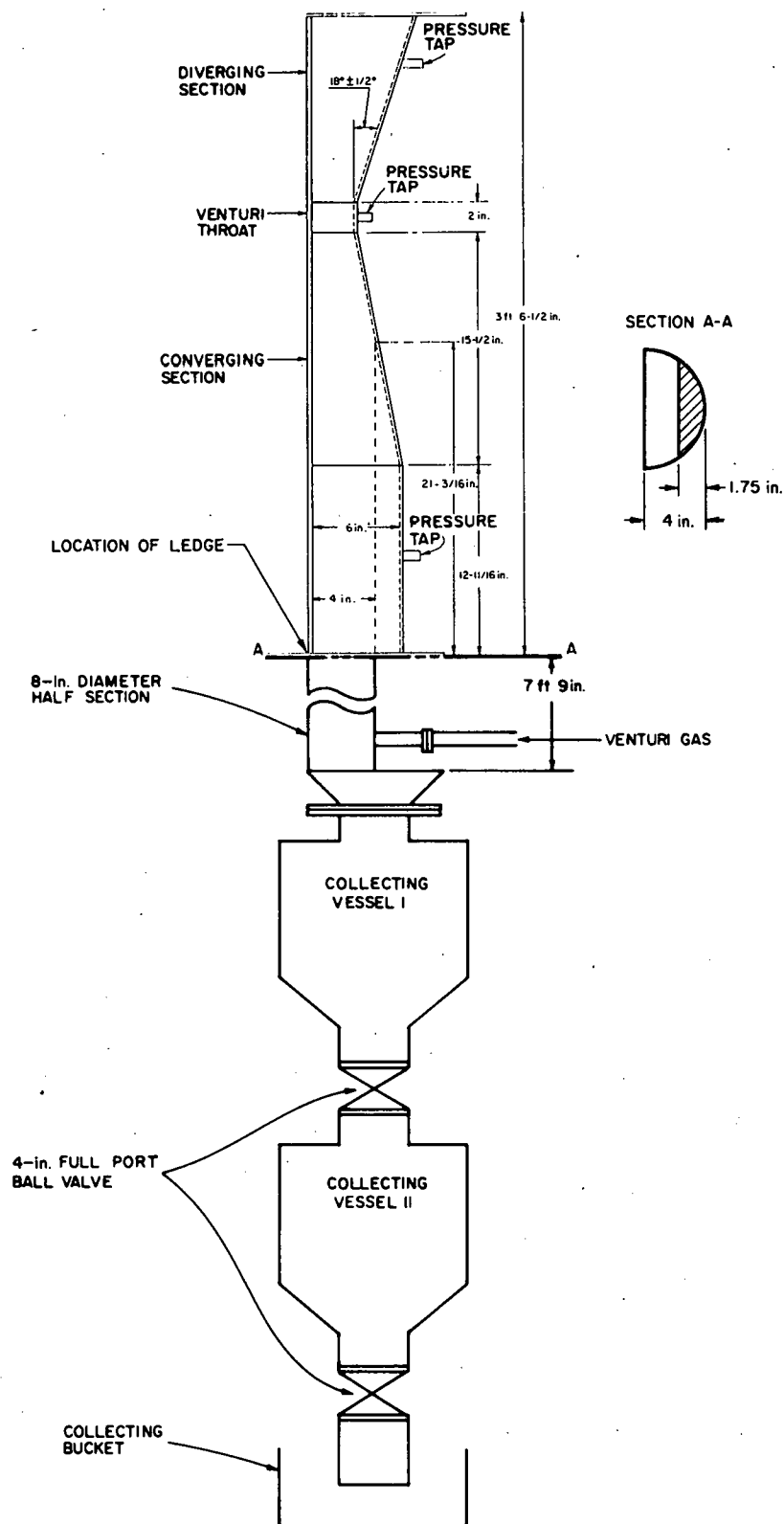
Figure B-17. AAG COLD MODEL: VIEW OF THE 6 x 2 in.
VENTURI SECTION WITH HALF-SECTION 21-3/16 in.
LONG AND 8-in. DIAMETER SLEEVE USED IN
RUNS 88-141

pipe of the same length and diameter. Also, two 2-in. ball valves in the discharge line were replaced by 4-in. ball valves to shorten the time needed to discharge material from Vessel I to Vessel II (Figure B-18). In Runs 98-101, a ledge as shown in Figure B-18 was installed at the bottom flange of the venturi section. The open area of the 8 in. half-diameter pipe at the ledge is 120% of the area of the 6-in. venturi throat.

Operation and Results

These runs, summarized in Table B-6, were to investigate the classification of agglomerates in total bed discharge. The poor classification in Runs 83 through 87 was thought to be because of inadequate (too low) gas velocities below the venturi throat. The velocities in Runs 83 through 87 were in the range of 5.6 to 8.7 ft/s in the 12-in. schedule 40 pipe below the venturi. Because this velocity is close to the bed material's terminal velocity, we felt we could re-entrain this material by using higher velocities in the pipe. We first modified the venturi by reducing the diameter of the half-section below the throat to 8 in. (Figure B-17). Runs 88 through 91 were conducted to determine if there was any improvement in classification. Although a slight improvement was detected, we decided to modify the discharge pipe below the venturi to an 8-in. diameter (Runs 92-93). This increased the velocity, for venturi velocities in the range of 40 to 70 ft/s, to 11-20 ft/s in the pipe. Excellent classification had been obtained in previous runs with lower pipe velocities above 10 ft/s. The tests conducted (Runs 92-93) with this new arrangement confirmed our hypothesis in that near perfect classification was obtained. The discharge rate of spheres was comparable with that obtained in Run 87, which was conducted at comparable conditions.

Test results from Run 93 showed improvement in total discharge rate, but with an unacceptable classification. An analysis of velocities in different sections below the venturi throat indicated that a velocity of 44.4 ft/s (at which Run 93 was conducted) produced a velocity of 24.7 ft/s in the 21-3/16 inch long converging zone and 12.4 ft/s in the 7 ft 9 inch long pipe. Because the velocity in this long section was very close to the terminal velocity of the bed material (acrylic), classification was not at an acceptable level. We hoped that replacement of the long 8-in. full section with a half-section will result in a better classification of spheres in the discharge by providing a greater distance to disengage the bed solids. Runs 94 through 97, conducted with this new configuration, did confirm this hypothesis, but total discharge rates were lower than those that occurred in Run 93 at comparable conditions. (Compare Run 95 with Run 93 in Figure B-19). In addition, Runs 96 and 97, both at higher superficial velocity than the latter runs and not to be directly compared with them, showed very little difference in



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Figure B-18. MODIFIED 4-FOOT DIAMETER HALF-SECTION COLD-FLOW MODEL WITH 7-ft 9-in. LONG HALF SECTION 8-in. DIAMETER PIPE AND LEDGE DETAILS (Runs 98-106 Only)

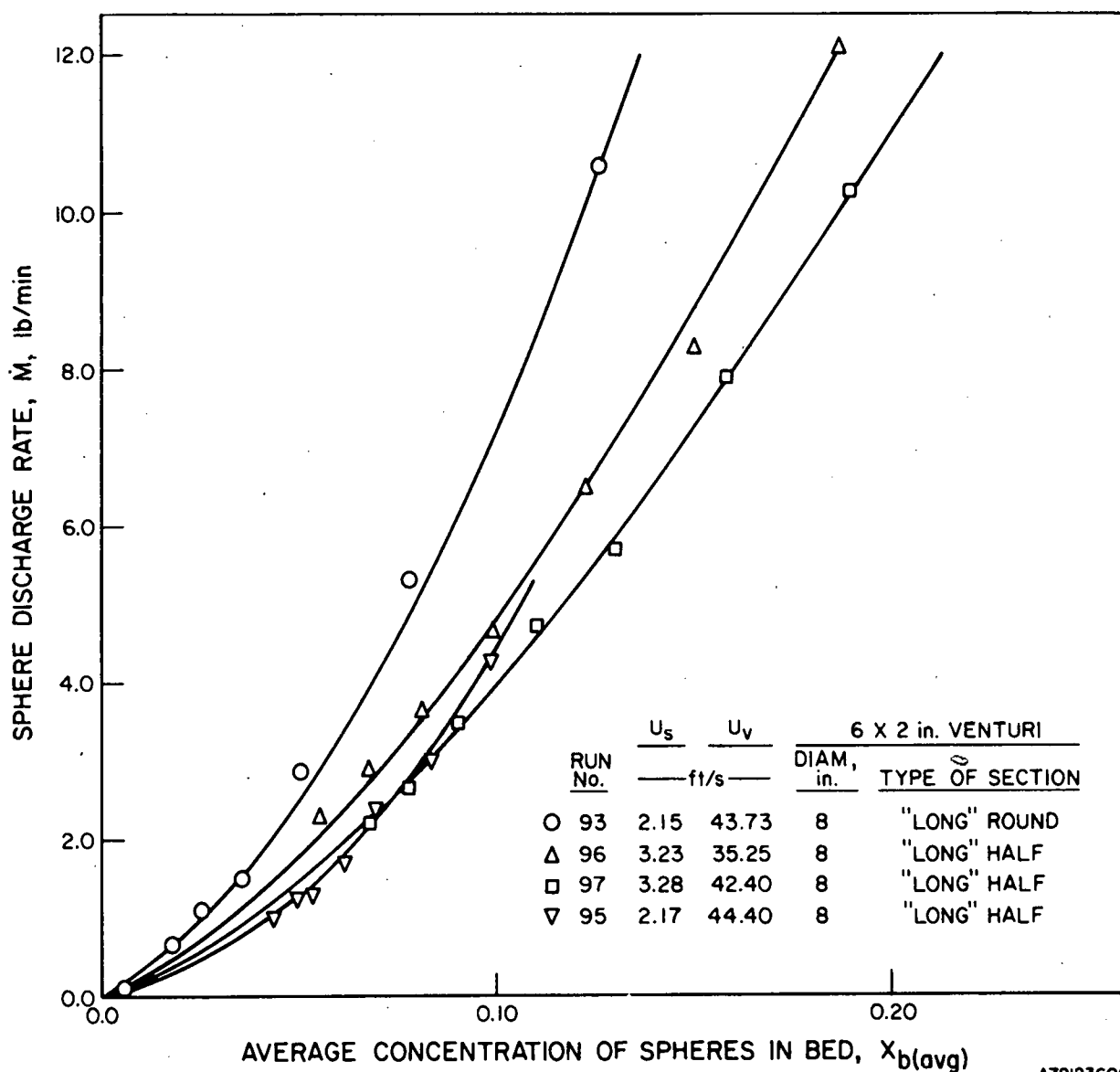
Table B-6. SUMMARY OF COLD-FLOW MODEL RUN CONDITIONS (Runs 88 Through 101)

Run No.*	6 X 2 in. Configuration, in.**	Venturi Velocity ft/s	Superficial Velocity	Weight Fraction of Spheres in Bed, X_b (avg)	Weight Fraction of Spheres in Discharge, X_d
88	8/12 (F)	63.4	3.74	0.019-0.129	0.200-0.711
89	8/12 (F)	58.4	3.28	0.067-0.266	0.200-0.790
90	8/12 (F)	59.6	3.29	--	--
91	8/12 (F)	59.6	3.30	0.030-0.282	0.053-0.803
92	8/8 (F)	58.7	3.27	0.028-0.265	0.917-1.000
93	8/8 (F)	43.7	2.15	0.004-0.214	0.333-0.963
94	8/8 (H)	50.2	2.26	0.021-0.181	1.000-1.000
95	8/8 (H)	44.4	2.17	0.040-0.098	1.000-1.000
96	8/8 (H)	35.3	3.23	0.055-0.231	0.600-0.856
97	8/8 (H)	42.4	3.28	0.066-0.243	0.933-1.000
98	8/8 (H-L)	43.5	2.16	0.038-0.244	1.000-1.000
99	8/8 (H-L)	36.1	3.25	0.059-0.272	0.889-1.000
100	8/8 (H-L)	50.8	2.24	0.063-0.263	0.972-1.000
101	8/8 (H-L)	37.0	2.14	0.037-0.206	0.800-0.991

* The following parameters were common at all the tests mentioned above:

Sphere Diameter = 0.152 in.
 Sphere Density = 157 lb/ft³
 Bed Particle Diameter = 0.0254 in.
 Bed Particle Density = 72.0 lb/ft³
 Bed Bulk Density = 36.0 lb/ft³

** The first number indicates the diameter of the half-section below the throat; the second number indicates the pipe diameter below the venturi; the letters F and H represent full and half sections, respectively. H-L represents half-section with the ledge.



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Figure B-19. AAG COLD MODEL: RATE OF SPHERE DISCHARGE
VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED

discharge rate, despite a 25% difference in venturi velocity. (Classification was high, greater than 93% for Run 97 and greater than 75% at X_b greater than 0.08 in Run 96.)

The ledge shown in Figure B-18 was installed at the bottom flange of the venturi section. The open area of the 8-in. half-diameter pipe at the ledge is 120% of the area of the 6-in. venturi throat. This configuration was employed in Runs 98 through 101. This configuration decreased the sphere discharge rates below the levels achieved in the previous configuration. (See next section.)

Conclusions

Initial tests of the 6 X 2 in. venturi configuration resulted in higher concentration of bed material in the discharge. Because this would imply loss of carbonaceous material and concomitant reduction in efficiency in the actual U-GAS[®] Process reactor, we modified the system to improve this aspect of its performance. We determined that inadequate velocities were being maintained in the discharge line below the venturi, and modified the system by reducing the pipe size from 12 to 8 in. in diameter. This resulted in restoring the classification to acceptable levels. The necessary velocity to accomplish this with the bed material was of the order of 10 ft/s.

Installation of an 8-in. diameter long half-section (used in Runs 94 through 97) produced a steady flow pattern at the throat which did not favor discharge of larger volumes of material from the expanding zone of the venturi section. In earlier runs (92 and 93), we used an 8-in. diameter full section below the venturi, which produced some type of obstruction to flows going to the venturi section. We hypothesize that this obstruction, in turn, created some unsteadiness at the venturi throat, and that because of this unsteadiness we got more discharge in Run 93 than in the remaining runs (94 through 97).

Installation of the ledge (Runs 98-101) caused the sphere discharge rate to drop below the rates achieved in previous runs. A possible explanation of this is that the ledge acts to break up clouds of spheres and bed material by direct impact, resulting in easier re-entrainment of the bed material.

c. Pressure Fluctuation Studies

Summary

Tests (Runs 101 through 115) at various conditions with several configurations were performed to determine the effect of the various configurations discussed above on discharge rate and classification of spheres. Initial tests (Runs 101 through 106) were performed to investigate the effect of a ledge in the shape of a segment of a circle (see

previous section) introduced below the venturi converging section to simulate the sudden transition from half to fully round pipe as employed in Runs 92 and 93. All of these runs (Runs 102 through 115) employed pressure-recording instrumentation to determine if the observable differences in discharge behavior between the configurations was reflected in the pressure fluctuations at various points in the unit. Runs 102 through 115 were replicates of earlier runs for which pressure-recording instrumentation was not available originally. Results of value in assessing the effect of superficial velocity were obtained.

Objective

The objectives of these studies are to determine the effect of an obstruction in the piping below the venturi on the discharge rate and classification of spheres, and to measure the pressure fluctuations in the system with the latter obstruction and with previously employed configurations.

Equipment Setup

Pressure recording instrumentation was employed in Runs 102 through 115 to record the pressure traces at and below the venturi. A Gould Brush Model 220 Recorder, Honeywell Model 218-1 Accudata Signal Conditioner, and Schaevitz 0-20 psig Range Transducer were used. For convenience the configuration with the 8-in. half-diameter pipe with the ledge will be referred to as the (H-L) configuration. The same setup without the ledge will be referred to as the (H) configuration. The setup originally used in Runs 92 and 93 with an 8-in. fully round pipe at the flange below the venturi section will be referred to as the (F) configuration.

Operation and Results

Runs 102 through 116 were performed at various conditions (Table B-7) to assess the magnitude of the configurational effect on the discharge rates. Thirteen of these runs were used for this purpose along with four earlier runs. These fifteen runs are divided into six sets: I-A, I-B, II-A, II-B, III-A, and III-B (Table B-8). Except for Set III-A, each set contains a run with each of the three configurations used since Run 92. The venturi velocity and superficial velocity within each set were maintained constant during each run and at substantially identical values. The discharge rate vs. average concentration of spheres in bed for each set appear as Figures B-20 through B-25.

In general it appears that the configuration below the venturi affects the discharge rates of spheres from the fluidized bed. In addition, the fully round configuration, with a few exceptions, gives a greater discharge rate.

Table B-7. SUMMARY OF COLD-FLOW MODEL RUN CONDITIONS (Runs 83 Through 116)

Run No.*	6 X 2 in. Configuration, in.**	Venturi Velocity	Superficial Velocity	Weight Fraction of Spheres in Bed, X_b (avg)	Weight Fraction of Spheres in Discharge, X_d
		ft/s			
102	8/8 (H-L)	35.1	3.25	0.083-0.219	0.964-1.000
103	8/8 (H-L)	35.4	2.18	0.045-0.261	0.955-1.000
104	8/8 (H-L)	34.6	3.24	0.067-0.220	0.952-0.979
105	8/8 (H-L)	49.6	2.20	0.066-0.254	0.988-1.000
106	8/8 (H-L)	43.6	2.07	0.047-0.261	0.975-1.000
107	8/8 (H)	34.6	3.19	0.064-0.263	0.875-0.902
108	8/8 (H)	34.6	2.09	0.020-0.217	0.333-0.876
109	8/8 (H)	43.2	2.11	0.051-0.208	0.941-1.000
110	8/8 (H)	49.6	2.22	0.041-0.208	1.000-1.000
111	8/8 (F)	33.3	3.13	0.042-0.228	0.250-0.715
112	8/8 (F)	44.5	2.10	0.044-0.266	0.750-0.937
113	8/8 (F)	49.5	2.14	0.067-0.285	0.727-1.000
114	8/8 (F)	49.4	3.16	0.943-1.000	0.100-0.276
115	8/8 (F)	34.2	3.27	0.026-0.252	0.127-0.594
116	8/8 (H)	39.9	2.91	0.116-0.279	0.973-1.000

* The following parameters were common at all the tests mentioned above:

Sphere Diameter = 0.152 in.
 Sphere Density = 157 lb/ft³
 Bed Particle Diameter = 0.0254 in.
 Bed Particle Density = 72.0 lb/ft³
 Bed Bulk Density = 36.0 lb/ft³

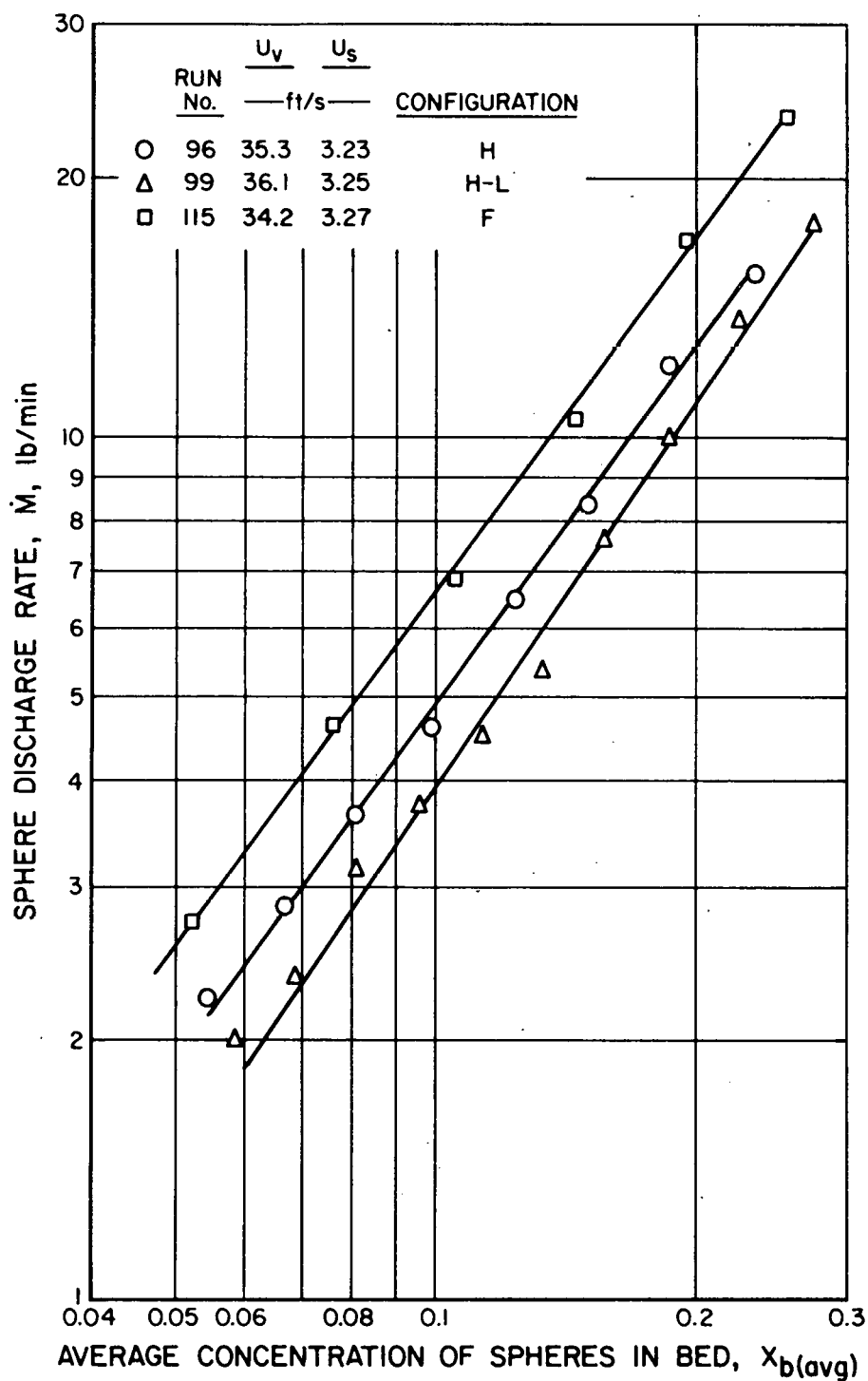
** The first number indicates the diameter of the half-section below the throat; the second number indicates the pipe diameter below the venturi; the letters F and H represent full and half sections, respectively. H-L represents half-section with the ledge.

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Table B-8. SUMMARY OF RUNS USED FOR DETAILED ANALYSIS OF PRESSURE FLUCTUATION EFFECTS

Set No.	Run No.	Venturi Velocity	Superficial Velocity	Configuration*
		ft/s		
I-A	99	36.1	3.25	H-L
	96	35.3	3.23	H
	115	34.2	3.27	F
I-B	104	34.6	3.24	H-L
	107	34.6	3.19	H
	111	33.3	3.13	F
II-A	98	43.5	2.16	H-L
	95	44.4	2.17	H
	93	43.7	2.15	F
II-B	106	43.6	2.07	H-L
	109	43.2	2.11	H
	112	44.5	2.10	F
III-A	100	50.8	2.24	H-L
	94	50.2	2.26	H
	--	--	--	F
III-B	105	49.6	2.20	H-L
	110	49.6	2.22	H
	113	49.5	2.14	F

* Notation is as in Table B-7.



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Figure B-20. SPHERE DISCHARGE RATE, $\dot{M}$, VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED, $X_{b(avg)}$, SET 1-A. EFFECT OF CONFIGURATION

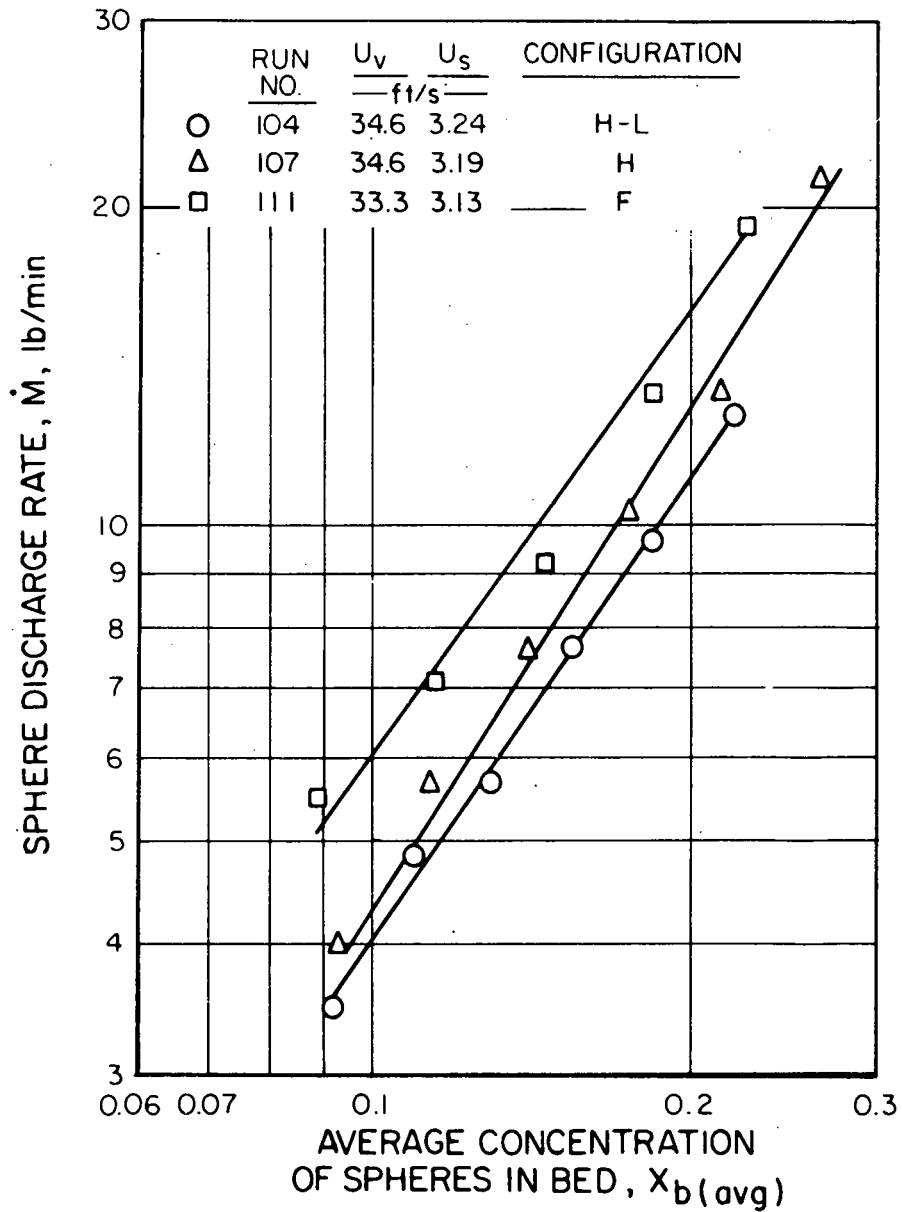
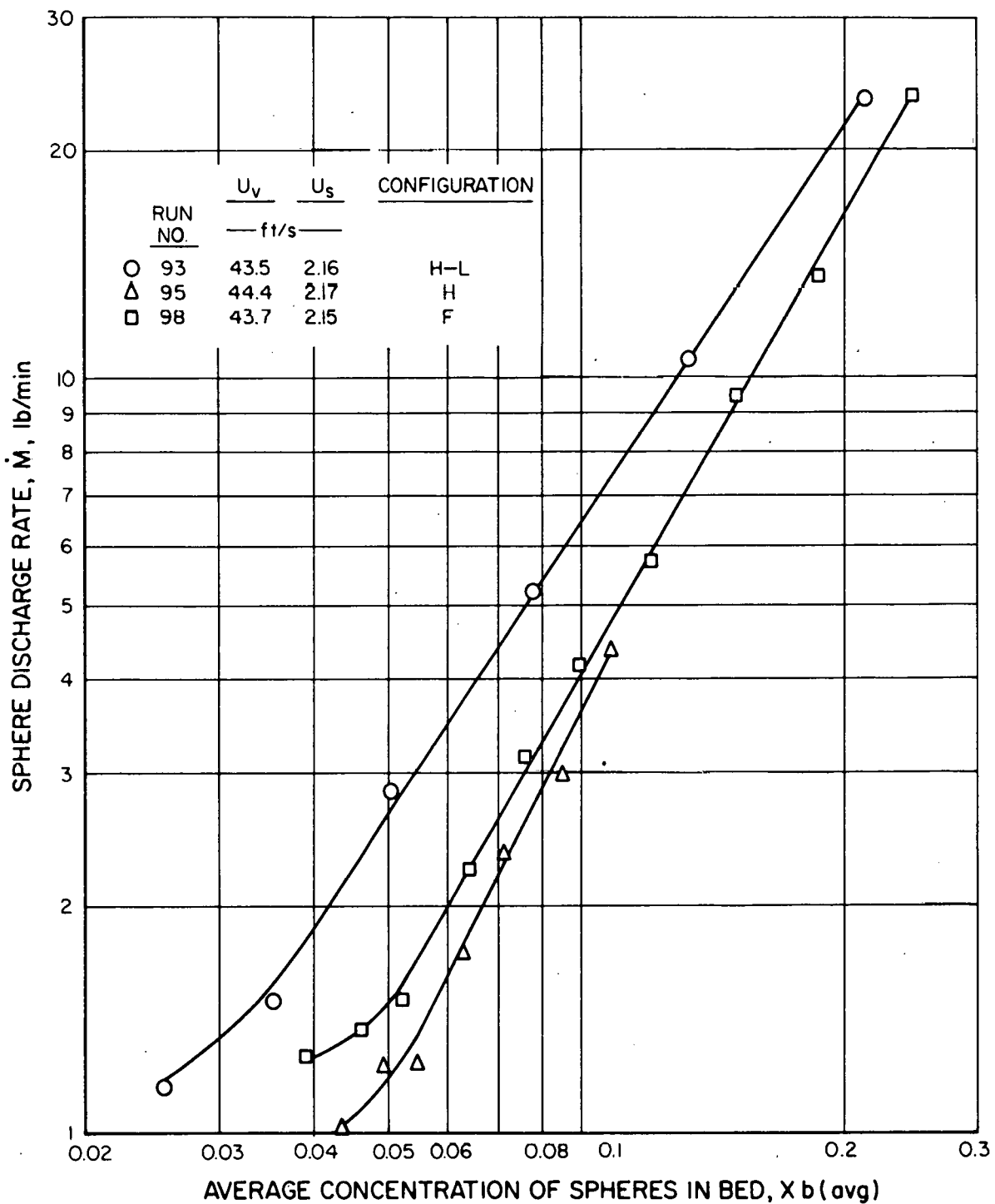
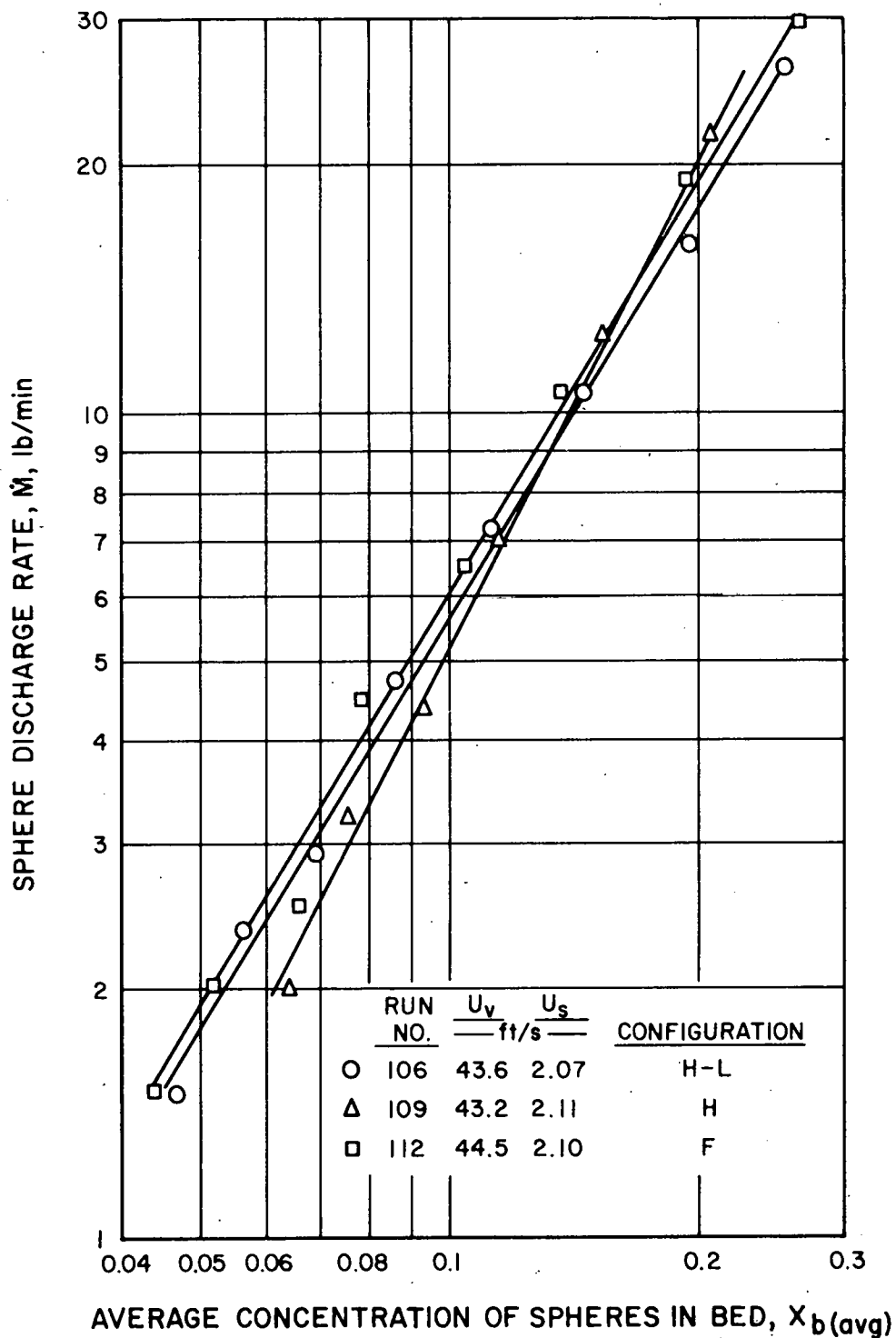


Figure B-21. SPHERE DISCHARGE RATE, $\dot{M}$, VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED, $X_{b(avg)}$, SET 1-B. EFFECT OF CONFIGURATION



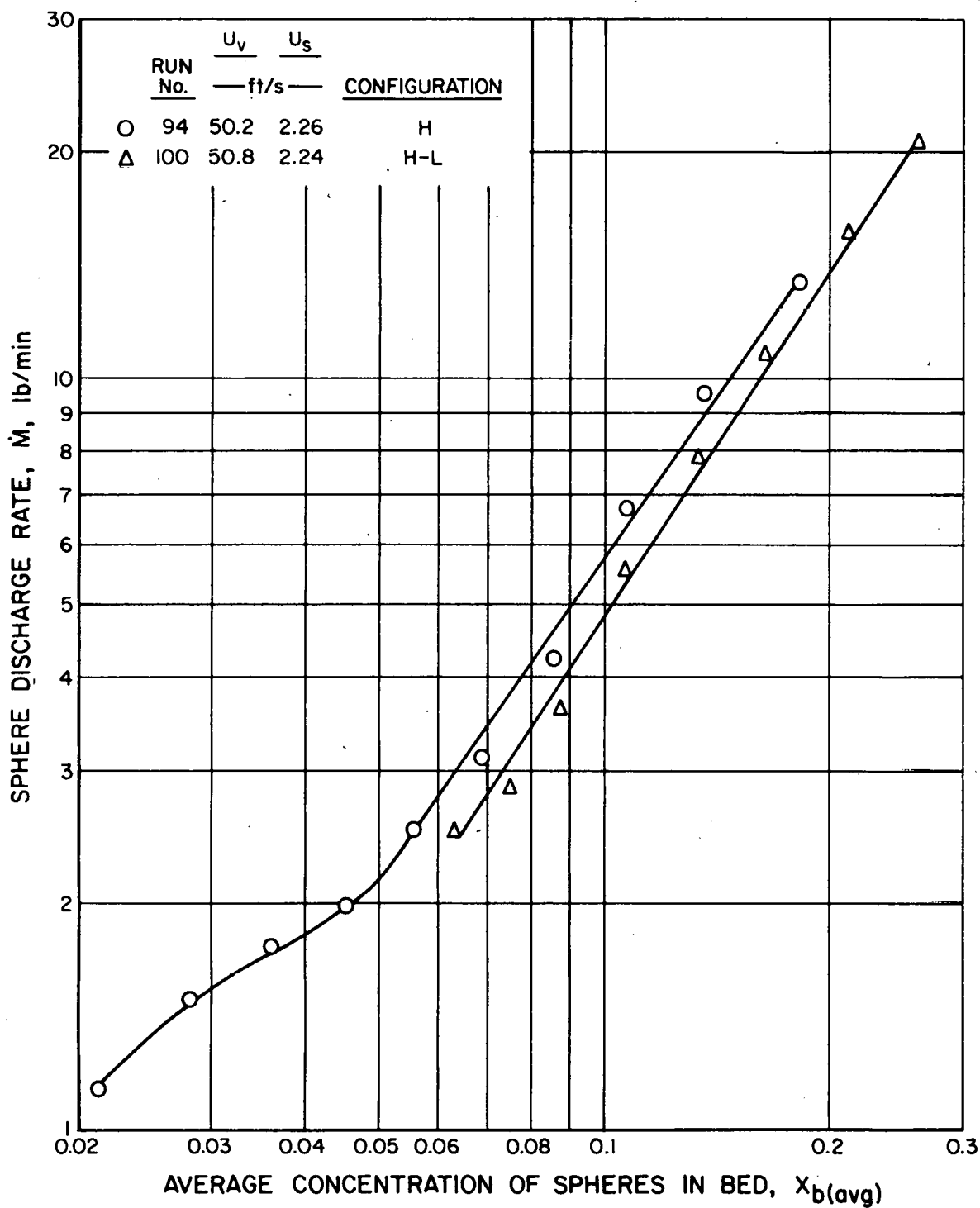
A79020247

Figure B-22. SPHERE DISCHARGE RATE, $\dot{M}$, VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED



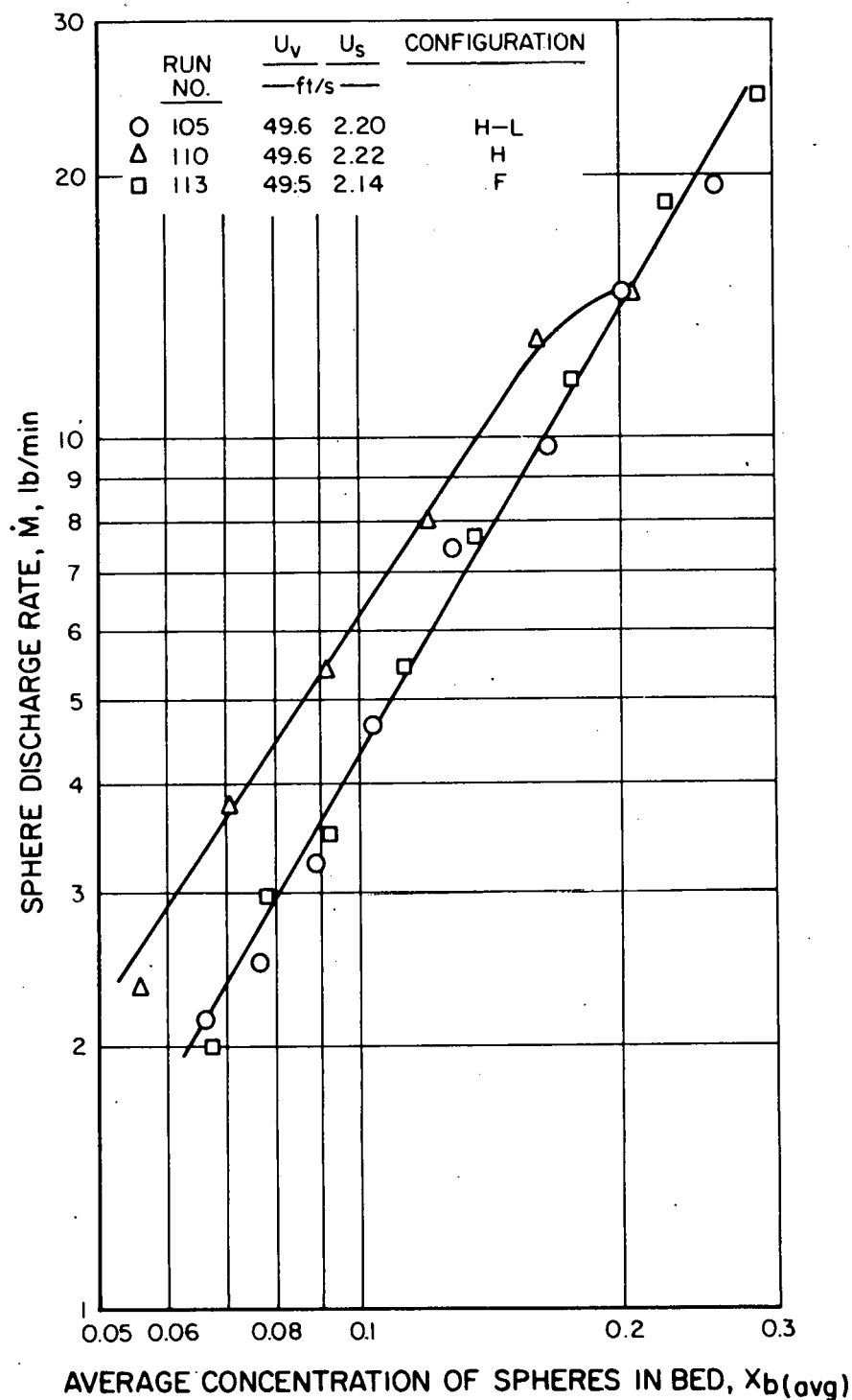
A79020250

Figure B-23. SPHERE DISCHARGE RATE, $\dot{M}$, VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED, $X_{b(avg)}$, SET II-B. EFFECT OF CONFIGURATION



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Figure B-24. SPHERE DISCHARGE RATE, $\dot{M}$, VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED



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Figure B-25. SPHERE DISCHARGE RATE, $\dot{M}$, VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED, $X_{b(avg)}$, SET III-B. EFFECT OF CONFIGURATION

Some information on the effect of superficial velocity (grid velocity) is available from the runs performed. Run 114 at $U_v = 49.4$ ft/s, $U_s = 3.16$ ft/s, may be compared with Run 113 at $U_v = 49.5$ ft/s, $U_s = 2.14$ ft/s. Both runs were made with the F configuration. Run 97 at $U_v = 42.4$ ft/s, $U_s = 3.28$ ft/s may be compared with Run 109 at $U_v = 43.2$ ft/s, $U_s = 2.11$ ft/s. These runs were in the H configuration. The $\dot{M}$ vs. $X_{b(avg)}$ data for these runs are shown in Figures B-26 and B-27. The data from these four runs indicate that increased grid velocity results in less discharge from the venturi.

The pressure-recording instrumentation was employed in Runs 102 through 116. A set of typical recorder traces for Runs 105, 110, 113 (Set III-B) is shown in Figure B-28. A trace for the blower discharge line with an empty bed is also shown.

Some attempt has been made to statistically relate the frequencies observed on the pressure traces to the variation of discharge rates at fixed values of bed concentration. No correlation could be found. The frequencies varied only between 1.6 and 2.7 s⁻¹; when averaged over an 8-second interval they are presented in Table B-9.

The concentration of spheres in the discharge was generally highest with the H-L configuration, with the H configuration higher than the F, all at comparable conditions. This is illustrated in Figure B-29, X_d vs. $X_{b(avg)}$ for Set II-B. Results for other sets followed this pattern. A possible explanation of the above is that the ledge acts to break up clouds of spheres and bed material by direct impact, resulting in easier re-entrainment of the bed material. The velocity in the lower pipe in the F configuration is only half that in the H or H-L configuration for fixed venturi velocity so that re-entrainment is more difficult. Concentrations of spheres in the discharge in the worst case, for the F configuration, were above 90% at $X_{b(avg)} > 0.10$ for venturi velocities of 43 and 49 ft/s in any case (corresponding to pipe velocities of 12 and 14 ft/s, respectively).

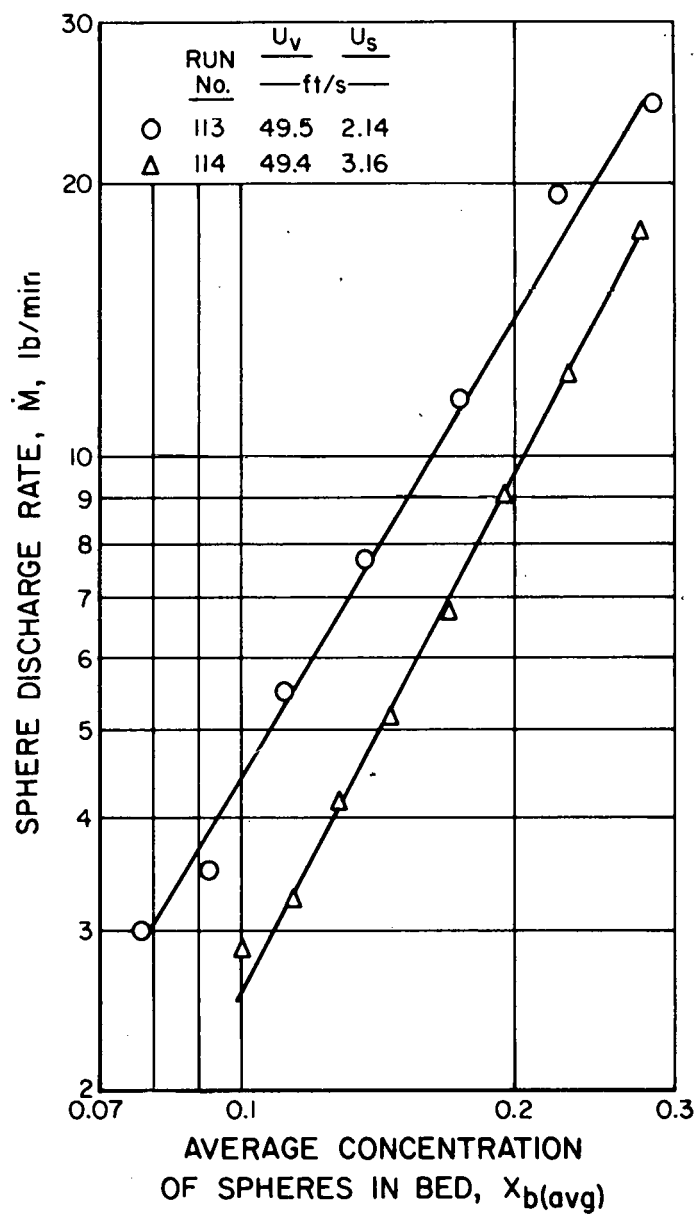
Conclusions

An appreciable effect of below-venturi configuration on discharge rate was observed. The particular values of venturi and superficial velocities employed appear to influence the relative behavior of the configurations with respect to sphere discharge rate.

Superficial velocity effects were examined, and it appears that increasing the superficial velocity (by increasing the grid flow at fixed venturi flow) results in a lessening of the discharge rate at any given bed concentration.

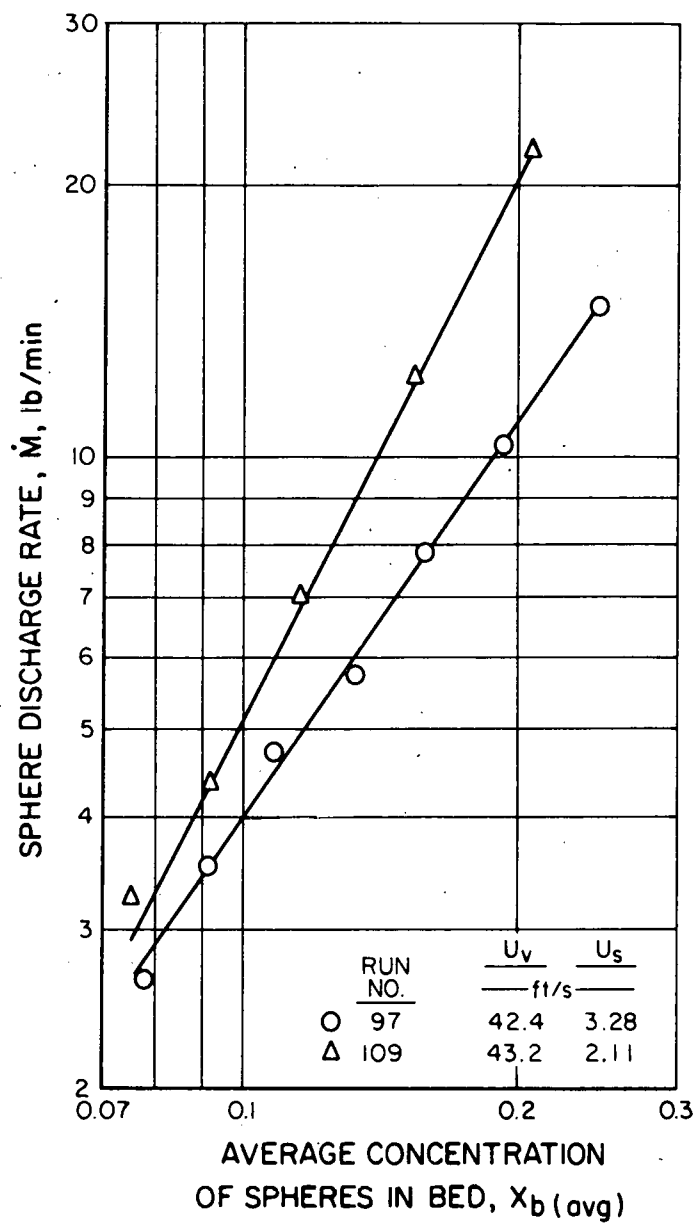
Table B-9. SPHERES DISCHARGE RATE, RUN CONDITIONS, VENTURI PRESSURE FLUCTUATION, AND FREQUENCY DATA FOR SELECTED RUNS

Run No.	Sphere Discharge Rate, $\dot{M}$, lb/min		Venturi Velocity	Superficial Velocity	Frequency, s ⁻¹	
	$x_b(\text{avg}) = 0.17$	$x_b(\text{avg}) = 0.18$			$x_b(\text{avg}) = 0.17$	$x_b(\text{avg}) = 0.18$
			ft/s			
104	8.8	9.5	34.6	3.24	2.50	2.38
105	11.1	12.5	49.6	2.20	2.00	2.50
106	13.2	14.3	43.6	2.07	2.00	1.75
107	10.01	10.8	34.6	3.19	2.75	2.25
109	15.4	17.0	43.2	2.11	2.00	1.9
110	14.7	16.43	49.6	2.22	2.13	2.13
111	12.0	13.10	33.3	3.13	1.63	1.88
112	15.6	17.14	44.5	2.10	2.00	1.84
113	11.1	12.5	49.5	2.14	2.13	2.13
115	13.9	15.2	34.2	3.27	2.13	2.25



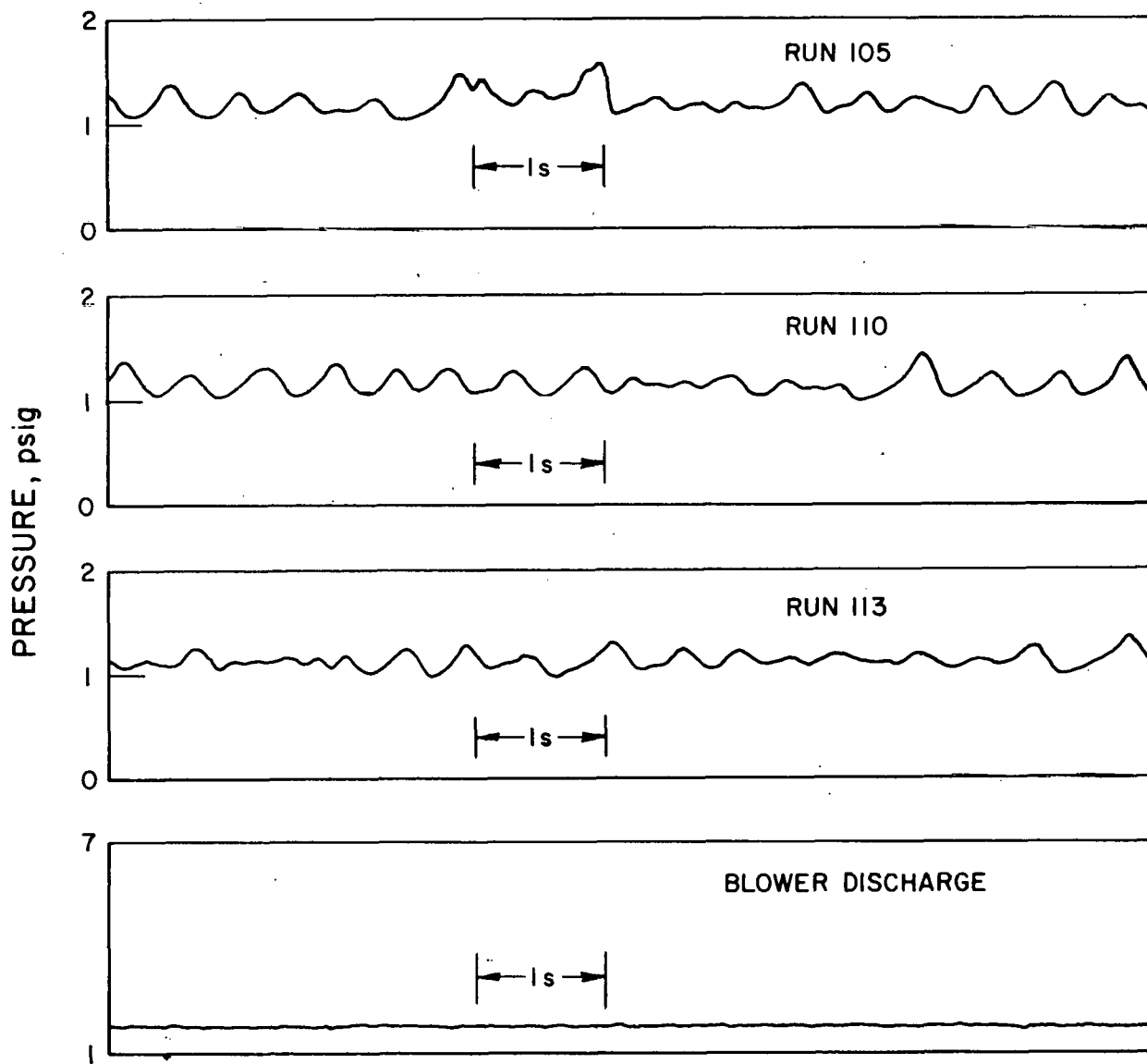
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Figure B-26. SPHERE DISCHARGE RATE, $\dot{M}$, VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED, $x_{b(avg)}$, FOR RUNS 113 AND 114. EFFECTS OF SUPERFICIAL VELOCITY



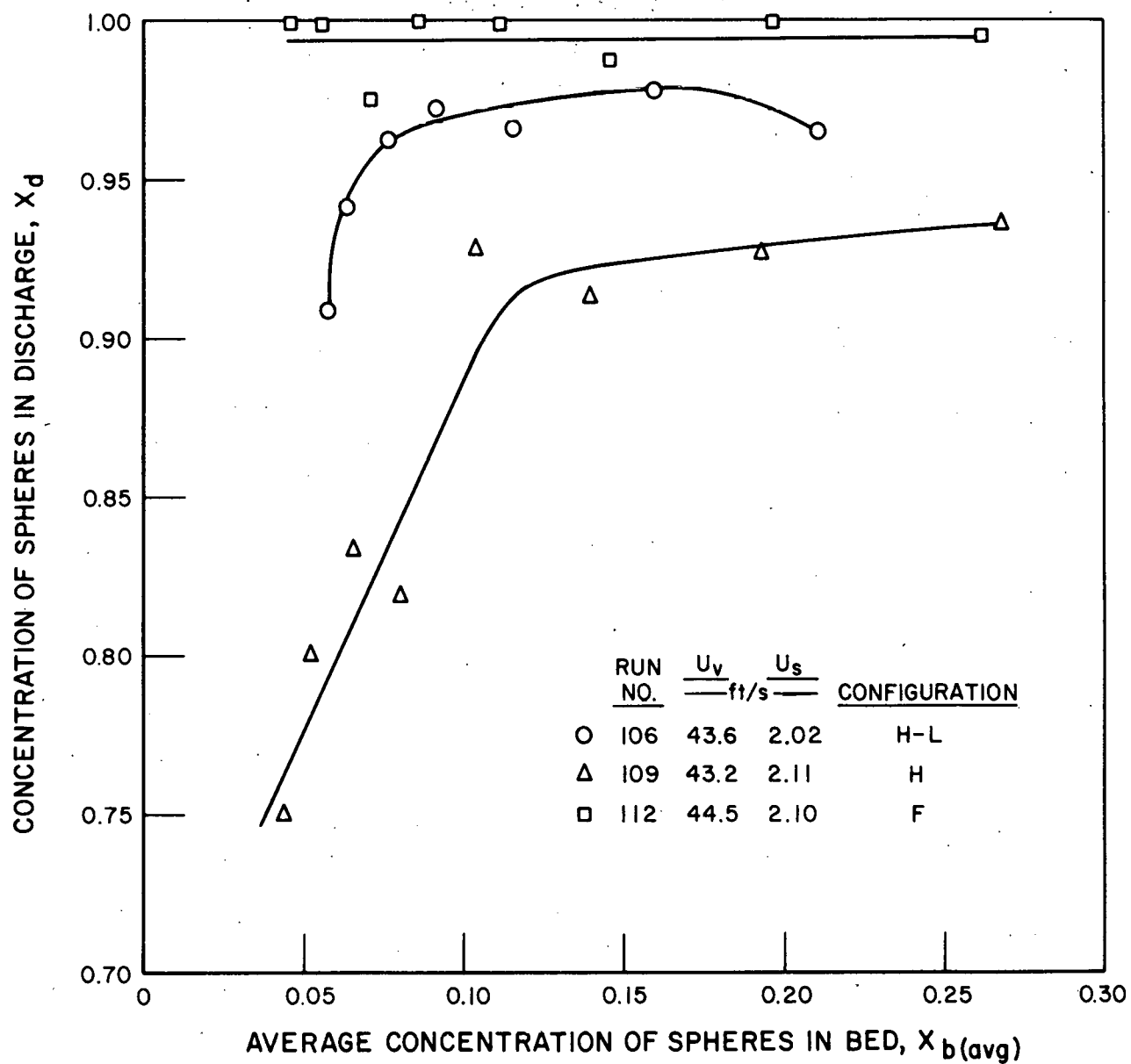
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Figure B-27. SPHERE DISCHARGE RATE, $\dot{M}$, VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED, X_b (avg), FOR RUNS 97 AND 109. EFFECT OF SUPERFICIAL VELOCITY



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Figure B-28. RECORDER TRACES FOR RUNS 105, 110, AND 113 AT $X_{b(avg)} \approx 0.20$



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Figure B-29. SPHERES CONCENTRATION IN DISCHARGE VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED, SET II-B. EFFECT OF CONFIGURATION

d. Discharge Rate Studies

Summary

A series of tests (Runs 116 through 132) were conducted at various conditions on the half-section cold-flow model with a 6-inch X 2-inch venturi configuration. All these runs used a 7-foot 9-inch-long, 8-inch-diameter half-section pipe below the venturi. Runs 116, 117, 118, 123, 124, 126, and 127 were conducted to investigate the effect of varying bed heights; Runs 119, 120, 121, 122, and 125 were conducted to further substantiate the effect of superficial velocity. Runs 128 through 132 were replicates of earlier runs. Table B-10 summarizes the range of variables studied in the above runs.

Objectives

The objectives of these runs were as follows:

1. To determine the effect of different bed heights on the rate of sphere discharge
2. To investigate further the effect of bed superficial velocity on the rate of sphere discharge
3. To find the degree of reproducibility of runs conducted at ostensibly similar conditions.

Equipment Setup

All the test runs (Runs 116 through 132) used the same experimental setup as that used in Runs 98 through 115 as shown in Figure B-30. Because the maximum bed height used in the tests was above the feed injection point, a blast connection was provided at the feed nozzle to force-inject bed material or spheres into the fluidized bed during the tests runs with higher bed heights.

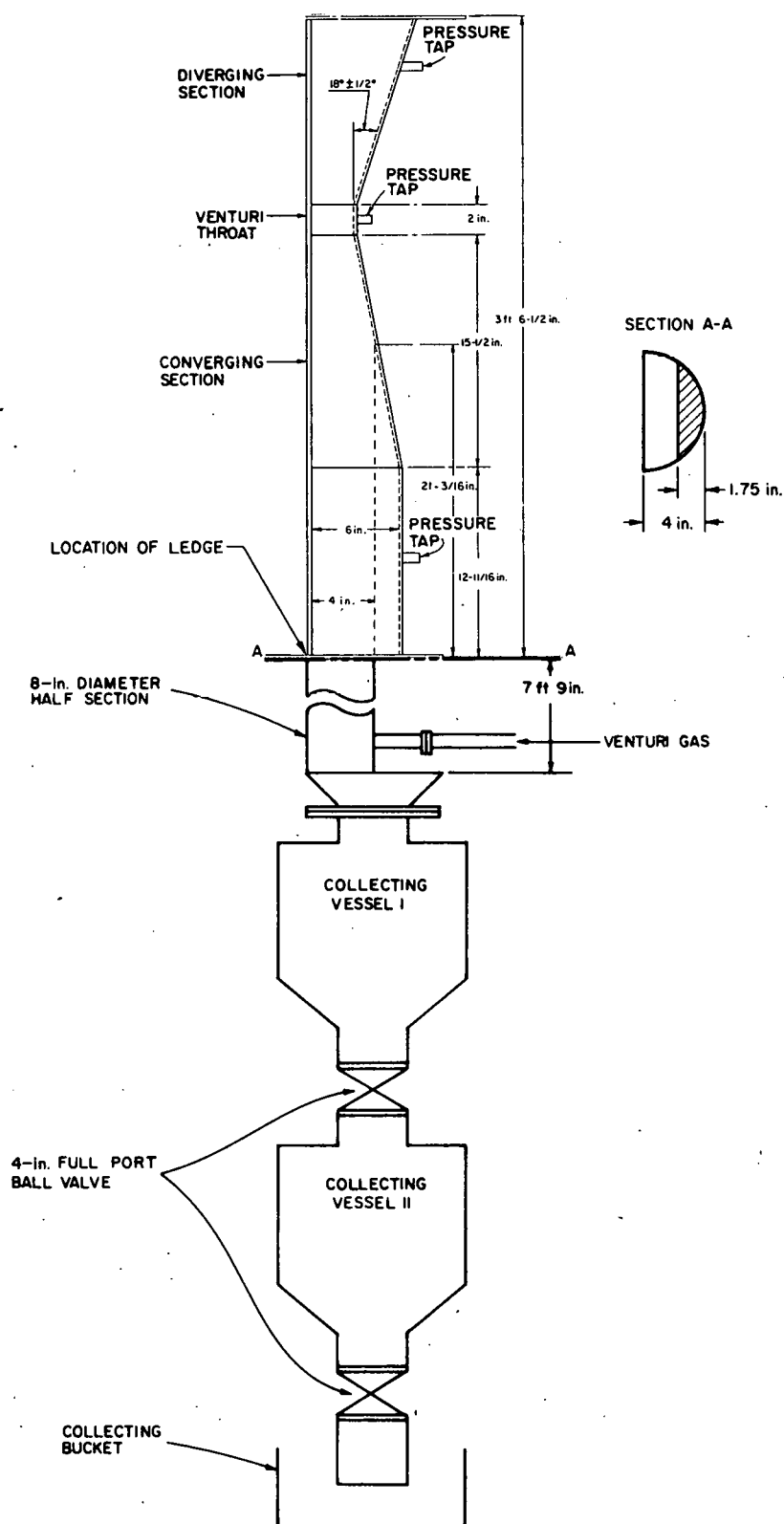
Operational Results

Runs 116, 117, 118, 123, 124, 126, and 127 were performed at bed heights varying from 6 to 48 inches and at approximately identical venturi and superficial velocities. Results of these runs are shown in Figure B-31. Figures B-32 and B-33 show the effect of superficial velocity on the rate of sphere discharge when all other conditions are maintained. Figures B-34 through B-37 show comparisons of results of replicate runs with those of previous runs at similar conditions. The figures also show 95% confidence limits of the regression lines developed from the data in each figure.

Conclusions

The test results show that there is a definite inverse correlation between sphere discharge rate and fluidized-bed height. Statistical analysis suggests the following type of correlation, as shown in Figure B-38:

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Figure B-30. MODIFIED 4-FOOT DIAMETER HALF-SECTION COLD-FLOW MODEL WITH 7-ft 9-in. LONG HALF SECTION 8-in. DIAMETER PIPE AND LEDGE MATERIALS (Runs 98-106 Only)

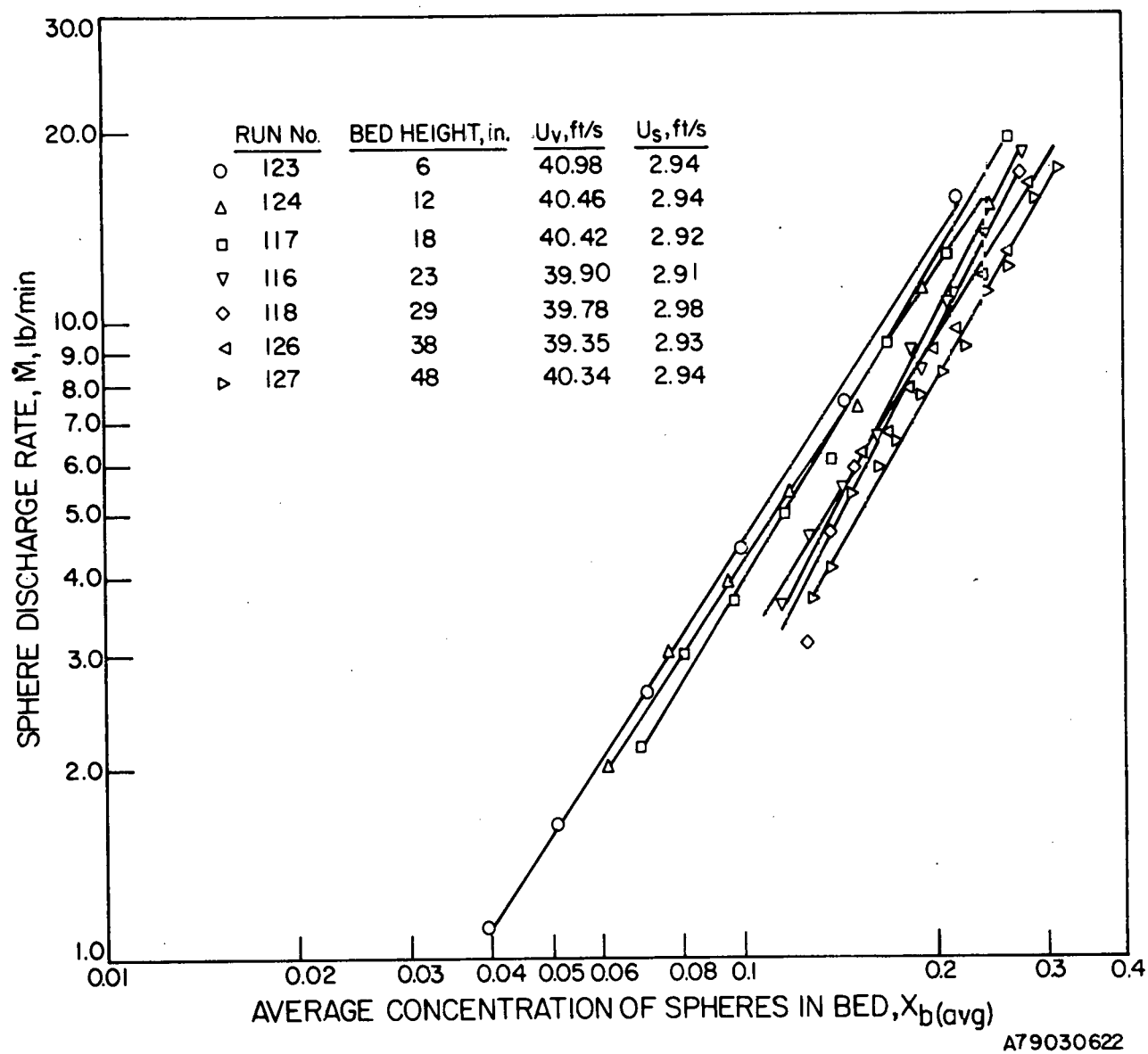


Figure B-31. SPHERE DISCHARGE RATE VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED:
EFFECT OF BED HEIGHT

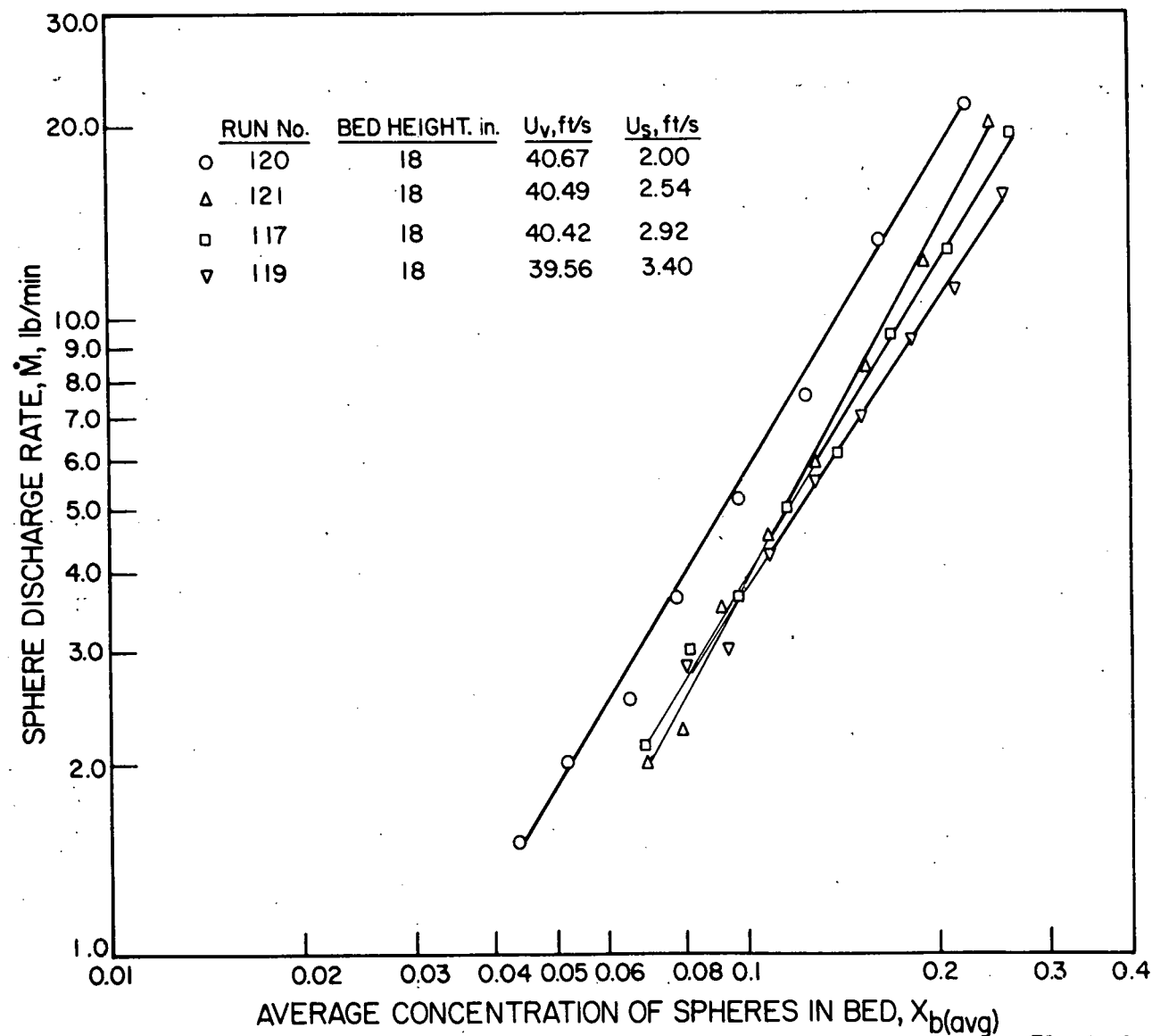
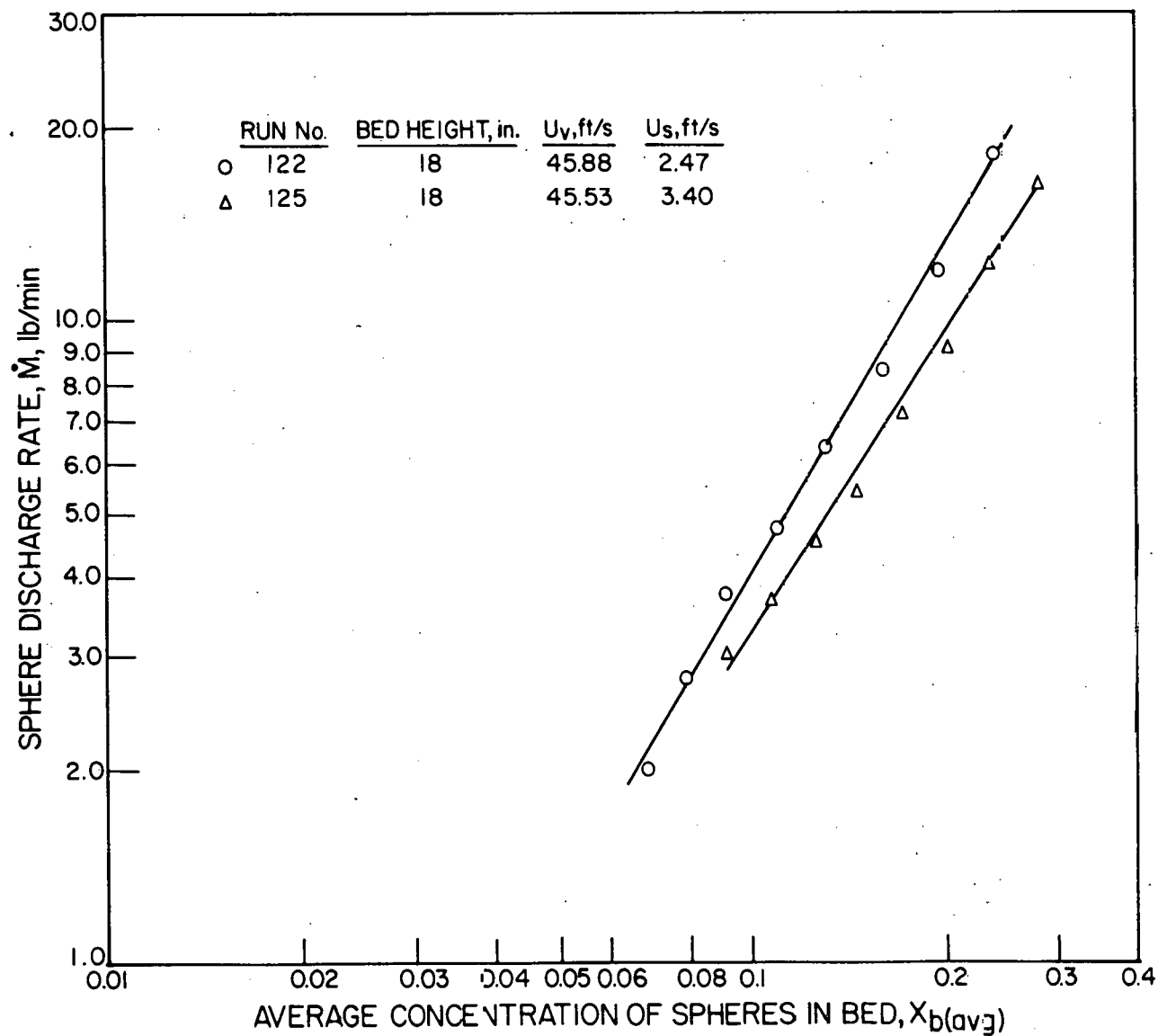
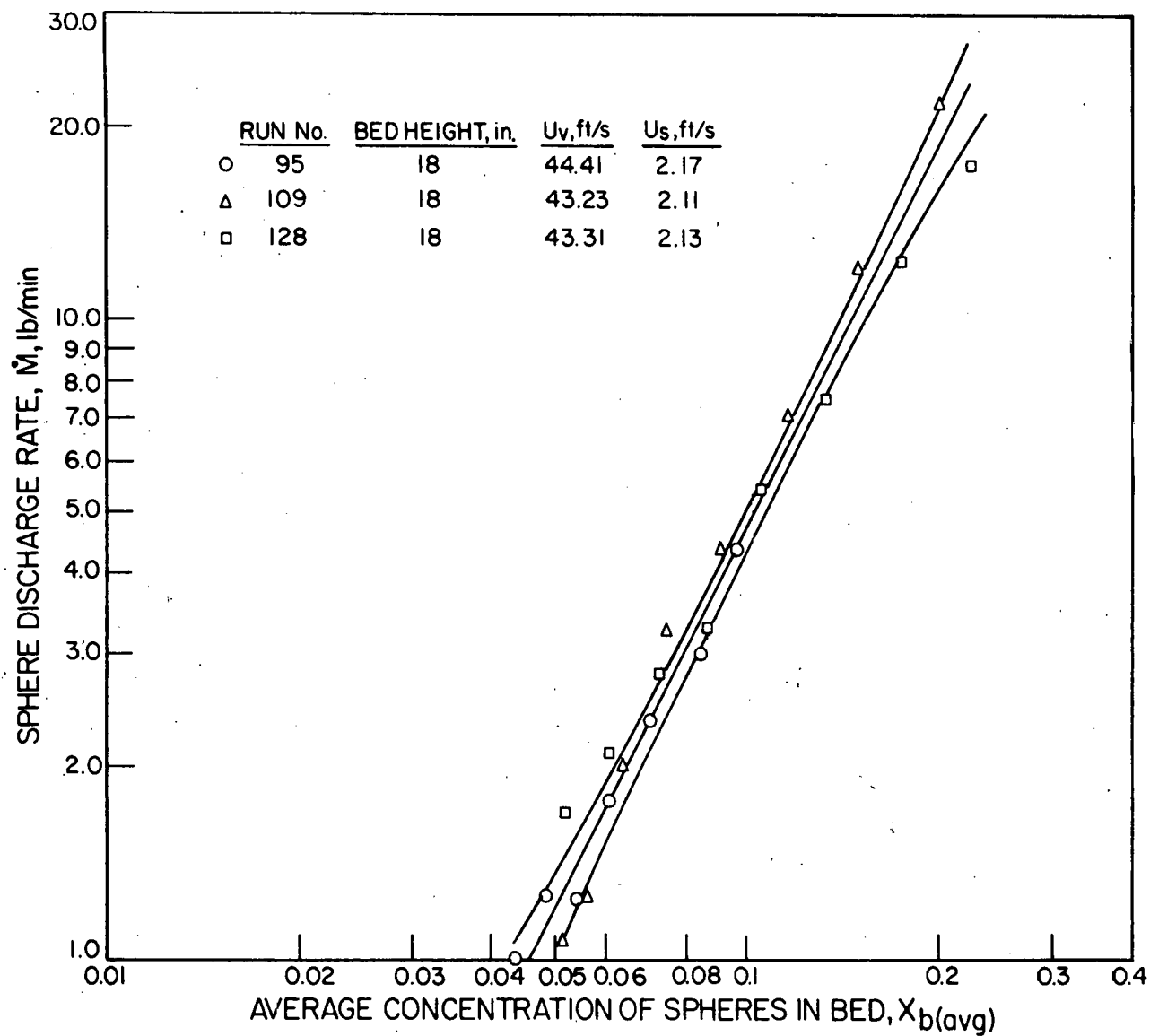


Figure B-32. SPHERE DISCHARGE RATE VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED:
EFFECT OF SUPERFICIAL VELOCITY



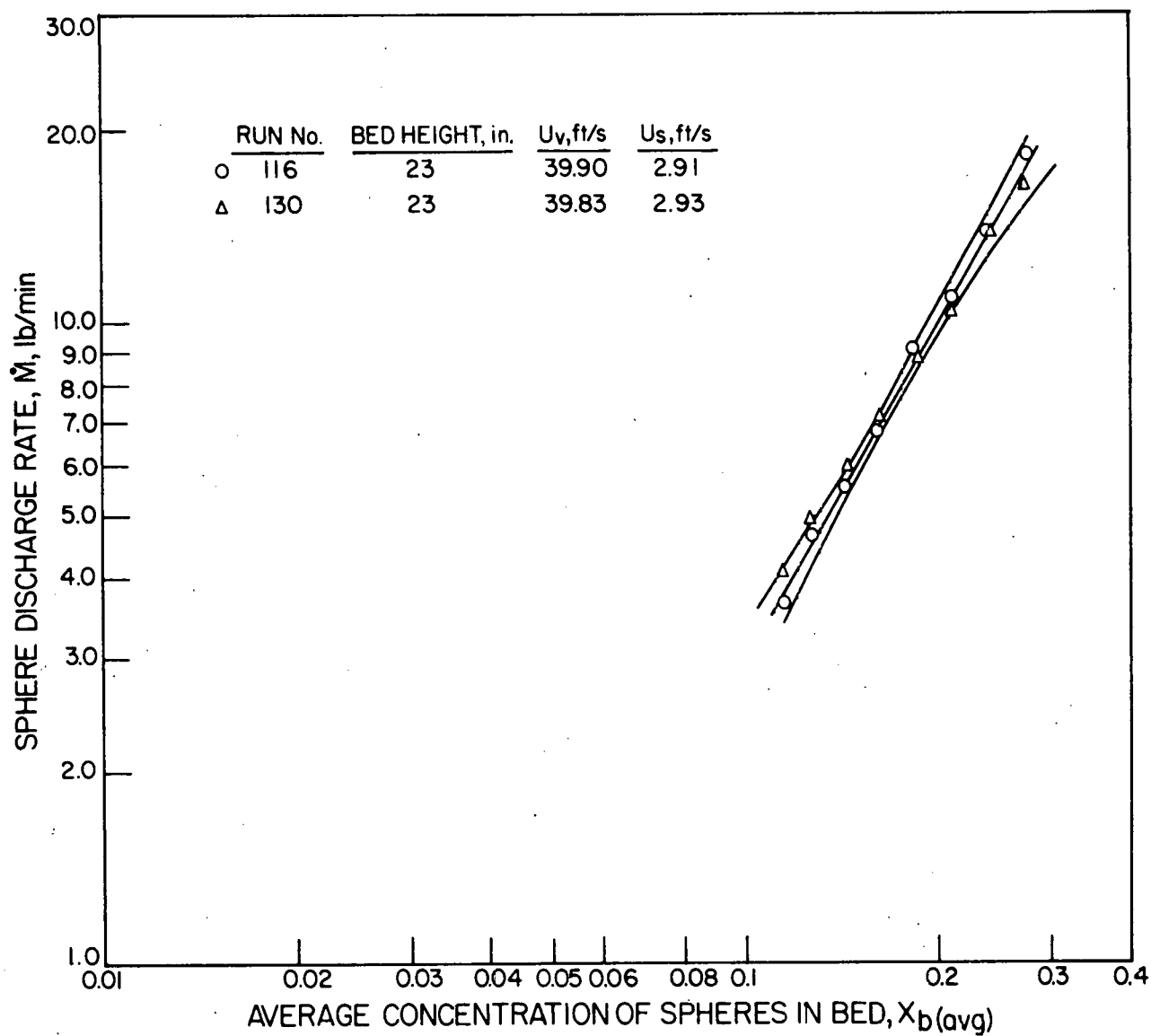
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Figure B-33. SPHERE DISCHARGE RATE VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED: REPLICATE RUNS



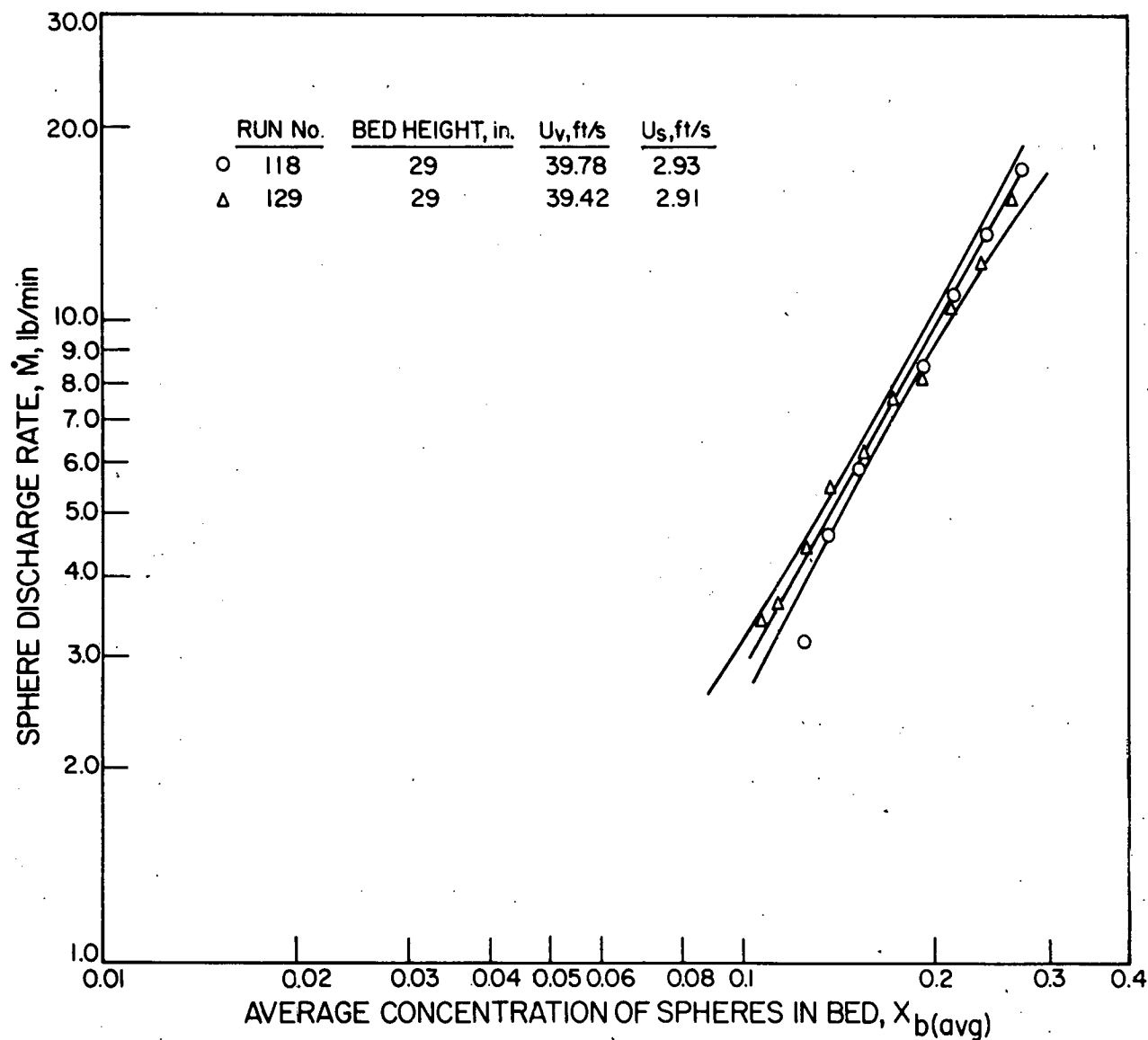
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Figure B-34. SPHERE DISCHARGE RATE VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED:
REPLICATE RUNS



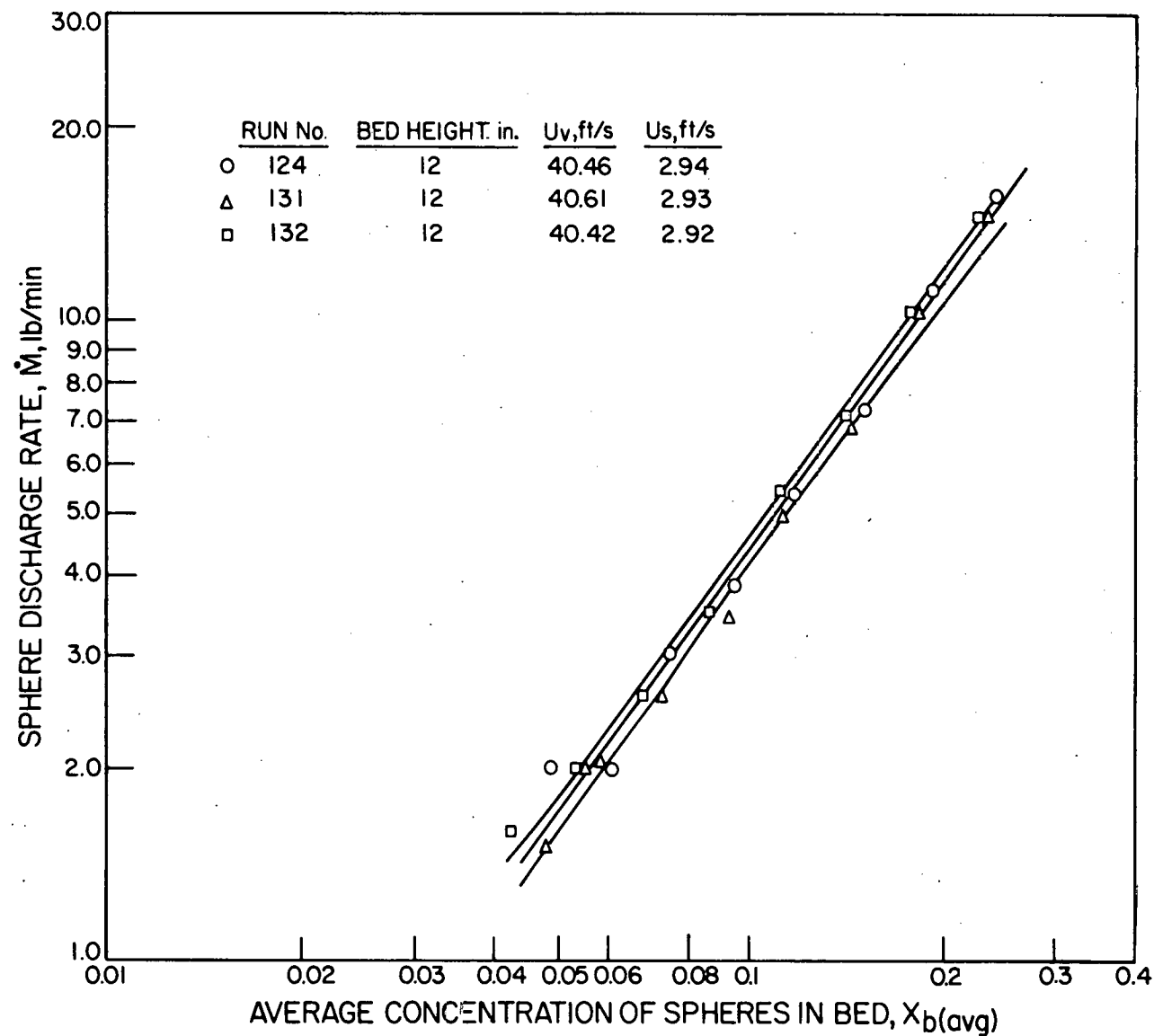
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Figure B-35. SPHERE DISCHARGE RATE VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED:
REPLICATE RUNS



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Figure B-36. SPHERE DISCHARGE RATE VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED:
REPLICATE RUNS



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Figure B-37. SPHERE DISCHARGE RATE VERSUS AVERAGE CONCENTRATION OF SPHERES IN BED:
REPLICATE RUNS

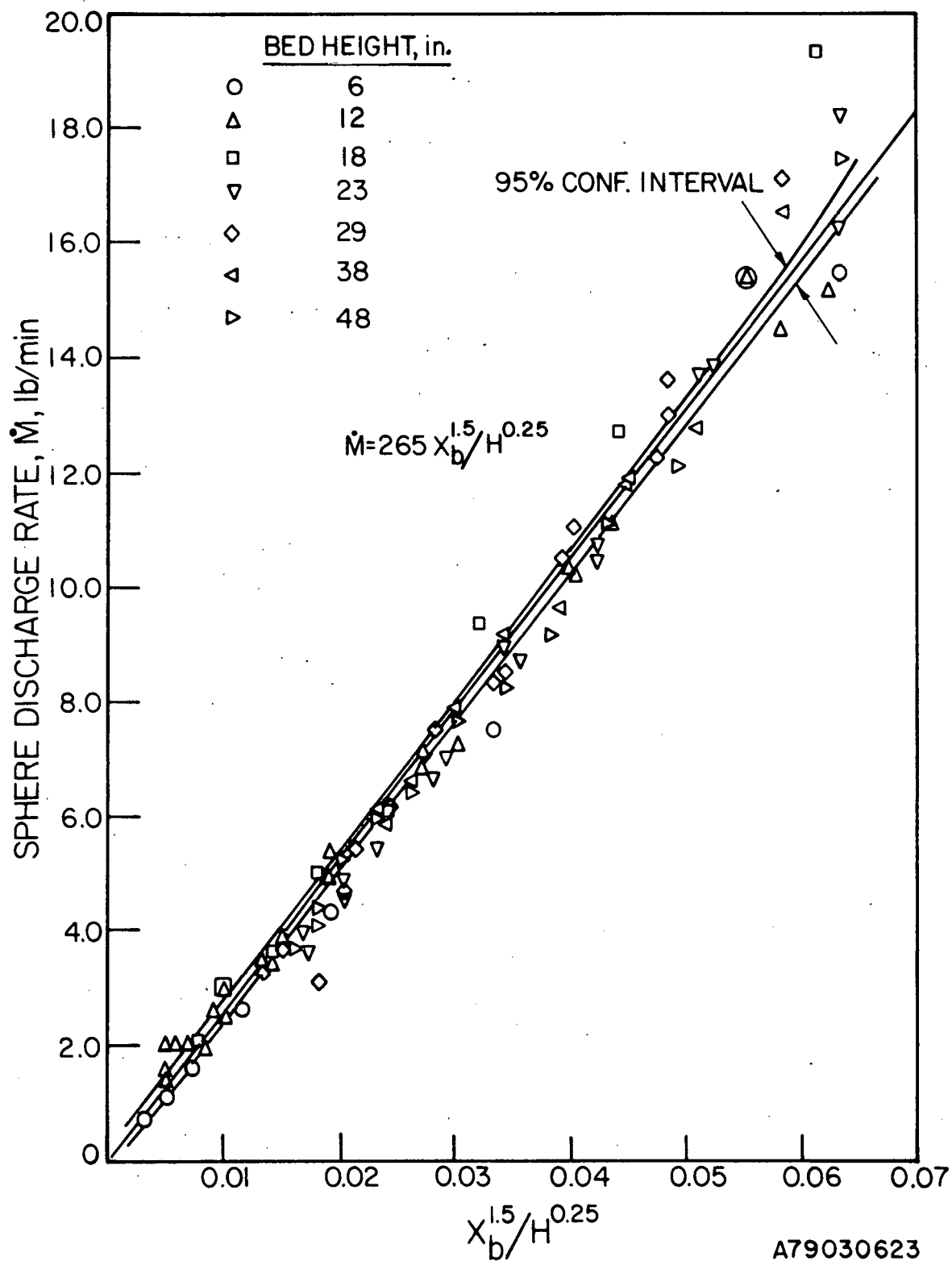


Figure B-38. RATIO OF AVERAGE SPHERE CONCENTRATION IN BED TO BED HEIGHT VERSUS SPHERE DISCHARGE RATE: EFFECT OF BED HEIGHT

$$\dot{M} = 265 X_b^{1.5} / H^{0.25} \quad (B-3)$$

where -

$\dot{M}$ = sphere discharge rate, lb/min

X_b = average concentration of spheres in bed

H = fluidized bed height, in.

The correlation coefficient of the regression fit in Equation B-2 can be further improved by incorporating the effect of the additional variables venturi velocity and grid velocity (flow to grid, ft³/s, divided by total bed area, ft²), respectively. The overall superficial velocity, U_s , presented in Table B-10 is related to the above variables by Equation B-3.

$$U_s = 0.150 U_v + 0.127 U_g \quad (B-4)$$

The correlation developed using the above additional variable is -

$$\dot{M} = 732 \frac{X_b^{1.55}(\text{avg})}{(H^{0.28})(U_s^{0.84})} \quad (B-5)$$

The correlation coefficient is in excess of 0.99 for the above regression. The values of U_s which were employed in the tests leading to the above equation were from 2.0 to 3.40 ft/s.

Examination of duplicated runs indicates that run data can be repeated within 5% accuracy.

Table B-10. SUMMARY OF COLD-FLOW MODEL RUN CONDITIONS.
(Runs 116 Through 132)

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Run No. *	Bed Height (H), in.	6 x 2 inch Configuration **	U_v , ft/s	U_g , ft/s	Weight Fraction of Spheres in Bed, $X_b(\text{avg})$	Weight Fraction of Spheres in Discharge X_d
116	23	8/8 (H)	39.90	2.91	0.1156-0.2785	0.9730-1.000
117	18	8/8 (H)	40.42	2.92	0.0692-0.2615	0.9600-1.000
118	29	8/8 (H)	39.78	2.93	0.1262-0.2731	0.9615-1.000
119	18	8/8 (H)	39.56	3.40	0.0802-0.2553	0.9888-1.000
120	18	8/8 (H)	40.67	2.00	0.0435-0.2217	0.9524-1.000
121	18	8/8 (H)	40.49	2.54	0.0694-0.2406	0.9756-1.000
122	18	8/8 (H)	45.88	2.47	0.0689-0.2423	0.9808-1.000
123	6	8/8 (H)	40.98	2.94	0.0246-0.2193	0.9459-1.000
124	12	8/8 (H)	40.46	2.94		
125	18	8/8 (H)	45.53	3.40	0.0917-0.2812	0.9600-1.000
126	38	8/8 (H)	39.35	2.93	0.1543-0.2866	0.9876-0.9946
127	48	8/8 (H)	40.34	2.94	0.1292-0.3146	0.9907-1.000
128	18	8/8 (H)	43.31	2.13	0.0521-0.2265	0.9934-1.000
129	29	8/8 (H)	39.42	2.91	0.1033-0.2646	0.9778-1.000
130	23	8/8 (H)	39.83	2.93	0.1152-0.2781	0.9849-1.000
131	12	8/8 (H)	40.61	2.93	0.0487-0.2354	0.9700-1.000
132	12	8/8 (H)	40.42	2.92	0.0441-0.2356	0.9700-1.000

* The following parameters were common to the tests mentioned above:
 sphere diameter = 0.152 inch; sphere density = 157 lb/ft³;
 bed particle diameter = 0.0254 inch; bed particle density = 72.0 lb/ft³;
 bed bulk density = 36.0 lbs/ft³.

** The first number indicates the diameter of the half section below the throat; the second number indicates the pipe diameter below the venturi; the letter H indicates half-section pipe below the venturi.

COAL, COKE, AND CHAR TESTS

Runs 133 Through 141

A series of tests (Runs 133 through 141) were performed at various conditions on the half-section cold model with a 6-inch x 2-inch venturi configuration with coke breeze (0 to 1/4 inch) and coal agglomerates (-10+20 mesh) from Run 133 of the AAG pilot plant. All these runs (except Runs 137, 138, and 141) used a 7-foot 9-inch long, 8-inch diameter half-section pipe below the venturi. Table B-11 summarizes the range of variables studied in the above runs.

Table B-11. SUMMARY OF COLD-FLOW MODEL RUN CONDITIONS
(Runs 133 Through 141)

Run No.*	Bed Material Used	Agglomerates Used	Venturi Velocity U_v , ft/s	Superficial Velocity, U_s , ft/s
133	Coke Breeze Feed, 1/4x0 in.	Coal Agglomerates, -10+20	43.0	2.16
134	Coke Breeze Feed, -10 mesh	Coal Agglomerates, -10+20	39.9	2.10
135	Coke Breeze Feed, -10 mesh	Coal Agglomerates, -10+20	48.1	2.10
136	Coke Breeze Feed, -10 mesh	Coal Agglomerates, -10+20	37.4	2.10
137	Final Bed (C-7) From Run 131	As Present in Bed Material	29.7	2.95
138	Final Bed (C-7) From Run 131	As Present in Bed Material	35.0	2.95
139	Coke Breeze Feed, -14 mesh	Coal Agglomerates, -10+20	37.7	2.00
140	Coke Breeze Feed, -14 mesh	Coal Agglomerates, -10+20	35.9	3.08
141	Final Bed (C-7), Run 152-55	Coal Agglomerates, -10+20	17.2	2.71

* The following parameters were common to all the tests mentioned above:

Agglomerates Particle Density = 140.0 lb/ft³

Agglomerates Bulk Density = 58.9 lb/ft³

Coke Particle Density = 78.0 lb/ft³

Coke Bulk Density = 44.5 lb/ft³.

Objective

The main objective of these runs was to study the discharge behavior of coal agglomerates from fluidized beds of coke or char.

Equipment Setup

All the test runs used the same experimental setup. (See Figure B-30.)

Operational Results and Discussion

Run 133 used coal agglomerates in the -10+20 mesh size range from Run 133 of the AAG pilot plant and coke breeze feed of 1/4 x 0-inch size. This run was primarily performed to find out the acceptable range of venturi velocity at which we could obtain reasonably high venturi discharge with satisfactory classification. This run was conducted in two parts: In Part 1, only coke breeze (1/4 x 0-inch size) was used and the venturi velocity was varied from 31.0 to 52.0 ft/s, maintaining superficial velocity constant at 2.90 ft/s. The venturi discharges at these two extreme conditions amounted to 43.0 and 0.5 pounds, respectively, over a period of 2 minutes. In Part 2, 40 pounds of coal agglomerates (-10+20 mesh) were added to the system and certain conditions in the range of those used in Part 1 were repeated. Because the classifications obtained were unsatisfactory, we decided to screen the coke breeze to -10 mesh to hopefully improve the classification of agglomerates.

Runs 134 and 136 were conducted with coal agglomerates (-10+20 mesh) and coke breeze (-10 mesh) at venturi velocities varying from 37.4 to 48.1 ft/s, but at a constant superficial bed velocity at 2.1 ft/s. Results of discharge and classification for Run 136 are shown in Figures B-39 and B-40. As shown in Figure B-39 classification was not complete at these conditions.

As shown by the horizontal distance of the data points from the 45 degree line, positive classification was obtained for all conditions of the run, however. It is interesting to note that a 50% agglomerate concentration in the discharge, for example, with a char bed ash with an average composition of 30% would yield a 65% ash-content discharge. Most of the data points in Figure B-40 indicate that an even higher ash content in the discharge would be possible.

An analysis of terminal velocity of coke and coal agglomerates and actual velocity available in the classification zone (area below venturi throat) showed that even when the system was operated at 48.1 ft/s, the velocity available in the classification zone (27 ft/s) was very close to the terminal velocity of coke particles of 10 mesh size. Table B-12 shows the calculated values of the terminal velocity of coke and coal

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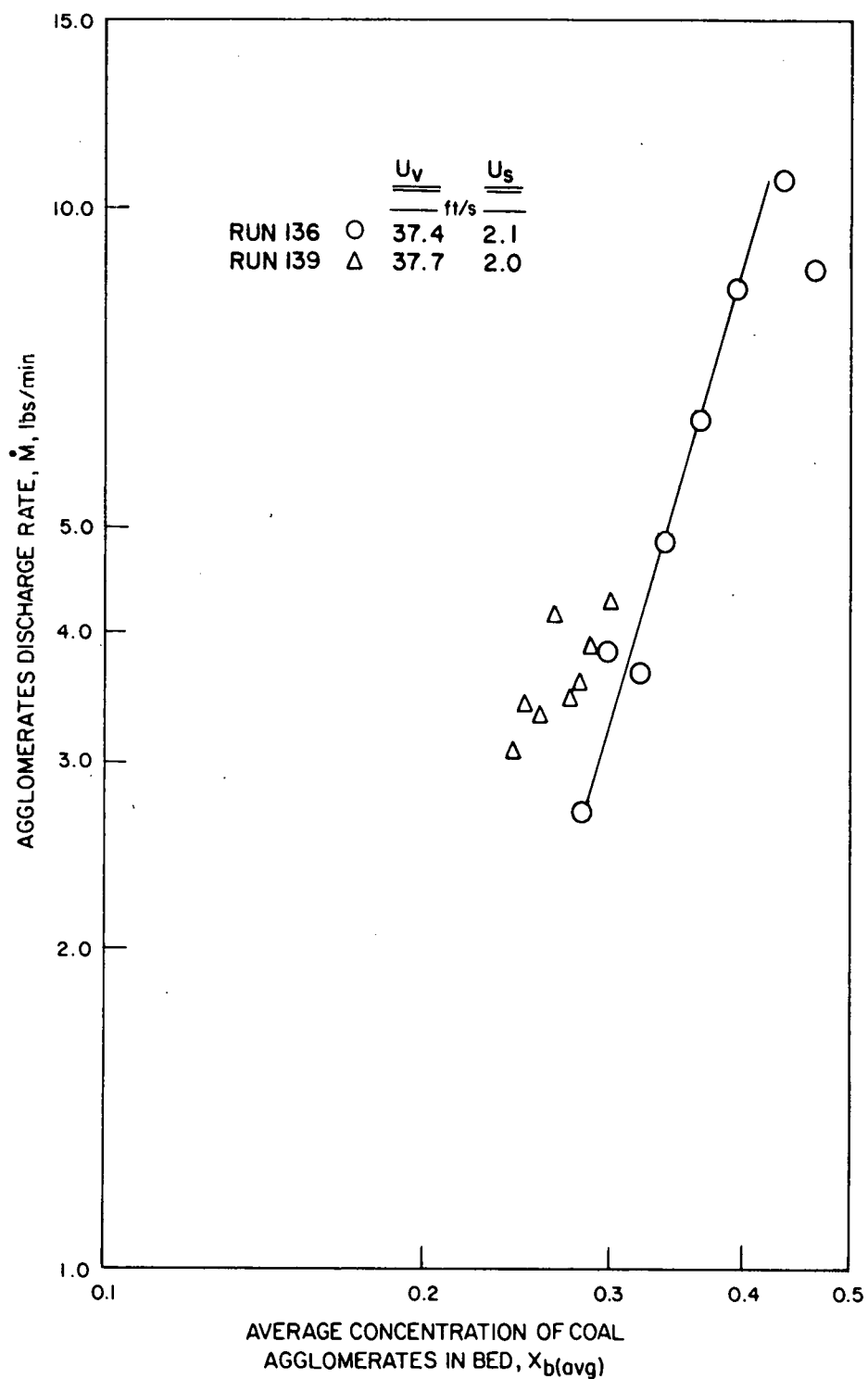


Figure B-39. COAL AGGLOMERATE DISCHARGE RATE, $\dot{M}$,
VS. AVERAGE CONCENTRATION OF AGGLOMERATES
IN BED, $X_{b(avg)}$: RUN 136 AND 139

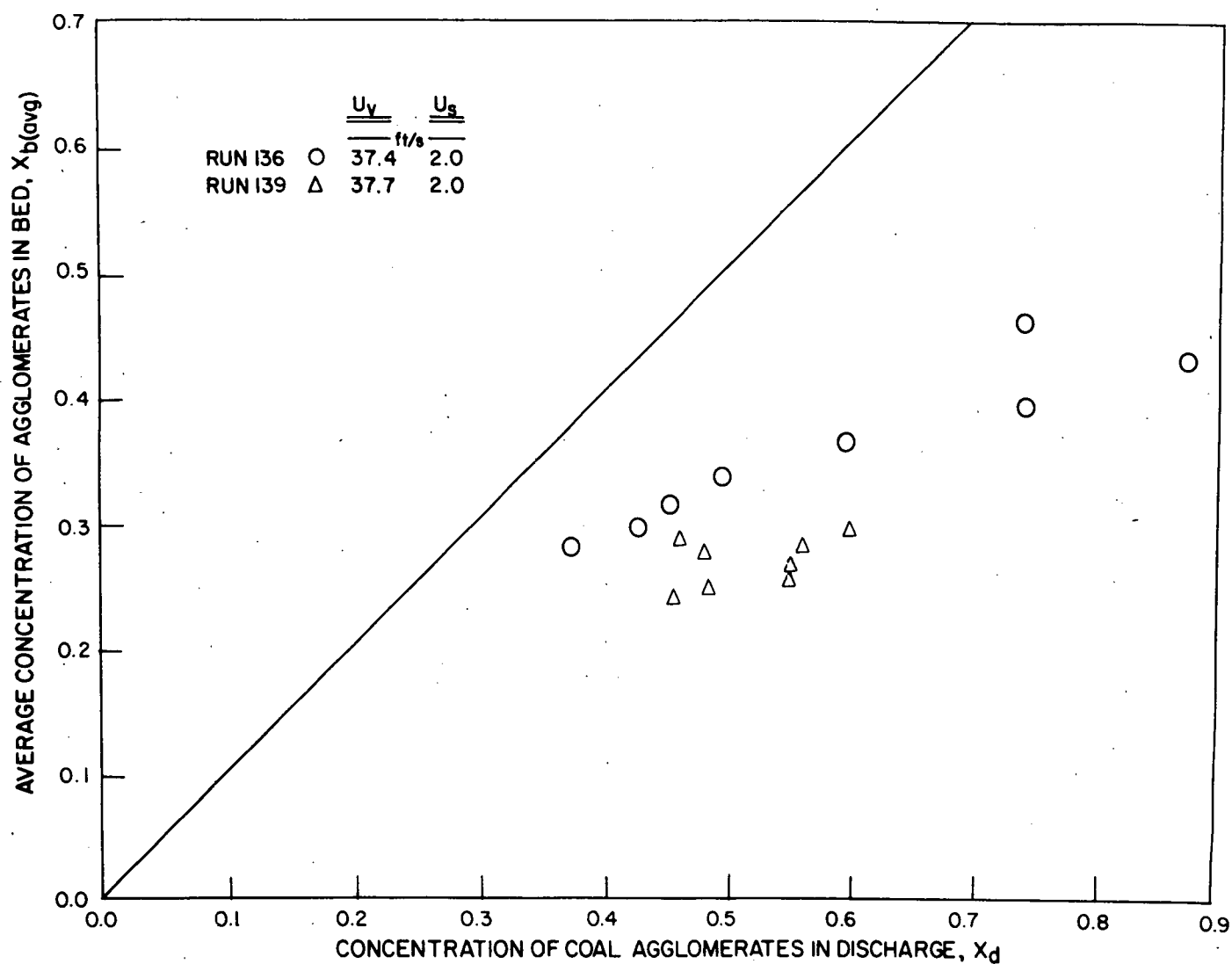


Figure B-40. CONCENTRATION OF AGGLOMERATES IN BED VS. CONCENTRATION OF AGGLOMERATES IN DISCHARGE (Runs 136 and 139)

agglomerate particles. This examination of terminal velocities led us to conclude that -14 mesh coke breeze should produce a more acceptable classification in the discharge. Runs 139 and 140 were conducted to test this hypothesis.

Table B-12. TERMINAL VELOCITY OF COKE AND COAL AGGLOMERATES

Material	Nominal Particle Size d_p , in.	Reynold's Number	Terminal Velocity* U_t , ft/s
Coke Breeze	0.25	>500	48.12
	0.20		43.04
	0.15		37.27
	0.10		30.43
	0.08		27.22
	0.07		25.46
	0.06		23.57
	0.05	<500	20.71
	0.04		16.56
	0.03		12.42
	0.02		8.28
Coal Agglomerates	0.0787 (10 mesh)	>500	36.19
	0.0331 (20 mesh)	<500	20.25

* Calculated from —

$$U_t = \left[\frac{3.1g(\rho_s - \rho_g)d_p}{g} \right]^{1/2} \quad \text{for } 500 < R_e < 200,000$$

$$U_t = \left[\frac{0.0178 (\rho_s - \rho_g)^2 g^2}{g^\mu} \right]^{1/3} \cdot d_p \quad \text{for } 0.4 < R_e < 500$$

where —

$$g = 32.2 \text{ ft/s}^2$$

$$\rho_s = \text{particle density of coal agglomerates/coke} = 140.0/78.0 \text{ lb/ft}^3$$

$$\rho_g = \text{density of fluidizing air at system temperature} = 0.07 \text{ lb/ft}^3$$

$$\mu = \text{viscosity of air at system temperature} = 1.3 \times 10^{-5} \text{ lb/ft-s.}$$

Because classification was not complete, it was difficult to separate coal agglomerates from total discharge in the size range of -10+20 mesh. We applied an ash-balance technique to determine the amount of coal agglomerates in the discharge. Ash

determinations were performed by the pilot plant analytical staff. Based upon ash balance, the equation is -

$$A = D \cdot \frac{X_D - X_C}{X_A - X_C} \quad (B-3)$$

where -

D = total discharge, pounds

A = weight of coal agglomerates in discharge, pounds

X_D = concentration of ash in discharge

X_A = concentration of ash in agglomerates

X_C = concentration of ash in coke breeze.

Subsequent tests used the ash-separation technique rather than the ash-analysis method above.

Run 137 was conducted at a venturi velocity of 29.7 ft/s and a superficial velocity of 2.95 ft/s. The material used in this test and in Run 138 was final bed from Run 131. This material was part of our inventory of various materials to be used in the tests. It was felt that if the material did represent an AAG pilot plant bed, tests using this material would be more nearly indicative of what could be expected under actual conditions. We achieved approximately 28 lb/min discharge (our highest to date in any run), but there was no classification as the ash level in the discharge was close to that in the bed. A second run with this material at a venturi velocity of 35 ft/s and a superficial velocity of 2.95 ft/s yielded discharges ranging from 7 to 10 lb/min, but again there was only nominal classification.

It was concluded that this material was not suitable for our tests. Close examination verified the absence of well-formed agglomerates in this material, and the measured ash level, approximately 25%, also substantiated this fact. No further tests were conducted with this material.

Previous tests using coal agglomerates with +1/4 x 0-inch and -10 mesh coke breeze feed material for the bed material did not produce acceptable classification of agglomerates and coke. It was suggested above that improvements would likely occur if coke was screened to -14 mesh. Coke feed material was screened to -14 mesh and Runs 139 and 140 were conducted. The venturi velocity in Run 139 was 37.7 ft/s, and the superficial velocity was 2.0 ft/s. The discharge rate for Run 139 is shown plotted against average concentration of coal agglomerates in the bed in Figure B-39. The results for

Run 136 are also shown in this figure. The discharge rate is higher for the same bed concentration in Run 139, but by an amount believed to be of a similar order of magnitude to the uncertainty in the data. The classification results in Figure B-40, also compared to Run 136, show an increasing degree of classification in the current run.

Run 140 was conducted at a lower venturi velocity than Run 139 (35.9 ft/s) and a higher superficial velocity (3.08 ft/s). Poor classification was achieved in this run, with some samples of ash discharge having slightly lower concentration of agglomerates than the bed average. Discharge rates were markedly higher than in Run 139, even though the velocity in the venturi was only 4% less than that in Run 139. The bed agglomerate concentrations were higher in Run 140, however, averaging 39%.

Run 141 was conducted to simulate the concentration and venturi velocity conditions employed in Run HP-4, Set No. 2 in the pressure cold model. The conditions of these runs are compared in Table B-13. The materials collected as final bed and samples from Run HP-4, performed first, were recombined, except for samples sent to the lab for ash-agglomerate fraction analysis, and half the quantity used was employed in Run 141. The essential differences between the run conditions are the pressure and initial bed agglomerate concentrations. If the effect of bed agglomerate concentrations is similar to what was previously observed for coal runs, then the difference cannot explain the observed differences in discharge rates.

Table B-13. COMPARISON OF HALF-SECTIONAL ATMOSPHERIC COLD-MODEL RUN CONDITIONS (Run 141) WITH PRESSURE COLD-MODEL RUN CONDITIONS (Run HP-4, Set No. 2)

Run No.	Pressure, psig	Temperature, °F	Venturi Velocity, ft/s	Superficial Velocity, ft/s	Agglomerate Discharge Rate, lb/min	Initial Bed Agglom. Concentrations
141	~1	140	17.2	2.71	39-43	~0.63
HP-4, Set No. 2	45	90	17.1- 19.3	2.53- 2.87	3.4-6.3	~0.54

Conclusions

Classification of coal agglomerates from coke can be achieved in the 6 x 2-inch venturi in the cold model provided that discharge pipe velocity is maintained high enough, i.e., terminal velocity of the bed material used is lower than discharge pipe velocity. There appears to be a relationship of the form —

$$\dot{M} = \beta X_b^n \quad (B-4)$$

where —

$\dot{M}$ = discharge rate of coal agglomerates

X_b = average concentration of agglomerates in bed

n = constant; slope of plot of $\log \dot{M}$ vs. $\log X_b$.

as shown in Figure B-38. This relation is the same form as that obtained with the acrylic-alumina system.

Preliminary conclusions as to the effect of pressure on discharge rate are that increasing pressure results in lowered discharge rate, all else being equal. Fundamental differences between semicircular units and fully circular units may account for the observed results, however.

Discharge rates as high as 43 lb/min (2580 lb/hr) of coal agglomerates were obtained in the cold model. These rates exceeded any obtained in the AAG Pilot Plant (only 200-300 lb/hr of ash are fed in this unit) and, as such, the cold model is enabling the investigation of conditions otherwise unobtainable in other equipment. These studies will be useful for detailed design of the venturi and gasifier, although much further work needs to be performed.

Ash Separation

Introduction

For purposes of determining the weight fraction of agglomerates in a composite sample of coal, char, coke breeze, or other carbonaceous material, a means for separating the agglomerates from the other carbonaceous material is required. We developed a rapid, inexpensive fluidized-bed technique for this purpose. This technique was employed in other activities of technical support in this program.

Discussion

A fluidized-bed technique was developed to separate agglomerates from other carbonaceous material. Initial tests were conducted in a 4-inch-diameter column. Subsequent work was performed on a 1-inch-diameter column, and the results in either case were acceptable separation of agglomerates and char. The technique separates these materials on the basis of their density. A sample of material is fluidized below the complete fluidization velocity. A vacuum nozzle is inserted just above the upper level of the fluidized bed. Materials vacuumed off in this fashion are collected and weighed. The material that is ultimately not fluidized and not removed is agglomerates or large shale particles. The latter has been confirmed both visually and by ash analysis of this material. Visual examination of the light material does not disclose any agglomerates or high-ash-content material.

Conclusion

A viable technique for separating ash agglomerates and bed or other lighter material (which may overlap in size with agglomerates) has been successfully developed and can be used for support of pilot plant and cold model test work where this procedure is necessary to obtain data on agglomerate concentrations.

Agglomerates, Char and Coke Elutriation Velocities

Introduction

This study was undertaken to examine the behavior of the venturi both in the AAG Pilot Plant Reactor and the cold model and was precipitated by the lack of perfect classification of coal agglomerates from char and coke breeze in the cold model.

Summary

In Runs 133 through 140, coke breeze feed was used as a substitute bed material. We examined this material by performing a screen analysis and comparing this with a screen analysis of bed material, the agglomerates of which had been removed by the techniques previously employed and described in this report. (See "Pressure Cold Model" section). The results of this screen analysis are shown in Figure B-41. The coke breeze is seen to be finer than the Run 133 bed material employed; it was felt that differences in the density and shape factor might account for this. Terminal velocities for the coal agglomerates and coke breeze were determined theoretically, and this information was used to determine a cut velocity in the discharge pipe that would allow the -10+20 mesh coal agglomerates to pass through but sweep back the coke breeze. This resulted in improved classification at higher venturi velocities in Run 139.

Four paired samples of ash discharge (C-4) and bed material from Run 133 of the AAG Pilot Plant were obtained. Three pairs of these samples were subjected to the elutriation testing, which produced data on the cumulative amount of material elutriated as a function of air velocity. Each sample was separated into agglomerates and char, and the test was performed on these fractions as well as the composite material. The venturi velocity prevailing in Run 133 was determined, and the results evaluated in light of those conditions.

Operations and Results

A sample of bed material from Run 133 was separated into char and agglomerates. A screen analysis of the char was performed. These results are shown in Figure B-41. The char is coarser than the coke breeze feed. Terminal velocities of the individual cuts of the coke breeze feed were compared to those of the coal agglomerates and are shown in Table B-12.

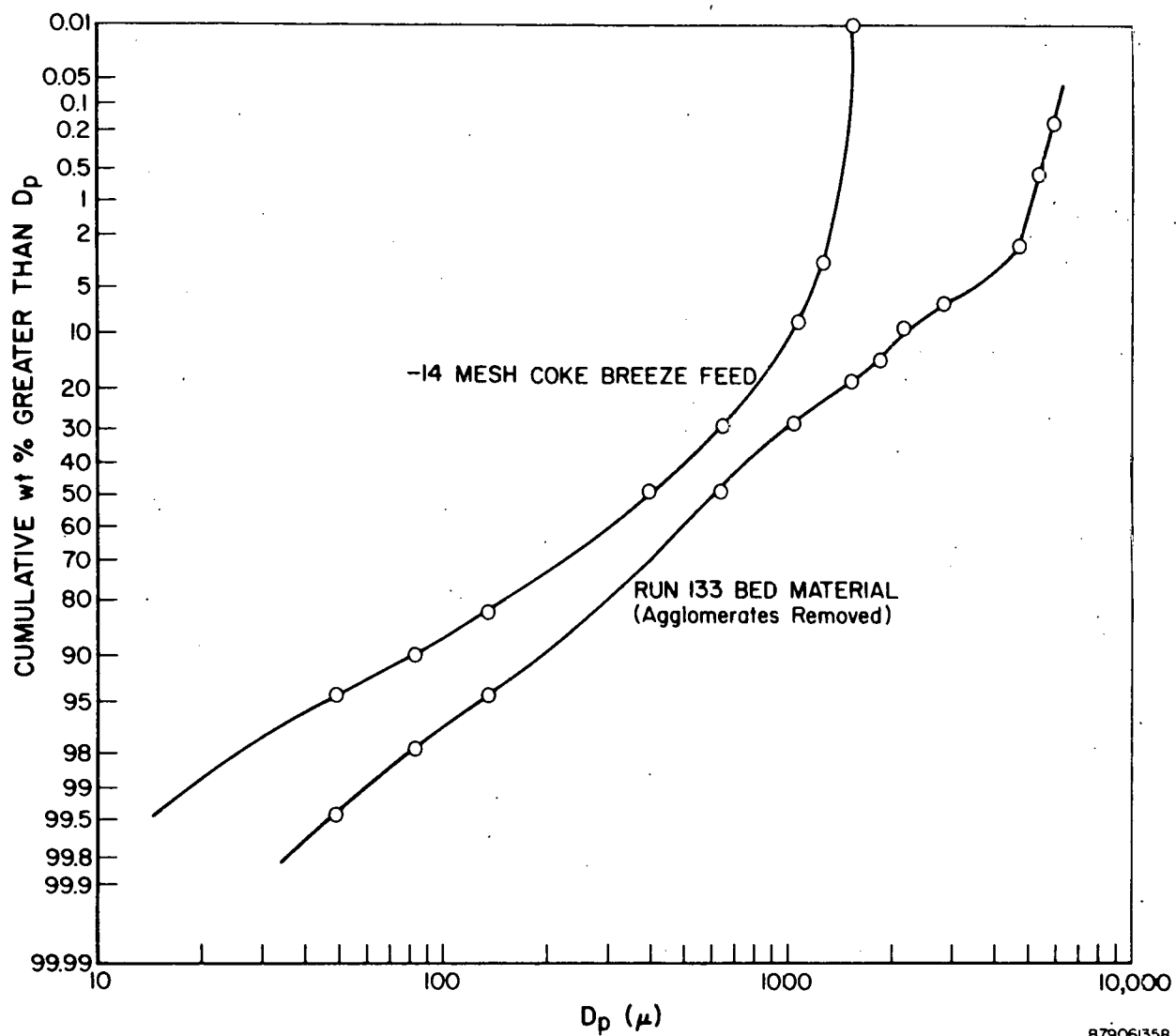


Figure B-41. COMPARISON OF COLD MODEL MATERIAL WITH RUN 133 BED MATERIAL (Char Only)

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Four pairs of samples from the ash-agglomerating gasifier were obtained from the inventory of samples maintained at the pilot plant. The samples used were from Run 133, and the designation of each sample and the time of removal from Run 133 is given in Table B-14.

The venturi velocities for each of these time intervals, or the closest time interval for which reduced data was available, was determined. A portion of each sample was separated into char and agglomerates via the fluidized-bed technique. The char, agglomerates, and composite samples were subjected to an elutriation test. In this test, the material was placed in a 1-1/2-inch-diameter fluidized-bed column, and a low air

Table B-14. LIST OF PAIRED SAMPLES FROM RUN 133 OF BED MATERIAL AND C-4 DISCHARGE FOR CHARACTERIZATION STUDY

Sample No.	Time of Removal	
	Date	Hours
1-A	2/22/79	2230
1-B	2/22/79	2300
2-A	2/23/79	1030
2-B	2/23/79	1100
3-A	2/23/79	1230
3-B	2/23/79	1300
4-A	2/24/79	0530
4-B	2/24/79	0600

velocity was commenced and held for several minutes. Materials entrained overhead were collected with a cyclone, removed, and weighed. The cumulative weight elutriated as the velocity was progressively increased was determined for Samples 1-A, 1-B, 2-A, 2-B, 3-A, and 3-B. Salient data from this analysis is presented in Table B-15.

Conclusions

The coke breeze feed used in cold model tests is finer than the char based on the sample of Run 133 char examined. Differences in particle size cannot, therefore, explain the poor classification obtained in some of the cold model tests. Density differences and shape factor differences remain to account for the observed results. Terminal velocity analysis was used to suggest a cut size (14 mesh), which improved classification in subsequent cold model tests with an appropriately high venturi velocity.

Elutriation tests performed with materials from pilot plant Run 133 led to the following conclusions:

- o Agglomerates that were in fact discharged had terminal velocities substantially in excess of the prevailing discharge pipe velocity.

Table B-15. SUMMARY OF ELUTRIATION TESTS ON BED AND ASH DISCHARGE SAMPLES
FROM RUN 133 (Samples 1A, 1B, 2A, 2B, 3A, and 3B)

	Ash Discharge Sample (1A)	Bed Sample (1B)	Ash Discharge Sample (2A)	Bed Sample (2B)	Bed Discharge Sample (3A)	Ash Sample (3B)
CWF Agglomerates with $U_t > U_a$	0.13-0.18	0.12-0.23	0.15-0.25	0.20-0.22	0.06	0.21
CWF Agglomerates with $U_t > U_p$	0.97-0.98	0.97-0.98	0.96-0.98	0.92-0.94	0.75	0.75
CWF Char with $U_t > U_a$	0	0.03-0.04	0.025	0.02-0.05	0.05	0.03
CWF Char with $U_t > U_p$	0.88-0.91	0.44-0.45	0.48-0.56	0.25-0.31	0.57	0.30
<u>Data from Run 133</u>						
Venturi Annulus Velocity, U_a , ft/s		32-35		34-36		36
Venturi Discharge Pipe Velocity, U_p , ft/s		14-16		15-16		16
Bed Height, ft		3.4-3.5		5.6-4.5		5.5
Bed Agglomerate Concentration, wt %		39.6		27.7		53.4
Ash Discharge Agglomerate Concentration, wt %		84.1		50.6		70.5

* Cumulative weight fraction.

- Additional material in the bed (approximately 20% of the bed) could potentially be discharged against the venturi velocity acting as a terminal velocity gate.
- Only 6% to 25% of the material in the discharge could potentially be discharged against the venturi velocity acting as a terminal velocity gate.
- A substantial fraction of the char that was discharged has a terminal velocity in excess of the discharge pipe velocity.
- Virtually none of the char that was discharged has a terminal velocity in excess of the venturi throat velocity.
- Agglomerate concentrations in the bed over a broad range of from 28% to 53% were maintained in Run 133 without adverse consequences.
- Ash agglomerate concentration in the discharge does not appear to be related to the ash agglomerate concentration in the bed.

The above conclusions are based on somewhat limited data and should be considered tentative. Further work along these lines appears to be useful in understanding the operation of the ash-agglomerating gasifier.

PRESSURE COLD MODEL

Objective

The objective of this portion of the technical support work was to obtain information on the behavior of the venturi discharge device that would be suitable for design of the demonstration-plant gasifier components.

Equipment

A test facility was constructed to accomplish the above objectives. The equipment was designed by IGT and consulting engineers from Salem Engineering Corp. Elements of a facility already under construction at the IGT pilot plant facility to conduct high-pressure cold-model studies for the HYGAS® SNG Process were incorporated to capitalize on accrued delivery schedule advantages and other work in progress. Considerable capital savings were also effected because the purchase of a compressor and structural steel were not required.

The vessel was fabricated by Imperial Steel Tank to IGT specifications. Vessel details are shown in Figure B-42. An instrumentation and control diagram of the system is shown in Figure B-43. The compressor is an ATLAS COPCO Model ZA6, which has a maximum delivery pressure of 200 psig. The cold-model system is being operated on a closed loop with nitrogen gas. Losses of nitrogen from the system during a run are minimal. Flows of nitrogen to the venturi, a central jet, and the grid plenum chamber are metered by orifice plates and differential pressure transmitters and are recorded separately. Additional data recorded are the vessel pressure, the compressor discharge temperature and pressure, the cyclone ΔP , the filter ΔP 's, the fluidized-bed height (ΔP) across 66 inches of bed, and the fluidized-bed density across 12 inches of fluidized bed.

The details of the 6-inch X 2-inch venturi, the grid, and the 1.5-inch Sched. 40 central jet are in Figure B-44. The grid is an inverted 45-degree frustum equipped with 270 1/4-inch hex-head bolts drilled through to 3/8 inch on a pitch of approximately 2.5 inches. A cyclone collects the fines that are elutriated out of the fluidized bed. Additional filters and strainers protect the compressor from harmful contamination.

A system of lockhoppers feeds materials into the bed and discharges material removed from the venturi into collecting containers. Feeding can be conducted at system pressure. A pneumatic conveying system can be used to convey feed and agglomerate material directly into the vessel provided the vessel pressure is kept below 35 psig.

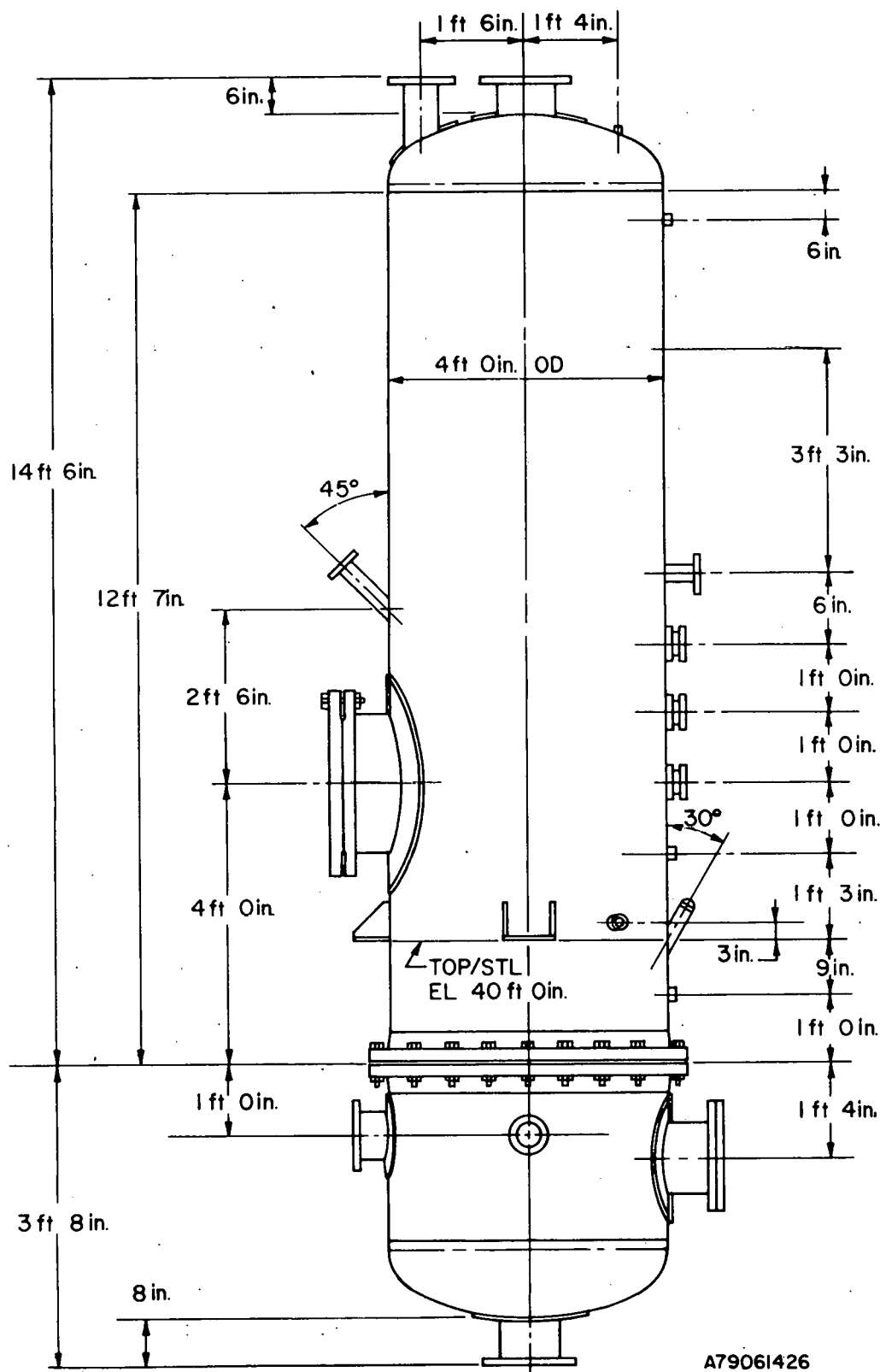
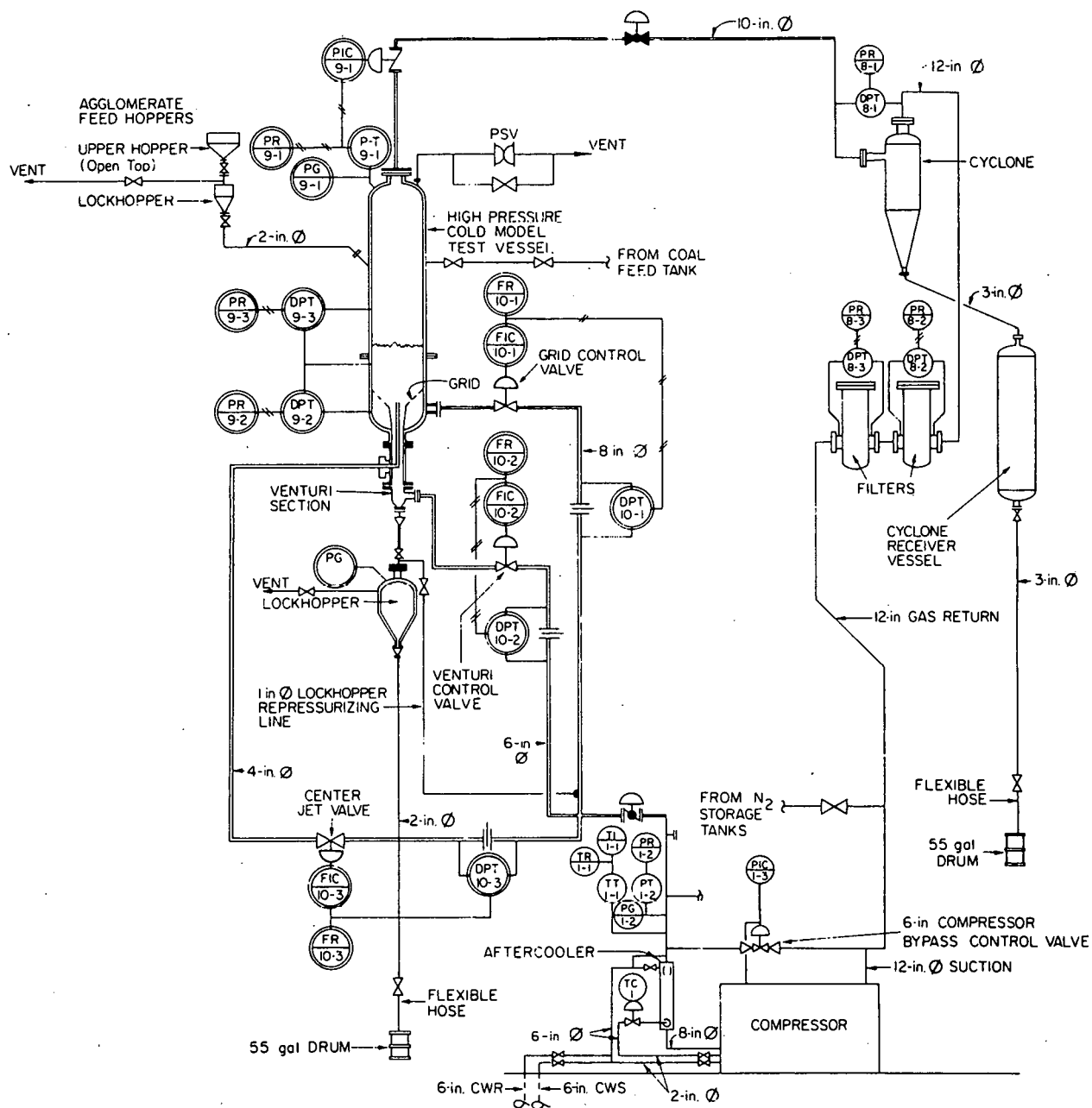


Figure B-42. PRESSURE COLD MODEL VESSEL DETAILS



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Figure B-43. PROCESS FLOW DIAGRAM FOR PRESSURE COLD MODEL

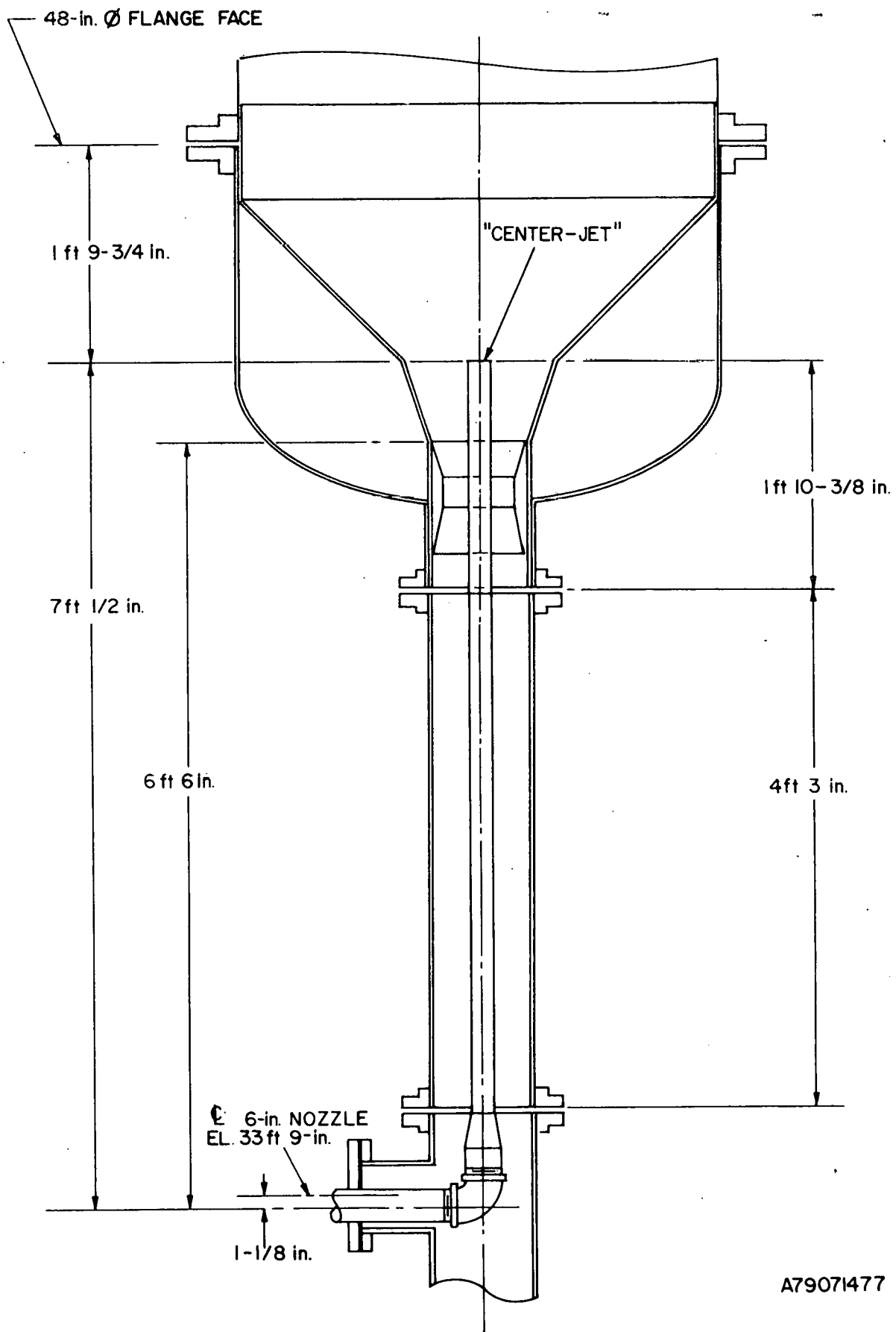


Figure B-44. DETAILS OF VENTURI GRID AND CENTRAL JET, PRESSURE COLD MODEL

Procedures

A detailed operating procedure manual was prepared for the pressure cold-model facility. This manual contains the detailed instrument specifications, calibrations, maintenance and safety instructions and checklists, and step-by-step operating procedures for start-up, run, and shutdown conditions. We will describe here only the operating procedures in a general manner.

Pre-weighed materials are loaded into the feed hopper using a portable conveyor. The compressor is started up, and flows through the cold-model vessel are initiated. A high velocity (greater than 50 ft/s) is maintained in the venturi to prevent the premature discharge of material. The velocity in the grid bolts is kept high enough to prevent weepage of solids into the plenum chamber below the grid. A pressure of approximately 8 psig, the lower limit, is set up in the vessel. An auxiliary flow of nitrogen is established in the conveying line from the feed hopper that enters the vessel at a point 4.5 feet above the grid. Alternatively, materials can be gravity fed from a lockhopper. Whichever method is used, when all materials have been fed, the feed system is isolated from the vessel and pressurization to the desired operating conditions is begun. The desired venturi flow, central jet flow, and grid flow are then set up. Initial material that has fallen into the collecting hopper from the venturi is discharged while material falling out at desired conditions collects above the now closed lockhopper valve. The collecting hopper is repressurized to venturi line pressure and the upper valve is opened to discharge the materials collected in the line above into the collecting hopper. When a set time elapses, generally 3 or 4 minutes, the upper valve is again closed.

The lockhopper is depressurized and emptied into collecting pails for subsequent weighing. The material collected now corresponds to a set time interval, and discharge rates can be determined. This procedure is repeated until an adequate amount of information at the flow and pressure conditions has been obtained. A new setting of flows or pressures can be made to obtain additional data if time permits. The bed conditions will, of course, have changed because of the discharge of material in prior sets. When all data have been obtained, the pressure is reduced, the venturi flow is adjusted to a low level, and the bed is discharged through the lockhopper.

Post-run data analysis consists of determining the concentration of agglomerates in the various discharge samples collected and the weights of all samples and the final bed. Any material found in the cyclone receiver vessel or the plenum chamber is collected and weighed. When these results are obtained, the initial bed weight and the agglomerate concentration is back calculated. Based on the facts and circumstances of

the run, appropriate assumptions as to the time of occurrence of cyclone or weepage losses are made. The flow recorder charts are examined, the orifice ΔP readings are adjusted for such things as pressure and temperature variations, and the velocities are calculated. Bed density and height ΔP readings are converted to calculated heights and densities. The processed data consisting of temperatures and pressures, mass discharge rates, degree of classification, bed agglomerate concentrations, and bed height are then collated and some correlation among the variables is sought.

Experimental Results

Summary

Five runs, HP-0 through HP-4, were conducted on the pressure cold model. Run HP-0 was essentially a debugging operation using acrylic bed material and alumina spheres. Runs HP-1 through HP-4 were conducted with ash agglomerates and the final bed produced in the ash-agglomerating gasifier (AAG) pilot plant in various trials with Western Kentucky coal. These runs were successful in that data were obtained.

Run HP-0

We fed acrylic material with the external feeding system and alumina simulated agglomerates with the lockhopper feeding system. The acrylic material was all elutriated before any discharge rate data could be obtained. Although a 50 ft/s velocity was maintained, agglomerates fell through the venturi. This was attributed to the absence of a bed, enabling the agglomerates to pick up considerable momentum and discharge immediately. As a result, the discharge information was not felt to be meaningful. A flow in the central jet was maintained in this test to check the instrumentation. Subsequently, the central jet was plugged off and the following tests did not employ a jet flow.

Run HP-1

Three separate settings were attained in this run. Set No. 1 was conducted at a pressure of 20 psig and a venturi velocity of 32 ft/s. Sets 2 and 3 were conducted at a pressure of 15-26 psig and 10-12 psig, respectively, and at venturi velocities of 21-27 ft/s and 22-23 ft/s, respectively. Fifteen measurements of discharge rate were obtained, which varied from 0.07 to 13 lb/min. A lengthy delay in the test program was caused by a compressor failure during this run. Further details of Run HP-1 and the other test runs are shown in Table B-16.

Table B-16. SUMMARY DATA - PRESSURE COLD-FLOW MODEL RUNS HP-1 THROUGH HP-4

Run No.	Temperature, °F	Pressure, psig	Venturi Velocity U_v , ft/s	Superficial Velocity U_s , ft/s	Maximum Agglomerate Conc. in Bed	Ash Agglomerate Concentration in Discharge (min-max)	Agglomerate Discharge Rate, lb/min
HP-1	90					NA*	
Set No. 1		11.2 - 19.5	32.1 - 38.7	1.8 - 2.2	0.62		0.07 - 1.75
Set No. 2		15.0 - 26.3	21.3 - 27.1	2.0 - 2.1	0.61		2.90 - 5.80
Set No. 3		10.1 - 12.4	22.0 - 23.2	2.1 - 2.2	0.58		11.20 - 13.00
HP-2	77	27.0 - 31.1	23.0 - 25.0	2.5 - 2.7	0.73	NA*	2.25 - 6.50
HP-3	90	30	15.4 - 19.2	2.2 - 2.6	0.81	0.68 - 0.95	3.21 - 15.08
HP-4	91						
Set No. 1		43.4 - 45.6	22.2 - 23.8	2.7 - 2.8	0.58	0.96 - 0.98	1.1 - 2.0
Set No. 2		43.4 - 51.9	16.6 - 19.3	2.5 - 2.9	0.54	0.71 - 0.92	3.4 - 6.6

* Assumed equal to 1.00 for calculation purposes because ash analysis values of discharge samples were 96% to 98%.

Run HP-2

Materials from AAG Runs No. 151 and 152 were used in this test. A venturi velocity of 23 to 25 ft/s, a superficial velocity of 2.48 to 2.69 ft/s at a pressure of 30 psi, and a temperature of 77°F were maintained during the run. No central jet flow was employed in these tests. Discharge rates of 2.2 to 5.1 lb/min were achieved.

Run HP-3

The same materials used in Run HP-2 were employed in this run except that fines losses in the cyclone were not reused. The superficial velocity in this test ranged from 2.2 to 2.6 ft/s. Venturi velocity was in the range of 15 to 19 ft/s. Discharge rates of 3 to 16 lb/min were attained, with good classification for the most part.

Run HP-4

Final bed and agglomerates used in Runs HP-1 through HP-3 were reused in Run HP-4. The final bed from AAG Run 155 was also fed to make up for char losses to the cyclone in previous runs. Two distinct conditions were maintained during this run. In the first part of the run (Set No. 1), the venturi velocity varied from 22 to 24 ft/s. In the second part of the run (Set No. 2), the venturi velocity was reduced to approximately 17 to 19 ft/s. Discharge rate data were obtained in both sets. In Set No. 1, discharge rates ranged from 1.1 to 2.0 lb/min. In Set No. 2, comparable values were 3.4 to 6.6 lb/min. This run was conducted at a pressure of 45 psig.

Discharge Composition Analyses

Individual samples from Runs HP-1 and HP-2 were analyzed for ash content. The results are presented in Table B-17. As indicated in the table, the discharge consisted almost entirely of ash (agglomerates). In Run HP-3, classification was good although not as high as in Run HP-2 (Table B-18). In our opinion, the lower average venturi velocity explains this difference. Figure B-45 summarizes the classification information we obtained. Points below the 45-degree line represent classification of agglomerates in the venturi. Several points in Run HP-3 represent a situation where the bed is richer than the discharge in agglomerates. Such a situation is obviously undesirable and appears preventable by using higher discharge line velocities.

Most of the discharge samples from Run HP-1 were screen analyzed; the results of this work are presented in Table B-19. Average particle size of the samples shows some degree of variation, but this variation was contained in the interval of -4+20 mesh. For most of the samples, at least 50% of the sample weight was in the size interval of -4+10 mesh.

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Table B-17. CLASSIFICATION OF AGGLOMERATES IN RUNS HP-1 AND HP-2*

<u>Sample No.</u>	<u>Discharge,</u>	<u>Uv,</u>
<u>Run HP-1, Set No. 1</u>	<u>wt %</u>	<u>ft/s</u>
1	97.0	32.1
2	99.5	33.2
3	98.0	34.0
4	81.5	34.7
5	92.0	35.6
6	98.5	36.2
7	98.5	36.8
8	98.0	38.7
Run HP-1, Set No. 2		
1	99.0	21.3
2	97.5	23.6
3	96.5	25.0
4	96.0	26.3
5	97.0	27.1
Run HP-1, Set No. 3		
1	96.5	22.0
2	96.0	23.2
Run HP-2		
1	97.0	23.3
2	97.0	24.6
3	97.5	25.0
4	95.5	24.4
5	94.5	23.4
6	96.0	23.0

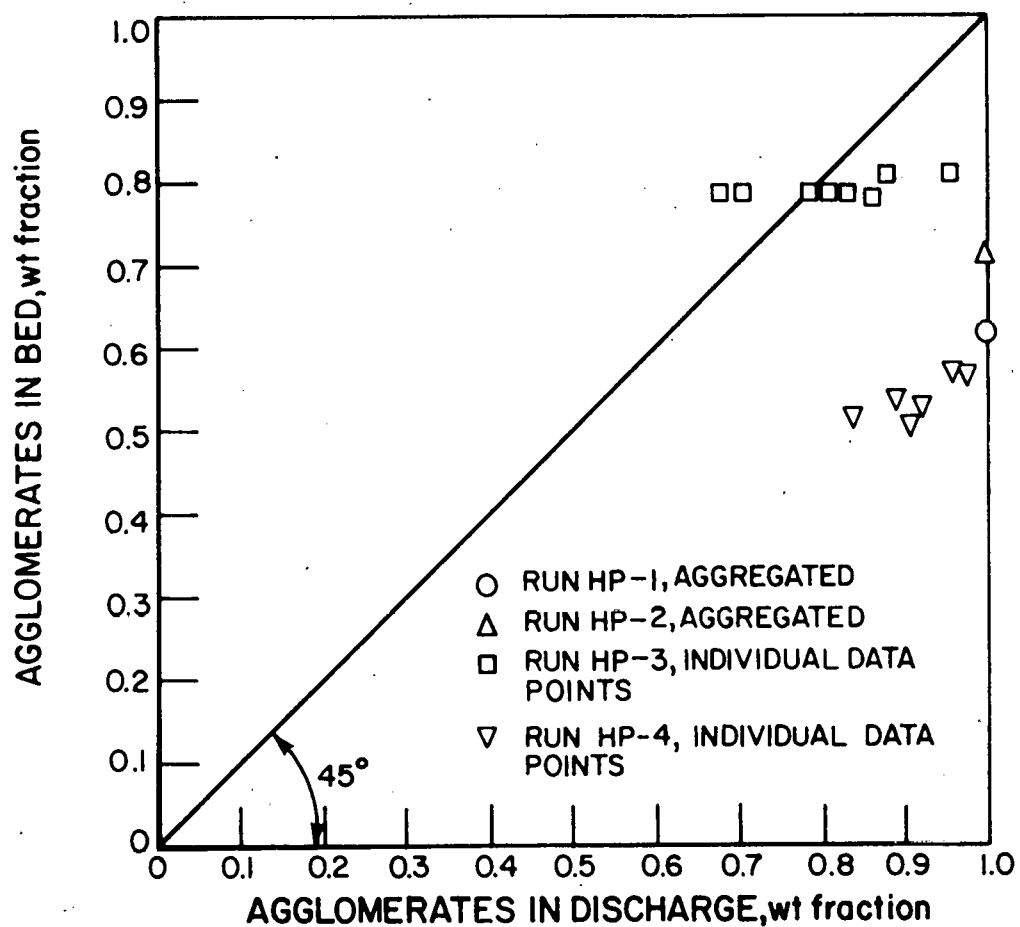
* wt% Ash in Final Bed = 79%

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Table B-18. CLASSIFICATION OF AGGLOMERATES IN RUN HP-3*

<u>Sample No.</u>	<u>Agglomerates in Discharge,</u> <u>wt %</u>	<u>U_v,</u> <u>ft/s</u>
1	95.0	16.2
2	87.5	16.7
3	81.8	18.0
4	84.8	19.2
5	79.5	15.4
6	85.7	16.6
7	70.7	17.3
8	67.6	17.3

*wt % Agglomerates in Final Bed = 79



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Figure B-45. CLASSIFICATION RESULTS FOR RUNS HP-1 THROUGH HP-4

Table B-19. RESULTS OF SCREEN ANALYSIS OF DISCHARGE SAMPLE FROM RUN HP-1

Sample No. USS Sieve No.	Set No. 1				Set No. 2	Set No. 3
	1	2	3	4	3	1
	% Retained					
4	—	—	—	—	—	—
10	0.585	0.766	0.550	0.616	0.568	0.406
20	0.313	0.177	0.441	0.164	0.425	0.561
40	0.018	0.019	0.007	0.085	0.003	0.027
60	0.002	0.011	—	0.056	—	0.004
80	—	0.005	—	0.030	—	0.001
100	—	0.004	—	0.011	—	—
200	—	0.005	—	0.023	—	—
325	—	0.004	—	0.006	—	—
Pan	—	0.001	—	0.004	—	—
Average Particle* Size, in.	0.08064	0.06299	0.08032	0.02997	0.08293	0.06807

* $\bar{d}_p = 1/\sum(x_i/\bar{d}_{p_i})$ x_i = wt fraction between screens of average diameter $\bar{d}_{p_i}$.

Discharge Rates Analyses

The discharge rate data were subjected to some elementary statistical procedures. The mean and standard deviation of all data in Runs HP-1 through HP-4 were calculated. (See Table B-20.) The standard deviation indicates the range about the mean within which approximately 68 percent of the measurements would be expected to lie if the data were random and had come from a normal distribution. The standard deviation is a measure of the variability of the data. The variability of our data increased when the discharge rates obtained increased. The standard deviation normalized by the mean discharge rate decreased progressively in the test series, possibly indicating improvements and refinements in operating the system and in taking data (that is, on learning curve effects). If all the data are assumed to be part of the same population, and the same statistical calculations are performed, a much higher standard deviation than in any of the individual sets results. This indicates that the conditions maintained in the sets are responsible for the differences in the data, not merely random fluctuations.

The mean discharge rates in Runs HP-3 and HP-4, Set No. 2 may be compared because the range of venturi velocity at which the data were obtained is similar. We see that a 6.5 lb/min difference exists between the mean discharge rate data. The pressure difference, which is considered to account for this variation, was 15.4 psi. Thus, the sensitivity of discharge rate to pressure is tentatively -0.42 lb/min/psi.

The differences in mean discharge rate between Runs HP-2 and HP-3 are attributed primarily to differences in venturi velocities. The differences in the mean discharge rates in Sets No. 1 and No. 2 of Run HP-4 are also caused by venturi velocity differences. In both cases, higher venturi velocity results in lower discharge: a rate of 1.2 (lb/min)/(ft/s) in the former case and a rate of 0.6 (lb/min)/(ft/s) in the latter case. The differences in these two sensitivities may be caused by pressure effects, but the imprecision of the data precludes a definitive assessment at this time.

Conclusions and Recommendations for Future Work

A venturi-type device can be used to classify agglomerates from a fluidized bed of agglomerates and char. At venturi velocities of 15 to 20 ft/s, good classification was achieved, and at higher venturi velocity, 20 to 25 ft/s, classification was excellent. Discharge rate data obtained to date are inconclusive. The large degree of variability observed in these data makes it difficult to definitively assess the effect of pressure, velocities, agglomerate concentrations, and bed height. Tentatively, it appears that increasing venturi velocity or pressure lessens the discharge rate.

Table B-20. STATISTICAL PARAMETERS OF DISCHARGE RATE DATA

Run No.	Sample Mean Discharge Rate, $\bar{M}$ lb/min	Sample Standard Deviation, σ_M lb/min	Sample Size, n
HP-1			
Set No. 1	0.43	0.58	8
Set No. 2	3.90	1.19	4
Set No. 3	12.1	—	2
HP-2	4.29	1.48	6
HP-3	12.31	5.67	8
HP-4			
Set No. 1	1.65	0.30	10
Set No. 2	5.78	1.18	8*

*One data point known to be in error was not included.

Additional testing will be required to obtain satisfactory information for design purposes. In addition, it does not appear that a single 6-inch X 2-inch venturi would be capable of discharging the required approximately 200 lb/min of ash from the gasifier at a velocity in the range of 15 to 25 ft/s. Lower velocities should be tried in the 6-inch X 2-inch venturi, and a larger venturi should be installed to determine the effect of venturi size on discharge. The effect of jet velocity, higher pressures, and higher superficial velocities (in the range of 3 to 4 ft/s) should be examined to better simulate the demonstration plant conditions.

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APPENDIX C. Bench-Scale Test Work and Ash Chemistry Studies

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BENCH-SCALE ASH AGGLOMERATION TESTS

Introduction

We conducted a series of experiments to determine the regimes of initiation of ash particle growth and sinter-free operation of Western Kentucky coal and coke in a fluidized-bed environment. The solids particles were fluidized at high temperature in the absence of oxygen to study the effect of variables under an unchanging environment. The reactor that was used in these experiments is shown in Figure C-1. The reactor is a 2.0-inch Schedule 40 316 SS pipe equipped with pipe caps that have tubing connections to serve as gas inlet and outlet ports and the thermocouple entrance port. A sintered metal porous plate serves as the gas distributor. A bed of about 250-300 grams of material is fluidized with N_2 at velocities of 2-5 ft/s. Other details of the experimental setup are shown in Figure C-2. The temperature of the bed was maintained by heating elements in a furnace that encloses the reactor. The N_2 was preheated in a separate furnace and/or in a coil encircling the reactor within the furnace, and the temperature of the N_2 as it enters the reactor is measured. Pressure taps immersed in the fluidized-bed zone are used to record a signal representing the density of the fluidized material. Any drastic decrease in this signal indicates the occurrence of a defluidized state in the reactor and the conditions pertaining to such a state are noted. The experiments were conducted at various superficial velocities and temperatures with several size ranges of material and with materials of differing ash contents. The material used was bed material from the AAG pilot plant. A sufficient quantity of a homogenous nature having the required characteristics was secured beforehand to ensure that all the tests could be conducted with that batch of material. The material removed from the bench-scale gasifier at the conclusion of the test is examined by visual means and screen analysis for any particle growth or agglomerate or sinter formation.

Preliminary and Feasibility Tests

Summary

Two gasification tests and a high-temperature fluidization test were conducted in the bench-scale test reactor with char that was obtained from the operation of the ash agglomerating gasifier (AAG Run 122) with coal. The objective of these tests was to determine the agglomeration characteristics of the char at the test reactor operation conditions shown in Table C-1. The feasibility of using gasifying environments other than N_2 , i.e., steam and oxygen, was also studied in these initial tests.

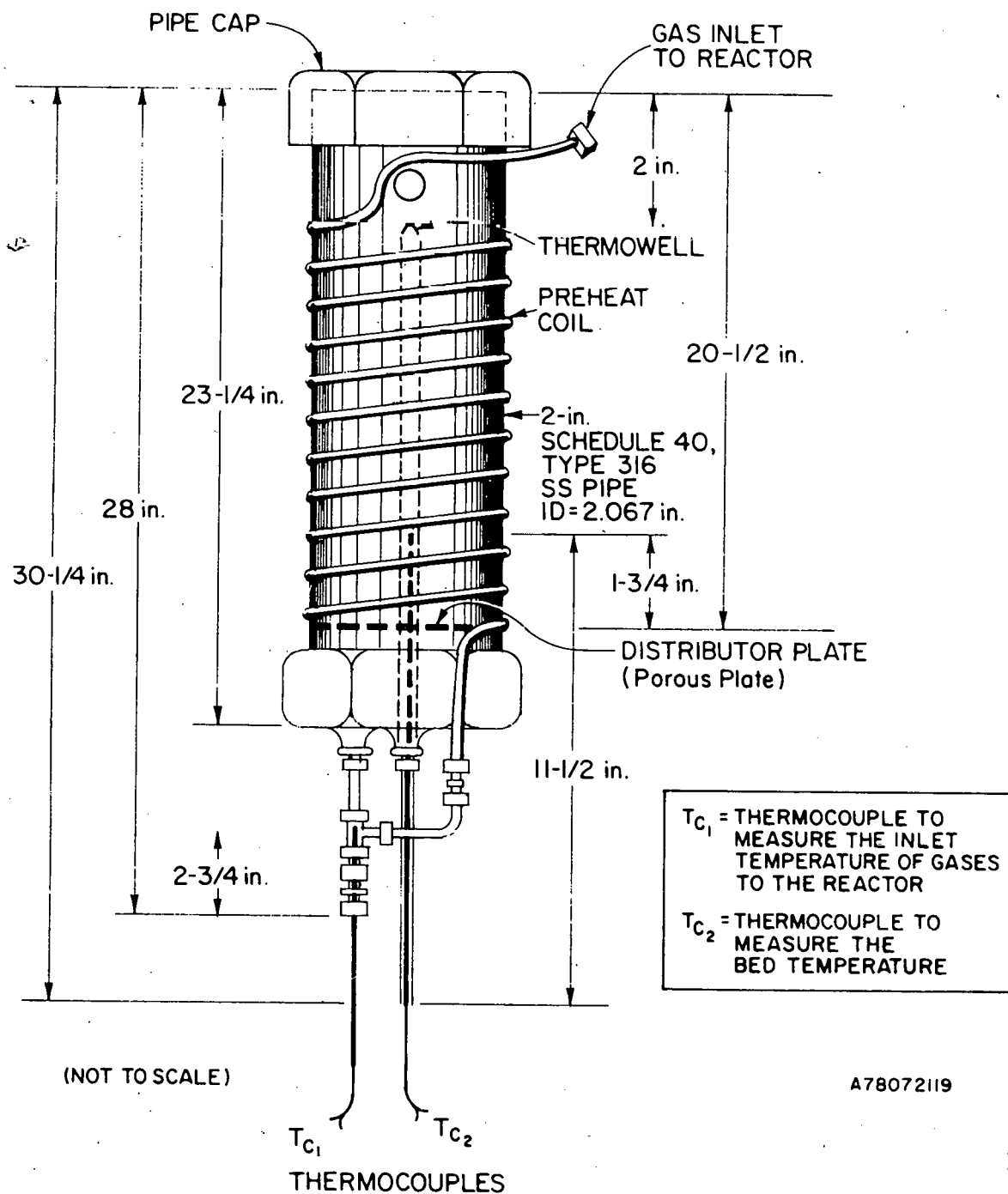
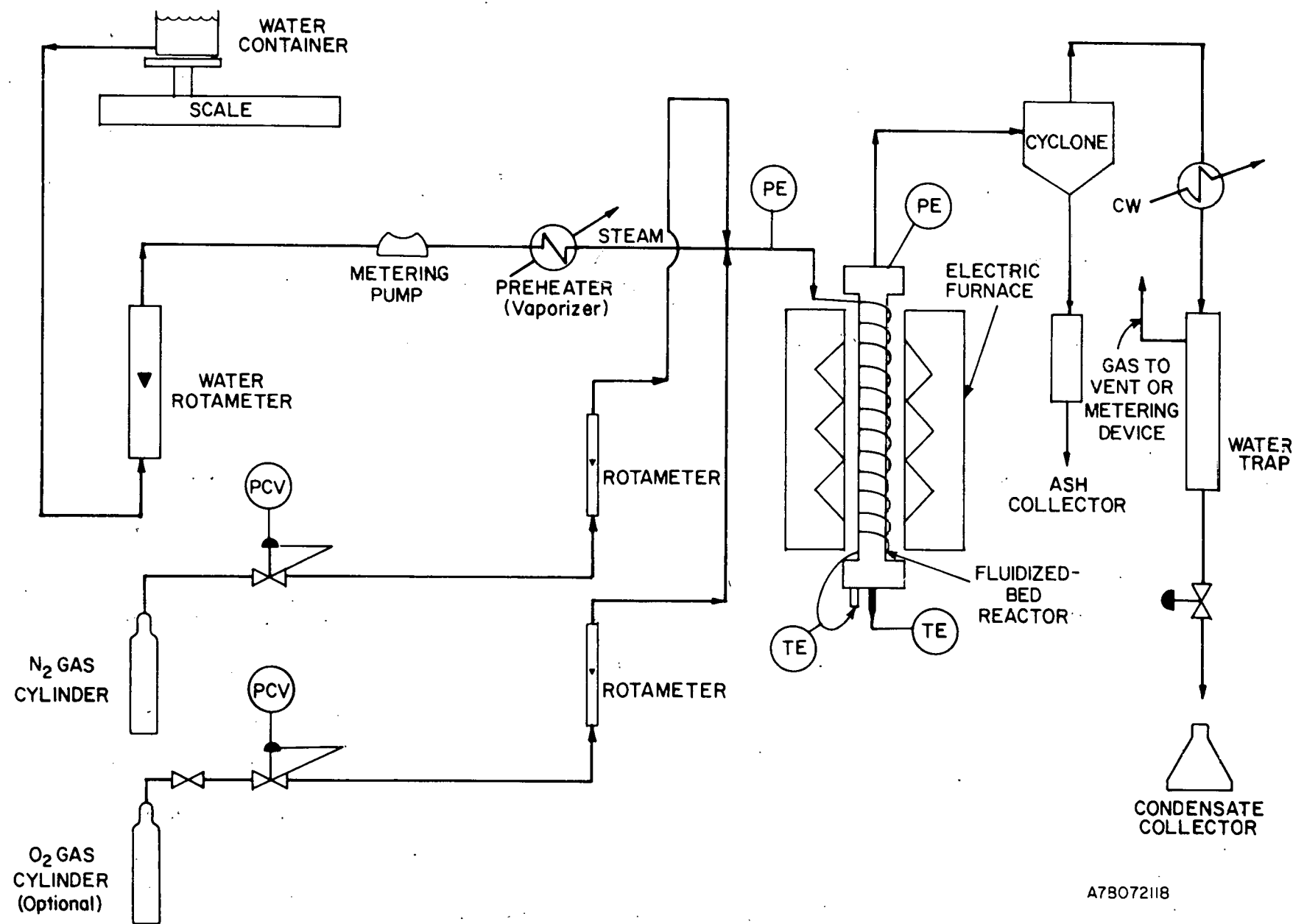


Figure C-1. BENCH-SCALE TEST REACTOR



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Figure C-2. DIAGRAM OF THE EXPERIMENTAL SET-UP

Table C-1. OPERATING CONDITIONS FOR AAG-CHAR GASIFICATION TESTS^a

Run No.	AGX-1	AGX-2	AGX-3
Nominal Feed Size, U.S.S.	-3-1/2+325	-3-1/2+325	-3-1/2+325
Reactor Temperature, °F ^b	1710	1812	1830
Reactor Pressure, psig	0-2	0-2	0-2
Weight of Feed, g	269	267.4	268.4
Bed Aspect Ratio, L/D	approx. 3	approx. 3	approx. 3
Nitrogen Rate, SCF/hr	20.5	9.5	7.9
Steam, lb/hr	--	0.397	0.401
Oxygen, SCF/hr	--	--	1.9
Superficial Gas Velocity, ft/s ^c	0.91-1.03	0.83-0.94	0.86-0.97
Steady-State Operating Period, min ^d	60	60	7

^a Conducted in a 2-inch Schedule 40 316 S.S. reactor fitted with a porous-plate distributor.

^b Average temperature during the steady-state period as based on the readings of the thermocouple, which is located 1-3/4 inches above the distributor plate.

^c N_2 + Steam + Oxygen (CF/S feed gas at reactor pressure and temperature)/cross-sectional area of reactor.

^d Started when the reactor attained the operating temperature.

Objective

The objective of the preliminary and feasibility tests was to determine if we could find conditions of operation in the bench-scale test setup which would enable us to fluidize a sample of coke, char, or coal at higher temperature and to identify needed equipment modifications and data requirements.

Operation and Results

A typical test was conducted as follows: Approximately 269 grams of the composite char (-3-1/2+325 U.S.S.) was placed in the reactor. Heat-up of the reactor and bed was accomplished by means of a furnace equipped with independently controlled electrical resistance heaters. During the reactor heat-up the char bed was kept fluidized by nitrogen. When the bed reached the operating temperature, the reacting gases (preheated steam for Run 1 and oxygen at 70°F for Run 3) were introduced into the bed. Each test except Run 3 was conducted for 1 hour. Because of sintering of the bed material, Run 3 was shut down after 7 minutes. After shutdown, screen analyses of the residue and cyclone fines were conducted for each test.

A summary of the operating results for the three tests is shown in Tables C-2 and C-3. Screen analyses of the feeds, residues, and the fines collected at the bottom of the cyclone were performed. For AGX Run 1 and 2, a comparison was based on the weight of +40 mesh material of feed and residue. Run 3 was not included in the analysis because of sintering of the char.

Conclusions From Preliminary and Feasibility Tests

The conclusions from these tests can be summarized as follows:

- There was no char agglomeration at the reactor operating conditions in AGX Runs 1 and 2.
- Char sintering was observed near the gas distributor in AGX Run 3, which was conducted with N₂, steam, and oxygen at 1830°F.

Western Kentucky Coal Tests - Non-Devolatilized Coal Tests, Fluidization Tests

Summary

These tests, Runs AGX-4 through AGX-10B were conducted with the available materials from Runs 122, 123, and 124 of the pilot plant reactor. We conducted these tests to determine the agglomeration and sintering behavior of these materials and, in addition, conducted atmospheric fluidization tests in a Plexiglas column to determine suitable operating conditions (minimum fluidization velocity, slugging velocity, bed density) for the high-temperature tests. The devolatilization of the coal in Runs AGX-5 through AGX-9 (40% ash coal) rendered quantitative data analysis impossible. As a

Table C-2. SUMMARY OF OPERATING RESULTS OF AAG-CHAR GASIFICATION TESTS

Run No.	AGX-1	AGX-2	AGX-3
Nominal Feed Size, U.S.S.	.-3-1/2+325	-3-1/2+325	-3-1/2+325
Test Period, min	60	60	7
Weight of Feed, g	269.0	267.4	268.4
Weight of Residual Char, g	224.4	205.8	193.8
Fines Collected, g	28.4	15.5	40.5
Losses, g	16.2	--	
Carbon Conversion, % **	--	50.0	22.7
Condensed Water, g	--	61.0	106.0

* Conducted with N₂, therefore no carbon conversion and no condensate.

** Calculated from the carbon balance.

wt of carbon gasified = wt of carbon in feed - (wt of carbon in residue + wt of carbon in fines)

$$\% \text{ carbon conversion} = \frac{(\text{wt of carbon gasified})}{(\text{wt of carbon in feed})} \times 100$$

Table C-3. CARBON AND ASH CONTENTS OF FEEDS, RESIDUES, AND FINES OF AAG GASIFICATION TESTS IN THE 2-INCH-DIAMETER REACTOR

Run No.	AGX-1			AGX-2			AGX-3		
	Feed	Residue	Fines	Feed	Residue	Fines	Feed	Residue	Fines
				wt%					
Carbon	--	--	--	34	16	80.7	36	23.5	72.0
Ash**	--	--	--	66	84	19.3	64	76.5	28.0

* This run was conducted with N₂, therefore no samples were given for analysis.

** Samples were analyzed at the pilot plant by the quick ash procedure.

result, a procedure of conducting a preliminary pyrolysis in the reactor by retention of the feed material at 1700°F was instituted. The sample is subsequently removed from the reactor, weighed, and screened and then used in an agglomerating test (e.g., Run AGX-10B). Tests with high ash (low-VM) char were successful in producing agglomeration as measured by increase of particle sizes in the +35 mesh range. Average particle size decreased, however, as a result of breakage and attrition, in some tests in which agglomeration was nonetheless felt to have occurred. Conditions leading to agglomeration were determined. Petrographic and microscopic examination of the materials produced in the agglomerating tests were made and are discussed below.

The current experimental equipment was then modified. This enabled operation at temperatures as high as 2100°F. The initial phase of the testing was conducted with N₂ only. Various combinations of temperatures and superficial velocities were used in the tests. A sample of higher ash material will also be prepared for use in the tests. Carbon and ash and screen analyses of all materials were performed and are presented in subsequent sections of this report.

Objective

The objective of our work was to determine the effect of temperature and ash level on the agglomeration of Western Kentucky coal char and to investigate the fluidization characteristics of the materials used in the tests.

Operations and Results

Three fluidization tests were conducted with AAG char from Run 123. These tests were conducted in equipment specifically designed for this purpose. The equipment is shown in Figure C-3. The fluidization column for these tests is a 2.5-inch ID Plexiglas tube with a porous plate distributor. (This is to be contrasted with the 2.0-inch ID stainless-steel reactor used in the agglomeration tests.) A typical test consisted of filling the fluidization column with a known weight of the char, fluidizing the bed for about 5 minutes, and then quickly turning off the feed gas to obtain a well-mixed bed. The static height of the bed runs obtained was noted, and the feed gas to the column was resumed. However, this time we started with low flow rates and gradually increased them. Pressure drops and various observations about the bed condition were recorded at each flow rate. When the bed started to slug violently, no further increase in the velocity was made, and the gas flow was decreased in small steps. The bed pressure drop, height, and condition were recorded for the decreased velocities until the gas flow was stopped. Tests were performed on a composite material having a size distribution as shown in Figure C-4 and with two size fractions (i.e., -12+40 and +12). Bed pressure drop

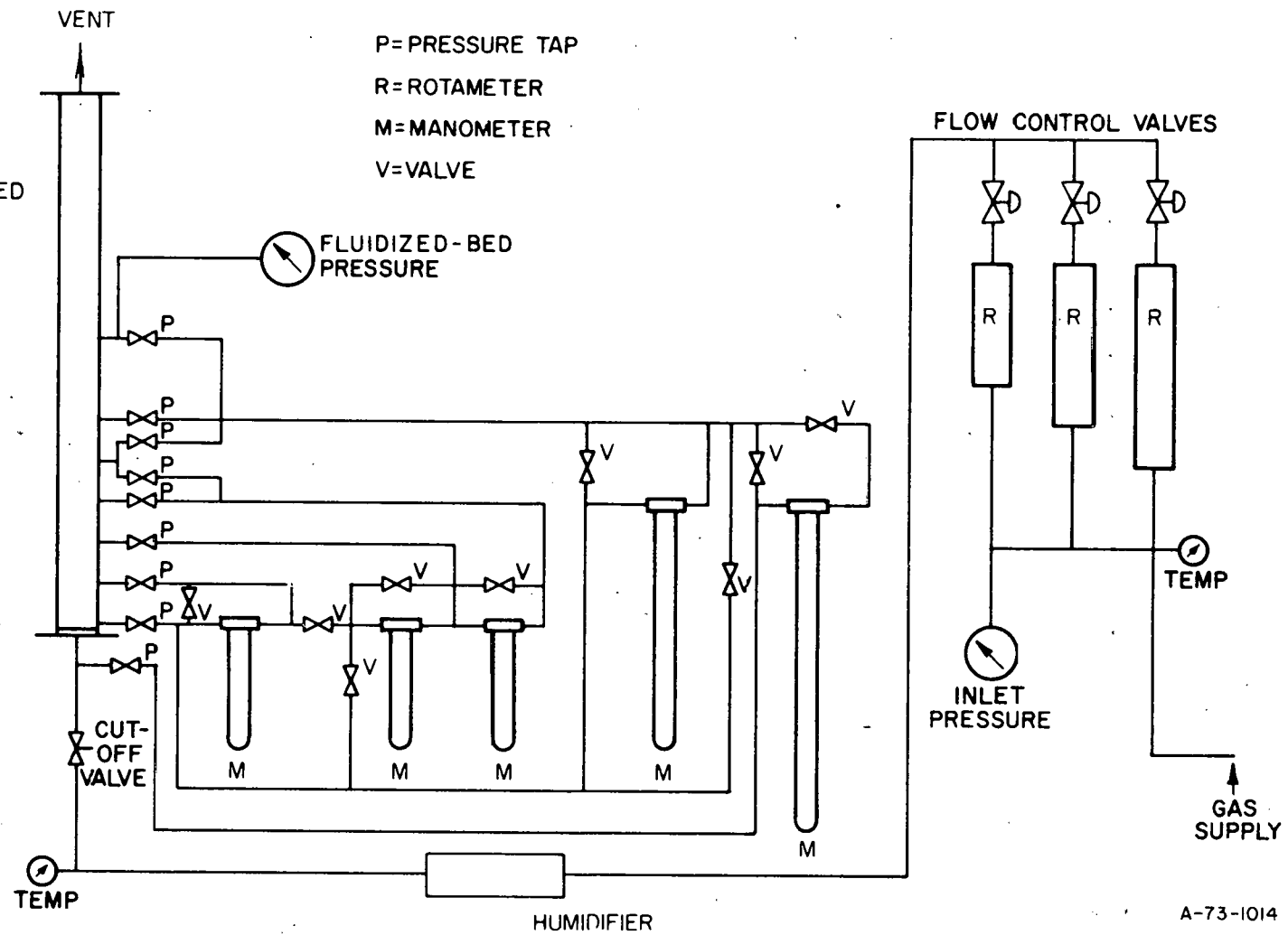
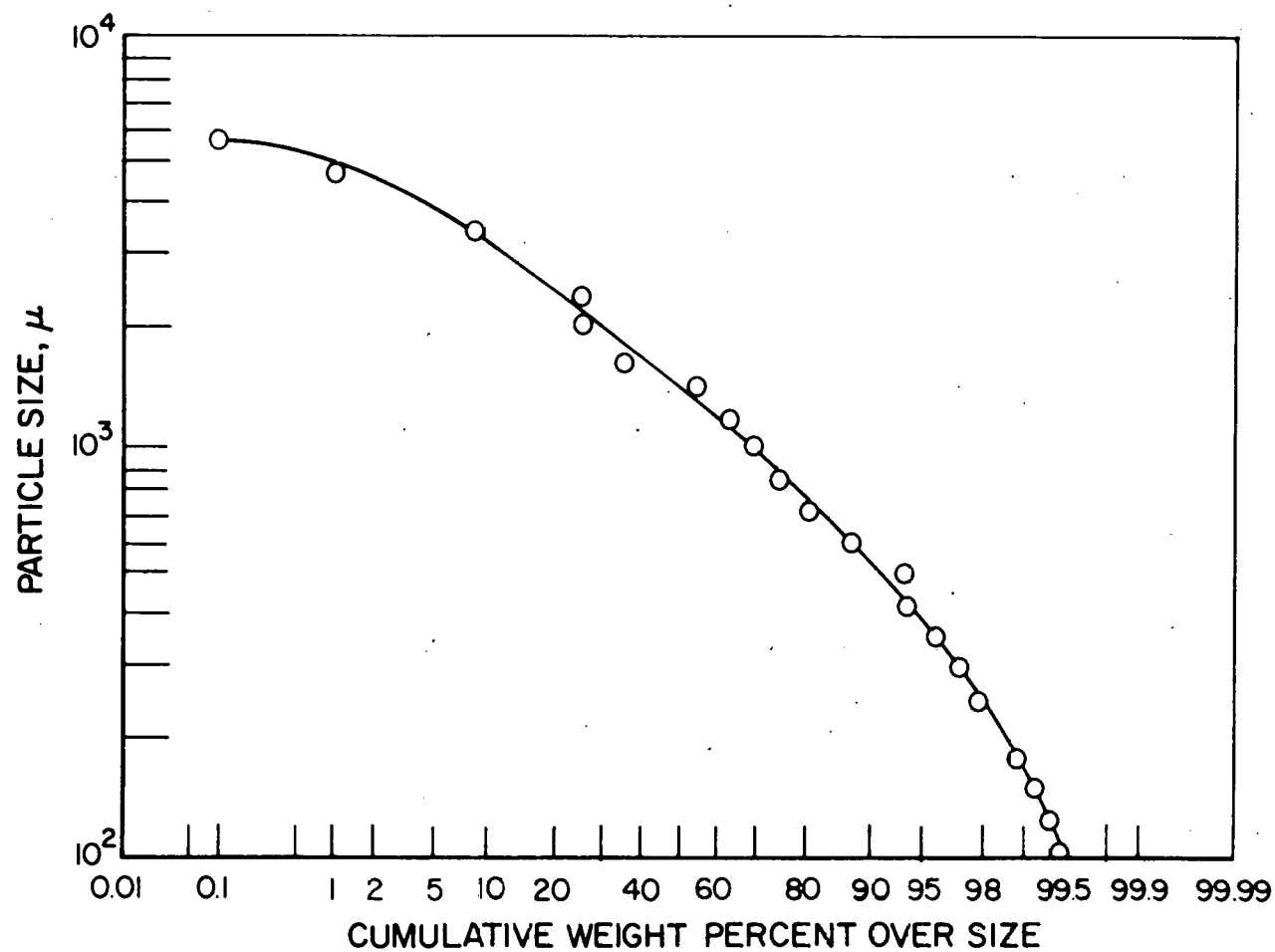


Figure C-3. SCHEMATIC FLOW DIAGRAM FOR FLUIDIZED-BED TESTS
IN THE 2.5-INCH-DIAMETER COLUMN

C-12

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Figure C-4. PARTICLE-SIZE DISTRIBUTION OF AAG CHAR USED IN FLUIDIZED-BED TESTS IN THE 2.5-INCH-DIAMETER COLUMN

versus superficial velocity and superficial velocity versus bed density were recorded and plotted over a wide range of conditions for these three size fractions and are shown in Figures C-5 through C-10.

Complete and minimum fluidization velocities were determined from the above plots and are shown as a function of particle size in Figure C-11. The minimum fluidization velocity was determined by the intersection of the lines through the packed- and fluidized-bed regions for decreasing velocity. The complete fluidization velocity was determined by visual observation; it is the minimum velocity at which top-to-bottom mixing of the solids is visible. The slugging velocity was also determined by visual observation; it is the velocity at which the bubble size in the bed equals the diameter of the column. Table C-4 presents this and other information concerning the tests. Figure C-11 indicates that the minimum and complete fluidization velocities are proportional to the 0.5 and 0.3 powers of average particle size, respectively. This information was used to determine suitable ranges for velocity in the tests in the 2.0-inch reactor.

We conducted seven additional test runs in the high-temperature agglomeration reactor. Three distinct materials were used in these tests — bed (C-7) discharge materials from Runs 122, 123, and 124. The first and last of these materials are quite similar in ash content and particle size. All of the runs conducted in this equipment are summarized in Table C-5. Runs AGX-5 through AGX-9 were performed with the nominal 40% ash content material. In all of these runs, large losses of material occurred. Although originally thought to be caused by leakage of oxygen into the system or by off-specification N_2 , these losses were later found to be caused by the residual high-volatile matter content of the char used. Run 10A was a precautionary devolatilization run. The negligible loss of material (10 grams) in this run is indicative of the success of this procedure, even though it was performed on high-ash material.

Runs AGX-4 and AGX-10, both conducted with high-ash material, resulted in a significant growth of material as shown in Figures C-12 and C-13. These figures are Rosin-Rammler plots of the feed and residue of Runs AGX-4 and AGX-10. The data from which these plots were developed are shown in Table C-6. The fact that the residue line lies below the feed line is indicative of the growth of the char under the conditions of the test.

Photomicrographs of the -3-1/2+4 and -10+12 fractions of feed and residue from Run AGX-4 are presented in Figures C-14 through C-18. Pronounced melting of the dense, light-colored shale particles in the feed (Figures C-14 and C-15) has yielded the rounded, vesicular agglomerates in the residue (Figures C-16 and C-17). The bubbles and

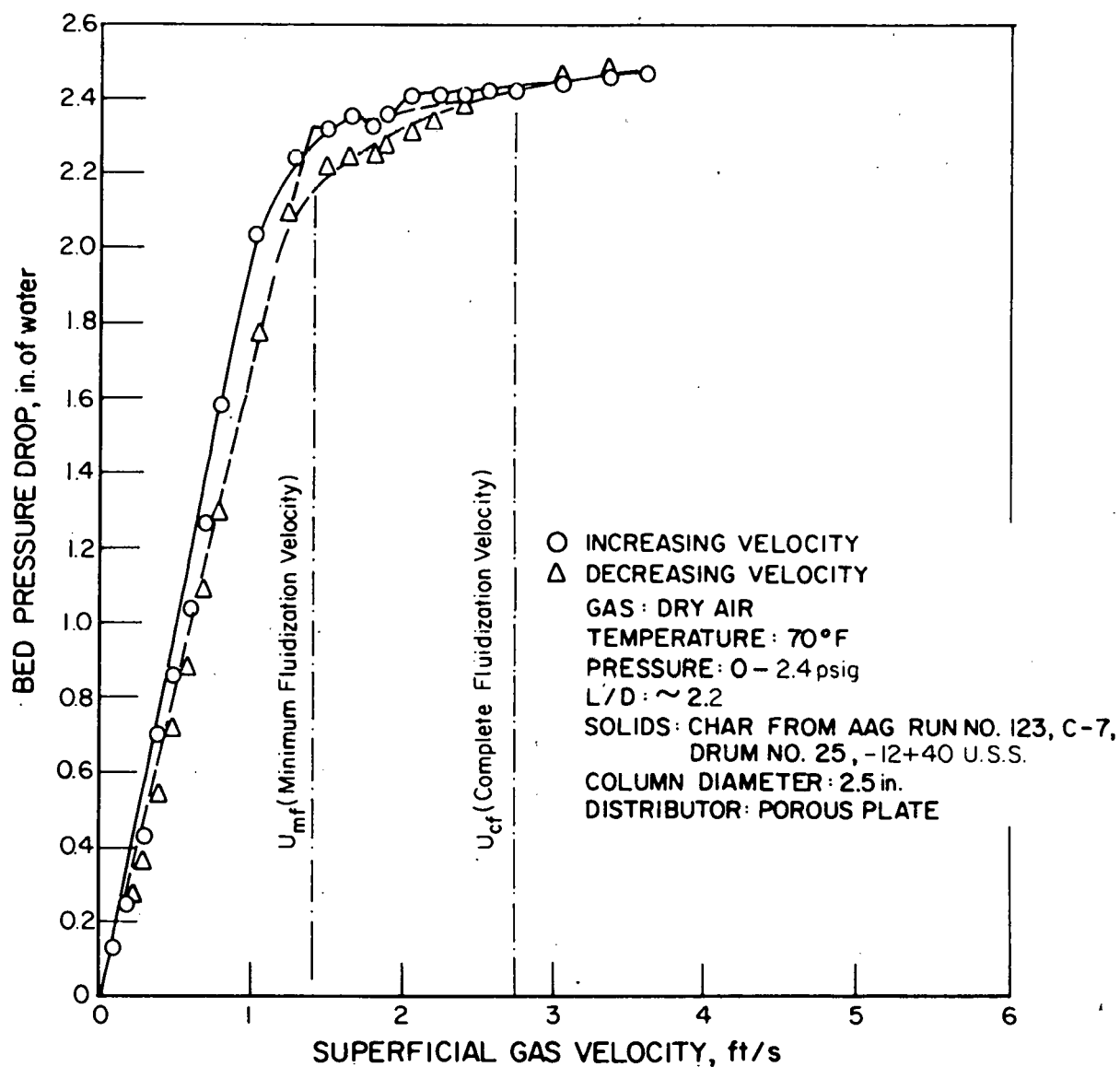


Figure C-5. EFFECT OF GAS VELOCITY ON BED PRESSURE DROP OF AAG CHAR
 (-12+40 U.S.S. Size Fraction)

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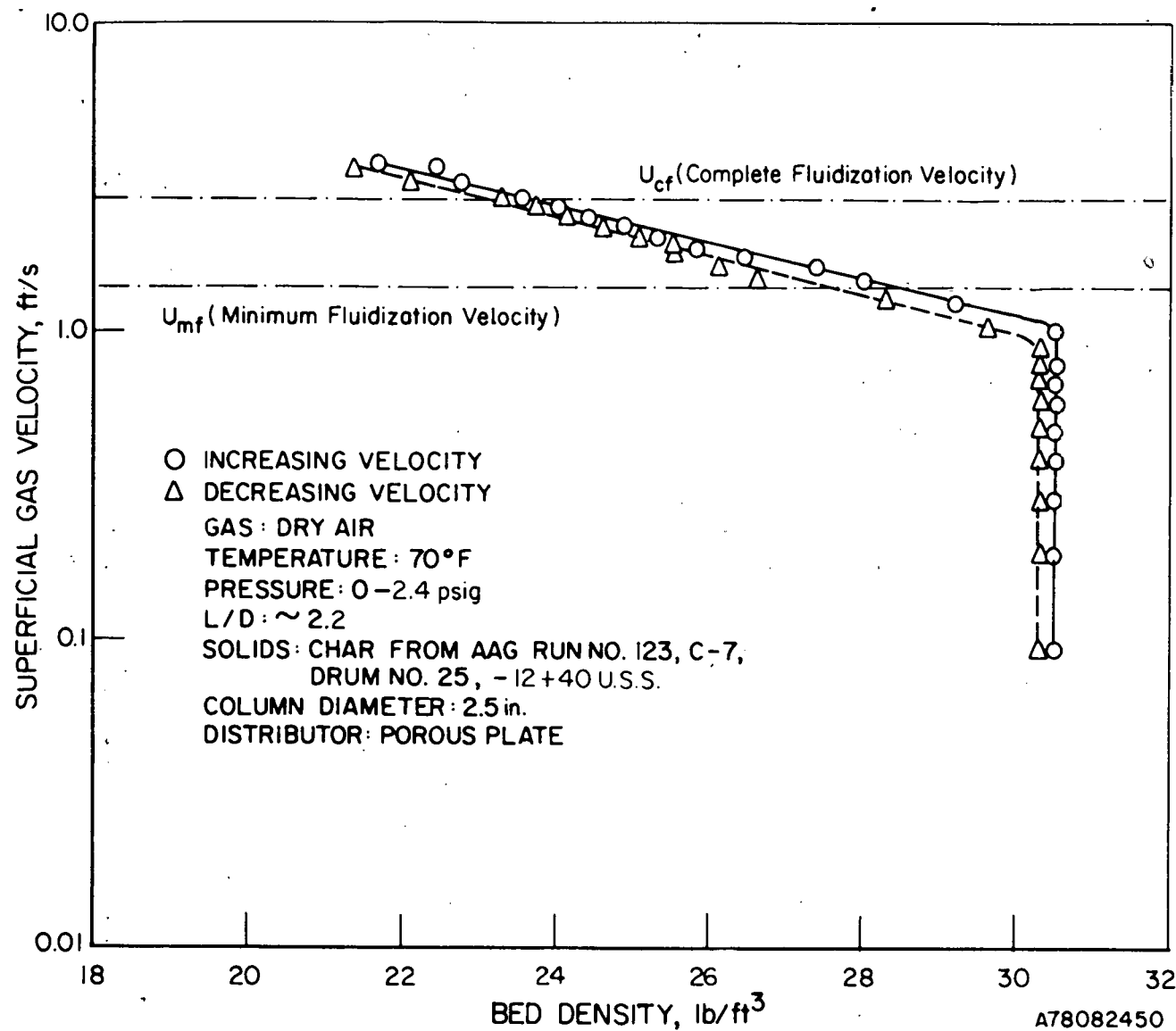


Figure C-6. EFFECT OF GAS VELOCITY ON BED DENSITY OF AAG CHAR
 (12+40 U.S.S. Size Fraction)

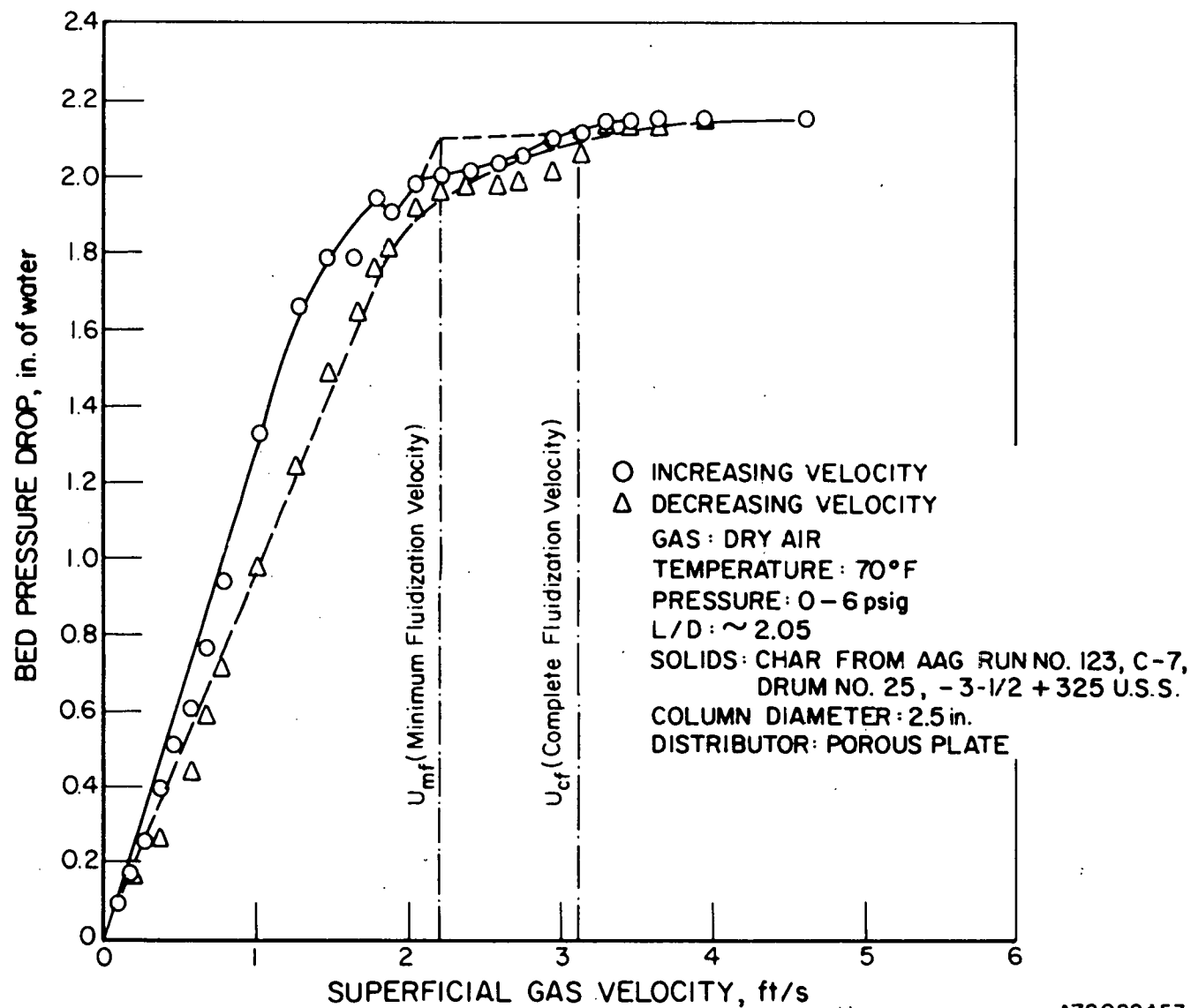
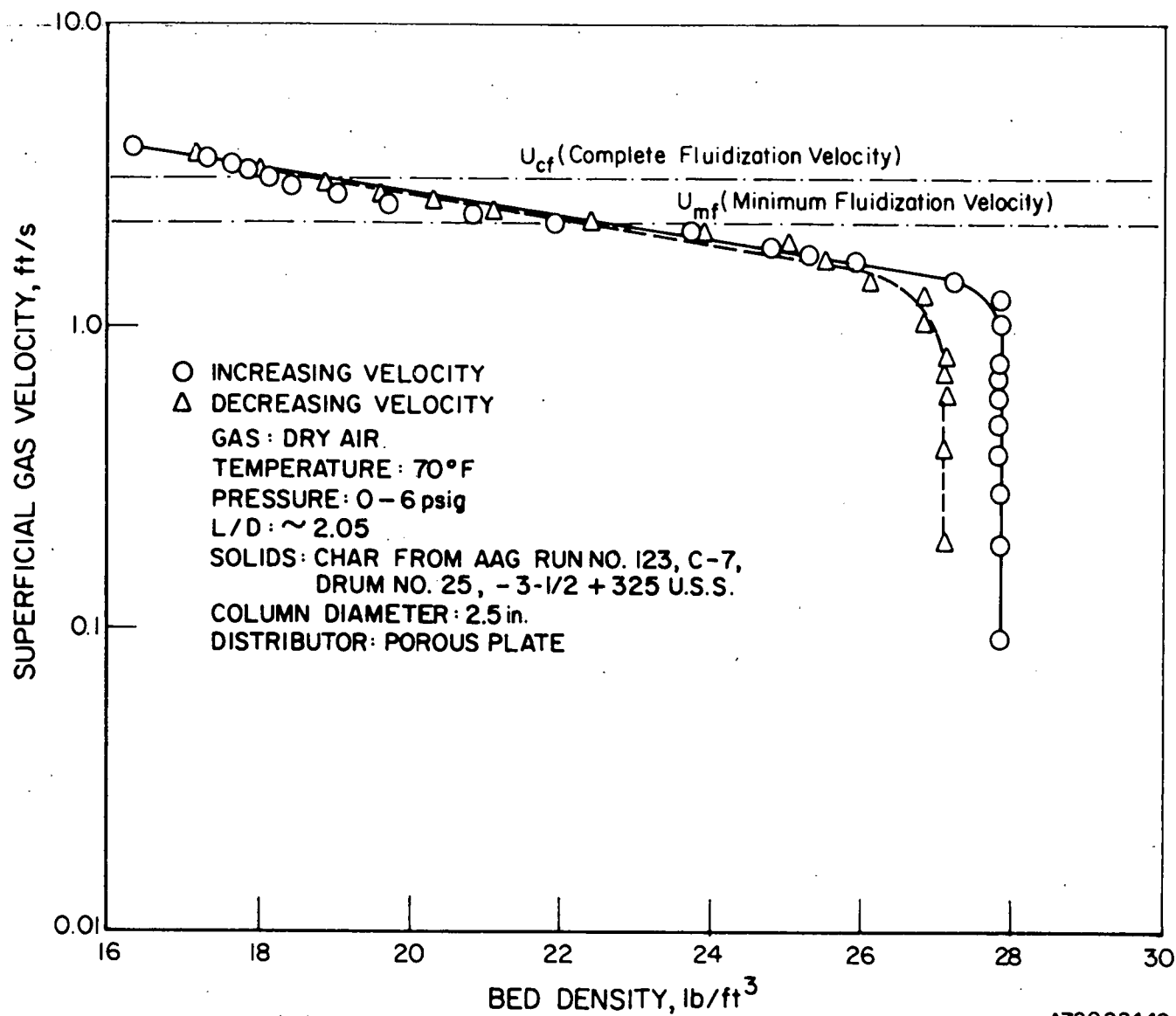


Figure C-7. EFFECT OF GAS VELOCITY ON BED PRESSURE DROP OF AAG CHAR
 (-3-1/2+325 U.S.S. Size Fraction)

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Figure C-8. EFFECT OF GAS VELOCITY ON BED DENSITY OF AAG CHAR
 (-3-1/2+325 U.S.S. Size Fraction)

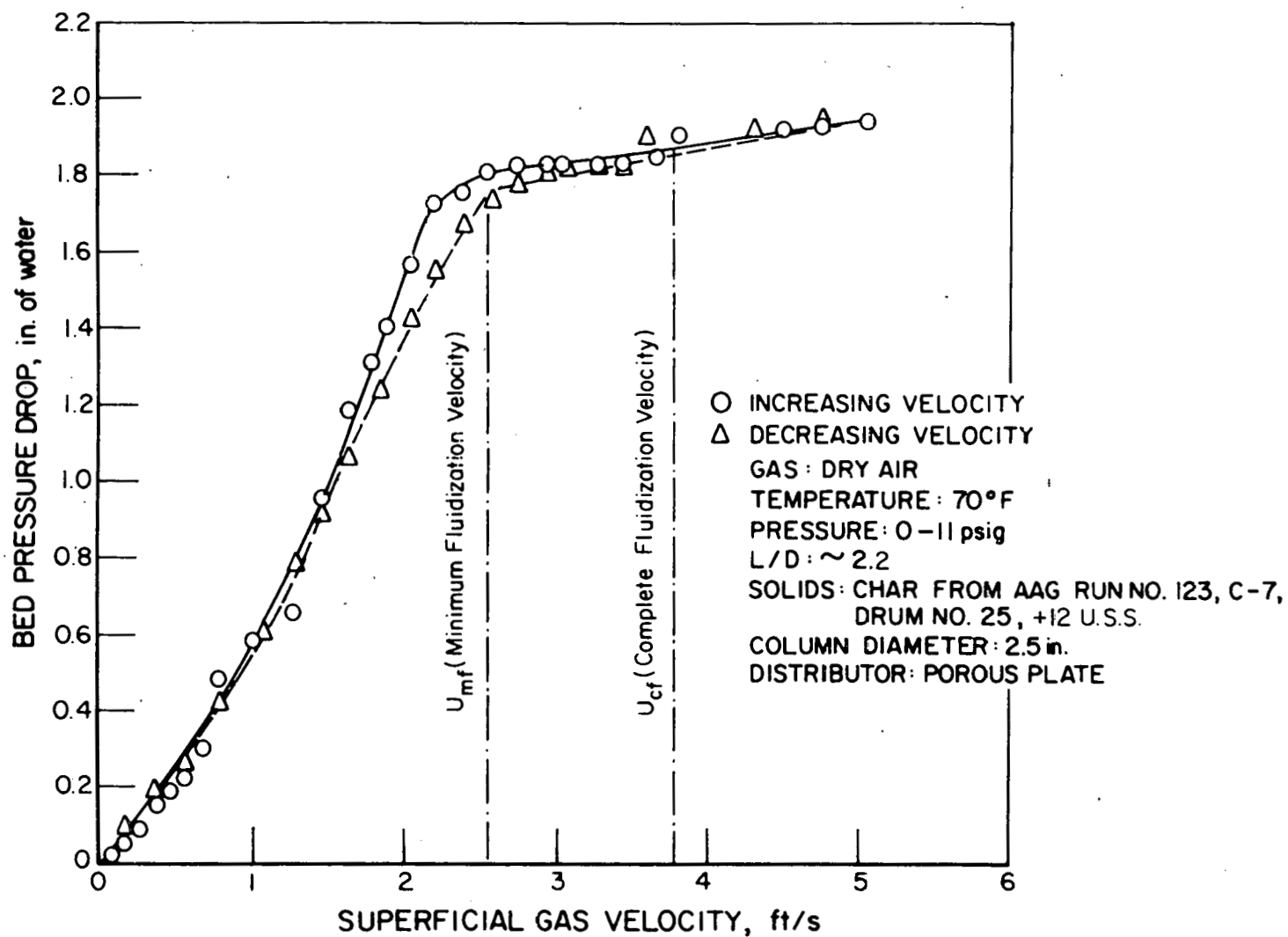


Figure C-9. EFFECT OF GAS VELOCITY ON BED PRESSURE DROP OF AAG CHAR (+12 U.S.S. Size Fraction)

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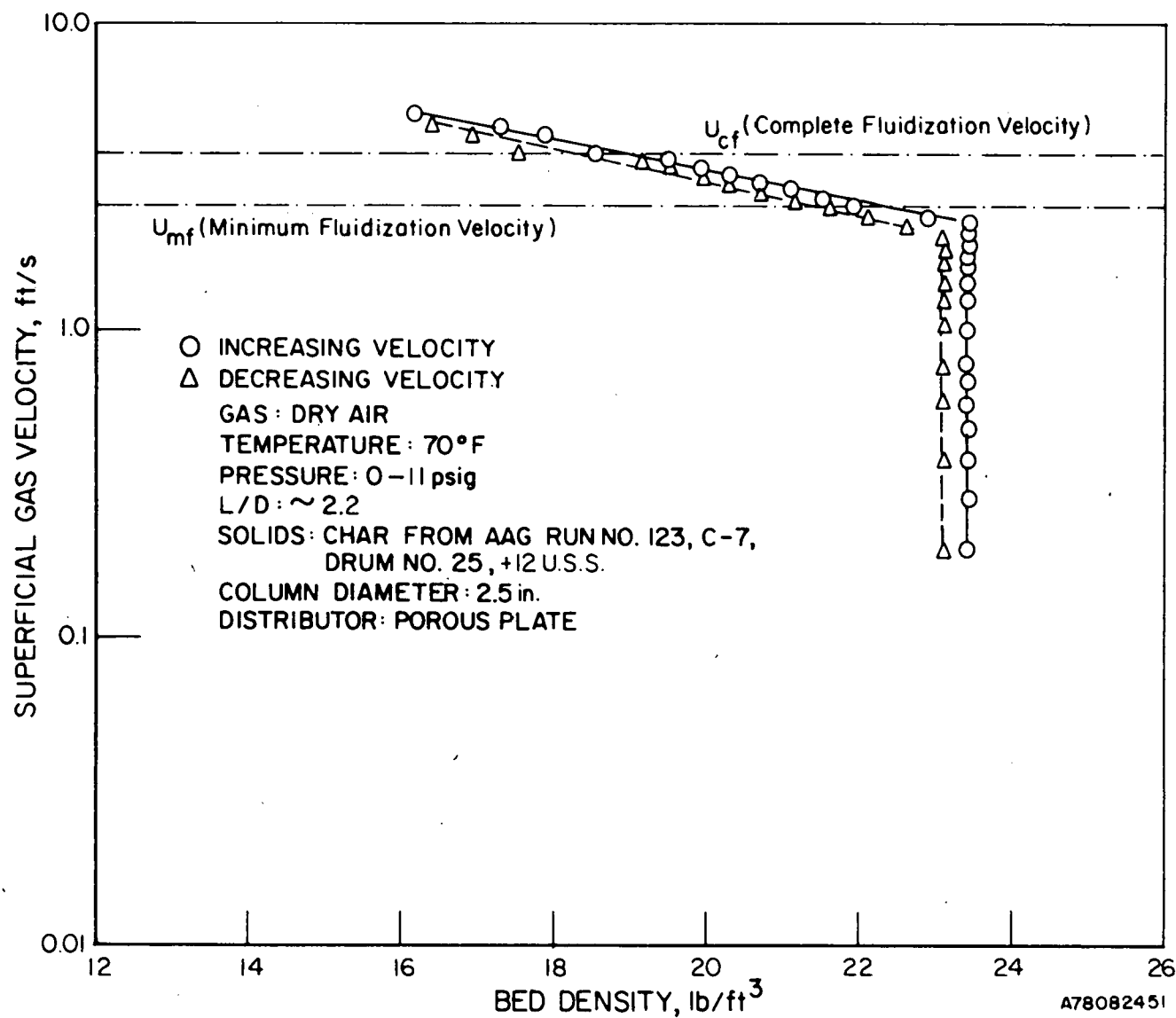


Figure C-10. EFFECT OF GAS VELOCITY ON BED DENSITY OF AAG CHAR
 (+12 U.S.S. Size Fraction)

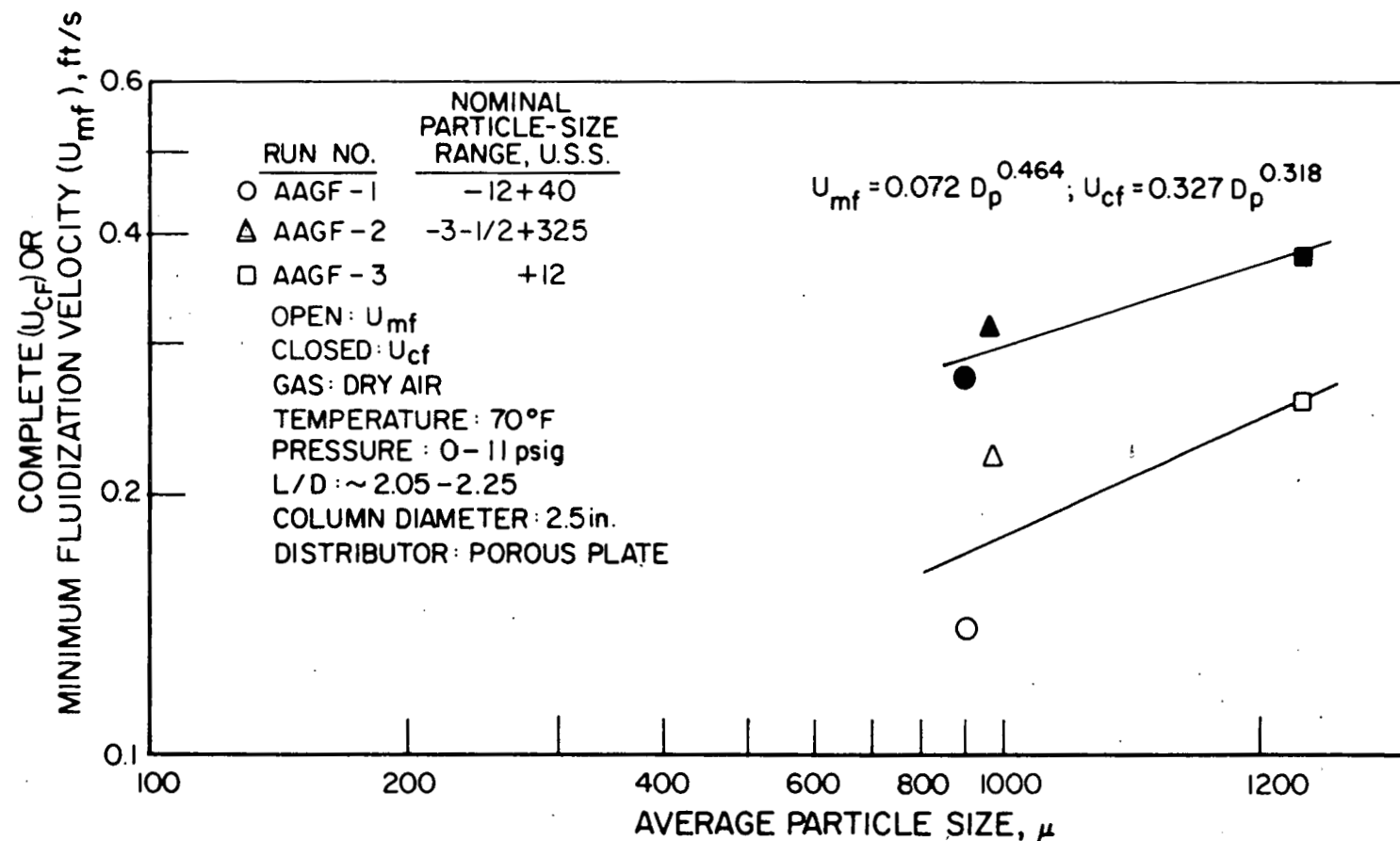


Figure C-11. EFFECT OF PARTICLE SIZE ON MINIMUM AND COMPLETE FLUIDIZATION VELOCITIES OF AAG CHAR

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Table C-4. MINIMUM FLUIDIZATION, COMPLETE FLUIDIZATION, AND SLUGGING VELOCITIES
FOR AAG CHAR IN THE 2.5-in.-DIAMETER COLUMN

Run No.	Gas	Pressure, psig	L/D (Nominal)	Particle-Size Range, U.S.S.	Average ^a Particle Diameter, μ	Superficial Gas Velocity ^b , ft/s		
						U_{mf}^c	U_{cf}^d	U_s^e
AAGF-1	Air	0-2.4	2.2	-12+40	913	1.4	2.71	No Slugging
AAGF-2	Air	0-6.0	2.05	-3-1/2+325	979	2.2	3.11	No Slugging
AAGF-3	Air	0-11.0	2.2	+12	2234	2.55	3.79	4.46

$$^a D_p = 1/\sum_i x_i/d_i$$

^b Based on the pressure at the top of the bed.

^c Point of intersection of the lines through the fixed and fluidized regions (decreasing flow).

^d Minimum velocity when top-to-bottom mixing of the solids is visible.

^e Slugging velocity.

Table C-5. SUMMARY OF BENCH-SCALE, PRELIMINARY & FEASIBILITY
HIGH-TEMPERATURE REACTOR TESTS WITH W. KENTUCKY COAL CHAR

Run No.	Sample Batch No.	Nominal Feed Size, U.S.S.	Ash Content	Feed, %	Residue, %	Fines, %	Losses, %	Test Period, min	Temp, °F	L/D	Fluidizing Gases	Sup. Gas Velocity, ft/s	Sinter	Agglomeration	Remarks
1	Run #122, Bed Discharge	-3-1/2 +325	65.0%	269	244.4	28.4	6.2	60	1710	3	N ₂	0.91-1.03		No	
2	Run #122, Bed Discharge	-3-1/2 +325	65.0%	267.4	205.8	15.5		60	1812	3	N ₂ , Steam	0.83-0.94		No	
3	Run #122, Bed Discharge	-3-1/2 +325	65.0%	268.4	193.9	40.5		7	1830	3	N ₂ , Steam, O ₂	0.86-0.97	Yes	--	
4	Run #122, Bed Discharge	-3-1/2 +400	65.0%	269	232.5	28.7	7.8	30	2000	3	N ₂	4	Yes	Yes	
5	Run #123, C-7, #25	-1/4" +400	40.0%	265.6	231.1	15.0	19.5	30	1900	3	N ₂	4	Yes	--	
6	Run #123, C-7, #25	-1/4" +400	40.0%	266.1	132.3	1.4	182.4	30	1900	3	N ₂	2	No	--	Possible weighing error in feed.
7	Run #123, C-7, #25	-1/4" +400	40.0%	265.7	204.8	1.5	59.4	30	1900	3	N ₂	2	No	--	
8	Run #123, C-7, #25	-1/4" +400	40.0%	265.7	211.5		54.2	30	1900	3	N ₂	2	No	--	Wt. of residue includes fines.
9	Run #123, C-7, #25	-1/4" +400	40.0%	265.0	209.3	3.4	52.3	30	1950	3	N ₂	4	No	--	
10A	Run #124, C-7, #95A		69.0%	417.4	371.9	33.3	12.2	60	1700	3	N ₂	1	No	--	Conducted to drive off volatile matter.
10B	Run #124, C-7, #95A			264.7	222.9	32.8	9.0	30	1900	3	N ₂	4	Yes	Yes	Residue & Fines of Run #10A used for feed.

C-23

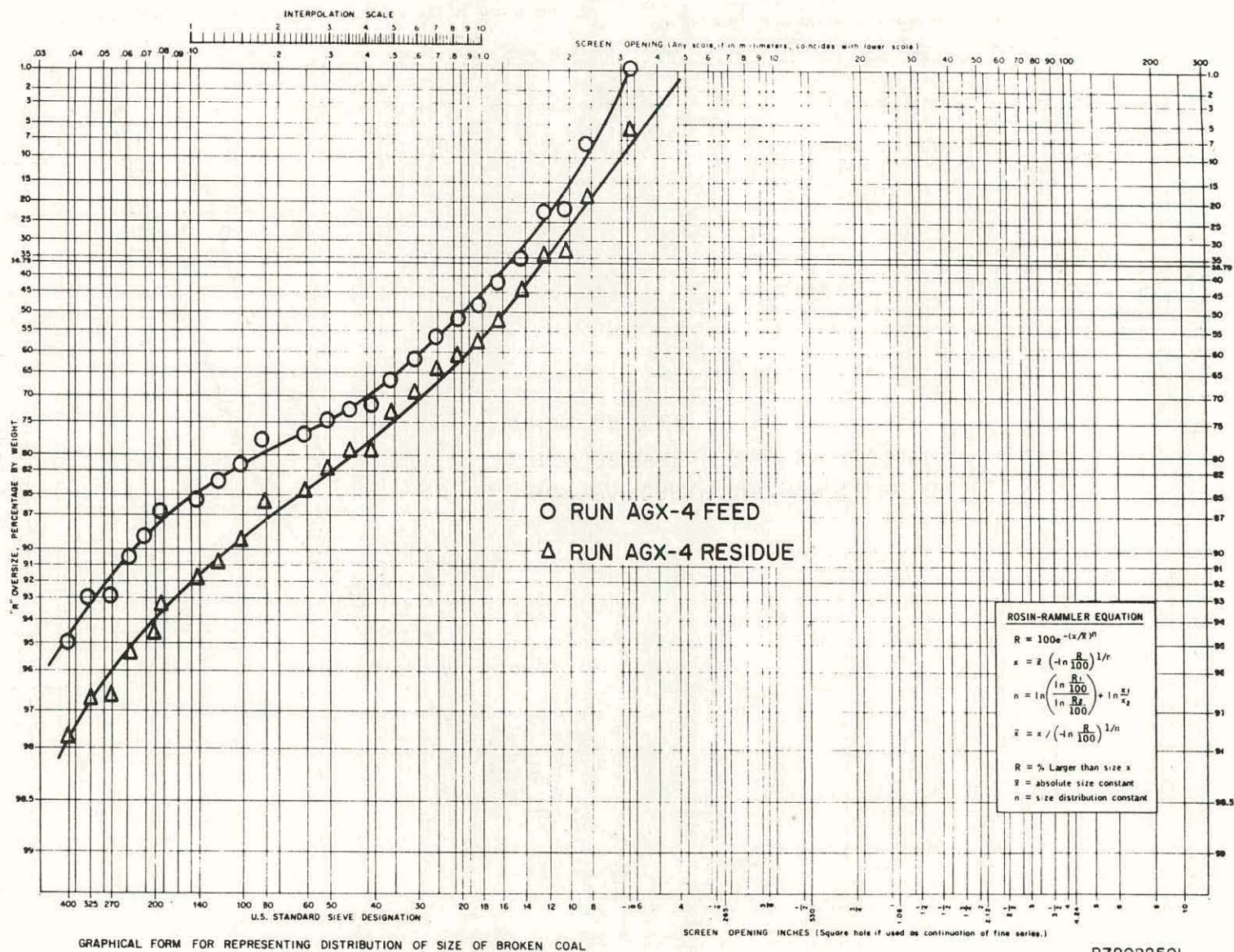
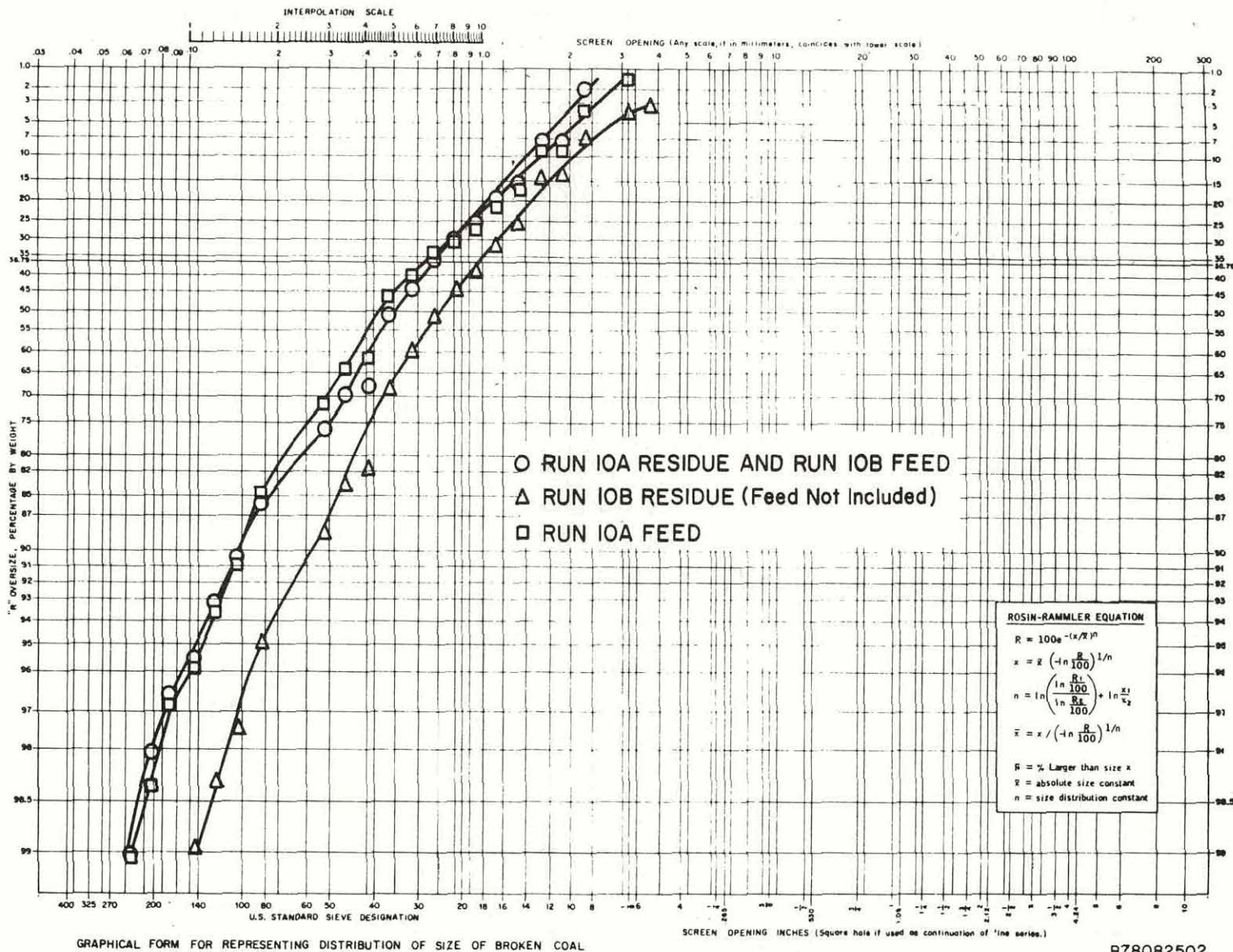


Figure C-12. ROSIN-RAMMLER PLOT OF FEED AND RESIDUE OF RUN AGX-4

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C-24



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Figure C-13. ROSIN-RAMMLER PLOT OF FEED AND RESIDUE OF RUN AGX-10

Table C-6. WEIGHTS OF VARIOUS SIZE FRACTIONS OF FEEDS
AND RESIDUES FOR RUNS AGX-4 AND AGX-10

Run No.	AGX-4		AGX-10	
	<u>Feed</u>	<u>Residue</u>	<u>Feed</u>	<u>Residue</u>
Sieve, U.S.S.	Weight, g			
+ 3-1/2	0.0	5.81	0.27	7.80
- 3-1/2+4	2.69	8.84	1.32	1.78
- 4+6	19.37	27.90	3.97	7.13
- 6+8	37.66	37.43	14.03	15.83
- 8+10	1.35	1.40	0.53	0.89
- 10+12	34.43	29.76	22.50	24.52
- 12+14	17.49	11.16	9.53	13.60
- 14+16	15.87	12.09	16.68	16.27
- 16+18	11.03	7.67	11.91	12.04
- 18+20	11.57	8.37	15.35	14.71
- 20+25	14.53	11.16	19.85	19.17



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Figure C-14. SHALE PARTICLES IN $-3-1/2+4$ MESH
CHAR FEED AT 5.4X MAGNIFICATION



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Figure C-15. AAG $-10+12$ MESH CHAR FEED AT
5.4X MAGNIFICATION



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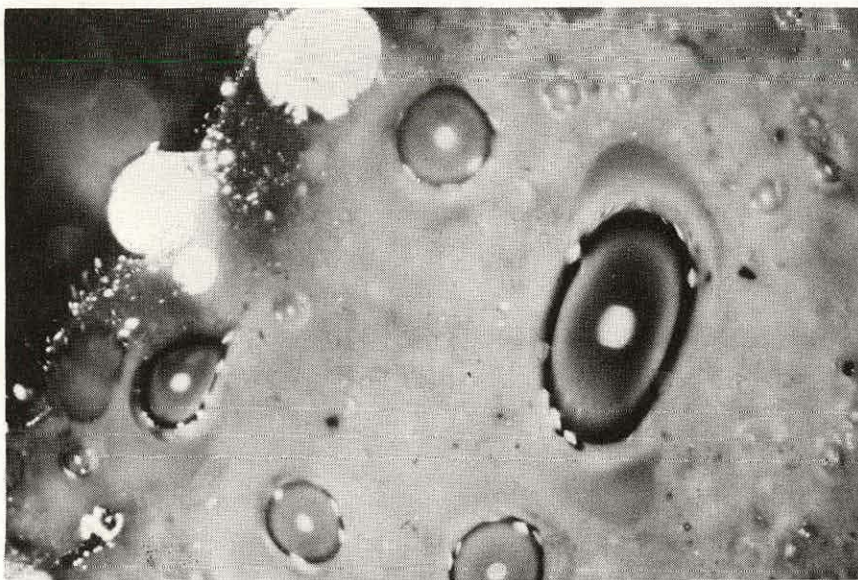
Figure C-16. AAG RESIDUE FROM - 3-1/2+4 MESH
CHAR FEED AT 5.4X MAGNIFICATION



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Figure C-17. AAG RESIDUE FROM -10+12 MESH
CHAR FEED AT 5.4X MAGNIFICATION

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Figure C-18. MELTED SHALE IN AAG CHAR RESIDUE
AT 141X MAGNIFICATION

uniform texture visible within some agglomerates at high magnification (Figure C-18) are an indication of the high degree of melting of the shale in the residue.

Conclusions From Non-Devolatilized Coal Tests, Fluidization Tests

Our test results show that agglomeration of the ash of Western Kentucky coal is possible in the bench-scale reactor. We have demonstrated this at temperatures of 1900° and 2000°F. A higher degree of agglomeration, as evidenced by greater relative growth, occurs at the higher temperature. This is shown by the comparisons of mean particle size of feed and residue in Table C-7. The percentage deviation of residue from feed mean particles sizes is 1.4 to 1.6 times greater for Run AGX-4 than for Run AGX-10, regardless of which methods of mean size is considered.

Coke Tests

Summary

Samples of raw and partially gasified Bethlehem Coke breeze were obtained from the pilot plant (Runs 124, 126) for a series of runs that were made in the bench-scale reactor. The conditions suitable for agglomerating this material were determined at the same residence time and superficial velocity as those used in the tests with Western Kentucky coal. We chose to investigate coke because it agglomerated readily in the pilot plant under a wider range of conditions.

We performed five additional agglomeration tests in the 2.0-inch ID reactor with coke samples obtained from the pilot plant (Run 124). Indications of agglomeration were obtained with the samples with ash contents of approximately 41% at 2000°F, but sintering was also evident in these cases.

Objective

The objective of this work was to determine the effect of temperature, ash level, and residence time on the agglomeration of Bethlehem coke breeze.

Operation and Results

The proximate and ultimate analyses of the eight coke samples from Run 124 were received and are shown in Table C-8. Coke with an ash content in the range of 16% to 42% was available from these samples. Six of these samples were used in this series of tests. In addition three tests, Runs AGX 11, 12, and 13, were performed with a raw coke from Run 126. The conditions of these tests are shown in Table C-9. The residue from Run AGX-16 indicated that agglomeration had taken place. A partial screen analysis of this run is shown in Table C-10. An approximate 7% growth of material above 30 mesh

Table C-7. MEAN DIAMETERS OF FEEDS AND RESIDUES OF BENCH-SCALE REACTOR TESTS

Run No.	AGX-4			AGX-10		
	Feed	Residue	Residue-Feed Feed X 100	Feed	Residue	Residue-Feed Feed X 100
Length Mean ^a	0.0424	0.0437	3.1	0.0382	0.0390	2.1
Surface Mean ^b	0.0464	0.0489	5.4	0.0404	0.0417	3.2
Volume Mean ^c	0.0518	0.0563	8.7	0.0436	0.0459	5.3
Sauter Mean ^d	0.0645	0.0746	15.7	0.0508	0.0556	9.4

^a Length Mean = $\frac{\sum_i (x_i / D_{pi}^2)}{\sum_i (x_i / D_{pi}^3)}$

^b Surface Mean = $(\frac{\sum_i (x_i / D_{pi})}{\sum_i (x_i / D_{pi}^3)})^{1/2}$

^c Volume Mean = $1 / (\sum_i x_i / D_{pi}^3)^{1/3}$

^d Sauter Mean = $1 / \sum_i x_i / D_{pi}$

 x_i = weight fraction of material in size class i.

 D_{pi} = average diameter of size class i.

Table C-8. ANALYSES OF COKE SAMPLES FROM RUN 124

Run No.	AGX-18	AGX-15	(NOT USED)	AGX-19	AGX-17	(NOT USED)	AGX-14	AGX-16
Date	5/28/78	5/28/78	5/31/78	5/31/78	6/1/78	6/1/78	6/1/78	6/1/78
Time	1700	2000	2200	2300	0400	1700	1900	2100
Proximate Analysis, wt % (as rec'd)								
Moisture	0.2	0.2	0.3	0.3	0.3	0.2	0.3	0.3
Volatile Matter	1.1	1.5	1.2	1.3	1.4	1.4	0.9	0.7
Ash	31.0	24.4	16.6	18.2	41.7	17.4	31.6	41.1
Fixed Carbon	67.7	73.9	81.9	80.2	56.6	81.0	67.2	57.9
Total	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Ultimate Analysis, wt % (dry basis)								
Ash	31.02	24.45	16.63	18.26	41.79	17.47	31.72	41.22
Carbon (total)	67.70	74.20	81.80	80.00	57.40	80.90	67.70	58.60
Hydrogen	0.24	0.28	0.24	0.19	0.22	0.26	0.21	0.18
Sulfur	0.56	0.55	0.56	0.59	0.27	0.60	0.50	0.52
Nitrogen	0.46	0.53	0.55	0.58	0.41	0.57	0.46	0.42
Oxygen (by diff.)	0.02	0	0.22	0.38	0	0.20	0	0
Total	100.00	100.01	100.00	100.00	100.00	100.00	100.59	100.94

Table C-9. SUMMARY OF BENCH-SCALE, HIGH-TEMPERATURE REACTOR TESTS WITH COKE

ACX Run No.	Sample Batch No.	Nominal Feed Size, U.S.S.	Approx Ash Content, %	Feed	Residue	Fines	Losses	Test Period, min	Temp, °F	L/D	Fluidizing Gases	Superficial Gas Velocity, ft/s	Sinter	Agglom- eration
				g										
11	Run No. 126, Raw Coke	-3-1/2+400	13.2	265.2	217.2	30.4	17.6	30	1900	3	N ₂	-4	No	No
12	Run No. 126, Raw Coke	-3-1/2+400	13.2	266.2	220.2	28.4	17.6	30	2000	3	N ₂	-4	No	No
13	Run No. 126, Raw Coke	-3-1/2+400	13.2	265.9	242.2	1.7	17.6	30	2100	3	N ₂	-4	No	No
14	Run No. 124, Coke Bed Sample	-3-1/2+400	31.2	262.5	208.0	53.0	1.5	30	1900	3	N ₂	-4	No	No
15	Run No. 124, Coke Bed Sample	-3-1/2+400	24.5	266.2	218.0	47.2	0.9	30	2000	3	N ₂	-4	No	No
16	Run No. 124, Coke Bed Sample	-3-1/2+400	41.2	249.6	229.7	17.6	2.3	30	2000	3	N ₂	-4	Yes	Yes
17	Run No. 124, Coke Bed Sample	-3-1/2+400	41.8	270.5	242.4	26.1	2.0	120	2000	3	N ₂	-4	Yes*	No
18	Run No. 124, Coke Bed Sample	-3-1/2+400	31.0	268.0	235.7	29.6	2.7	30	2000	3	N ₂	-4	No	No
19	Run No. 124, Coke Bed Sample	-3-1/2+400	18.2	--	--	--	--	20	2100	3	N ₂	-4	No	Yes

*Extensive sintering occurred, preventing determination of agglomeration.

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Table C-10. FEED AND RESIDUE ANALYSES FOR BENCH-SCALE
REACTOR TEST AGX-16

	<u>Absolute Weight, g</u>	
	<u>Feed</u>	<u>Residue</u>
+3-1/2	0	0
+4	1.0	0.9
+6	13.5	11.9
+8	20.2	22.0
+10	9.2	9.0
+12	10.5	11.3
+14	11.0	10.6
+16	11.0	12.4
+18	9.0	9.0
+20	10.2	13.1
+25	13.7	16.1
+30	11.2	13.1
SUM above 30 U.S.S.	120.5	129.4 $\Delta = 8.9$ $\% \Delta = 7.4$

(U.S.S.) was found. The distribution of material in the residue was such, however, that an approximate 6% decrease in particle size (50% cumulative size) of the residue occurred. This was primarily caused by the loss of some +6 mesh material from the feed, possibly the result of attrition or breakage, which compensated, to some extent, for the net growth of material in the -20+30 mesh range. Run AGX-17 was conducted with a coke of similarly high ash content as that used in Run AGX-16, but for a longer test period. At the termination of the test, considerable sinter formation was found at the top of the fluidized bed and adhering to the internal wall of the reactor. As a result, we did not perform a screen analysis of the remaining material. A very small degree of sintering was observed in Run AGX-16, but the amount was insignificant, and data analysis could still be performed.

Run AGX-19 was the last test to be conducted using the Run 124 coke samples. (The analysis of Run 124 coke samples was presented on p.C-31 of this report.) The sample used was that from Run 124 (5/31/78, 2300 hours), which had an ash content of 18.3 weight percent (dry basis of ultimate analysis). This run, conducted at 2100°F, a 20-minute retention time, and 4 ft/s superficial velocity, resulted in particle growth. This is shown by the screen analysis for the +35 mesh U.S.S. material (Table C-11). A net growth of approximately 5 grams out of approximately 48 grams of initial feed of +35 mesh material occurred. There was an approximate 3% decrease in average particle diameter, however, as shown in Table C-11. This is the only run performed with these AAG Run 124 coke samples that agglomerated, other than Runs AGX-16 and AGX-17. The latter two runs utilized material of approximately 42% ash content. Thus, higher temperatures can produce agglomeration in materials having ash contents that do not agglomerate at lower temperatures. This is further demonstrated in Figure C-19, which shows the results of Runs AGX-11 through AGX-19. A tentative line of demarcation between the agglomerating runs and the non-agglomerating runs has been indicated in the figure. This line may change either as more data are obtained or as a function of the fluidizing velocity or retention time used in the tests.

Runs AGY-1, 2, and 3 were performed with samples obtained from C-7 discharge of Run 127 (9/25/78, 1900 hours). An analysis of this material is given in Table C-12, and the conditions of the three runs are indicated in Table C-13. Lower superficial velocities were used in these tests than in previous tests. Because there is very little difference in run conditions between these three runs, they can serve to measure the degree of reproducibility of the treatment producing agglomeration. Certain size distributions are shown graphically on Rosin Rammler plots in Figures C-20, C-21, and C-22.

Table C-11. RESIDUE AND FEED SCREEN ANALYSIS AND AVERAGE PARTICLE SIZES FOR RUN AGX-19

Pan No.	U.S. Sieve No.	U.S. Sieve Size, in.	Average Diameter, in.	Residue		Feed	
				Weight Fraction	Cum. Weight Fraction	Weight Fraction	Cum. Weight Fraction
1	-3-1/2+4	-0.2230+0.1870	0.2050	0.0041	0.0041	0.0040	0.0040
2	-4+6	-0.1870+0.1319	0.1594	0.0516	0.0557	0.0620	0.0660
3	-6+8	-0.1319+0.0929	0.1124	0.1225	0.1782	0.1266	0.1926
4	-8+10	-0.0929+0.0787	0.0858	0.0607	0.2389	0.0659	0.2585
5	-10+12	-0.0787+0.0669	0.0728	0.0770	0.3159	0.0746	0.3331
6	-12+14	-0.0669+0.0551	0.0610	0.0642	0.3801	0.0667	0.3998
7	-14+16	-0.0551+0.0465	0.0508	0.0870	0.4671	0.0864	0.4862
8	-16+18	-0.0465+0.0394	0.0429	0.0630	0.5301	0.0626	0.5488
9	-18+20	-0.0394+0.0335	0.0364	0.0849	0.6150	0.0812	0.6300
10	-20+25	-0.0335+0.0280	0.0307	0.1188	0.7338	0.1101	0.7401
11	-25+30	-0.0280+0.0236	0.0258	0.0895	0.8233	0.0851	0.8252
12	-30+35	-0.0236+0.0197	0.0217	0.1767	1.0000	0.1748	1.0000
				Mean Particle Size, in.		Mean Particle Size in.	
				Mean: 0.03959		Mean: 0.04038	
				Median: 0.05757		Median: 0.05949	
				Mass Mean: 0.04668		Mass Mean: 0.04905	

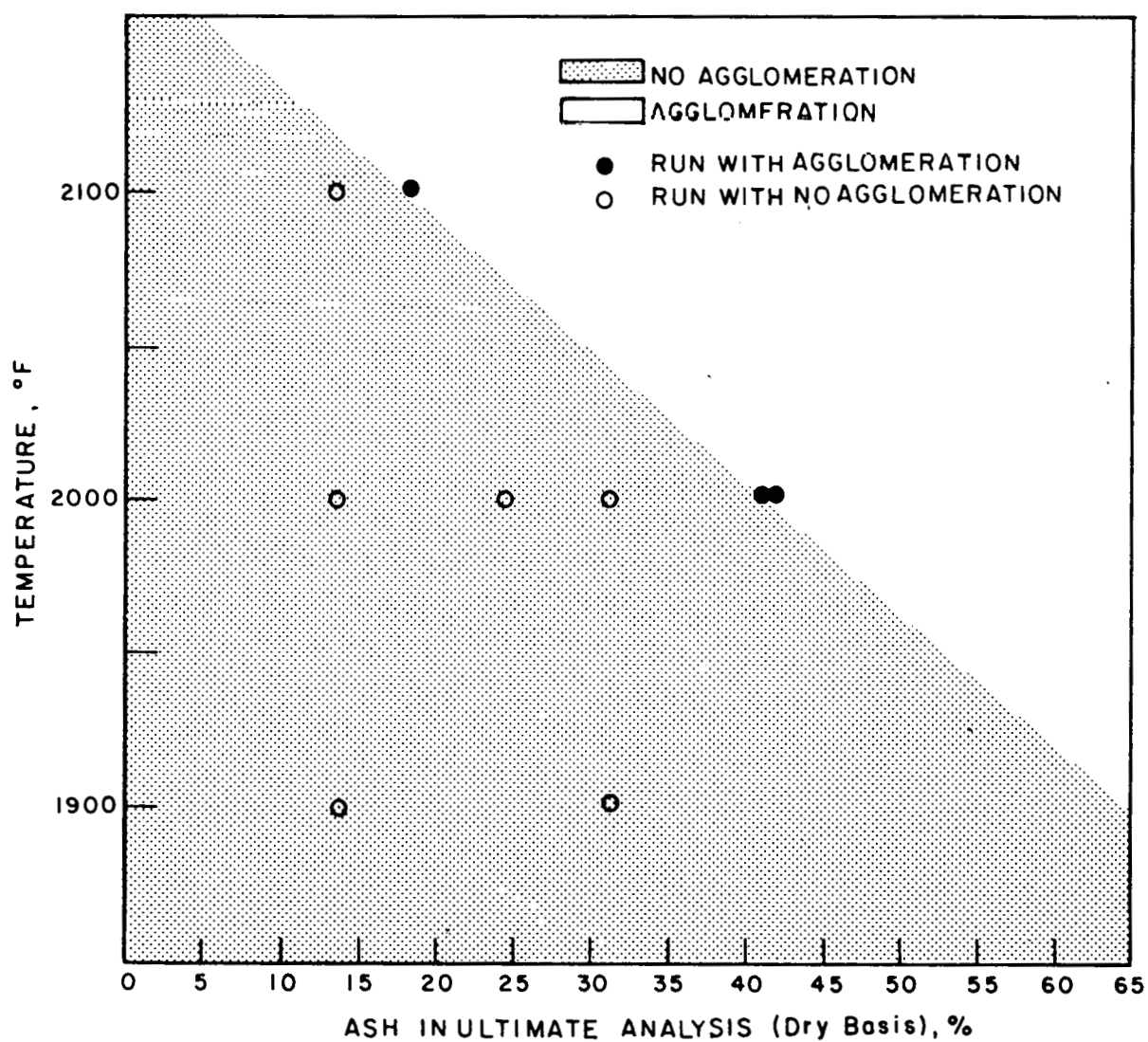


Figure C-19. AGGLOMERATING REGIONS OF BETHLEHEM COKE

A78113291

Table C-12. ANALYSIS OF RUN 127 C-7 DISCHARGE MATERIAL
(Runs AGY-1, 2, and 3)

Proximate Analysis (As Received)

Moisture	0.1
Volatile Matter	0.6*
Ash	47.4
Fixed Carbon	<u>51.9</u>
Total	100.0

Ultimate Analysis (Dry Basis)

Ash	47.46
Carbon (Total)	52.10
Organic Carbon	0.00
Hydrogen	0.13
Sulfur	0.44
Nitrogen	0.24
Carbon Dioxide	0.00
Oxygen (By Diff.)	<u>0.00</u>
Total	100.37

* 0.6 dry, 1.2 dry ash-free basis.

Table C-13. SUMMARY OF BENCH-SCALE, HIGH-TEMPERATURE REACTOR TESTS

<u>AGX Run No.</u>	<u>Sample Description</u>	<u>Approximate Ash Content, %</u>	<u>Time at Max Temperature, min</u>	<u>Maximum Temperature, °F</u>	<u>Superficial Gas Velocity, ft/s</u>	<u>Sinter</u>	<u>Agglomerate</u>
1	Run 127 C-7 Discharge (9/25/78, 1900 hours)	47.5	60	2000	1.11	Yes	Yes
2	Run 127 C-7 Discharge (9/25/78, 1900 hours)	47.5	60	2000	1.38	Yes	Yes
3	Run 127 C-7 Discharge (9/25/78, 1900 hours)	47.5	60	2000	1.23	Yes	Yes

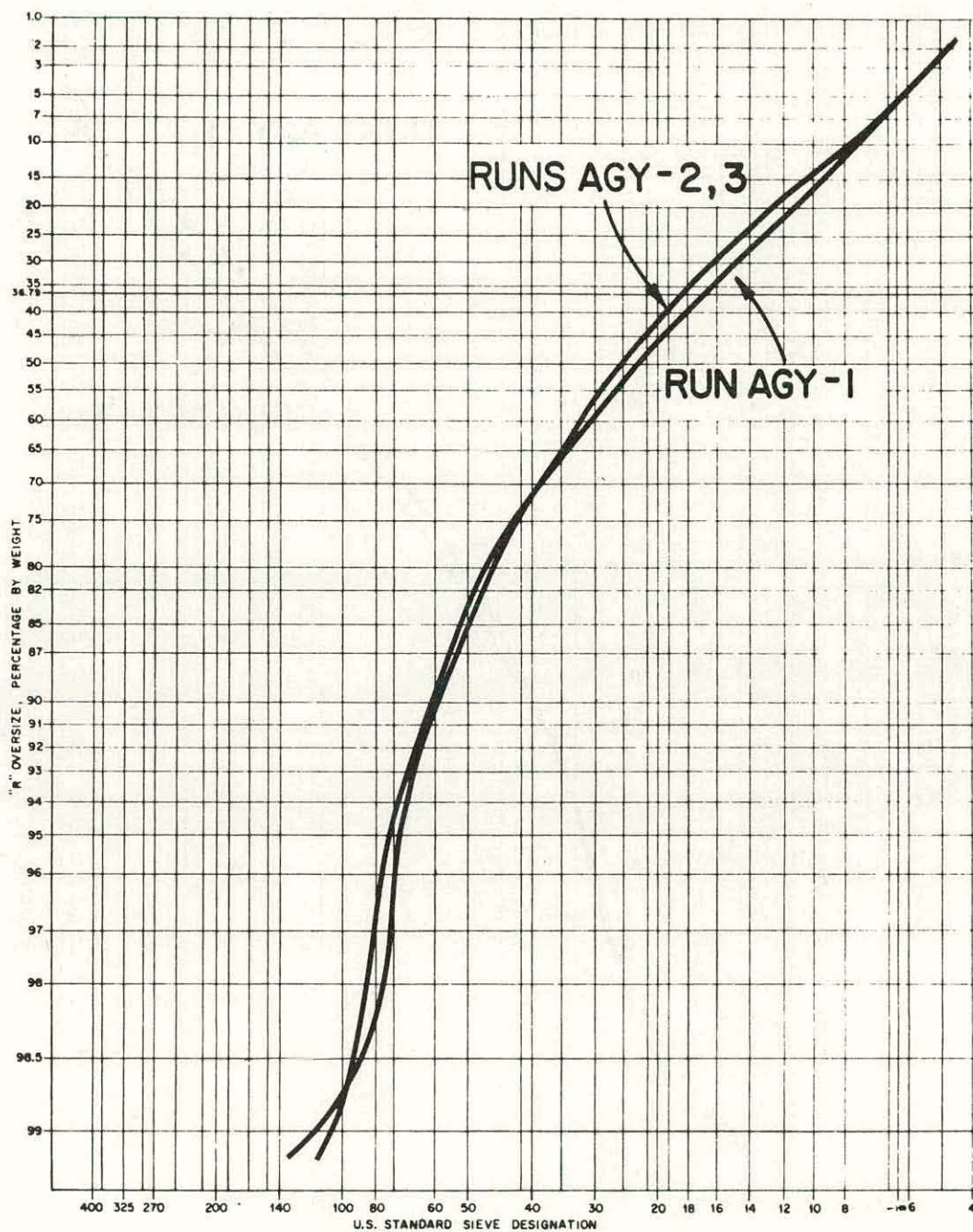


Figure C-20. SIZE DISTRIBUTION OF FEED (Runs AGY-1, 2, and 3)

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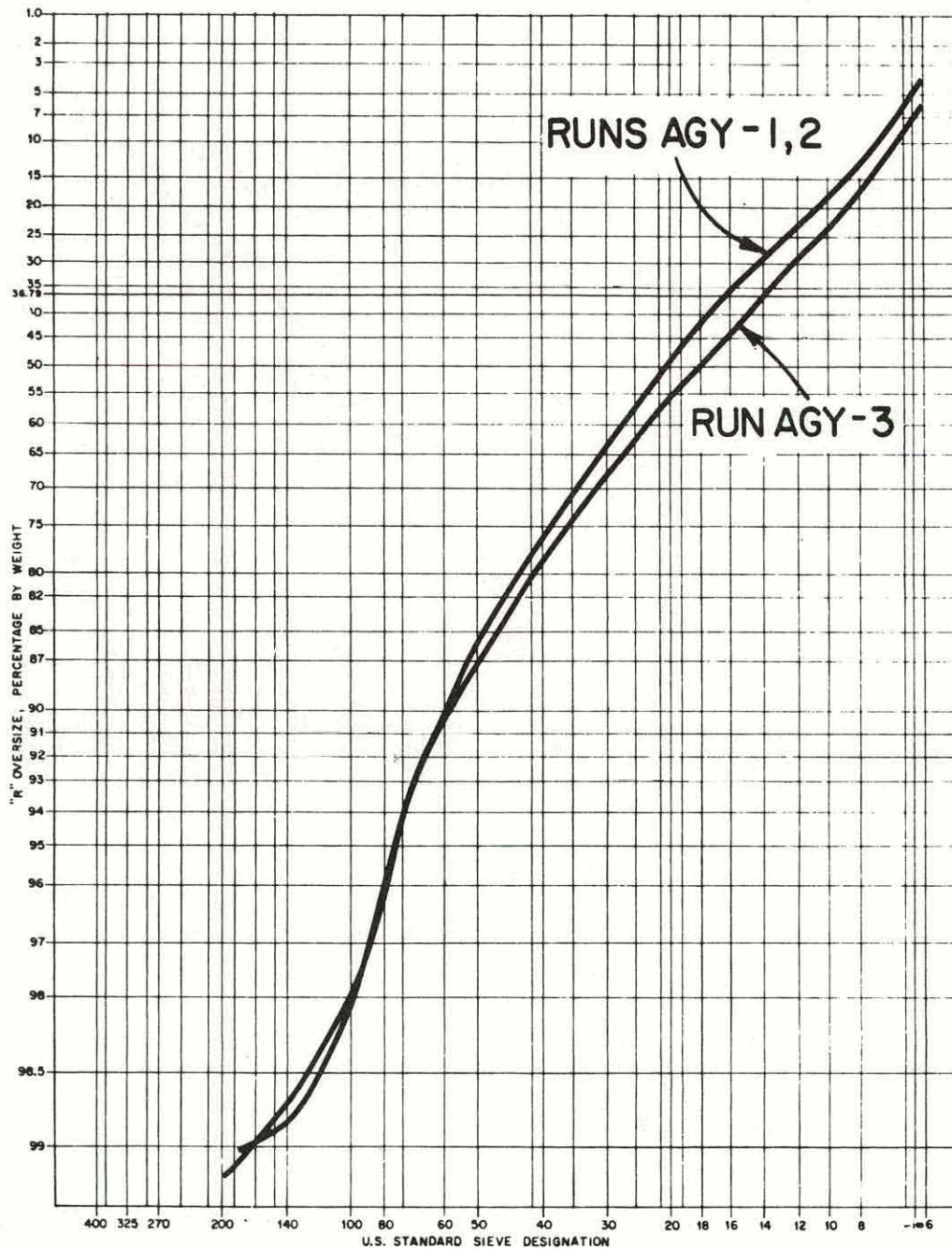


Figure C-21. SIZE DISTRIBUTION OF RESIDUE PLUS FINES
(Runs AGY-1, 2, and 3)

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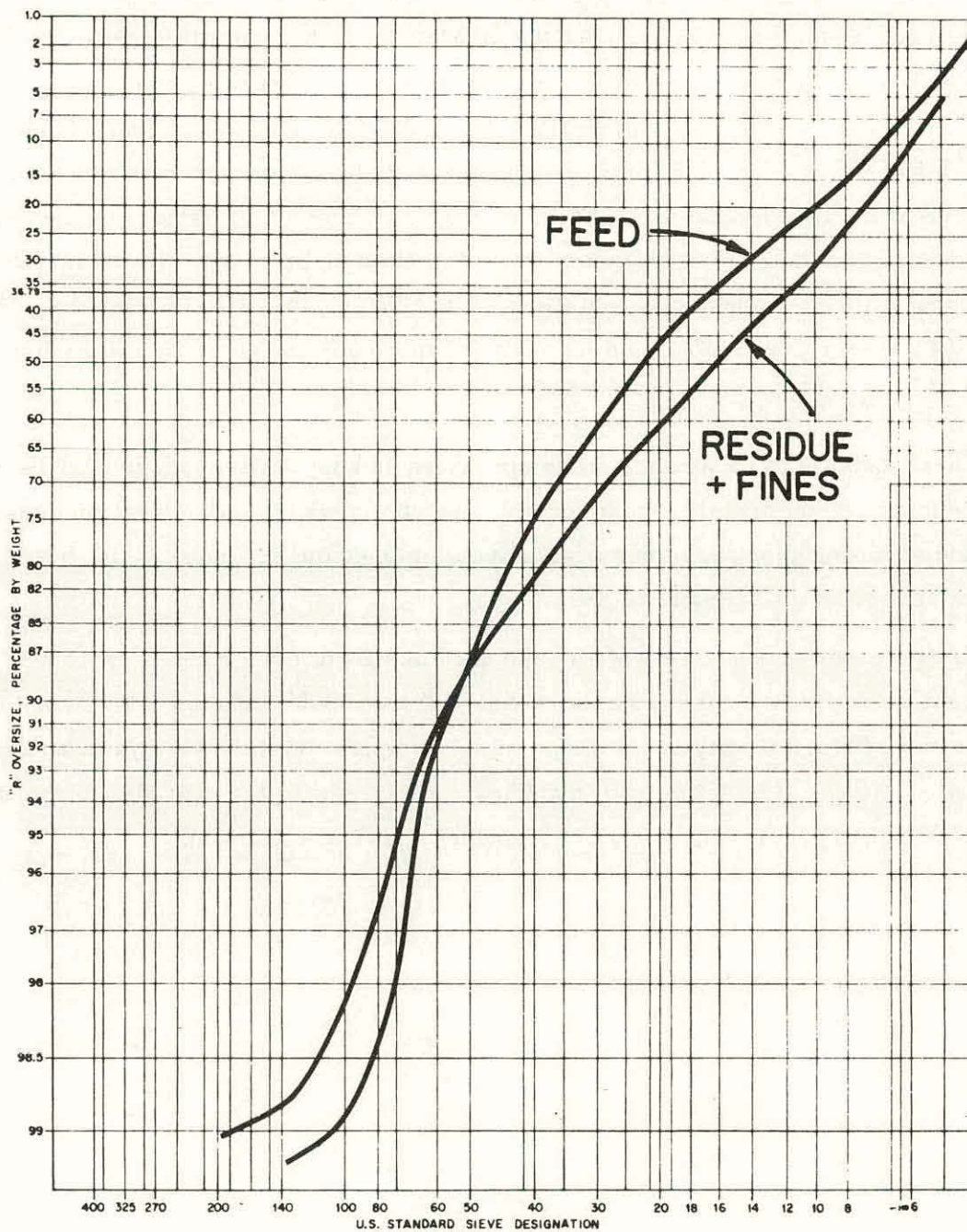


Figure C-22. SIZE DISTRIBUTION OF FEED AND RESIDUE PLUS FINES FROM RUN AGY-3

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The size distributions of feeds for Runs AGY-1, 2, and 3 are shown in Figure C-20. The distributions for AGY-2 and AGY-3 are almost indistinguishable on the plot, whereas the distribution for AGY-1 shows some deviations from the results. The residue plus fines distributions shown in Figure C-21 show distributions from Runs AGY-1 and AGY-2 to be indistinguishable, but the distribution of AGY-3 residue plus fines shows deviations from those results. The overall deviation between residue plus fines and feed, plotted for Run AGY-3 (Figure C-22), shows deviation of a larger magnitude than those found between feeds and residue plus fines. It appears that the difference between feeds and residue plus fines distribution is not merely because of random variations in the data. The magnitude of the latter variation seems large enough, however, to warrant further replications of these experiments to determine the magnitude of the variations. The results of further experiments could be analyzed more accurately if this were done.

Conclusion of Coke Tests

There appears to be a region of temperature and ash level conditions of Bethlehem coke in which agglomeration takes place and a region in which agglomeration does not take place. These regions have been tentatively mapped out in Figure C-19, based on the results of runs AGX-11 through AGX-19.

Preliminary results of runs with coke obtained from AAG Run 127 with a 47.5% ash content indicate that agglomeration takes place at 2000°F and 60 minutes retention time at a superficial velocity in the range of 1.1 to approximately 1.4 ft/s in a nitrogen atmosphere. We found indications of significant variations in the size distributions of feeds and residues plus fines in runs of ostensibly identical conditions.

ASH CHEMISTRY (PETROGRAPHIC STUDIES AND COAL AND COKE CHEMISTRY STUDIES)

Experimental Methods

Petrographic samples were mounted, sectioned, and polished according to methods used in coal petrography. Most of the microscopic observations were made with 40X Achromat and 16X Antiflex oil immersion objectives on a Zeiss Universal microscope.

Iron oxides and sulfides can be identified by appearance, especially when more than one species of oxide or sulfide is present. Ferrous sulfide is brilliant, although not as brilliant as pyrite, has a yellowish cast, and is distinctly anisotropic. Ferrous oxide is dead white; magnetite is darker and blue-gray in color. X-ray emission in a scanning electron microscope, conducted at the Illinois Institute of Technology Research Institute or at McCrone Associates, was relied upon for initial identification of glass and crystal phases such as ferrous aluminum silicate, ferrous silicate, ferrous aluminate, calcium aluminum silicate, and quartz. These phases can usually be recognized under the optical microscope by reflectance, which varies roughly with the iron content, and morphology. Quartz has a reflectance similar to that of the epoxy resin mounting medium and is only slightly anisotropic; thus, it is not easy to identify under the optical microscope. Its most distinctive characteristic in ash is the presence of fractures from thermal shock.

Ash samples were analyzed by the lithium metaborate fusion method of Boar and Ingram⁴ with atomic absorption finish for aluminum, calcium, iron, silicon, and titanium. Sodium and potassium were determined by flame emission.

Sulfide content in chars and coke was determined by refluxing the ground sample with 6N hydrochloric acid, trapping evolved hydrogen sulfide in alkaline cadmium chloride solution, reacting this sulfide with standard iodine solution, and titrating the excess iodine with standard thiosulfate. It was assumed that the sulfide in coke and chars was present as ferrous sulfide, although it is possible that calcium also could form a sulfide. The percentage of the iron present as sulfide was calculated from this determination and from the determinations of ash and total iron in the ash.

Analyses and ObservationsAsh From Coke

Previous work on agglomerates from coke and FMC char was reported in a paper presented at the American Chemical Society Fuel Chemistry Division meeting of May-June 1977.⁶ Ash agglomerate beads from coke were shown to consist of unmelted ash particles embedded in a matrix believed to be molten in the gasifier. This is shown in the photomicrograph in Figure C-23, which was taken with oblique illumination at low magnification. The unmelted, unassimilated particles appear lighter than their background. The beads also contain many vesicles: Some are empty and appear black, others are filled with the mounting medium and thus have the same tone as the outside of the beads.

With vertical illumination, residual coke appears as highly reflectant particles. Only a small amount was present as small particles on the outside surface of the bead or on the inside surface of vesicles in the beads. The molten ash has a higher surface energy than the coke, and thus tends not to wet it.¹⁰ The low carbon content of the ash agglomerates was confirmed by analysis of a hand-picked sample in which only 1.9% carbon was found.

Under higher magnification, the matrix in most regions of the agglomerates appears as a glass containing small crystals (Figure C-24). X-ray emission in the scanning electron microscope indicated that the glass contains iron, aluminum, and silicon; the crystals are ferrous aluminate. The lack of the graininess typical of sintered clay and the presence of well-formed crystals within the matrix indicated that the matrix had been molten in the gasifier. A few regions are present that also contain small amounts of one or more other kinds of crystals: iron silicate, ferrous oxide, magnetite, and (probably) calcium aluminum silicate. These regions were observed in greater abundance in agglomerates from some of the later coke runs. A few small particles of ferrous sulfide are also present in the matrix and unassimilated ash.

Ash composition (major and minor oxides) and fusibilities of some coke feeds, bed samples, and product streams are shown in Table C-14. Additional analyses* on coke used in Runs 102 to 126 were obtained by X-ray fluorescence on the whole coke sample, because

* These analyses were obtained from the Illinois State Geological Survey.

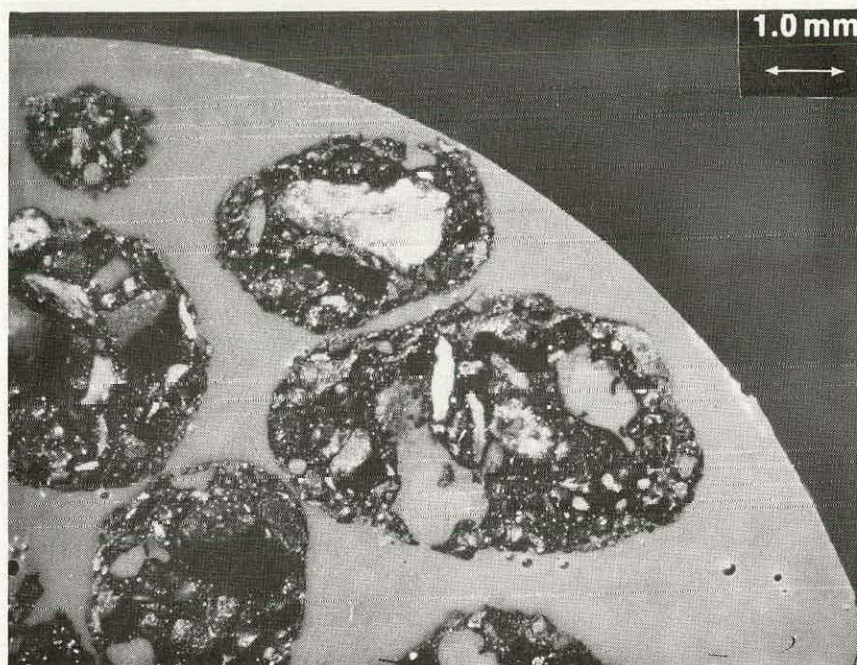


Figure C-23. PHOTOMICROGRAPH OF AGGLOMERATES FROM COKE



Figure C-24. PHOTOMICROGRAPH OF GLASSY SILICATE MATRIX WITH FERROUS ALUMINATE CRYSTALS IN COKE AGGLOMERATES

Table C-14. COMPOSITION AND FUSIBILITY OF ASH OF COKE AND ITS PRODUCTS

Sample	Feed	Feed	Feed ^a	Feed	Feed ^a	Feed ^a	C4 ^a	Bed ^a	Bed ^a	Bed ^a	Bed ^a	Avera	
AAG Run & Sample No.	111-39	111-40	116-22	125-32	127-18	127-71	118-16	124-	124-	124-	124-		
Hour	430	830	2400	1600	1400	2400	1630	2000	3000	400	1900		
Date	10/12/77	10/12/77	12/13/77	6/26/78	9/23/78	9/26/78	1/23/78	5/28/78	5/31/78	6/1/78	6/1/78		
Ash Composition							wt %						
SiO ₂	42.4	46.8	47.2	40.6	39.3	38.6	--	--	--	--	--	44.3	
Al ₂ O ₃	16.6	19.4	23.4	20.0	21.7	19.3	--	--	--	--	--	20.9	
Fe ₂ O ₃	30.0	25.2	18.2	21.8	21.9	23.1	--	--	--	--	--	24.3	
TiO ₂	0.8	0.8	1.10	1.0	0.96	0.82	--	--	--	--	--	0.95	
CaO	6.1	5.0	4.02	5.1	5.69	6.79	--	--	--	--	--	5.68	
MgO	1.7	1.6	1.49	1.6	1.66	1.67	--	--	--	--	--	1.69	
Na ₂ O	0.1	0.1	0.66	0.74	0.75	0.70	--	--	--	--	--	0.53	
K ₂ O	1.1	1.2	1.45	1.8	2.05	1.99	--	--	--	--	--	1.66	
SO ₃	--	--	1.93	4.4	3.81	4.83	--	--	--	--	--	--	
Total	98.8	99.7	99.5	97.0	98.8 ^b	98.8 ^c	--	--	--	--	--	100.0	
Ash Fusibility							°F						
Reducing Atmosphere													
IT	--	--	--	--	2070	2085	1960	2000	1980	2010	2020	2018	
ST	--	--	--	--	2180	2185	2065	2170	2240	2190	2180	2173	
HT	--	--	--	--	2290	2285	2180	2230	2300	2230	2220	2248	
FT	--	--	--	--	2395	2390	2295	2360	2420	2450	2420	2390	
Oxidizing Atmosphere													
IT	--	--	--	--	2300	2305	2345	2300	2300	2330	2310	2313	
ST	--	--	--	--	2400	2380	2415	2450	2450	2470	2450	2431	
HT	--	--	--	--	2490	2450	2480	2500	2500	2520	2500	2491	
FT	--	--	--	--	2575	2520	2545	2650	2590	2580	2610	2581	

^a Analysis and ash fusibility determined by Commercial Testing and Engineering. Other analyses determined at IGT.

^b Includes 0.12% BaO, 0.56% Mn₃O₄, and 0.38% P₂O₅.

^c Includes 0.10% BaO, 0.58% Mn₃O₄, and 0.44% P₂O₅.

^d Composition normalized to SO₃-free basis.

some question arose as to the origin of the coke. Results of these analyses are shown in Table C-15. Magnesium oxide could not be determined. The analyses have been normalized to the following basis: the sum of oxides found plus 1.6% MgO = 100. The analytical recovery, or sum of oxides found (including MgO) as a percent of the amount of ash found in the proximate analysis, is also given in Table C-15. The ash compositions are quite unlike those of coke made from Appalachian coals at the Bethlehem plant in Burns Harbor, Indiana, which the supplier has indicated to be the source. The 6% to 13% of calcium oxide in the ash of the coke feeds is typical of Illinois Basin coals, in contrast to 0.5% to 3% in Appalachian coals.

Pyrite in coal decomposes to ferrous sulfide (FeS_{1+x}) in coking. An analysis of three samples of coke fed to Run 127 (a definitive run on coke) showed that iron in the form of sulfide ranged from only 1.5% to 1.9% of the total iron. Thus, conversion of iron sulfide to oxide is not a factor in the agglomeration of coke ash, as it is shown to be for coal in data presented later.

Ash From Illinois Coal

A Run 116 C-4 discharge sample collected after the gasifier had run for 10 hours on Illinois No. 6 coal from the Peabody No. 10 mine was examined petrographically to determine whether agglomerates were being formed from coal ash or wholly or in part from residual coke ash still retained in the bed. The agglomerates found in the C-4 discharge consisted of unmelted clay and a few other mineral grains in a melted or well-sintered matrix that tended to assume a rounded form. The agglomerates were not as rounded as those from coke, but this can be ascribed to a smaller ratio of liquid phase to unmelted mineral grains. It is highly probable that the matrix material was formed from coal ash: The most convincing evidence is the prevalence of small particles of ferrous sulfide distributed throughout the matrix material and in some of the unassimilated grains; as discussed above, very little ferrous sulfide is present in the coke feed or, consequently, in ash agglomerates from coke. A second distinctive characteristic of the matrix material from this coal is the pattern of the polished surface as seen under the optical microscope (Figure C-25); many of the crystals exsolved in cooling are lath-shaped and have been identified in the scanning electron microscope as a calcium aluminum silicate, probably anorthite ($\text{CaO-Al}_2\text{O}_3\text{-2SiO}_2$). Iron silicate and calcium iron silicate crystals were also present.

Ash From Kentucky Run-of-Mine Coal

The composition and fusibility of the ash of samples of run-of-mine Kentucky coal and some product streams from it are shown in Table C-16. The average fusibility

Table C-15. COMPOSITION OF COKE ASH BY X-RAY FLUORESCENCE

Run and Sample No.	102-10	103-64	105-27	108-1	113-9	114-10	115-17	116-22	117-4	118-22	120-33	121-6	122-9	123-35	124-96	125-13	126-8
Hour	0400	0600	1200	1200	1800	0800	0800	2400	1800	1600	1800	0800	0200	2000	1800	1800	0800
Date	6/2/77	6/10/77	7/8/77	9/13/77	11/3/77	11/14/77	11/23/77	12/13/77	1/9/78	1/23/78	3/7/78	3/17/78	3/30/78	5/9/78	6/1/78	6/21/78	7/11/78
Ash Composition*	wt %																
SiO ₂	43.4	42.6	43.9	48.8	43.4	43.9	45.3	44.9	42.0	42.5	47.3	37.0	42.1	41.7	40.4	43.5	42.1
Al ₂ O ₃	21.4	21.9	19.9	18.2	22.1	23.0	23.4	22.3	21.1	18.7	24.5	16.6	19.1	18.7	19.2	21.2	20.8
Fe ₂ O ₃	24.1	24.2	24.9	14.9	23.9	24.7	19.0	21.4	27.9	26.2	17.9	29.4	22.1	25.2	24.2	22.2	22.8
TiO ₂	1.36	1.32	1.14	0.98	1.33	1.34	1.43	1.31	1.24	1.08	1.48	1.04	0.98	1.18	1.22	1.24	1.23
P ₂ O ₅	0.06	0.05	0.05	0.08	0.04	0.04	0.05	0.04	0.04	0.04	0.05	0.05	0.05	0.05	0.07	0.07	0.07
CaO	6.30	6.62	6.77	13.38	5.90	3.48	6.98	6.20	3.83	7.27	4.62	12.7	11.1	9.22	11.2	7.61	8.91
MgO	1.6																
Na ₂ O	0.10	0.19	0.19	0.69	0.29	0.27	0.45	0.66	0.67	1.08	0.59	0.21	0.6	0.49	0.51	0.38	0.28
K ₂ O	1.65	1.70	1.72	1.47	1.62	1.79	1.88	1.69	1.63	1.57	1.97	1.45	2.35	1.86	1.63	2.19	2.18
Analytical Recovery	103.2	105.8	104.9	102.0	105.1	111.9	111.7	106.5	104.5	91.4	102.3	96.6	101.9	101.9	98.4	105.2	105.5

* Normalized to 1.6% MgO, SO₃-free, basis.

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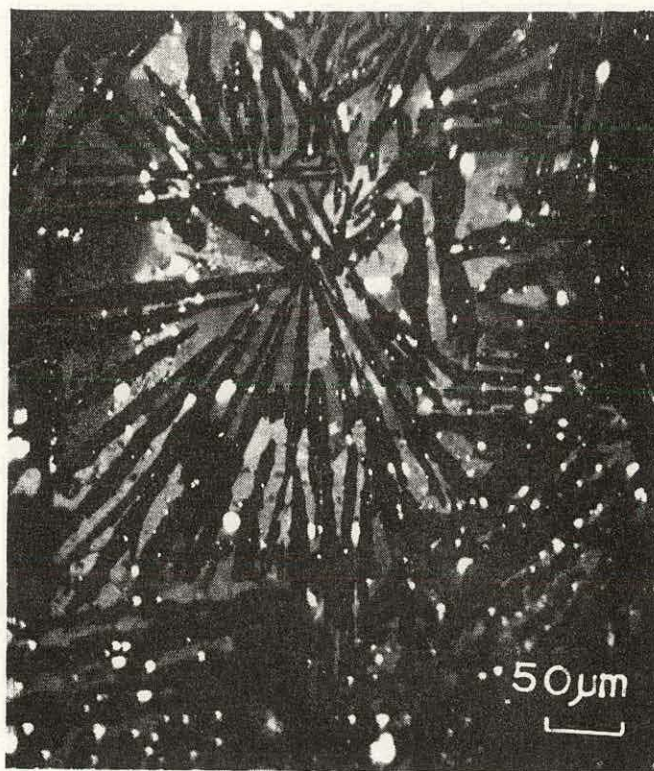


Figure C-25. PHOTOMICROGRAPH OF CRYSTALS IN AGGLOMERATES
FROM ILLINOIS COAL

Table C-16. COMPOSITION AND FUSIBILITY OF RUN-OF-MINE KENTUCKY COAL

Run and Sample Hour Date	Feed*	Feed*	Feed	Feed	Feed	Feed*	C4*	C4	C4	C7	C7	Second Stage Cyclone Dust		Avg ⁺
	(Sampled at the Mine)	(Grab Sample) 1/5/78	118-24 0400 1/24/78	121-7 1200 3/17/78	122-11 1600 3/30/78	(Grab Sample)	118-18 0430 1/24/78	122-51 1630 3/29/78	122-7 1630 3/30/78	122-13 1730 3/30/78	122-14 1830 3/30/78	122-20 1600 3/30/78	122- 2000 3/29/78	
Ash Composition	wt %													
SiO ₂	46.9	--	45.1	44.6	45.7	43.2	48.1	40.2	47.4	43.3	45.5	45.7	45.1	46.9
Al ₂ O ₃	18.7	--	17.4	15.9	15.9	17.5	17.8	13.7	16.1	15.0	15.6	16.7	18.3	17.2
Fe ₂ O ₃	21.0	--	21.2	21.3	23.4	23.5	21.7	33.3	23.4	26.0	26.0	23.3	21.8	24.8
TiO ₂	0.81	--	0.84	0.8	0.8	0.82	0.73	0.67	0.83	0.67	0.75	0.75	0.85	0.81
P ₂ O ₅	0.14	--	0.06	--	--	0.21	0.06	--	--	--	--	--	--	--
CaO	6.02	--	5.40	5.3	6.4	6.30	5.30	4.86	5.10	8.40	7.70	4.97	5.02	6.14
MgO	1.08	--	1.11	0.85	0.90	1.16	1.13	1.30	1.23	1.06	1.08	0.98	0.97	1.11
Na ₂ O	0.60	--	0.70	0.67	0.67	0.72	0.60	0.43	0.52	0.38	0.42	0.59	0.84	0.62
K ₂ O	2.50	--	2.45	0.85	0.88	2.60	1.50	2.72	3.48	2.48	2.57	3.04	2.95	2.43
SO ₃	1.96	--	4.34	--	--	3.80	1.23	3.0	2.8	2.1	1.4	4.4	3.7	--
Total	99.7	--	98.6	90.5	95.6	99.8	98.2	100.2	100.9	99.4	101.0	100.4	99.5	100.0
Ash Fusibility	°F													
Reducing Atmosphere														
IT	2025	1900	--	--	--	2025	1950	--	--	--	--	--	--	1975
ST	2155	2110	--	--	--	2110	2050	--	--	--	--	--	--	2106
HT	2295	2160	--	--	--	2185	2155	--	--	--	--	--	--	2199
FT	2455	2360	--	--	--	2240	2260	--	--	--	--	--	--	2329
Oxidizing Atmosphere														
IT	--	2160	--	--	--	--	2285	--	--	--	--	--	--	--
ST	--	2300	--	--	--	--	2375	--	--	--	--	--	--	--
HT	--	2350	--	--	--	--	2465	--	--	--	--	--	--	--
FT	--	2520	--	--	--	--	2565	--	--	--	--	--	--	--

* Analysis and ash fusibility determined at Commercial Testing & Engineering. Other analyses determined by IGT.

+ Composition normalized to an SO₃- and P₂O₅-free basis.

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temperatures of run-of-mine coal are slightly lower than those of coke, but the difference is probably not significant.

A mineralogical analysis of a sample of run-of-mine Kentucky coal was obtained from the Illinois State Geological Survey. The sample yielded 24.09% low-temperature ash (mineral matter) compared with 19.2% high-temperature ash. Results of the analysis are as follows:

<u>Component</u>	<u>Mineral Matter, wt %</u>	
Pyrite	24	
Quartz	12	
Calcite	8	
Kaolinite	13	Ratio determined on -2 μ m fraction. Coarser particles of clay, pre- sent in minor amounts, may differ in compo- sition.
Illite Clay	30	
Expandable (Mixed Layer) Clay	13	
Total	100	

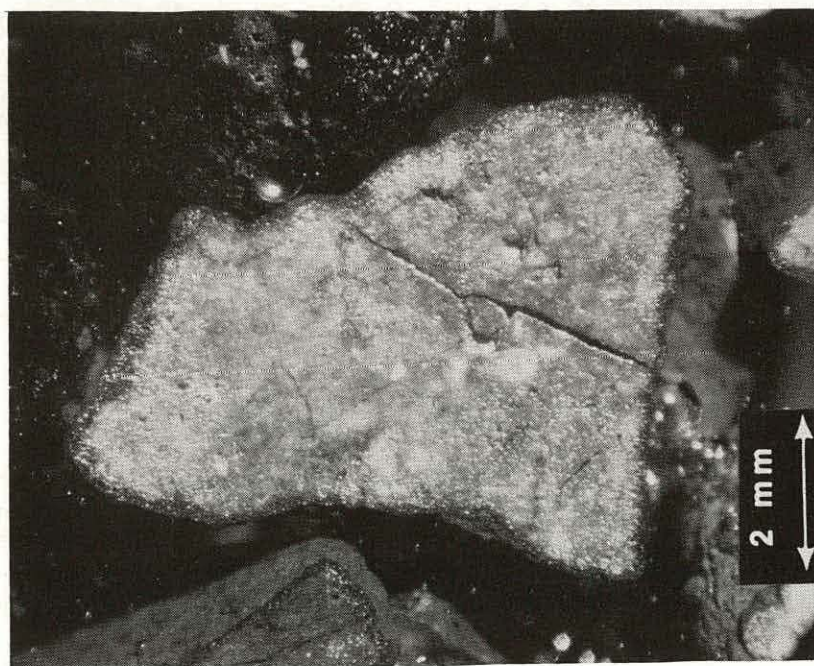
This distribution is typical of the mineral matter found in Illinois Basin coals, except that illite is higher and expandable clays lower than average.¹¹

The form of iron in the feeds was also investigated. Iron is most commonly present in coal as pyrite (FeS_2) but also occurs as carbonate and as part of the structure of illite clay. Iron in the latter two nonpyritic forms is soluble in hydrochloric acid, but pyrite is not. Determinations by this method were made on a number of samples of the coal fed to Run 124; results are shown in Table C-17. The nonpyritic iron content varied from 16% to 36% of the total iron and averaged 23.5%. If the iron is present in illite clay, the clay is likely to melt or sinter at lower temperatures than it would otherwise.

Samples of C-4 discharge from Run 122, obtained during periods when run-of-mine Kentucky coal was fed, were examined petrographically. The particles have dispersed ferrous sulfide in the iron-rich, low-melting matrix. Some other structural features of the particles differ from those of Illinois coal, as well as from those of coke. About three-quarters of the particles have an iron-rich, low-melting coating on the outside of an iron-poor, aluminosilicate-rich interior (Figure C-26). The uncoated particles and the interior regions of coated particles are similar: Some are uniformly light in color, others are uniformly light in color except for dark streaks or cracks that suggest the boundaries of agglomerated subparticles, and still others are uniformly dark or mottled (Figure C-27). Most of these particles or regions appear to be well sintered or even partially melted. A few particles of char and particles consisting only of iron sulfide and/or oxide are also

Table C-17. AMOUNT AND FORM OF IRON IN RUN-OF-MINE
KENTUCKY COAL FED TO RUN 124

AAG Sample No.	Hour/Date	Ash Content, wt %	CaO in Ash, wt %	Fe ₂ O ₃ in Ash, wt %	Nonpyritic Iron, % of Total Iron
124-198	1600/6-2	22.2	5.85	23.1	36.4
124-199	400/6-3	16.4	-	22.5	22.5
124-237	800/6-3	15.1	-	24.2	24.3
124-200	1600/6-3	16.4	6.81	20.6	23.6
124-201	400/6-4	16.3	-	22.4	22.4
124-202	1600/6-4	16.5	7.05	20.7	15.5
124-203	2000/6-5	18.2	6.38	23.5	22.7
124-204	400/6-6	16.9	-	23.5	28.4
124-205	1200/6-6	16.3	5.48	23.9	24.9
124-206	400/6-7	17.8	-	22.8	24.0
124-207	1200/6-7	16.5	6.18	21.2	21.6
124-208	400/6-8	19.5	-	24.5	21.6
124-238	800/6-8	17.2	-	24.1	22.5
124-209	1200/6-8	14.4	6.22	21.5	19.9
124-210	2000/6-8	18.3	-	23.0	20.3
	Avg.	17.2	6.28	22.8	23.4

Figure C-26. PHOTOMICROGRAPH OF CROSS-SECTION OF A SHALE
PARTICLE WITH AN IRON-RICH COATING IN THE C-4
DISCHARGE FROM RUN 122

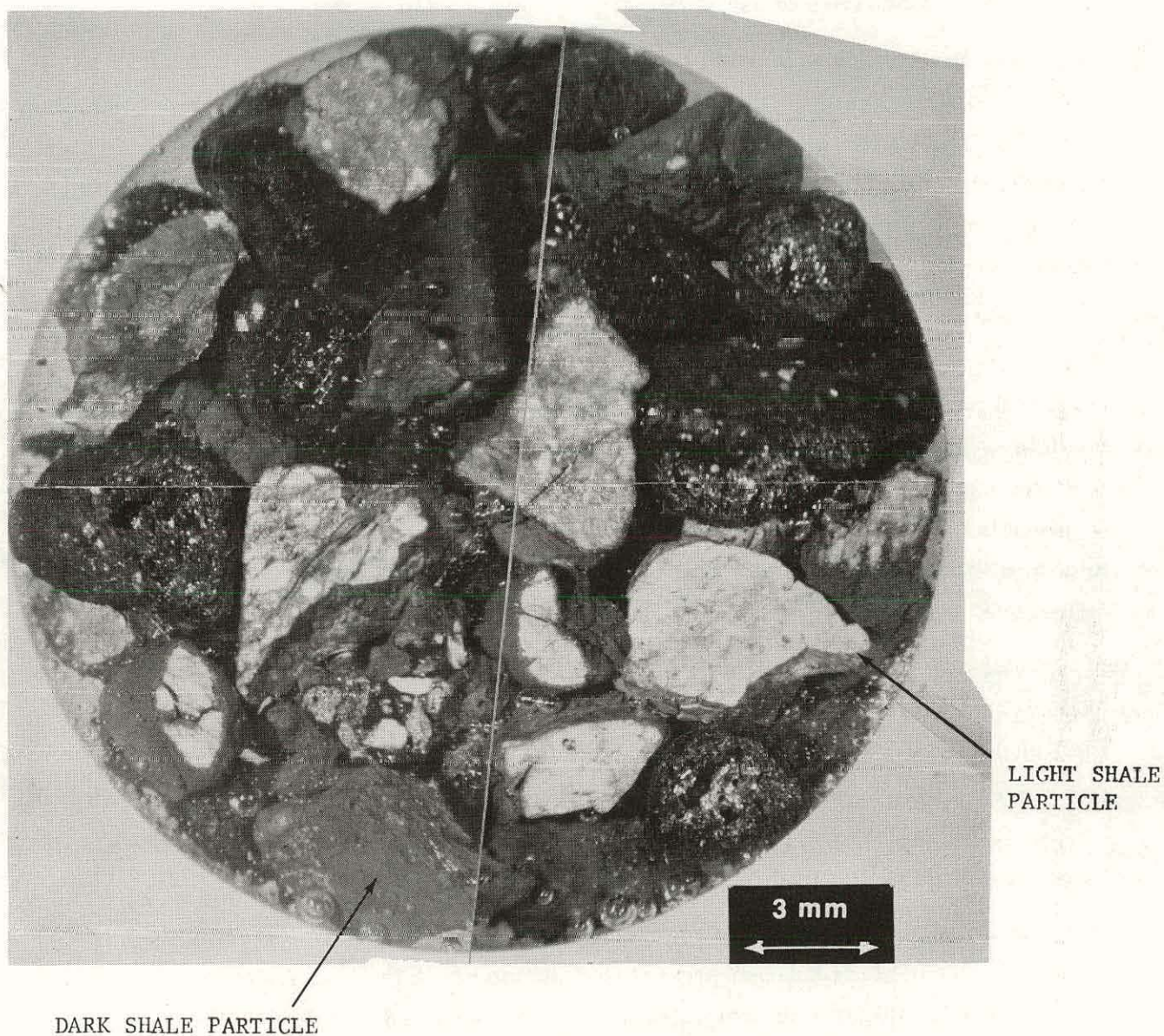


Figure C-27. PARTICLES IN THE 6-12 MESH FRACTIONS OF C-4
DISCHARGE FROM RUN 122
(The particles have been mounted
and sectioned petrographically.)

present. The same kind of particles are also present in the side discharge (C-7) samples. Very few particles are composed of a continuous iron-rich, low-melting matrix containing small unmelted particles of ash.

The iron-rich, low-melting coating of the particles contains mostly crystals of calcium aluminosilicate, iron silicate, and ferrous sulfide, with lesser amounts of calcium iron silicate and magnetite (Figure C-28). These compounds are the same as those found in agglomerates from Illinois coal; however, the crystal habit pattern is different in that few calcium aluminosilicate blades are present in the coatings on the particles from the Kentucky coal.

The uniformity within and the diversity among particles of the aluminosilicate or shale-derived parts of the C-4 discharge particles have been noted. Some coarse (+6 mesh) particles of the coal spread on a plate were ashed so that ash from a single particle could be observed. The ash from a single particle is typically uniform in color, but varies greatly from particle to particle. This is believed to account for the particle-to-particle diversity of the discharge particles. Examination of the particles by electron microprobe did not show any difference in composition that would account for the color difference.

C-7 discharge samples from a number of coal-feeding periods of Run 124 were examined petrographically. Very few of the large shale particles that were present in the Run 122 C-4 discharge were found, and only a small amount of silicate matrix phase was observed on the outside of some of the ash particles that were present. Most of the ash particles consist of only partially rounded noncrystalline aluminosilicate (shale) cores surrounded by ferrous sulfide (Figure C-29).

A layer of ferrous sulfide on an aluminosilicate core has also been noted frequently in samples from operation with FMC char.⁶ A layer inside of an iron silicate matrix layer is also not uncommon, as shown in Figure C-28. (The outermost layer in this photomicrograph is magnetite.) Thus, it appears that formation of a ferrous sulfide layer is a precursor to formation of some or all of the low-melting matrix phase. It might be formed by oozing of the ferrous sulfide out of porous shale particles to the surface or by spreading of ferrous sulfide after contact between a shale particle and a ferrous sulfide particle. It is also possible that ferrous sulfide itself can act as an agglutinating agent, as suggested by the morphology of a few particles in the Run 124 samples. These particles consist of distinctly separate shale regions joined by ferrous sulfide, as shown in Figure C-30.

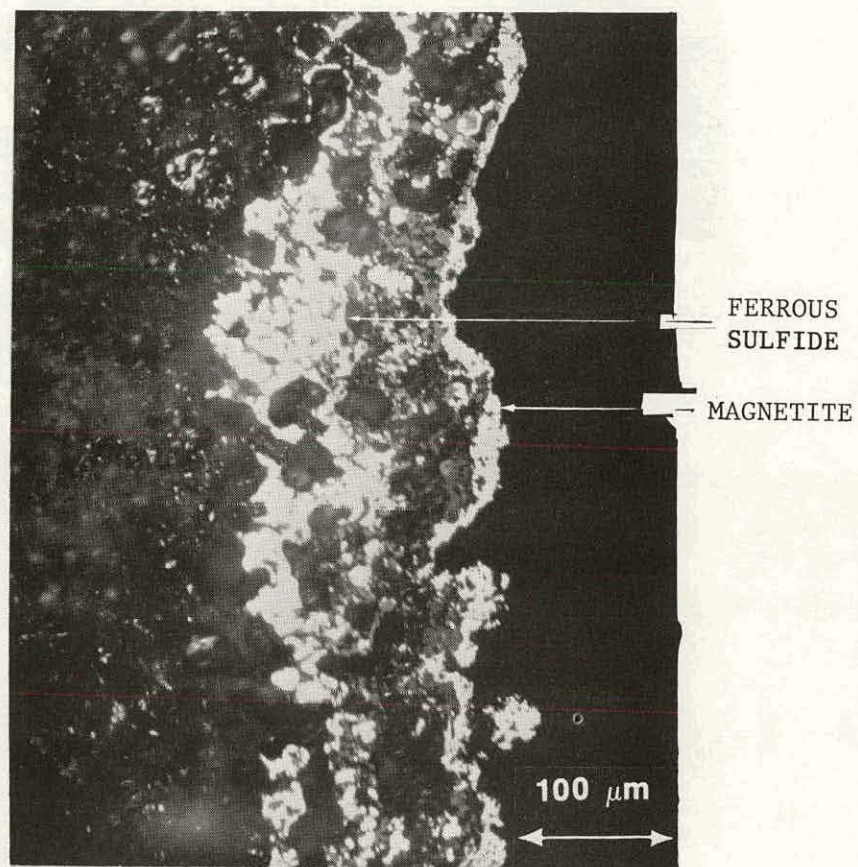


Figure C-28. CRYSTALS IN THE IRON-RICH COATING FOUND ON SOME PARTICLES IN C-4 DISCHARGES FROM RUN 122

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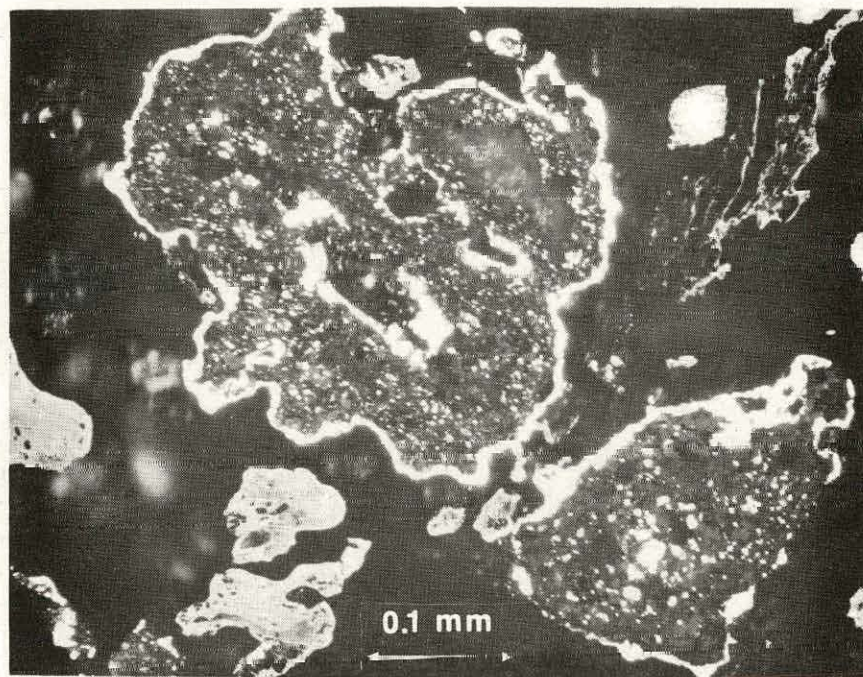
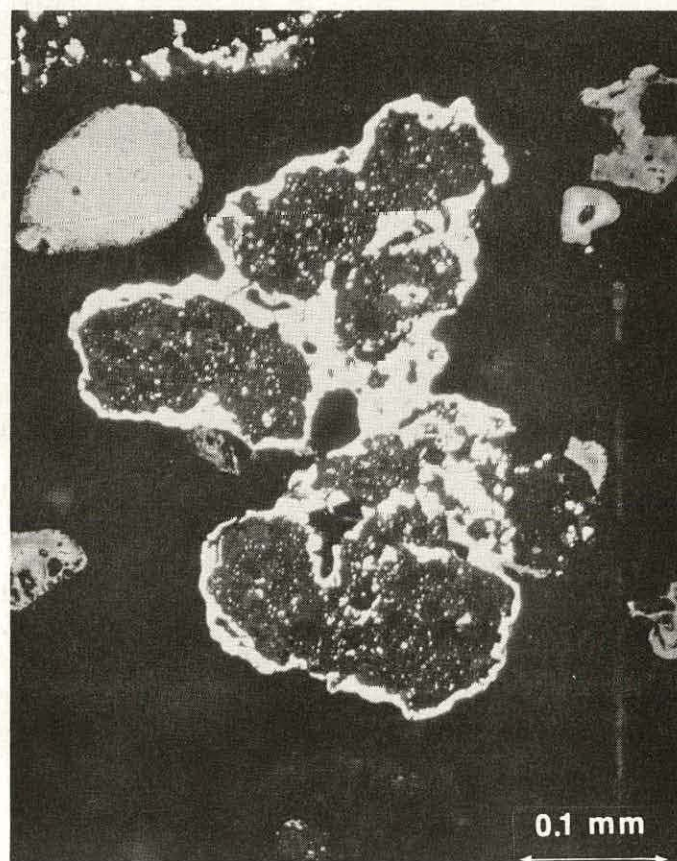


Figure C-29. PHOTOMICROGRAPH OF TYPICAL ASH PARTICLES
IN C-7 DISCHARGE FROM RUN 124



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Figure C-30. FeS JOINING SHALE PARTICLES IN C-7 DISCHARGE
FROM RUN 124

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The spreading of ferrous sulfide on the surface of aluminosilicates indicates that the sulfide has a lower surface energy than the aluminosilicates. However, the ferrous sulfide very likely has a higher surface energy than the char and probably does not wet it.

Thus, a Godel mechanism¹⁴ might be a mode of contact between the sulfide and aluminosilicate (shale) particles; that is, there is a possibility that ferrous sulfide in the interior of a char particle becomes molten and exudes to the surface of the char particle where it may be picked off when touched by an ash particle. To investigate this possibility, a sample of +20 mesh particles of a Run 124 C-7 discharge sample with 73% of its iron in the form of ferrous sulfide was chosen. Particles were mounted in the scanning electron microscope for examination as is, that is, without sectioning. Only a single ferrous sulfide particle (and that one angular rather than rounded as is characteristic of the Godel mechanism) was detected on 1 of 15 char particles that were examined; therefore, such a mechanism appears very unlikely.

Bed discharge (C-7) samples from most of the steady-state periods of Run 124 on run-of-mine Kentucky coal were chemically analyzed for iron sulfide. Very little discharge through the venturi occurred during these periods; thus, the C-7 was representative not only of the bed but also of the total ash product, if that removed as cyclone dust is neglected. The percentage of total iron present as sulfide was calculated on the assumption that all of the hydrogen sulfide came from iron sulfide and none from calcium sulfide. The treatment with hydrochloric acid dissolved about 90% of the iron and practically all of the calcium; thus, the method appears adequate for the purpose of this program, although small amounts of iron sulfide occluded in sintered clay may not be measured. Results are shown in Figure C-31. The earlier periods, up to June 5, 1978, show little or no conversion of sulfide, because the coal feed already had 23% nonsulfidic iron. Later periods show increased conversion, especially during operation with steam and oxygen on June 8, 1978.

Two samples each of second- and third-stage cyclone dust were analyzed for sulfide, with the following results:

	<u>0800 hours on 6/3/78</u>	<u>0800 hours on 6/8/78</u>
	----- Sulfidic Iron, % of total iron -----	
Second-Stage Dust	81	66
Third-Stage Dust	41	28

These samples also showed less sulfidic iron in the latter period.

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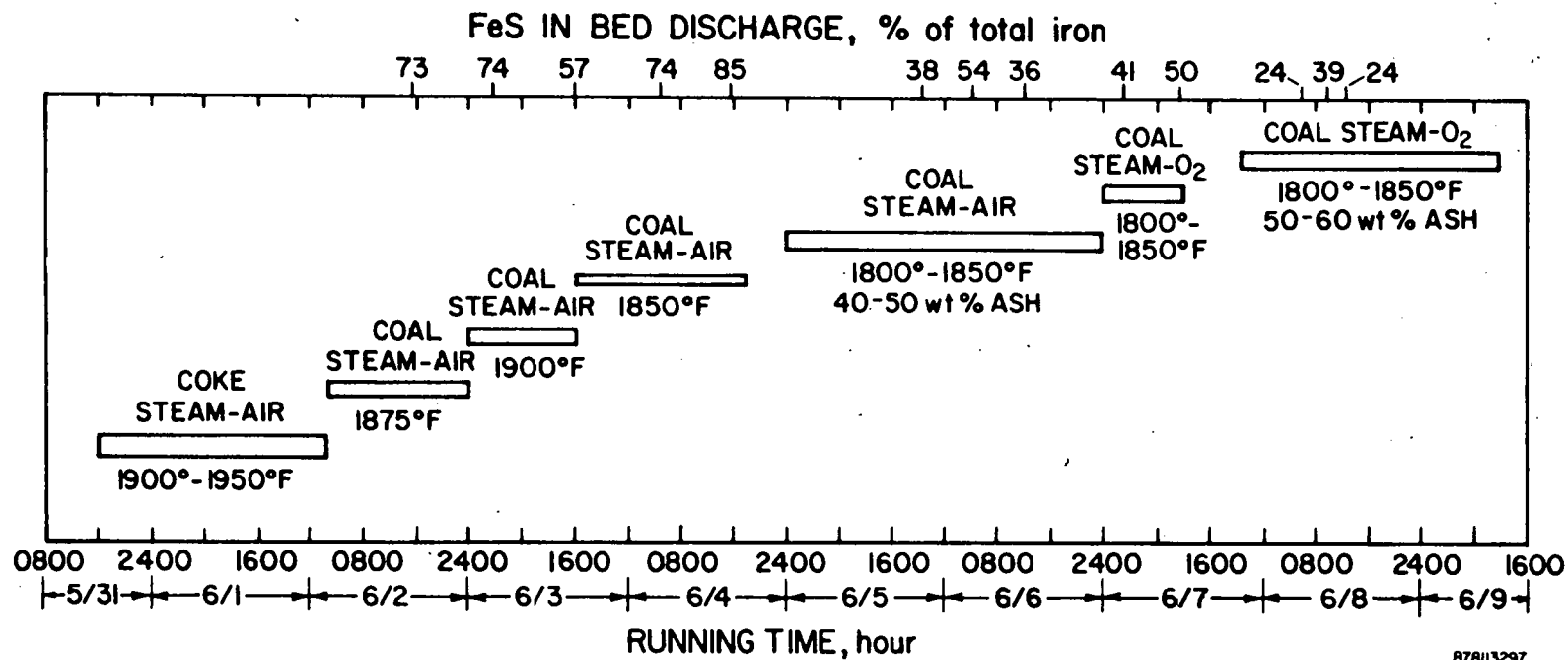


Figure C-31. IRON SULFIDE AS PERCENT OF TOTAL IRON IN BED DISCHARGE DURING RUN 124

Methods of determining the size distribution and density of ash particles in bed and bed discharge samples were investigated. Tests were made on C-7 bed discharge samples from Run 124 by ashing 50 grams or more of the sample at a temperature low enough (about 1400° to 1500°F) to preclude further sintering or agglomerating of the ash; a sieve analysis was then obtained on the ash. A similar test was performed on coal samples, but at a lower furnace temperature (about 1100°F) because coal burns faster than the bed discharge and tends to become hotter than the ashing furnace.

Iron sulfide is converted to ferric oxide in the ashing. It is suspected that the sulfide conversion might tend to disrupt particles or make them more friable and thus liable to lose material to fines during the sieving. Accordingly, a different method of removing carbon — treatment with steam at 1850°F for 3 to 6 hours — was investigated. Observation under a low-power microscope showed that no further agglomeration occurred during the steam treatment.

Results from the ashing and steam gasification tests are shown in Table C-18. Ashing tests on coke and coal, which are shown in the first columns, will be discussed later in this report. Duplicate ashing tests were performed on three C-7 samples, identified as 1400 hr/6-3, 2400 hr/6-6, and 1600 hr/6-6. Agreement between duplicates is within 3% for all but a few values; this is reasonable in view of the small sample sizes (24 to 45 grams). Similar agreement is shown for a duplicate gasification test on the 2400 hr/6-5 test. One of these gasification tests was run for 3 hours and the other for 6 hours. Only 0.8% carbon remains in the residue from the 6-hour test; the agreement between the tests shows that 3 hours of gasification was probably sufficient. However, subsequent tests were run for 6 hours.

A comparison of ashing tests with gasification tests shows that significantly less coarse particles and more fines are obtained in the ashing tests. These results were interpreted to mean that fines were produced from the coarse particles in the course of the ashing and sieving of the ash and, therefore, that the ashing tests are less satisfactory.

These tests were performed primarily to characterize the ash particles in the gasifier bed and to determine comparative amounts of agglomeration in different periods of operation. It was concluded that the gasification test could, in general, be used for this purpose, although one would want to be able to increase the size of the sample. To characterize the ash of the original coal, a large sample would have to be gasified or otherwise decarboned because the ash content of the coal is much less than that of the bed discharge samples. However, some conclusions can be drawn from the results of the

Table C-18. SIZE DISTRIBUTION AND PARTICLE DENSITY OF ASH PARTICLES

Run No.	118	118	124	124	122	124	124	124	124	124	124	124	124	124	124	124
Sample	Coke Feed	Coal Feed	Coal Feed	Coal Feed	C7	C7	Bed	C7	C7	C7	C7	C7	C7	C7	C7	C7
Hour and Day	--	--	1200/6-4	1600/6-6	1230/3-30	1830/3-30	1400/6-3	1600/6-3	2400/6-5	1600/6-6	600 to 1210/6-8	1600/6-6	600 to 1210/6-8	1600/6-6	600 to 1210/6-8	600 to 1210/6-8
Method	Ashing*	Ashing	Ashing	Ashing	Ashing	Ashing	Ashing	Ashing	Gasif.	Ashing	Ashing	Ashing	Gasif.	Ashing	Ashing	Gasif.
Sieve No., USS	cumulative wt %															
on 6	5.7	9.7	--	--	5.1	4.3	--	--	--	--	--	--	2.5	--	2.4	0.3
12	14.4	20.4	15.1	23.8	17.0	17.6	0.9	1.3	2.5	0.4	15.2	11.6	15.2	11.3	7.8	9.2
20	23.1	29.8	26.1	35.0	35.0	42.6	5.7	5.9	14.1	5.5	25.8	23.2	33.3	30.9	17.3	18.8
40	31.0	38.2	35.0	41.9	63.9	81.4	29.1	29.7	45.5	18.0	39.9	37.7	49.8	52.0	29.3	31.8
70	40.4	47.8	47.6	50.8	73.0	91.4	63.3	64.1	84.1	51.0	--	--	78.5	78.7	59.9	63.4
140	57.0	60.3	61.7	61.0	83.4	93.5	74.7	74.8	94.6	73.0	--	--	92.4	92.4	75.6	79.2
200	--	78.7	69.4	67.5	--	--	79.4	79.7	96.8	81.0	--	--	95.9	--	80.7	84.5
230	--	--	72.7	72.2	--	--	82.0	83.0	--	--	83.7	85.5	96.5	--	82.5	86.9
270	92.5	--	--	--	88.3	95.0	--	--	--	87.6	--	--	--	97.3	--	--
Pan	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0	100.0
Ash in C7, % of total					35.0		72.3								84.0	
Particle Density, [†] lb/ft ³	--	--	--	--	--	--	119.0	--	108.0	--	--	--	120.0	--	--	128.0

* Coke, C7, and bed samples ashed at 1400°F furnace temperature; coal at 1100°F.

† Gasification with steam at 1850°F for 3 hours; all other samples for 6 hours.

† Determined on -6 +100 mesh fraction.

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ashing tests run on the coal as shown in Table C-18. A substantial amount of the mineral matter of the coal is present as relatively coarse particles; 35% or more of the ash is coarser than 40 mesh. This conclusion is also supported by results of a sink-float test on the +40 mesh fraction of the coal; at a density of 1.6 g/cm^3 , about one-third of the ash of the original coal was found in the sink sample.

No evidence for agglomeration is found by comparing ashing results on the coal with ashing results on bed discharge samples, especially if the amounts of coarse fractions in the latter are normalized to the original coal basis according to the amount of ash appearing in the bed discharge. However, this correction is predicated on the supposition that none of the coarse ash particles are attrited sufficiently to make a significant contribution to the ash in the cyclone dust. It is also possible that the coarse ash particles of the bed discharge had shrunk because of sintering in the gasifier. Particle densities were obtained on -6+100 mesh sieve fractions of some gasification and ashing residues from bed discharge samples (Table C-18), but none were obtained on ash of the feed.

Ash From Washed Kentucky Coal

Data on the composition and fusibility of the ash of washed Kentucky coal fed to Run 130 are presented in Table C-19 together with analyses obtained on a sample taken from the piled coal of the first shipment, average values for the run-of-mine coal fed to previous runs, and analyses of solid product samples from Run 130. Similar data on feed and products from Run 131 are presented in Table C-20.

The ash of the washed coal is very similar to that of the run-of-mine coal except that it appears to have about 3% less ferric oxide. Note that the amount of calcium oxide obtained on the coal pile sample is quite low (approximately 3%) in comparison to that of samples taken during Run 130, and that in the latter values there is more variation than for other elements. Also note that the C-4 (venturi) discharge samples are greatly enriched in calcium oxide while the cyclone dust samples are depleted. We tentatively attribute the latter effect to occurrence of the calcium oxide in the coal as particles of calcite that are large enough to be selectively retained in the bed even after their decomposition to oxide. The variability in the amount of calcium found in the coal can be attributed to the occurrence of calcite as linings of cleat (fracture) surfaces that are unevenly distributed in the coal bed. Segregation of these particles in the coal pile may also occur.

Nonpyritic iron in the coal feed samples ranged from 19.3% to 28.8% of the total iron. The average amount is about equal to that found in the run-of-mine coal. The feed coals from Run 131 were analyzed for carbon dioxide to determine whether the nonpyritic iron might be

Table C-19. COMPOSITION AND FUSIBILITY OF ASH IN RUN 130 FEED AND PRODUCTS

Run And Sample No. Sample Hour Sample Date	ROM Coal, avg	Washed Coal					C4			Second-Stage Cyclone Dust		
		Coal Pile		130-93 2000 11/4/78	130-96 2400 11/5/78	130-99 2200 11/6/78	130-83 1830 11/6/78	130-85 0200 11/7/78	130-86 0600 11/7/78	-12+20 Fraction 1000 11/7/78	130-42 2200 11/6/78	130-45 1400 11/7/78
		(IGT Analysis)	(CTE Analysis)									
Ash Content of Sample	--	7.9	8.0	8.5	8.4	7.4	55.0	73.9	81.6	95.4	19.8	18.7
Ash Composition												
SiO ₂	46.9	45.1	47.0	44.0	45.4	46.9	45.0	44.2	43.2	41.3	52.3	51.4
Al ₂ O ₃	17.2	19.8	22.2	17.7	18.2	18.6	18.4	18.0	17.4	16.7	19.3	19.0
Fe ₂ O ₃	24.8	21.4	19.8	21.7	21.0	21.6	20.8	24.1	25.4	25.8	18.7	18.8
TiO ₂	0.81	1.0	1.03	0.9	0.9	0.9	0.6	0.5	0.5	0.6	1.2	1.2
CaO	6.14	2.9	3.00	7.5	6.6	4.8	13.5	12.6	14.2	14.5	3.2	3.4
MgO	1.11	0.84	1.02	0.9	0.9	1.0	1.1	1.0	0.9	0.9	0.9	0.9
Na ₂ O	0.62	0.46	0.53	0.4	0.4	0.4	0.3	0.3	0.2	0.2	0.6	0.6
K ₂ O	2.43	2.3	2.55	2.3	2.4	2.5	1.1	0.9	0.7	0.5	3.1	3.2
SO ₃	--	3.0	2.69	4.3	3.4	2.7	1.5	1.2	1.1	1.0	2.2	2.2
Total	100.0	96.8	99.98*	99.7	99.2	99.4	102.3	102.8	103.6	101.5	101.5	100.7
					% of total iron							
Nonpyritic Iron	23	10	--	28.8	19.3	23.3	--	--	--	--**	--	--
Ferrous Sulfide	--	--	--	--	--	--	11	4	3	1.3	22	37
Ash Fusibility												
Reducing Atmosphere												
IT	1975	2055	1990	2040	1960	2010	--	--	--	--	--	--
ST	2106	2170	2170	2105	2060	2095	--	--	--	--	--	--
HT	2199	2285	2220	2165	2160	2200	--	--	--	--	--	--
FT	2329	2405	2405	2230	2250	2295	--	--	--	--	--	--
Particle Density, lb/ft ³	***		--	--	--	--	--	--	--	159/0	--	--

* Includes 0.19% P₂O₅, 0.01% BaO.

** On total C4.

*** Determined on a sample ashed to remove char particles.

Table C-20. COMPOSITION AND FUSIBILITY OF ASH IN RUN 131 FEED AND PRODUCTS

Run and Sample No. Sample Hour Sample Date	Washed Coal			C-4	Bed	Second-Stage Cyclone Dust
	131-66 1800 12/9/78	131-71 1800 12/10/78	131 2200 12/10/78	131-80 1830 12/10/78	131-62 1800 12/10/78	131-89 1800 12/10/78
	wt %					
Ash Content of Sample *	13.2	10.6	12.7	80.4	59.9	18.9
CO ₂ Content of Coal	0.88	0.51	0.76	--	--	--
Ash Composition						
SiO ₂	42.8	47.2	40.4	43.4	45.3	52.5
Al ₂ O ₃	16.4	17.2	15.0	16.6	17.6	21.2
Fe ₂ O ₃	20.0	19.6	26.3	24.2	22.2	12.5
TiO ₂	1.0	1.1	0.9	0.8	0.7	1.5
CaO	10.2	6.1	9.2	10.7	10.2	3.1
MgO	1.0	1.0	1.4	1.5	1.2	1.0
Na ₂ O	0.8	0.8	0.7	0.8	0.7	1.2
K ₂ O	2.2	2.5	2.0	1.6	1.6	3.9
SO ₃	4.6	3.6	4.6	2.2	1.8	2.2
Total	99.0	99.1	100.5	101.8	101.3	99.1
	% of total iron					
Nonpyritic Iron	26.5	15.1	26.0	--	--	--
Ferrous Sulfide	--	--	--	5.5	4.5	8.4
	% of total calcium					
CaO as Carbonate	83	100	83	--	--	--
Ash Fusibility	°F					
Reducing Atmosphere						
IT	1980	2000	1970	--	--	--
ST	2100	2160	2130	--	--	--
HT	2130	2190	2150	--	--	--
FT	2170	2340	2190	--	--	--

* As ashed to prepare ash analysis sample.

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present in the form of carbonate. If all the carbon dioxide is attributed to calcium carbonate it accounts for 83% to 100% of the total calcium in the coal. Similar results were obtained from carbon dioxide determinations on some feed samples from Run 130. Thus, it does not appear likely that the nonpyritic iron is present as carbonate. Instead, we think it may be oxide and sulfate formed during the drying of the coal at the pilot plant.

The ash analyses of the product streams indicate that potassium is depleted in the C-4 discharge and is enriched in the cyclone dust. This effect can be attributed to the volatility of potassium oxide.

A sample of C-4 (venturi) discharge from Run 130 on washed Kentucky coal was examined. The carbon, hydrogen, and ash contents of sieve fractions from the sample are shown in Table C-21. Agglomerated ash is present in the form of small beads concentrated in the -12+20 mesh sieve fraction. Analysis of the ash of this fraction is shown in Table C-22. Petrographic examination of the beads confirms the absence of the large shale particles that were present in ash from run-of-mine Kentucky coal. The crystal pattern in the iron-rich silicate regions, as shown in Figure C-32, resembles that in the coatings on shale particles of the C-4 discharge from Run 122 on run-of-mine Kentucky coal. The crystals are composed mainly of ferrous silicate, ferrous calcium silicate, and calcium aluminum silicate; small grains of ferrous sulfide and ferrous oxide are also present. Analysis of the total C-4 discharge showed that only about 1% of the iron was present as sulfide; thus, almost all of the iron had been converted to oxide and silicate.

Analysis of C-4 samples from Run 131 (Table C-20) for hydrogen sulfide evolved on treatment with acid confirm that most ferrous sulfide (or other sulfide if present) has been oxidized; the amount of sulfide remaining corresponds to 3% to 11% of the total iron in the sample. Sulfide remaining in the second-stage cyclone dust corresponds to 22% to 37% of the total iron in the dust sample.

We also investigated the size distribution and density of ash agglomerates in the reactor bed. A C-7 bed discharge sample taken at 1465 hours on 12-7-78 of Run 131 was chosen for method development; at this time, the reactor was on the verge of being in ash balance with only a minimal rate of bed discharge. The sample was first scalped to remove the fraction coarser than 10 mesh; the -10 mesh fraction was placed in a fluidization column and partially fluidized with air. At velocities between 1.5 and about 2 ft/s the sample separated into two distinct levels: an upper, fluidized one containing char, and a lower, quiescent one containing agglomerates. After cutting off the air the upper fraction was vacuumed off, thus

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Table C-21. COMPOSITION OF C-4 DISCHARGE FROM RUN 130
(Sample Taken at 1000 Hours on 11/7/78)

	Sieve Fraction, U.S.S.					Total
	+6	-6+12	-12+20	-20+40	-40	
	wt %					
Amount	4.5	19.0	70.1	5.7	0.3	100.0
Composition						
Ash	32.05	51.88	95.43	87.63	--	83.45
Carbon	65.6	47.7	5.9	13.2	--	16.9
Hydrogen	0.35	0.21	0.04	0.07	--	0.09

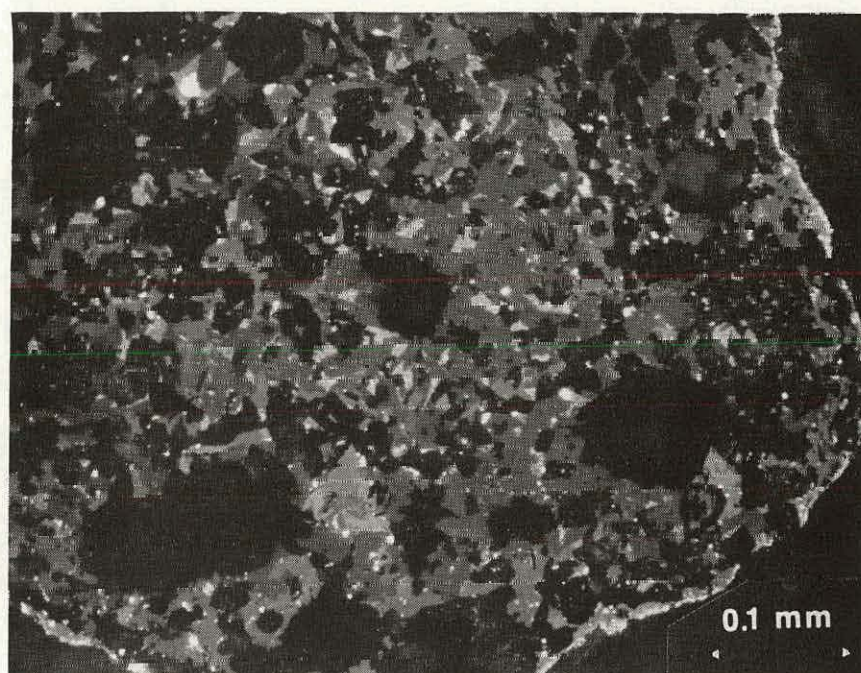


Figure C-32. PHOTOMICROGRAPH OF IRON-RICH SILICATES IN
SMALL ASH AGGLOMERATES FROM WASHED KENTUCKY COAL

Table C-22. CHARACTERISTICS OF REACTOR BED PARTICLES DISCHARGED
IN RUN 131 AT HOUR 1465 ON 12-7-78

Fraction of Total Sample	Total Sample	+10 Mesh Fraction	-10 Mesh Fraction				Loss (Fines)
			Light Fraction		Heavy Fraction		
			-10+20 Mesh Fraction	-20 Mesh Fraction	-10+20 Mesh Fraction	-20 Mesh Fraction	
			wt %				
	100.0	12.1	20.4	24.0	19.5	20.6	3.4
Sieve Analysis							
Sieve No., USS							
on 6	3.6	--		0.0		0.0	--
12	18.6	--		9.8		2.8	--
20	33.7	--		45.5		39.7	--
40	35.6	--		32.1		40.8	--
70	7.3	--		11.9		1.5	--
140	1.1	--		0.5		0.1	--
270	0.3	--		0.0		0.0	--
Pan	0.1	--		0.2		0.1	--
Composition							
Ash	57.79	30.31	22.0	39.3	95.6	100.1	--
Carbon	41.5	68.4	76.0	59.1	5.5	1.3	--
Hydrogen	0.13	0.17	0.25	0.19	0.04	0.02	--
Bulk Density	48.7	29.6	26.1	lb/ft ³			
Particle Density	--	--	--	41.3	80.6	94.6	--
				98.3	157	170	--

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accomplishing separation into light and heavy fractions. The two fractions were further separated into fractions coarser than and finer than 20 mesh. Particle size, composition, and density data on the fractions are presented in Table C-23.

A cursory petrographic examination indicated that the light fraction contained very few agglomerates and that the heavy fraction contained very little char; the latter conclusion is also shown by the ash and carbon contents. Thus, a sharp separation was obtained.

Agglomeration Tests

Three samples of ash particles (agglomerates from coke, agglomerates from washed Kentucky coal, and shale particles found in the C-4 discharge from Run 122 on run-of-mine Kentucky coal) were subjected to agglomeration tests. In these tests a small (1 to 2 centimeters in diameter) heap of the material in a small alumina or porcelain boat is inserted in a hot furnace and left there for about 10 minutes, which is sufficient time for the test heap to attain furnace temperature but not to sinter appreciably if it does not melt. Tests were run in both a reducing atmosphere (60:40 ratio of carbon monoxide to carbon dioxide) and in air. The agglomeration tests are designed to show the temperatures required for the ash particles or agglomerates to 1) become stuck together and 2) run together and form a rounded agglomerate or bead in the gasifier. The two temperatures are termed "agglomerated" and "fused", respectively, in Table C-23, which shows the results of the test.

Agglomeration (sticking) temperatures in the reducing atmosphere for the three samples are essentially equal at about 1920°F. The temperature for fusion or rounding of beads is higher (about 1980°F) and is also higher than the bed temperature for both coke and coal. Thus it appears that agglomeration can occur in the bed but that the formation of rounded beads can occur only in higher temperature regions of the reactor.

Table C-23. AGGLOMERATION TEMPERATURES OF ASH PARTICLES

Sample	Reducing Atmosphere (60:40 CO/CO ₂)		Air	
	Agglomerated	Fused	Agglomerated	Fused
	°F			
Ash Beads From C-4 Discharge, Run 127 (Coke Feed)	1910	1983	1910	2125
Ash Particles From C-4 Discharge, Run 122 (Run-of-Mine Kentucky Coal)	1925	2160	1930*	2225
C-4 Discharge, -12+20 Mesh, Run 130 (Washed Kentucky Coal)	1920	1980	1900	2030

* Some particles were sticking at temperatures of 1800°F and higher. At 1930°F particles with iron silicate coating stuck to each other and to uncoated particles as well.

The behavior of the C-4 ash particles from Run 122 run-of-mine Kentucky coal in the air atmosphere was surprising. Some dark particles stuck together at temperatures as low as 1800°F. These were probably iron sulfide particles that, on insertion in the furnace, were partly oxidized to yield a liquid mixture of ferrous sulfide and oxide; the phase diagram shows a eutectic between ferrous sulfide and ferrous oxide melting at 1724°F.¹ The fact that the agglomeration temperature of the coal ash agglomerates (Run 130) was slightly lower in air than in the CO/CO₂ mixture may be the result of a similar effect of ferrous oxide and ferrous sulfide in a silicate mixture. The eutectic of a mixture of the two compounds with ferrous silicate melts at 1688°F.⁸

Reactor Deposits

Two ash deposits that were found in the reactor on different occasions were analyzed. The first, a mushroom-cap shaped deposit with a hole through the center, grew around a grid nozzle during Run 124. (Similar deposits have also been formed in runs with coke feed.) The second, which appeared as scales about 1/8-inch thick, was found in the cyclone dipleg after Run 128; a white coating was present on one side of these scales. Some of this coating was removed by grinding or cutting with a saw and was analyzed separately; two samples of the coating-free deposit were also analyzed. Results from the analyses of the deposits are shown in Table C-24.

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Table C-24. COMPOSITION OF ASH DEPOSITS

Composition	Mushroom Cap, Run 124	Cyclone Deposit, Run 128		
		White Coating Sample	Samples With White Coating Removed	
		wt %		
SiO ₂	35.0	9.4	21.8	21.2
Al ₂ O ₃	14.2	63.8	16.1	16.0
Fe ₂ O ₃	34.6	18.3	46.5	48.2
TiO ₂	0.6	0.8	0.8	0.8
CaO	11.2	6.6	9.0	9.0
MgO	1.3	0.7	1.7	1.7
Na ₂ O	0.33	0.3	0.3	0.3
K ₂ O	1.0	0.5	0.3	0.3
SO	1.1	0.1	0.2	0.2
Total	99.33	100.5	96.7	97.7
Ash Fusibility				

°F

Reducing Atmosphere

IT	1910	- -	- -	1910
ST	2010	- -	- -	2030
HT	2050	- -	- -	2080
FT	2340	- -	- -	2150

Oxidizing Atmosphere

IT	2240	- -	- -	2250
ST	2500	- -	- -	2430
HT	2580	- -	- -	2490
FT	2690	- -	- -	2660

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Both deposits are composed of ash, but are enriched in iron: 34.6% Fe_2O_3 in the mushroom cap compared with 23% in the ash of the feed coal, and 47% in the cyclone deposit compared with 24% in the ash of the coke feed. Oxides other than iron are present in about the same ratio as in the feeds, except that the amount of calcium is about twice as high as expected. Sulfide corresponding to about 7% of the iron was present in the mushroom cap, indicating its origin from coal ash. The sulfide content of the original deposit may have been much higher. In both cases it appears that the deposits were formed from ash particles that had higher than average iron contents. The white coating on the cyclone deposit is high in alumina, confirming that the deposit formed on the alumina refractory in the cyclone.

DiscussionFormation of Low-Melting Silicates

The minerals that commonly occur in coal in the greatest amount are aluminosilicates (clay), which may also contain varying amounts of other elements in their crystal structure. Iron is commonly the third element of major importance in ash of Eastern U. S. coals, especially in high-sulfur coals. (Many Western U.S. coals have more calcium plus magnesium oxides than iron oxide.) Iron is usually present as sulfide (pyrites) but may also occur as carbonates and as part of the structure of some clays, especially illite.

The melting behavior of the ash of Eastern U. S. coals is dominated by the content of the three most abundant oxides – silica, alumina, and iron oxide – and also depends greatly on the oxidation state of the iron. The effect of iron content on initial deformation and softening temperatures from the ASTM ash fusibility tests on ashes of typical U.S. coals is shown in Figure C-33 for both reducing and oxidizing atmospheres.⁶ In the mildly reducing atmosphere of the test, in which the iron is in ferrous form, the temperatures decrease sharply with increasing iron content. The effect is much less in an oxidizing atmosphere in which the iron is present as ferric oxide or in a strongly reducing atmosphere in which the iron is reduced to metal. The phase diagram for $\text{FeO-Al}_2\text{O}_3\text{-SiO}_2$ in Figure C-34 shows a quadrilateral low-melting region at about 48% FeO, 13% Al_2O_3 , and 39% SiO_2 by weight.⁹ The minimum liquidus temperature in the presence of hercynite ($\text{FeO}\cdot\text{Al}_2\text{O}_3$) and fayalite ($2\text{FeO}\cdot\text{SiO}_2$), shown at the lower left corner of this quadrilateral, is $1990\pm 9^\circ\text{F}$. However, the other component of the eutectic, iron cordierite ($2\text{FeO}\cdot 2\text{Al}_2\text{O}_3\cdot 5\text{SiO}_2$), tends not to crystallize in the absence of seed crystals, and a metastable eutectic of fayalite, hercynite, and tridymite may form at $1963\pm 9^\circ\text{F}$.⁷ The diagram presented as Figure C-34 applies to the oxides when they are in equilibrium with metallic iron; at 1990°F this is equivalent to a steam/hydrogen mole ratio of 0.8. Note that the system is a pseudoternary one; that is, some ferric iron is present in the liquid even when it is in equilibrium with metallic iron. The eutectic melting temperature decreases with an increase in oxidizing potential to a minimum of approximately 1920°F at a steam/hydrogen mole ratio of about 25 (10^{-10} atm of oxygen). Of greater importance for ash mixtures is the solubility in the liquid of the minor components of calcium, magnesium, sodium, and potassium oxides and the resulting probable decrease in melting temperature; it is also possible, and indeed very likely according to petrographic observations, that supercooled liquid may form in some composition ranges.

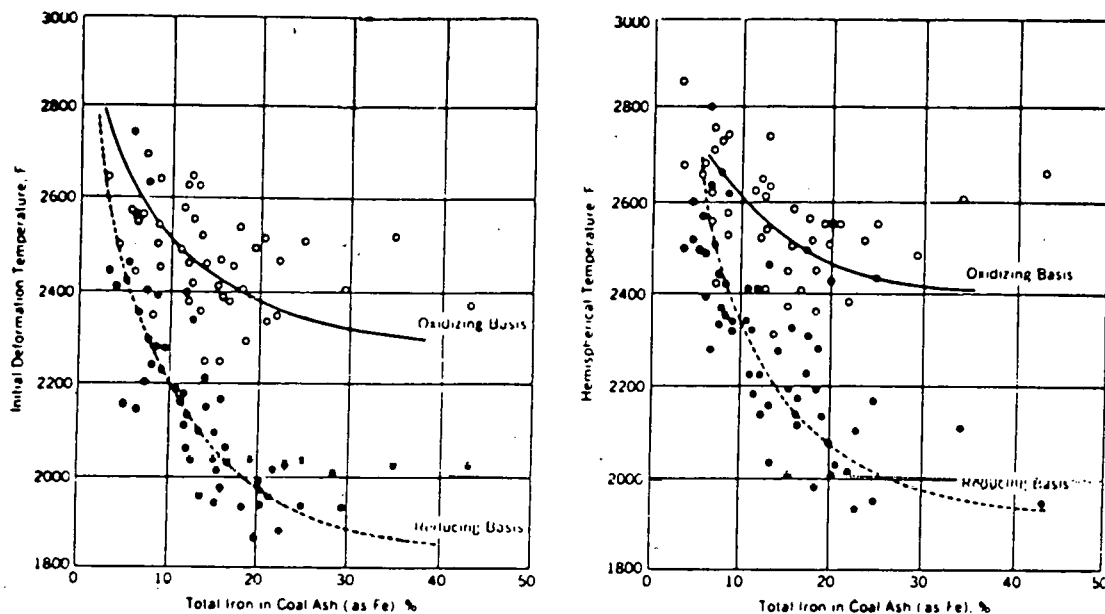


Figure C-33. EFFECT OF IRON CONTENT ON FUSIBILITY OF COAL ASH²

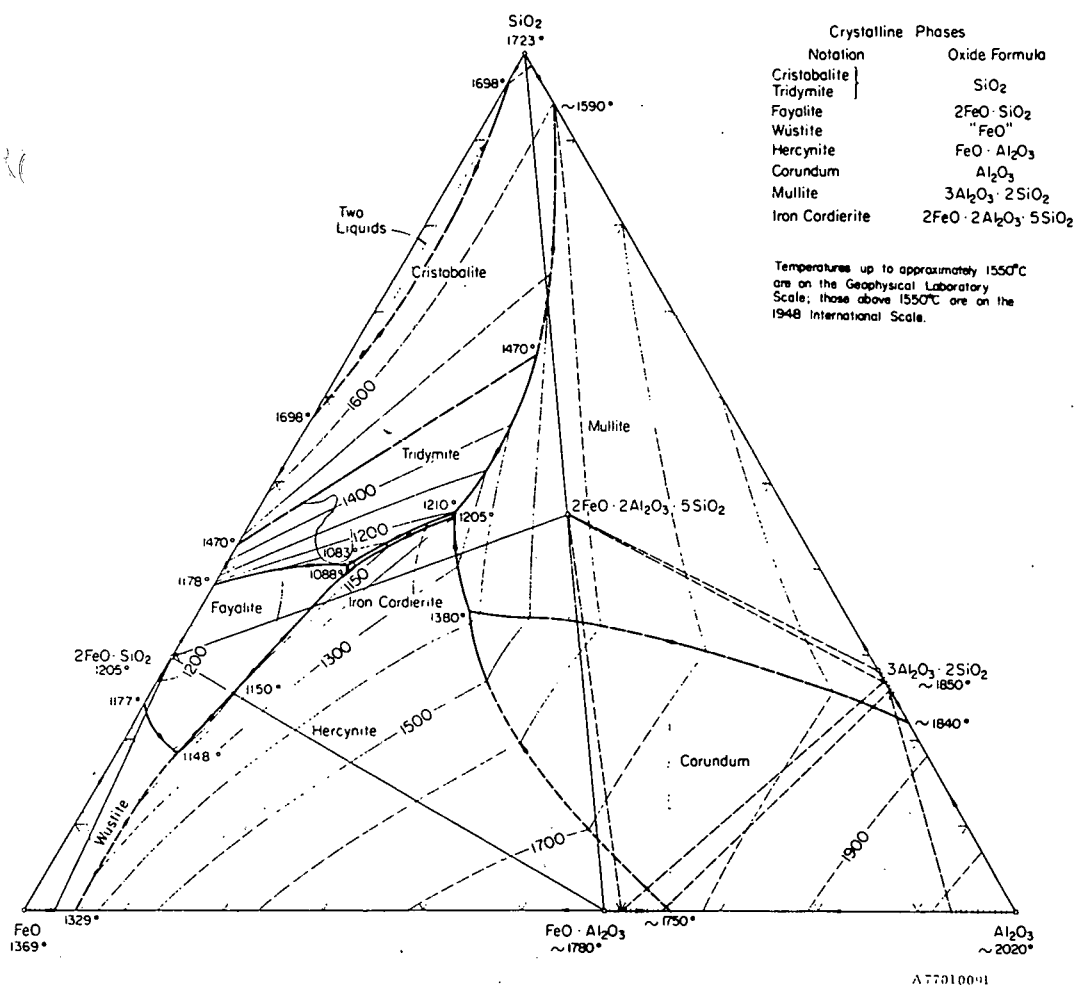


Figure C-34. PHASE DIAGRAM OF THE FeO-Al₂O₃-SiO₂ SYSTEM⁹
(Temperatures in °C and Compositions in wt %)

Formation of a low-melting silicate mixture from coal ash may be envisioned as follows: pyrite (FeS_2) in the coal readily decomposes to ferrous sulfide (FeS) when heated to reactor bed temperatures; the ferrous sulfide must then be oxidized to ferrous oxide, wustite ($\text{Fe}_{0.947}\text{O}$), as discussed later in this report. When a particle of the ferrous oxide comes into contact with a particle of clay, a reaction occurs that yields compositions at the particle junction ranging from pure ferrous oxide on one side to pure clay on the other; on the phase diagram these compositions lie along a line from the left (FeO) corner to a point on the Al_2O_3 - SiO_2 leg of the triangle. The position of the line terminus there depends on the alumina/silica ratio of the clay. This ratio ranges from about 0.3 to 0.9 by weight in different clay minerals; higher ratios are typical of kaolinite, and the lowest are typical of montmorillonite.⁵ Those of illite, the most prevalent clay in coals of the Illinois Basin, are intermediate. Thus, the reaction line falls close to or some distance to the right of the quadrilateral, depending on the silica/alumina ratio. At ferrous oxide contents above 40 weight percent this places the line in the hercynite (ferrous aluminate) field, thus accounting for the presence of hercynite crystals in the matrix phase of some agglomerates. In the ash of coal and coke feeds, the ferrous oxide content is about 27% on the silica plus alumina plus ferrous oxide basis. Thus, reaction of the ferrous oxide with all of the clay and quartz in the coal or coke ash would place the product in the mullite field; the eutectic is then at the right end of the quadrilateral, melting at 2200°F. It appears that reaction of the iron with only a fraction of the aluminosilicates and quartz in the ash is essential for production of a low-melting matrix phase and agglomeration. It is believed that, at equivalent temperatures, a greater amount of reaction occurs in the ASTM fusibility test than in agglomerate formation because of the slower heating rate and finer particle size (and thus greater surface area) in the ASTM test. This may also account for the "fusion" temperature of 1989°F obtained in IGT's agglomeration tests (Table 5) on agglomerates, compared with the ASTM softening temperature of 2200°F or higher obtained on the coke and coal ash (Table C-14 and C-16).

Silica occurs in coal as quartz (SiO_2) as well as in aluminosilicates. A search was made for particles of quartz in agglomerates from Run 127, made with coke feed. A sufficient number were found immersed in the matrix phase to indicate that quartz is much less reactive than the aluminosilicates of the coke or coal ash.

Calcium oxide is the next most abundant oxide in coal ash, especially in coals of the Illinois Basin. It also is a flux for aluminosilicates but, in general, is not as effective as ferrous oxide. For example, the lowest melting eutectic on the $\text{CaO-Al}_2\text{O}_3\text{-SiO}_2$ phase diagram melts at 2138°F.⁹ This is also shown (for large amounts of calcium oxide)

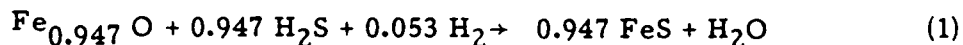
oxide) in Sondreal and Ellman's correlation for ash softening temperature under a reducing atmosphere (Figure C-35).¹³ Parameters of this correlation are percent by weight of basic constituents and dolomite ratio of the ash. The basic constituents are $\text{Fe}_2\text{O}_3 + \text{CaO} + \text{MgO} + \text{Na}_2\text{O} + \text{K}_2\text{O}$; these and the acidic constituents $\text{SiO}_2 + \text{Al}_2\text{O}_3 + \text{TiO}_2$ make up the total. The dolomite ratio is the weight percent of $\text{CaO} + \text{MgO}$ in the basic constituents. The figure shows that as calcium oxide and magnesium oxide replace ferric oxide (which is actually ferrous oxide in a reducing atmosphere) the softening temperatures rises. However, the effect of small amounts of calcium oxide is not readily apparent from the available phase diagrams. It seems likely that the addition of calcium oxide to the three-component eutectic lowers the eutectic melting temperature and on cooling tends to produce a different set of solid compounds: calcium aluminum silicate (anorthite, $\text{CaO} \cdot \text{Al}_2\text{O}_3 \cdot 2\text{SiO}_2$), iron silicate, and calcium iron silicate, instead of glass.

Role of Sulfur

Most of the iron in coal occurs as pyrite, which readily decomposes to ferrous sulfide in the reactor bed. Oxidation of the ferrous sulfide by oxidizing gases is therefore a prerequisite for formation of the low-melting silicate mixture and for agglomerate formation. The melting behavior of ferrous sulfide may also be a factor in the formation of agglomerates.

Rosenqvist's phase diagram of the iron-sulfur system is shown in Figure C-36.¹² The stable form of ferrous sulfide at elevated temperature is pyrrhotite, labeled $\text{FeS}_{1+x}(\beta)$ in the diagram. A range of sulfur content above stoichiometric FeS is shown; the equilibrium amount is dependent on the pressure of sulfur vapor or, equivalently, on the hydrogen sulfide/hydrogen mole ratio. The maximum melting point of ferrous sulfide is 2174°F , but at lower $\text{H}_2\text{S}/\text{H}_2$ ratios it forms liquid mixtures with metallic iron. The eutectic, at about 30 weight percent sulfur, melts at 1810°F (988°C). The effect of the $\text{H}_2\text{S}/\text{H}_2$ ratio is shown in Figure C-37, which was also obtained from Rosenqvist.¹²

Ferrous oxide also forms liquid mixtures with ferrous sulfide, yielding a eutectic that melts at 1724°F , as shown in Figure C-38.¹ However, no experimental equilibrium data are available for conversion reactions. With steam the equation is as follows:



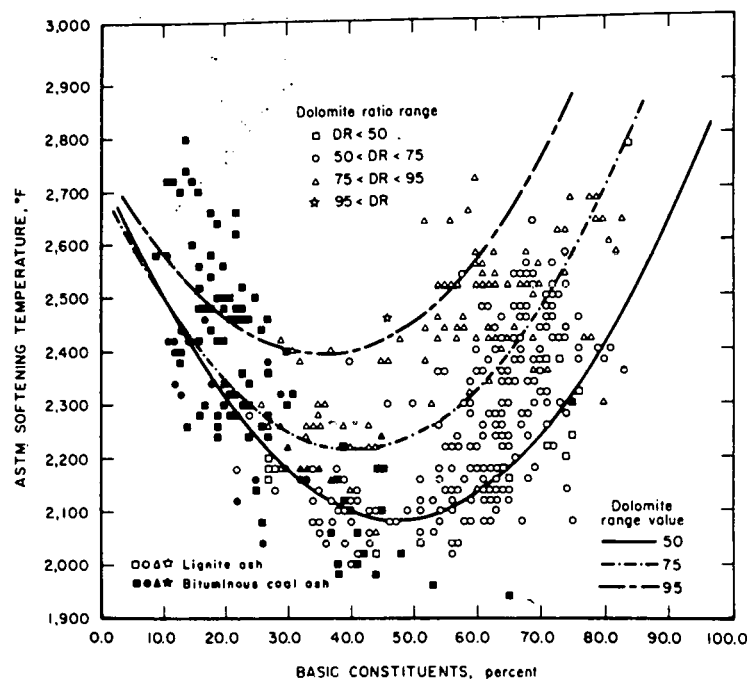


Figure C-35. SONDREAL AND ELLMAN CORRELATION FOR FUSIBILITY OF COAL ASH¹³

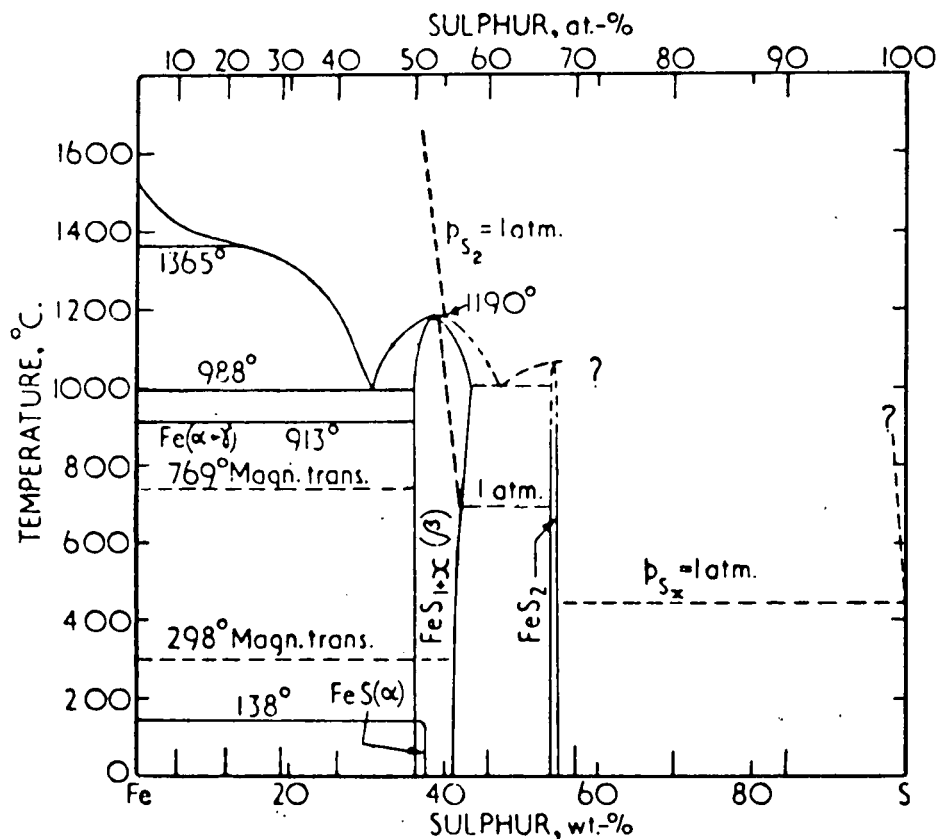


Figure C-36. PHASE DIAGRAM OF THE IRON-SULFUR SYSTEM¹²

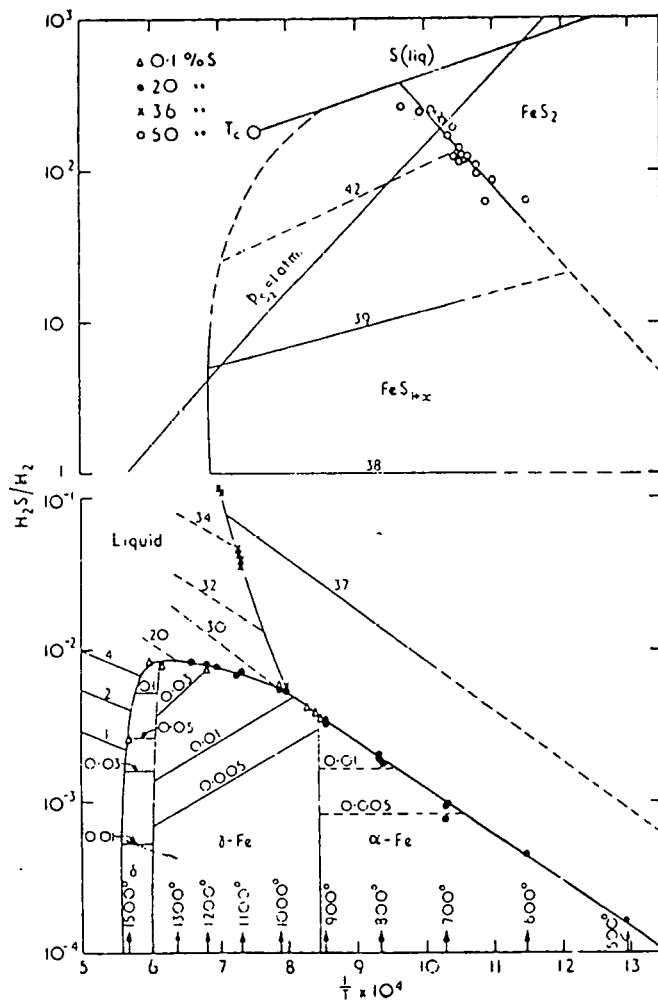


Figure C-37. EFFECT OF H_2S/H_2 RATIO OF THE IRON-SULFUR SYSTEM¹²

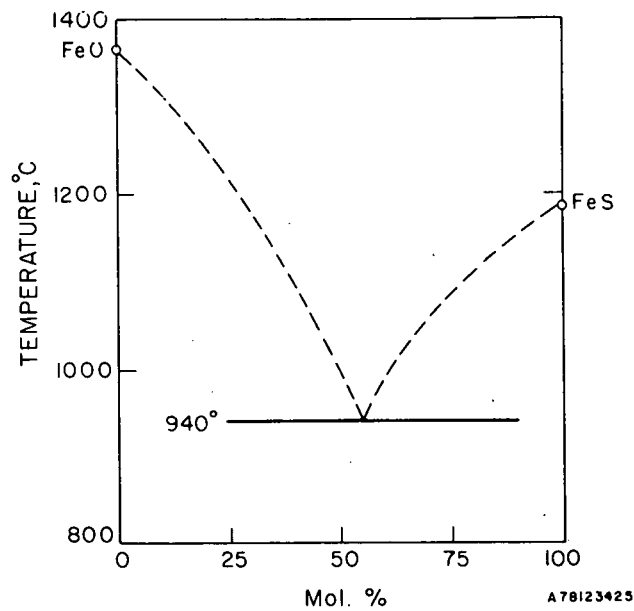


Figure C-38. PHASE DIAGRAM OF THE FeS-FeO SYSTEM¹

The equilibrium constant is expressed as follows:

$$K = \frac{a_{\text{FeS}}}{a_{\text{FeO}}} \frac{P_{\text{H}_2\text{O}}}{(P_{\text{H}_2\text{S}})^{0.947} (P_{\text{H}_2})^{0.053}} = \frac{a_{\text{FeO}}}{a_{\text{FeS}}} \frac{P_{\text{H}_2\text{O}}/P_{\text{H}_2}}{(P_{\text{H}_2\text{S}}/P_{\text{H}_2})^{0.947}} \quad (2)$$

Stoichiometric FeS was assumed, as little deviation from this is expected at the temperatures of interest. a_{FeO} and a_{FeS} are activities of ferrous oxide ($\text{Fe}_{0.947}\text{O}$) and ferrous sulfide, respectively; $P_{\text{H}_2\text{S}}$, etc., are partial pressures that can also be expressed as mole ratios of the gases $\text{H}_2\text{S}/\text{H}_2$, etc. The equilibrium relations were estimated by combining Rosenqvist's equilibrium data for the reaction –



with Baran and Knacke's thermodynamic data for the formation of ferrous oxide from metallic iron and steam.³ Rosenqvist's results for the solid up to the eutectic point were expressed as –

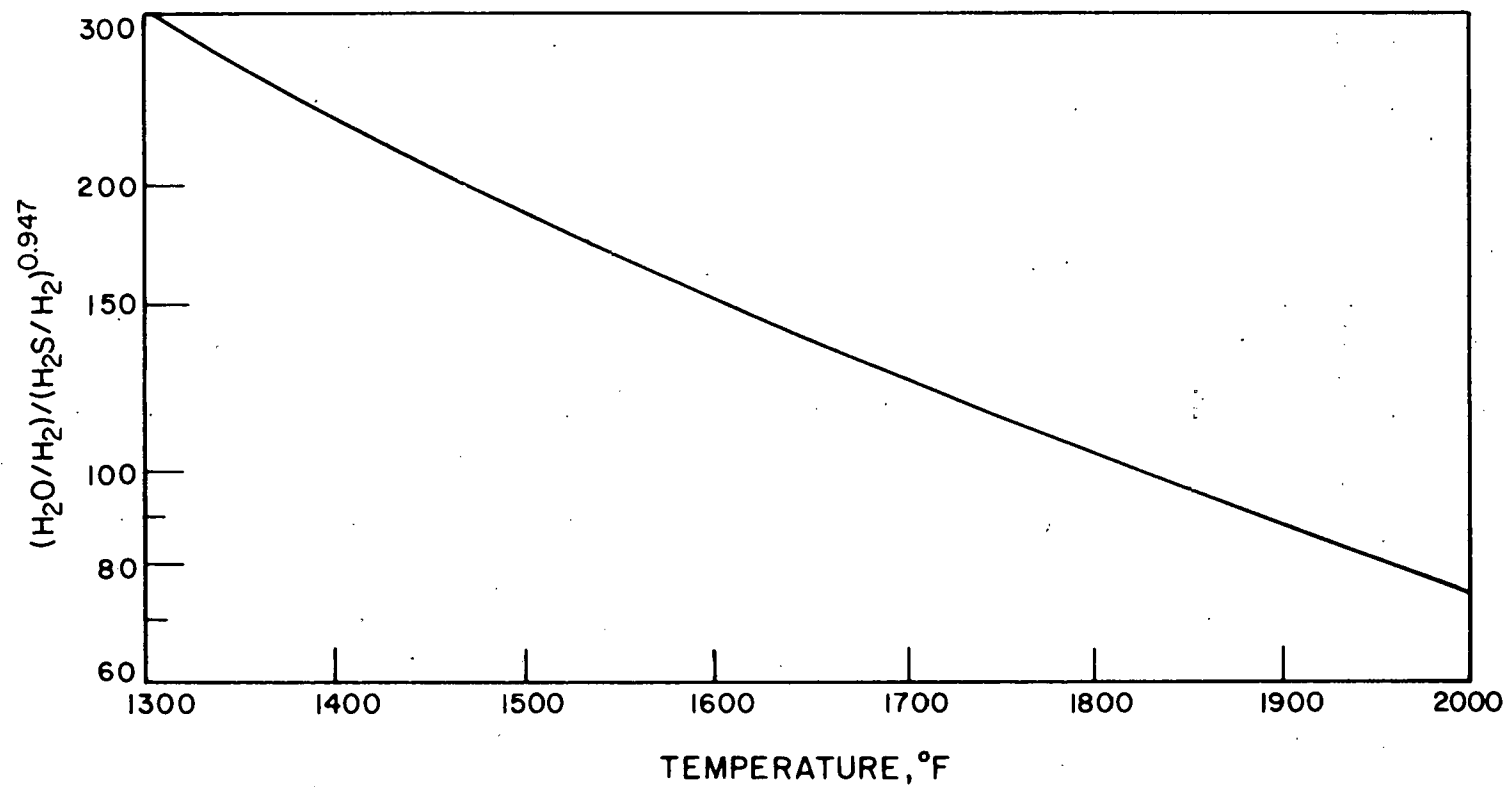
$$\log (P_{\text{H}_2\text{S}}/P_{\text{H}_2}) = \frac{-3100}{T} + 0.179 \quad (4)$$

where T is in $^{\circ}\text{K}$. IGT extrapolated this equation beyond the eutectic temperature to 1900°F to yield the equilibrium constant values shown in Figure C-39.

To estimate the range of gas mole ratios at 1800° and 1900°F within which liquid mixtures of ferrous oxide and ferrous sulfide are stable, it was assumed that the liquid mixtures are ideal solutions. Then, using the right-hand solubility curve of Figure 16, the activity of ferrous oxide at a given temperature is the ratio of its mole fraction given on the right-hand curve to its mole fraction given on the left-hand curve, where it is in equilibrium with solid ferrous oxide and its activity is unity. A similar relation applies to the ferrous sulfide. Results of the calculation are shown in Figures C-40 and C-41. In the central parts of the figures, the solubility of metallic iron in the sulfide-oxide mixture affects the relations. It should be possible to estimate the effect from the binary phase diagrams. However, this has not been done yet; thus, the position and shape of the dashed curves are conjectural.

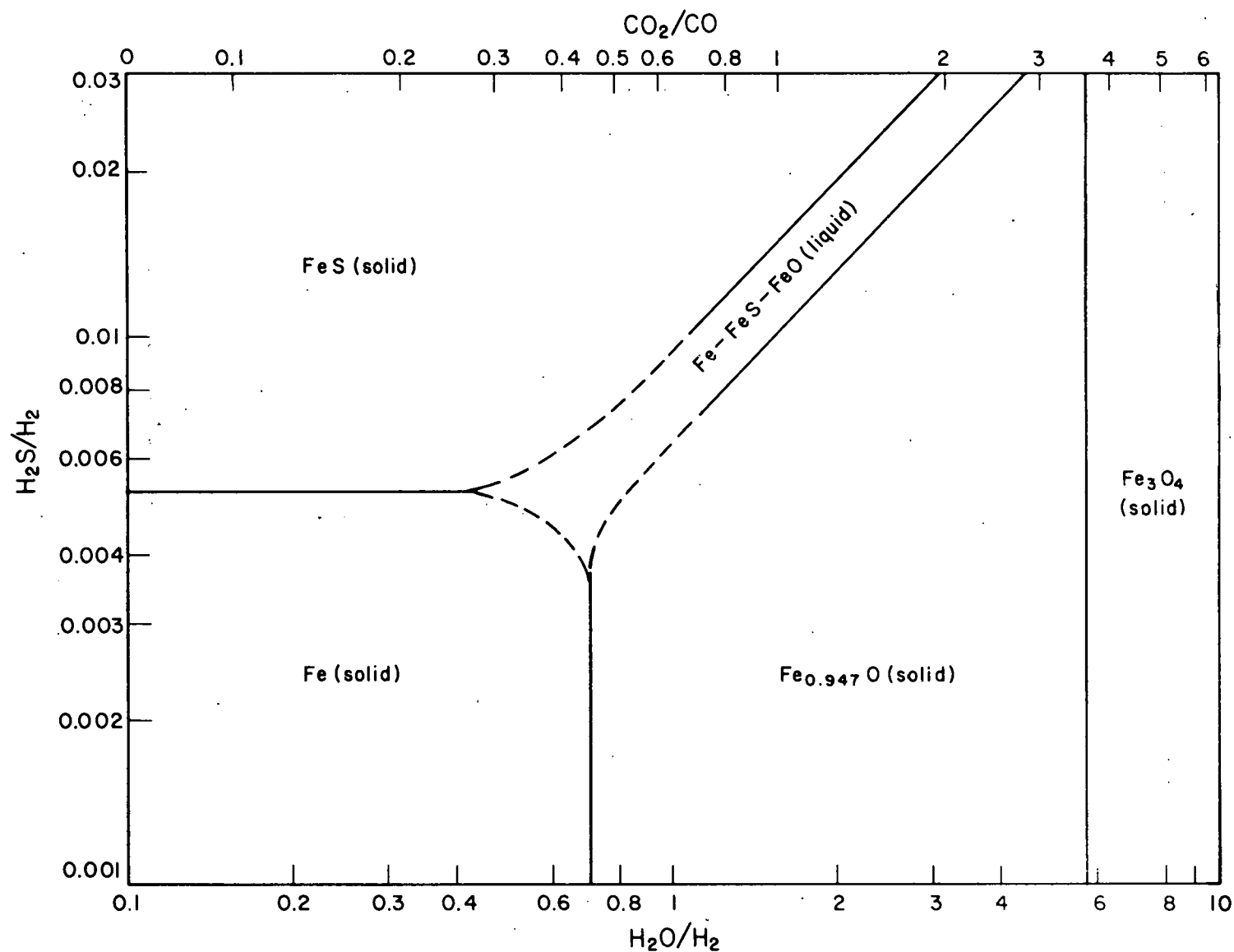
The relevant gas ratios in the product gas from steady-state periods of Runs 124 and 130 have been calculated (Table C-25). Hydrogen sulfide in the gas was calculated from a sulfur balance; feed sulfur, less the sulfur in the solid products, was assumed to appear in the gas as hydrogen sulfide plus carbonyl sulfide. Carbonyl sulfide was estimated from the equilibrium of the following reaction:³





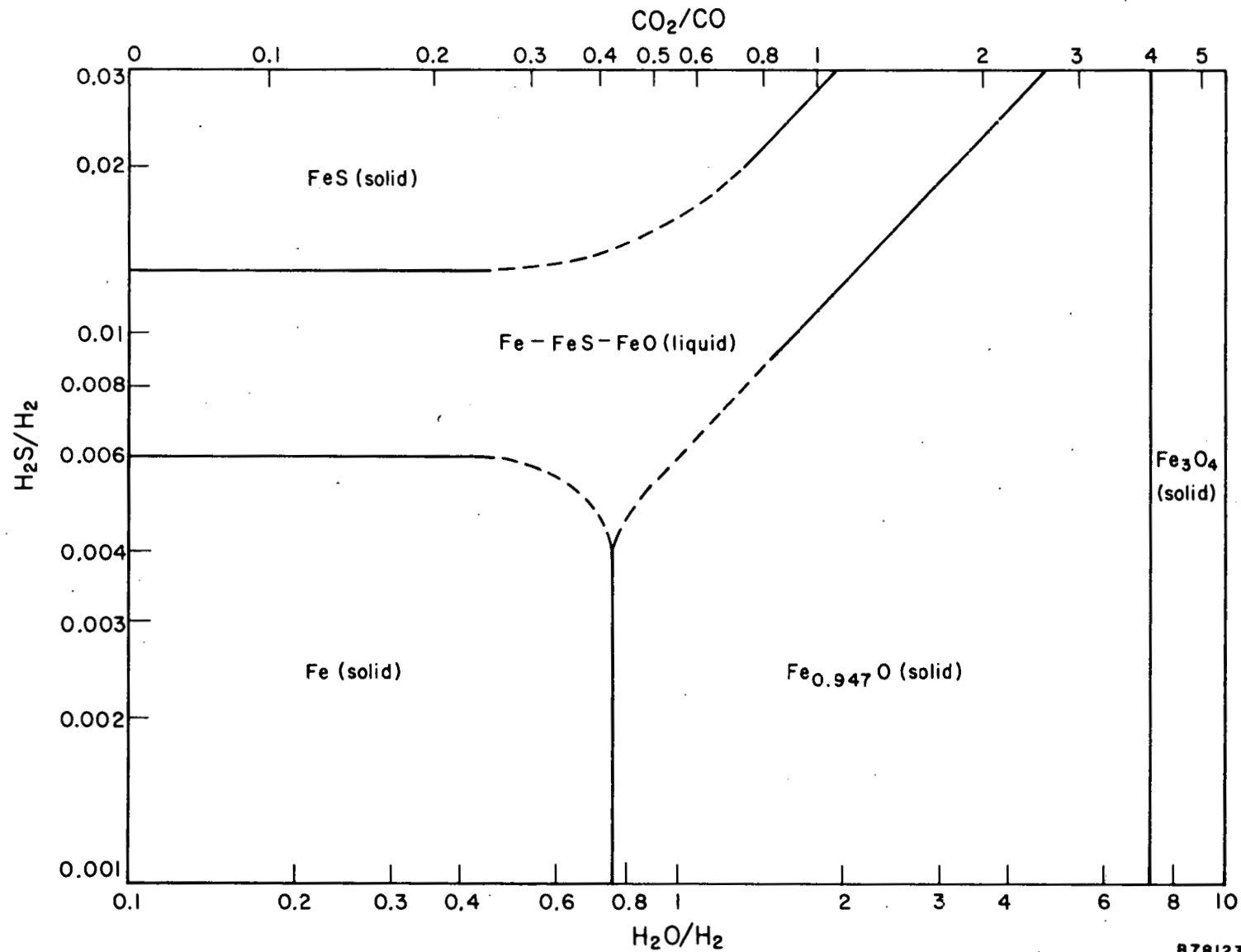
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Figure C-39. EQUILIBRIUM CONSTANT FOR OXIDATION OF FERROUS SULFIDE BY STEAM



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Figure C-40. EFFECT OF ATMOSPHERE ON THE Fe-S-O SYSTEM AT 1800°F



878123624

Figure C-41. EFFECT OF ATMOSPHERE ON THE Fe-S-O SYSTEM AT 1900°F

Table C-25. EFFECT OF PRODUCT GAS ATMOSPHERE ON FERROUS SULFIDE

Run No.	124		130			
Steady-State Period						
Start, (hr, date)	0200, 6/3/78	0200, 6/8/78	1600, 11/4/78	2000, 11/5/78	2200, 11/6/78	1400, 11/7/78
Stop, (hr, date)	1600, 6/3/78	2200, 6/8/78	0600, 11/5/78	0400, 11/6/78	0800, 11/7/78	2000, 11/7/78
Bed Temperature, °F	1891	1815	1855	1874	1870	1871
Product Gas Rates, mol/hr						
CO	30.55	19.61	29.53	34.44	33.87	48.10
CO ₂	12.96	15.15	21.30	18.41	24.27	34.37
H ₂	34.47	31.27	39.89	39.04	47.10	68.92
H ₂ O	11.44	57.67	28.83	32.69	63.63	30.80
H ₂ S + COS*	1.26	1.10	1.35	1.31	1.48	1.51
Product Gas Ratios						
H ₂ O/H ₂	0.33	1.84	0.72	0.84	1.35	0.45
CO ₂ /CO	0.42	0.77	0.72	0.53	0.72	0.72
H ₂ S/H ₂	0.033	0.035	0.032	0.032	0.030	0.020
Stable Iron Phase	Solid FeS	Solid FeS	Solid FeS	Solid FeS	Solid FeS	Solid FeS

* This is taken to be equal to the sulfur in the feed less the sulfur in the solid products. The H₂S/H₂ ratio was corrected for COS according to its equilibrium with CO₂ and H₂O.

The amount of carbonyl sulfide ranged from 2% to 9%. The measured composition of the gas is expected to differ from the composition in the bed by water-gas shift during cooling. However, use of either the steam/hydrogen or carbon dioxide/carbon monoxide ratio of the measured compositions places all of the points in the solid ferrous sulfide field. Thus it appears that ferrous sulfide cannot be oxidized to ferrous oxide in the upper part of the fluidized bed. The hydrogen sulfide/hydrogen ratio is likely to be lower and the oxidizing potential higher, both being more favorable to the oxidation, in the bed below the coal entry level. However, it appears likely that most of the oxidation occurs in the vicinity of the oxygen reaction region.

A different reaction route to form the ferrous aluminosilicate is theoretically possible. Its formation from ferrous oxide is thermodynamically favored, and thus a direct reaction of aluminosilicate, ferrous sulfide, and oxidizing gas might be possible in atmospheres unfavorable to formation of intermediate ferrous oxide. However, this reaction would involve diffusional transport of oxygen through one or another of the solid or liquid phases, and such transport is likely to be very slow.

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Conclusions

Results of laboratory observations, analyses, and tests made in support of pilot plant gasification tests on coke, run-of-mine Kentucky coal, and washed Kentucky coal are reported here. Agglomerated ash in the form of rounded beads was produced in pilot plant tests on coke and washed Kentucky coal. With both feeds, formation of the agglomerates is dependent on production of an iron-rich, relatively low-melting matrix in which a substantial portion of other ash is embedded. The matrix, which appears to have been molten in the gasifier, consists predominantly of a ferrous aluminosilicate mixture with small amounts of other oxides.

The aluminosilicate of the matrix is produced from aluminosilicate (clay) minerals in the coal, including also coal feed to coking in the case of coke. Most of the iron occurs in the Kentucky coal as pyrite, FeS_2 , with small amounts present in other forms, probably as carbonate or as a constituent of the clay minerals. Ash of the coke feed to the pilot plant had an equal amount of iron, but very little of it was present as sulfide. (Low-sulfur, and thus low-pyrite, coal is used in coking.)

Pyrite readily loses half its sulfur to form ferrous sulfide, FeS , when heated to reactor bed temperature. The ferrous sulfide in the coal char must then be oxidized before the iron can react with the aluminosilicates to form the low-melting matrix material. IGT's analyses show that oxidation of ferrous sulfide and formation of the agglomerating matrix material was achieved in Runs 130 and 131 on washed Kentucky coal but not in runs on run-of-mine coal. In Run 124 with run-of-mine coal, sulfide iron in the bed discharge ranged from about 70% of the total iron when gasifying with air to about 40% of the total iron when gasifying with oxygen; petrographic and other tests show that in neither case was there appreciable formation of the low-melting matrix material or agglomeration of the ash. However, it is probable that successful production of agglomerated ash with washed Kentucky coal was more the result of differences in mode of operation than of the character of the feed. The main difference was the introduction of most of the oxygen into the central venturi area in the runs with washed coal, instead of dispersion through the grid. Details of reactor operation in Runs 122 and 124 on unwashed coal and for Run 130 on washed coal are presented in Appendix A of this report.

Thermodynamic data indicate that ferrous sulfide cannot be oxidized to an oxide in the product gas atmosphere of Run 124 or Run 130. Conditions may be more favorable in the bed below the coal entry level, but it appears likely that most of the sulfide

oxidation occurs in the vicinity of the oxygen reaction region.

The temperature at which ash agglomerates from coke or from washed coal stick firmly together was determined to be about 1920°F in a reducing gas atmosphere. The temperature at which the agglomerates run together (presumably that which is required to form rounded beads in the reactor) was about 1980° under a reducing atmosphere. This is higher than bed temperature in both the coal and coke runs, which indicates that formation of agglomerates occurs only in the vicinity of the central jet where temperatures are higher than that of the bed.

From phase diagrams and other evidence, it is concluded that the iron content of the matrix material that is responsible for agglomeration of ash is probably 35 weight percent (calculated as ferric oxide) or more. Fusion temperatures of ash containing 22 to 24 weight percent ferric oxide, the amount present in the coal and coke ash, are substantially higher than those found above. Thus, reaction of the iron with only part of the aluminosilicates in the ash is an essential feature for production of the low-melting material. Low-melting high iron-content material is also involved in the formation of deposits in the reactor; a deposit formed on the grid during Run 124 contained 35 weight percent ferric oxide, and one formed in the internal cyclone during Run 128 contained about 47 weight percent ferric oxide.

The effects of substantial variation in ash composition, especially the amounts of slagging oxides, and the effects of such variables as the type of clay mineral, the size and dispersion of clay granules and pyrite, and the association of minerals with each other have not been considered in this report. In future studies some information on the effect of type of clay mineral can perhaps be obtained from clay mineral technology. However, pilot plant operation on a variety of coals appears necessary to establish the range of coal ash composition and mineralogy over which ash agglomeration can be achieved.

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APPENDIX D. Design Data for the Demonstration Plant

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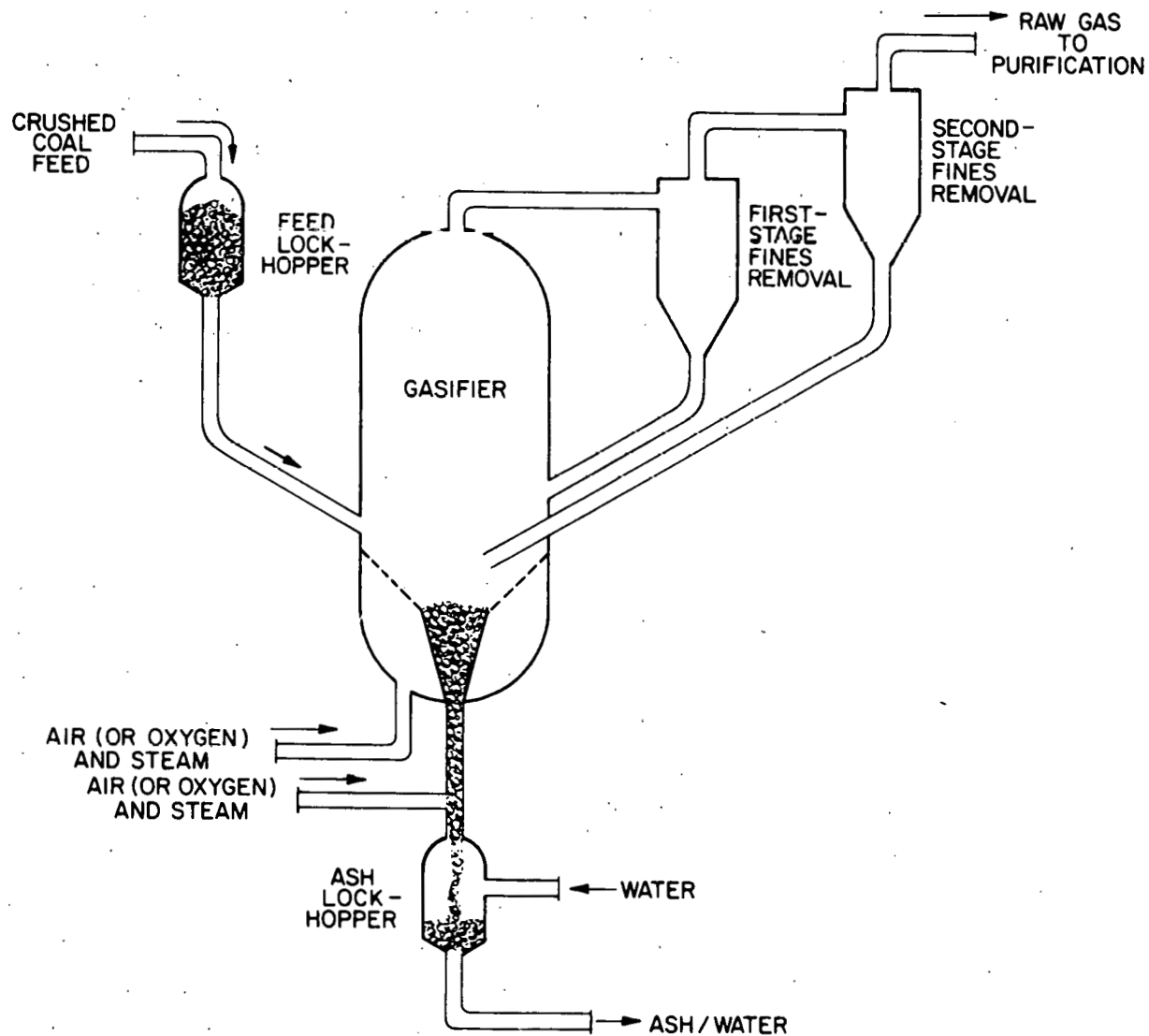
PROCESS DESCRIPTION

The U-Gas® Process has been developed by the Institute of Gas Technology (IGT) to produce a low- to medium-Btu (150 to 300 Btu/SCF) fuel gas from coal in an environmentally acceptable manner. The primary use of the gas is as an industrial fuel replacing natural gas and fuel oil, thereby reducing the requirement for importing those premium hydrocarbons. In the form of synthesis gas, the product may be used as a chemical feedstock for production of hydrogen, ammonia, or methanol. Pressurized, the gas is suitable for advanced electric power generation systems such as a combined gas turbine/steam turbine cycle, a fuel cell, or a magnetohydrodynamic (MHD) generator.

The U-GAS Process accomplishes four important functions in a single-stage, fluidized-bed gasifier (Figure D-1): It decakes coal, devolatilizes coal, gasifies coal, and agglomerates and separates ash from char. Other characteristics of the process are shown in Table D-1.

In the process, washed coal (1/4-inch X 0) is dried only to the extent required for handling purposes. It is pneumatically injected into the gasifier through a lockhopper system. Within the fluidized bed, coal reacts with steam and oxygen (air can be substituted for oxygen) at a temperature of 1750° to 1900°F. The temperature of the bed depends on the type of coal feed and is controlled to maintain nonslagging conditions for ash. The operating pressure of the process depends on the ultimate use of product gas and may vary between 50 and 350 psi. The pressure must be optimized for a particular system; for production of an industrial fuel, a minimum pressure of 80 to 100 psi is desirable. At the specified conditions, coal is gasified rapidly, producing a gas mixture of hydrogen, carbon monoxide, carbon dioxide, and a smaller fraction of methane. Because reducing conditions are always maintained in the bed, nearly all of the sulfur present in the coal is converted to hydrogen sulfide.

Simultaneously with coal gasification, the ash is agglomerated into spherical particles and separated from the bed. Part of the fluidizing gas enters the gasifier through a sloping grid. The remaining gas flows upward at a high velocity through the ash agglomerating device and forms a hot zone within the fluidized bed. The temperature within the hot zone is greater than at other locations in the bed. High ash content particles agglomerate



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Figure D-1. U-GAS GASIFIER

under these conditions and grow into larger and heavier particles. Agglomerates grow in size until they can be selectively separated and discharged from the bed into water-filled ash hoppers where they are withdrawn as a slurry. In this manner, the fluidized bed achieves the same low level of carbon losses in the discharge ash that is generally associated with the ash-slugging type of gasifiers.

Coal fines elutriated from the fluidized bed are collected in two external cyclones. Fines from the first cyclone are returned to the bed, and fines from the second cyclone are returned to the ash agglomerating hot zone, where they are gasified, agglomerated with bed ash, and discharged with ash agglomerates. The raw product gas is significantly free of tar and oils, thus simplifying ensuing heat recovery and purification steps dictated by the end use of the product gas.

Table D-1. U-GAS PROCESS CHARACTERISTICS

- High conversion of coal to gas using the ash agglomeration technique
- Capability to gasify all ranks of coal
- Ability to accept fines in coal feed
- Simple design and safe, reliable operation
- Easy to control and able to withstand upsets
- Product gas significantly free of tar and oils
- No environmental problems

PROCESS DEVELOPMENT

Most of the process development work has been accomplished on a pilot plant put into operation in 1974. The development work can be divided into three separate phases: Phase I, during which the process feasibility was demonstrated using metallurgical coke and char as feed; Phase II, during which the pilot plant was modified to feed coals, and trial tests were made with coal; and Phase III, during which process feasibility was proved using coal as feed and data were developed for scale-up of the process and design of a demonstration plant. A variety of feedstocks have been used in the pilot plant to determine process sensitivity to variables such as ash content, particle size, reactivity, and free-swelling index of caking coals.

Phase I Development

The pilot plant (also called the Ash Agglomerating Gasifier) was constructed in 1974 and operated until mid-1976 with funding from the Office of Coal Research (OCR) and the American Gas Association (A.G.A.) as part of the HYGAS project. The objective of the program was to demonstrate high carbon utilization in fluidized beds by applying the technique of ash agglomeration. A total of 67 runs were made — about 50 with metallurgical coke and 17 using chars prepared by Project COED Illinois and Western Kentucky bituminous coals.

All tests were performed near atmospheric pressure, and the operating temperature ranged between 1850° and 1960°F. Gasifying agents included air-steam, oxygen-steam, and air-oxygen-steam mixtures. The first nine test runs were used for equipment shakedown and instrument calibrations. For Runs 10 to 31, operating conditions, reactor geometry, fluidization regimes, and flow parameters were established for stable gasification and agglomerate production. These runs were also used to develop a method to reinject fines collected in the external cyclone back into the bed and to agglomerate the ash component. The next 20 runs were conducted to produce design data and to study the effects of operating variables on the process. A long-term duration test of 10 days was conducted to establish the feasibility of continuous ash agglomeration. The run was voluntarily terminated, and examination of internals indicated there was no ash buildup. The 17 tests conducted with COED char produced good quality gas, but it was difficult to maintain ash-balanced operation because of the unsatisfactory quality of ash agglomerates.

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During these tests, several process and mechanical changes made to the pilot plant resulted in an improved design. Two of the significant modifications were changes in the geometry of the sloped grid and the location of a central oxygen jet within the ash agglomerating section of the gasifier. The latter was necessary to operate with steam-oxygen gas mixtures. Milestones during this period were -

- Demonstrating the operability of the gasifier system and first large-scale U.S. production of ash agglomerates
- Perfecting the technique of ash agglomeration and entrained fines recycle on metallurgical coke during 67 pilot plant tests
- Demonstrating the feasibility of achieving high carbon conversions in the range of 95+% by utilizing the ash agglomeration technique
- Operating the gasifier both with steam-air producing low-Btu gas and with steam-oxygen producing medium-Btu gas.

Phase II Development

As a result of encouraging results, in July 1976, the U.S. Energy Research and Development Administration (ERDA), under a new contract, sponsored a modification of the pilot plant facilities to enable feeding of coal to the gasifier and conducting extended period tests. The major modifications were increased feed-storage capacity, a coal dryer, a novel raw coal-injection system, an additional quencher and stainless-steel cyclones for recycling of hot entrained coal fines, and liquid oxygen storage capacity. The modifications were completed in April 1977 and, subsequently, 18 tests were conducted up to the end of the year.

The first four runs were made with metallurgical coke to shake down the modified pilot plant and reestablish operating conditions for gasification and agglomeration. During these tests the new coal feed injection system was debugged, and operating experience was gained. Both agglomeration and fines recycle were achieved on the modified gasifier system. Five tests were conducted with Montana subbituminous coal from the Coal Creek Mine. This coal was successfully gasified and good quality gas produced without evidence of any organic condensables present in the product gas. However, because of the high reactivity of this coal and the mechanical limitation of the coal dryer and oxygen feed to the gasifier, the bed temperatures were limited to a range of 1600° to 1700°F. Therefore, agglomeration of subbituminous coal

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was not achieved during these runs. Nevertheless, in two test runs, a high coal utilization efficiency (90% to 92%) with coal fines recycle was attained.

In Runs 112 through 117, Illinois No. 6 bituminous coal was tested in the pilot plant. Despite its caking tendency, the coal feed injection system was able to freely feed this coal directly into the gasifier, and the bed remained fluidized and free of caked coal deposits. During these tests, the operating mode, temperature, feed size, and fluidizing velocities were varied to establish the conditions required for stable fluidization and ash agglomeration. Both 6 mesh and 1/4-inch size coal were used, the temperature ranged between 1800° and 1875°F, and the superficial velocity ranged between 2 and 4 ft/s. Although ash agglomerates could be produced from this coal, a stable fluidized bed could not be maintained because of ash sintering and clinkering. However, the conditions required to form agglomerates were established, but how to discharge sufficient ash to control the ash level of the fluidized bed remained a problem. Achievements during this period included —

- Gasification of Montana subbituminous coal with recycle of fines and coal utilization exceeding 90%
- Establishing the technical feasibility of feeding Illinois No. 6 caking coal directly into the gasifier without pretreatment.

Phase III Development

In late 1977 the U.S. Department of Energy (DOE) selected Memphis Light, Gas and Water Division's (City of Memphis) proposal to design, construct, and operate an industrial fuel gas demonstration plant based on the U-GAS Process. The pilot plant began operations in January 1978 in support of this program. The objective was to provide mechanical, operating, and process data for the preliminary design of the demonstration plant using Western Kentucky No. 9 coal. A total of 16 test runs were conducted with the two coals. The final run (over 150 hours) was conducted at a pressure of 30 to 45 psig.

As is typical in pilot plant operations, several problems — both mechanical and process — were encountered during this period. One by one, solutions were found for each of them and by the end of the test program (March 1979), process feasibility was firmly established and a strong data base was obtained for the preliminary design of the demonstration plant. Out of the total 15-month program; pilot plant tests were conducted for about 10 months; the other period was downtime due to plant modifications.

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In the initial tests (Runs 118 through 122), the objective was to determine the operating conditions necessary for stable gasification and agglomeration of the unwashed Western Kentucky coal. These tests revealed that a higher superficial velocity in the range of 4 ft/s is necessary to maintain stable fluidization conditions in the bed when operating at 1850° to 1900°F with the bituminous coals. This velocity was almost double the value used for design of the gasifier and all the peripheral equipment. Operations at these conditions, therefore, imposed a severe strain on the coal feed lockhopper system, the cyclones, and the downstream equipment, resulting in several mechanical problems. To overcome the limitation, the pilot plant gasifier internal diameter was reduced from 4 feet to 3 feet after the first five tests.

In the next series of tests (Runs 123 through 126), successful stable gasification for about 170 hours was obtained. Both steam-air and steam-oxygen gas mixtures were used for gasification. However, during these tests sufficient ash could not be agglomerated and withdrawn from the fluidized bed to maintain an overall ash balance. Several attempts to increase the ash agglomeration rate by increasing operation temperature (up to 1950°F) and the bed ash content (up to 70%) were not successful. Therefore, after Run 126, the mechanical design of the slopping grid and the ash-agglomerating section were changed to provide more flexibility in introducing oxygen to the system and, at the same time, to remove ash agglomerates from the fluidized bed. During this period, all pilot plant changes necessary to operate the gasifier at its design pressure of 50 psig were completed. In addition, a product gas water scrubber was installed to provide information for the design of the demonstration plant and also to detect trace quantities of impurities in the product gas.

The next series of seven tests (Runs 127 through 133) was highly successful. The changes in configuration and mode of operation completely eliminated the problem of ash sintering or clinkering and, after seven tests, the grid and the fluidized bed area were free of any ash buildup. Two tests, lasting about 250 hours, were conducted with metallurgical coke to establish baseline operating conditions of the gasifier after the configuration changes to the ash agglomerating section. The next five tests were conducted with a washed Western Kentucky coal. Washing the coal reduced the ash content from

about 19% to 12%, but the size consist and the fines content of the coal were relatively unchanged. Both unwashed and washed coal feed sizes were nominal (1/4 inch). These tests were all of extended duration with actual operating time on coal of 75 to 150 hours in each test and were planned to yield specific design information for the demonstration plant. During these tests, repeated stable gasification and ash agglomeration was achieved with both steam-air and steam-oxygen mixtures. The raw product gas heating value was 255 to 285 Btu/SCF. Coal fines from both the first and the second cyclone were recycled, yielding coal gasification efficiency of up to 99% and coal utilization efficiency of about 95%.* The latter is limited because of the constraints of pilot plant cyclone fines collection efficiency. The coal feed-injection system operated without any problems with the highly caking coals and was optimized to reduce the level of tar and oils in the product gas to trace levels. The scrubber was operated continuously during the latter tests and showed no buildup of any tar or oils.

The last test in the program, Run 133, was performed at 45 psig, the design pressure of the pilot plant. All previous tests were at lower pressures of 5 to 10 psig. During the test, the system pressure was varied to provide information on gasification and ash agglomeration. At the elevated pressure, the system operated more smoothly and was more easily controlled, and fluidization was improved. There was no problem in forming or discharging ash agglomerates at the higher pressure, although the coal feed increased to about 1800 lb/hr. The bed operating temperature and the superficial velocity also did not require any changes. Excellent heat and material balance enclosures were obtained during these long-duration tests, and the results verified the computer model prediction of bed volumes, oxygen requirement, and gas composition.

The highlights of test operations were as follows:

- The pilot plant tests firmly established process feasibility and provided a strong data base for completing the preliminary demonstration plant design.

* Coal gasification efficiency accounts for loss of carbon in agglomerates; coal utilization efficiency accounts for loss of carbon in ash discharge and coal fines.

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- Four consecutive, extended period tests were conducted during which good quality raw product gas (285 Btu/SCF) and high-ash-content (80 to 90 weight percent) ash agglomerates were produced from Western Kentucky coal.
- A technique of feeding caking coals directly into the gasifier without pretreatment was perfected. Over 400 tons of caking coal with a free-swelling index (FSI) of 4 to 7 were fed.
- Stable operability of the gasifier while recycling entrained coal fines back into the gasifier under continuous agglomerating conditions was demonstrated.
- Data related to environmental aspects of the U-GAS Process, particularly wastewater characteristics, which indicated the presence of only trace quantities of tar and oils, were provided.
- The pilot plant operated for more than 100 hours at pressures of up to 60 psia (gasifier design pressure is 65 psia) proving applicability of the ash agglomeration technique and the ash agglomeration discharge mechanism at moderate elevated pressures.
- A broad operating window for the major operating variables of temperatures, superficial velocity, and bed ash content was established.
- Suitable materials of construction were tested, and the design for the internal components of the gasifier was established.

In addition to the pilot plant activities, support studies have been conducted on 1) ash chemistry to better understand the principle of ash agglomeration, 2) bench-scale tests to determine the main operating variables affecting the formation of ash agglomerates, 3) cold-flow model tests to define the mechanism of selective separation of agglomerates and obtain scale-up information, and 4) computer modeling to predict the performance of the gasifier.

FEED CHARACTERISTICS

The gasifier is suitable to operate on char, coke, or any rank of coal. However, a representative coal, namely Western Kentucky No. 9 bituminous coal has been chosen for the basis of design. The typical analysis of the coal is given in Table D-2. The coal is crushed to less than 1/4-inch size before introducing it into the gasifier. The typical size consist of the feed and its physical properties are shown in Tables D-3 and D-4, respectively. The data have also been gathered for the range of variability of Western Kentucky No. 9 coal moisture, ash, sulfur, and chlorine contents for both cleaned and run-of-mine coal. These data are summarized in Table D-5. The as-received coal moisture content should be reduced to less than 5% to improve its flowability and packing characteristics.

Table D-2. CHARACTERISTICS OF THE COAL FEED

Type of Coal: Washed Western Kentucky No. 9

Analysis of Coal

Ultimate Analysis		wt %
Moisture		2.50
Ash		13.15
Carbon		66.93
Hydrogen		4.71
Nitrogen		1.10
Chlorine		0.22
Sulfur		3.83
Oxygen		<u>7.56</u>
Total		100.00
Proximate Analysis		
Moisture		2.50
Ash		13.15
Volatiles		38.78
Fixed Carbon		<u>45.57</u>
Total		100.00

Limitations: Moisture 5%

Table D-3. SIZE OF COAL FEED TO THE GASIFIER

<u>Retained</u>	<u>Microns</u>	<u>Cumulative wt %</u>
+1/2	6350	1.8
+4 mesh	4760	12.4
+6	3360	32.1
+12	1680	58.6
+40	420	84.4
+70	210	90.8
+140	105	94.2
+200	74	95.1
+270	53	96.5
-270	--	100.0

Limitations: +1/2, 2%; -100 mesh, 10%

Table D-4. PHYSICAL PROPERTIES OF COAL FEED

HHV, Btu/lb	12,222
FSI	4-6
Bulk Density, lb/ft ³	44.4
Particle (Apparent) Density, lb/ft ³	80.0
Complete Fluidization Velocity, ft/s	2.5
Fluidized Bed Density, (At complete fluidization), lb/ft ³	29.0
Poured Angle of Repose	34 deg
Angle of Internal Friction	45 deg

Table D-5. DESIGN BASIS VARIABILITY OF MOISTURE, ASH,
SULFUR, AND CHLORINE IN ISLAND CREEK MINE KENTUCKY NO. 9 COAL

Washed, or Cleaned Coal			
<u>Coal Component</u>	<u>Minimum Weight %</u>	<u>Maximum Weight %</u>	<u>Average Weight %</u>
Moisture	7.5	9.0	8.0
Ash*	9.0	10.0	9.3
Sulfur*	2.9	3.2	3.05
Chlorine*	0.18	0.22	0.20
Raw, or Run-of-Mine Coal			
<u>Coal Component</u>	<u>Minimum Weight %</u>	<u>Maximum Weight %</u>	<u>Average Weight %</u>
Moisture	6.7	8.2	7.0
Ash*	16.0	17.5	16.5
Sulfur*	4.0	4.4	4.3
Chlorine*	0.11	0.16	0.13

* These components are reported on a moisture-free basis.

START-UP MATERIAL

Because of the possibility of tar evolution from coal at temperatures below 1200°F, it has been found advantageous to start the gasifier with carbonaceous material having a low volatile content. Bethlehem coke breeze has been found to be a suitable material for this. However the process is not limited to start-up with this particular type of coke. Any other material having similar properties can be substituted. The start-up material, like coal, should be crushed to less than 1/4-inch size before it is fed to the reactor. Typical analysis size consist and the physical properties of Bethlehem coke breeze are shown in Tables D-6, D-7, and D-8, respectively. Again the start-up material should be dried to less than 5% moisture for good flowability from the lockhoppers.

Table D-6. CHARACTERISTICS OF START-UP MATERIAL

Type of Coal: Bethlehem Coke Breeze

Analysis of Coke Breeze

Ultimate Analysis	wt %
Moisture	2.5
Ash	10.7
Carbon	84.3
Hydrogen	0.4
Nitrogen	0.6
Sulfur	0.8
Oxygen	<u>0.7</u>
Total	100.0
Proximate Analysis	
Moisture	2.5
Ash	10.7
Volatiles	3.0
Fixed Carbon	<u>83.8</u>
Total	100.0

Limitations: Moisture 5%

Table D-7. SIZE OF THE COKE FEED

<u>Retained</u>	<u>Microns</u>	<u>Cumulative wt %</u>
+6	3360	4
+12	1680	28
+20	850	55
+40	420	76
+70	210	86
+140	105	93
+270	53	97
-270	--	100

Table D-8. PHYSICAL PROPERTIES OF COKE

HHV, Btu/lb	12,569
Bulk Density, lb/ft ³	50
Particle (Apparent) Density lb/ft ³	95
Complete Fluidization Velocity, ft/s	
Fluidized Bed Density (At complete fluidization), lb/ft ³	
Poured Angle of Repose	
Angle of Internal Friction	

GASIFIER MATERIAL BALANCE

The heat and material balance for a U-GAS gasifier based on the design coal at full capacity is given in Table D-9. The medium for gasification is steam and oxygen. The conditions for the inlet feed streams for a gasifier operating at 90 psia are given in Table D-10. The balance utilizes the data from the successful pilot plant tests. The basis for the balance are also summarized in Table D-10.

Tables D-11 and D-12 depict the material balances at 75% and 65% of the capacity. There are several ways in which the gasifier can be turned down. These are described in a later section. In the present case, the 75% material balance is obtained by reducing the bed height to 44%, whereas 65% balance is obtained by maintaining 44% bed height while reducing the overall superficial velocity to 86%. The merits of methods of turn down are also described in a later section.

Table D-9. BASIS OF HEAT AND MATERIAL BALANCE

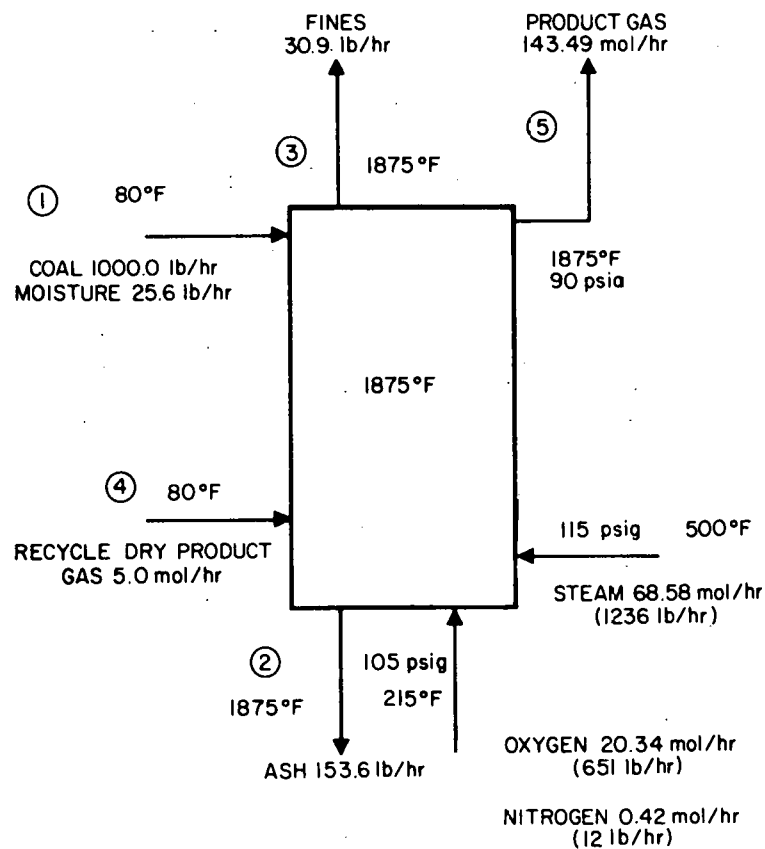
Feed Condition

Steam	500°F 115 psig
Oxygen	215°F 105 psig Purity - 98% O ₂ , 2% N ₂
Coal	80 °F 75 psig

Basis of Calculation

1. Ash Agglomerate Composition	85% Ash
2. Fines Losses	0.03 lb/lb of coal feed
3. Heat Losses	150,000 Btu/ton of dry coal
4. Product Recycle	0.1 lb/lb of coal

Table D-10. GASIFIER BALANCE; OPERATION AT 100% CAPACITY
(Basis: 1000 lb/hr Dry Coal Feed)

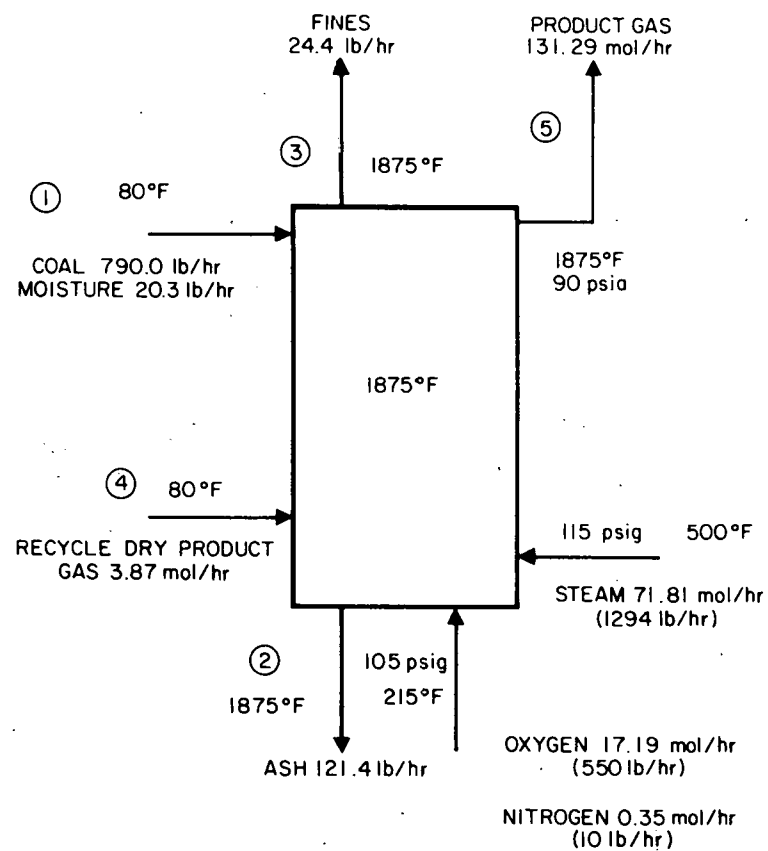


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Stream Number	①		②		③	
	Coal Feed		Residue		Fines	
Temperature, °F	80		1875		1875	
	lb/hr	wt %	lb/hr	wt %	lb/hr	wt %
C	686.4	68.64	21.5	13.99	23.5	75.98
H ₂	48.3	4.83	0.0	0.01	0.1	0.25
O	77.5	7.75	--	--	--	--
N ₂	11.2	1.12	0.5	0.35	0.2	0.71
S	39.3	3.93	1.0	0.65	0.4	1.36
Ash	137.3	13.73	130.6	85.00	6.7	21.70
Total	1000.0	100.00	153.6	100.00	30.9	100.00
Moisture	25.6					

Stream Number	④		⑤	
	Recycle Dry Product Gas		Product Gas	
Temperature, °F	80		1875	
	mol/hr	mol %	mol/hr	mol %
CO	1.08	27.90	21.23	16.17
CO ₂	0.99	25.44	19.37	14.75
H ₂	1.58	40.76	31.01	23.625
H ₂ O	--	--	54.19	41.29
CH ₄	0.19	5.01	3.82	2.91
H ₂ S	--	--	0.90	0.68
COS	--	--	0.04	0.03
N ₂	0.03	0.89	0.66	0.51
NH ₃	--	--	0.05	0.035
Total	3.87	100.00	131.27	100.00

Table D-11. GASIFIER BALANCE; OPERATION AT 75% CAPACITY



A79091973

Stream Number

Temperature, °F

C

H₂

O

N₂

S

Ash

Total

Moisture

Stream Number

Temperature, °F

CO

CO₂

H₂

H₂O

CH₄

H₂S

COS

N₂

NH₃

Total

①

Coal Feed

80

lb/hr wt %

542.3 68.64

38.2 4.83

61.2 7.75

8.8 1.12

31.0 3.93

108.5 13.73

790.0 100.00

20.3

②

Residue

1875

lb/hr wt %

17.0 13.99

0.0 0.01

-- --

0.4 0.35

0.8 0.65

103.2 85.00

121.4 100.00

③

Fines

1875

lb/hr wt %

18.5 75.98

0.1 0.25

-- --

0.2 0.71

0.3 1.36

5.3 21.70

24.4 100.00

④

Recycle Dry Product Gas

80

mol/hr mol %

0.98 29.27

0.81 24.35

1.35 40.39

-- --

0.17 5.10

-- --

-- --

0.03 0.89

-- --

3.34 100.00

⑤

Product Gas

1875

mol/hr mol %

19.02 17.67

15.83 14.70

26.23 24.375

41.84 38.87

3.32 3.08

0.77 0.71

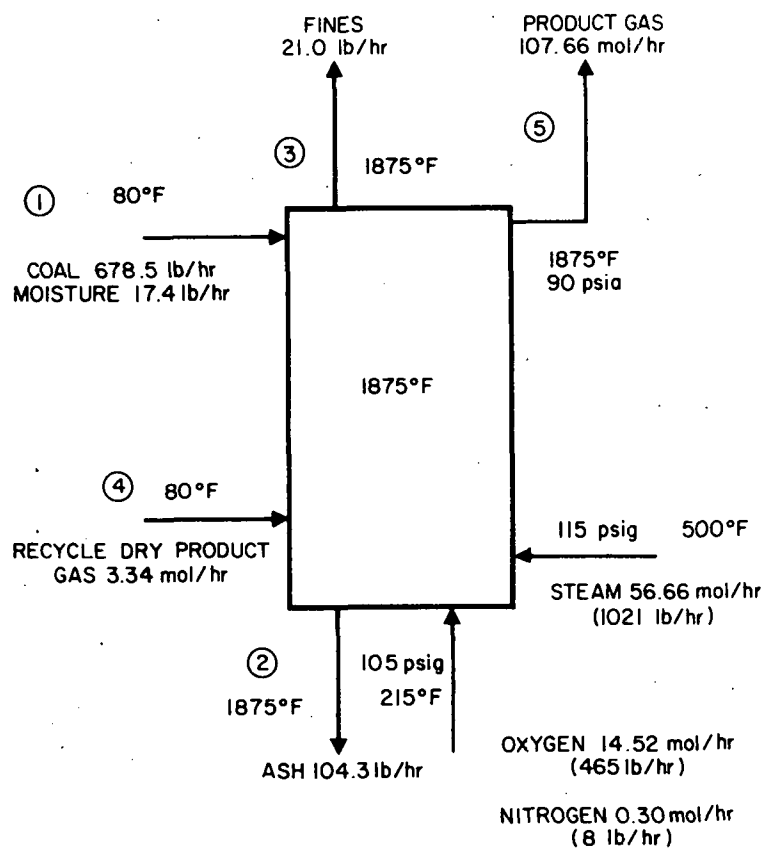
0.03 0.03

0.56 0.53

0.04 0.035

107.64 100.00

Table D-12. GASIFIER BALANCE; OPERATION AT 65% CAPACITY



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Stream Number

Temperature, °F

C

H₂

O

N₂

S

Ash

Total

Moisture

Stream Number

Temperature, °F

CO

CO₂

H₂

H₂O

CH₄

H₂S

COS

N₂

NH₃

Total

①

Coal Feed

80

lb/hr wt %

465.6 68.64

32.8 4.83

52.6 7.75

7.6 1.12

26.7 3.93

93.2 13.73

678.5 100.00

17.4

②

Residue

1875

lb/hr wt %

14.6 13.99

0.0 0.01

-- --

0.4 0.35

0.7 0.65

88.6 85.00

104.3 100.00

③

Fines

1875

lb/hr wt %

15.9 75.98

0.1 0.25

-- --

0.1 0.71

0.3 1.36

4.6 21.70

21.0 100.00

④

Recycle dry
Product Gas

80

mol/hr mol %

1.63 32.59

1.08 21.55

1.98 39.63

-- --

0.27 5.34

-- --

-- --

0.04 0.58

-- --

5.00 100.00

⑤

Product Gas

1875

mol/hr mol %

30.85 21.50

20.42 14.23

37.52 26.155

47.58 33.16

5.06 3.53

1.13 0.79

0.05 0.03

0.81 0.57

0.05 0.035

143.47 100.00

GASIFIER OPERATING CONDITIONS

Range of Operating Conditions

The U-GAS Process has operated over a wide range of temperatures, pressures, and superficial velocity. Table D-13 lists the range of conditions under which the operation of the process has been free from unwarranted sinter and clinker formation in the reactor and undesired accumulation of ash in the reactor and which have produced near design quality of the medium-Btu gas. The clinker formation is defined as the uncontrolled growth of ash agglomerates. It differs from sinter formation in the sense that sinter is formed by cohesive forces between the ash particles. Both the clinker formation and sinter formation are undesirable for the process. The clinker and sinter formation are both functions of bed temperature, bed ash concentration, superficial velocity, and the local oxygen concentration. The recommended operating conditions are the result of extensive pilot plant test work aimed to alleviate this undesirable aspect of the process. Depending upon the feedstock and gas load, a judicial adjustment of temperature ($\pm 25^{\circ}\text{F}$) may be required to achieve ash balanced operation.

Gasifier Turndown

The capacity of the gasifier can be reduced in several ways. At constant pressure, three operating variables can be adjusted to achieve the load following. These variables are temperature, bed height, and superficial velocity. The gasification rate is an exponential function of the temperature, a linear function of the superficial velocity and about the $1/3$ power of the bed height. Therefore, a large turndown is possible by a slight change in the bed temperature. However, if the lower temperature is out of the range for ash agglomeration, other means of turndown should be employed. The reduction in the superficial velocity should be limited so that it is above its limitation of preventing the sinter formation. The variation in the bed height should be utilized as the last alternative for achieving the turndown. If constant pressure is not the desired process criterion, reduction in the reactor pressure will also achieve a linear turndown in the gasification unit. Table D-14 summarizes the results of four cases at different capacities and shows important flowstreams and operating conditions.

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Table D-13. RANGE OF OPERATING CONDITIONS

Bed Temperature	1825° - 1900°F
Reactor Pressure	90 psia
Superficial Velocity	3 - 4 ft/s
Bed Height	25 ft
Bed Ash	40% - 60%

Table D-14. GASIFIER TURNDOWN

	Base Case	Case 1	Case 2	Case 3	Case 4
Feed Coal, tons/hour	37.11	33.85	30.27	27.38	22.32
Steam, mph	4964	5058	3582	3681	3827
Oxygen, mph	1472	1332	1155	1059	915
Nitrogen, mph	30	27	24	22	19
Product Gas, mph	9839	9665	7697	7399	6878
Ash, tons/hour	5.56	5.07	4.53	4.10	3.34
Fines, lb/hr	2235	2038	1822	1655	1344
Temperature, °F	1875	1850	1875	1850	1850
Superficial Velocity, ft/s	4.0	4.0	3.0	3.0	3.0
HHV, Btu/SCF	287.8	282.7	293.7	287.9	274.9
Total Heating Value, 10 ⁹ Btu/day	17.0	15.6	14.1	12.7	10.1
Total Heat Loss, 10 ⁶ Btu/hr	5.5	5.5	5.5	5.5	5.5
Bed Height, ft	24.4	24.9	24.5	25.0	12.4
Capacity, %	100	92	83	75	59
Gas Composition (Dry, H ₂ S free), Mol %					
CO	32.59	29.87	34.15	31.60	27.57
CO ₂	21.56	23.07	20.01	21.79	25.08
H ₂	39.63	40.90	39.51	40.30	41.38
CH ₄	5.34	5.30	5.47	5.45	5.10
N ₂	0.88	0.86	0.86	0.86	0.87
Cold Gas Efficiency,* %	78.1	78.6	79.4	79.1	77.1

* Based on total thermal output of the gasifier.

FINES CHARACTERISTICS

Nominally the process gasifies and agglomerates ash from all the fines returned to the gasifier. However, the amount of fines ultimately returned to the gasifier depends upon the collection efficiencies of the cyclones. For achieving maximum coal conversion efficiency it is recommended that two-stage fines collection be employed and fines collected in both the stages returned to the gasifier. The fines loading and the recycle ratio from the two stages of fines collection depends largely on the superficial velocity and the bed temperature. Tables D-15 through D-19 show the characteristics of fines to be expected at various stages of the process.

For the cases when the fines are recycled, the design numbers are fairly conservative so that the pipings to return the fines and in the gas cleanup sections can be adequately designed to handle the cases when the upstream equipment may malfunction.

Table D-15. FIRST-STAGE FINES (Always Recycled)

Cyclone Loading: 1.1 lb/lb of coal feed

Recycle Rate: 1.08 lb/lb of coal feed

Size of the First-Stage Fines		
<u>Retained</u>	<u>Microns</u>	<u>Cumulative wt %</u>
+70	210	33.0
+140	105	82.9
+200	74	87.8
+270	53	92.3
-270	--	100

 Physical Properties of First-Stage Fines

Bulk Density, lb/ft³Particle (Apparent) Density, lb/ft³

Complete Fluidization Velocity, ft/s

Fluidized Bed Density

(At complete fluidization), lb/ft³Dipleg Density, lb/ft³

26

Poured Angle of Repose

Angle of Internal Friction

Heat Capacity, Btu/lb-°F

$$0.1517 + 3.95 \times 10^{-4}t - 2.13 \times 10^{-7}t^2 + 4.25 \times 10^{-9}t^3$$

Table D-16. SECOND-STAGE FINES (Not Recycled)

Cyclone Loading: 0.025 lb/lb of coal feed

Size of the Second-Stage Fines (When Not Recycled)

<u>Retained</u>	<u>Microns</u>	<u>Cumulative wt %</u>
+70	210	--
+140	105	38.0
+200	74	47.0
+270	53	54.7
-270	--	100

Physical Properties of Second-Stage Fines

Bulk Density, lb/ft ³	37.5
Particle (Apparent) Density, lb/ft ³	68
Complete Fluidization Velocity, ft/s	0.06
Fluidized Bed Density (At complete fluidization), lb/ft ³	27.0
Dipleg Density, lb/ft ³	--
Poured Angle of Repose	--
Angle of Internal Friction	42 deg
Heat Capacity, Btu/lb-°F	$0.1517 + 3.95 \times 10^{-4}t - 2.13 \times 10^{-7}t^2 + 4.25 \times 10^{-9}t^3$

Table D-17. SECOND-STAGE FINES (Recycled)

Cyclone Loading*: 0.43 lb/lb of coal feed

Recycled Rate: 0.4 lb/lb of coal feed

Size of the Second-Stage Fines (When Recycled)

<u>Retained</u>	<u>Microns</u>	<u>Cumulative wt %</u>
+200		29.0
+230		34.5
+50 microns		48.3
+40		77.1
+20		92.0
+10		98.8
+5		100.0

Physical Properties of Second-Stage Fines (With Recycle)Bulk Density, lb/ft³Particle (Apparent) Density, lb/ft³

Complete Fluidization Velocity, ft/s

Fluidized Bed Density

(At complete fluidization), lb/ft³Dipleg Density, lb/ft³

18

Poured Angle of Repose

Angle of Internal Friction

Heat Capacity, Btu/lb-°F $0.1517 + 3.95 \times 10^{-4}t - 2.13 \times 10^{-7}t^2 + 4.25 \times 10^{-9}t^3$

* Conservative estimate for designing dipleg sizes adequately.
Actual figures may be lower.

Table D-13. FINES LOSSES (Without Recycle)

Rate of Fines Losses: 0.001 lb/lb of coal feed

Size of the Fines Escaping the Second-Stage
Cyclone With No Recycle

<u>Retained</u>	<u>Microns</u>	<u>Cumulative wt %</u>
+70	210	0
+140	105	8.2
+200	74	12.8
+270	53	16.5
-270	--	100

Physical Properties of Fines Escaping Second-Stage
Cyclone With No Recycle

Bulk Density, lb/ft ³	14
Particle (Apparent) Density, lb/ft ³	68
Complete Fluidization Velocity, ft/s	
Fluidized Bed Density (At complete fluidization), lb/ft ³	
Dipleg Density, lb/ft ³	
Poured Angle of Repose	
Angle of Internal Friction	
Heat Capacity, Btu/lb-°F	$0.1517 + 3.95 \times 10^{-4}t - 2.13 \times 10^{-7}t^2 + 4.25 \times 10^{-9}t^3$

Table D-19. FINES LOSSES (With Recycle)

Rate of Fines Losses*: 0.03 lb/lb of coal feed

<u>Size of Fines Escaping Second Cyclone With Recycle</u>		
<u>Retained</u>	<u>Microns</u>	<u>Cumulative wt %</u>
	50	4.0
	40	10.0
	20	21.5
	10	67.0
	5	100.0

Physical Properties of Fines Escaping Second Cyclone
With Recycle

Bulk Density, lb/ft ³	14
Particle (Apparent) Density, lb/ft ³	68
Complete Fluidization Velocity, ft/s	
Fluidized Bed Density (At complete fluidization), lb/ft ³	
Dipleg Density, lb/ft ³	
Poured Angle of Repose	
Angle of Internal Friction	
Heat Capacity, Btu/lb-°F	$0.1517 + 3.95 \times 10^{-4}t - 2.13 \times 10^{-7}t^2$ $+ 4.25 \times 10^{-9}t^3$

* Conservative estimate for designing handling equipment.
Actual losses may be lower.

ASH AGGLOMERATES

For process purposes, the ash agglomerates are defined as high-ash-bearing material, generally greater than 75% ash, discharged from the gasifier through the venturi annulus. Although this definition is applicable regardless of the materials being processed in the gasifier, it is not unusual to find the ash agglomerates in a variety of shapes and sizes corresponding to a given feed material and operating conditions. Generally, the higher temperatures promotes the growth of ash agglomerates. Tables D-20 and D-21 show the typical size and physical properties of ash agglomerates produced from the Western Kentucky No. 9 coal. Table D-22 is the analysis of ash to indicate a typical mineral content of the ash.

Table D-20. SIZE OF ASH AGGLOMERATES

Coal Type: Western Kentucky No. 9 Coal

<u>Retained</u>	<u>Microns</u>	<u>Cumulative wt %</u>
+6	3360	4.5
+12	1680	23.5
+20		93.6
+40	420	99.3
-40	--	100

Table D-21. PHYSICAL PROPERTIES OF ASH AGGLOMERATES

Coal Type: Western Kentucky No. 9 Coal

Bulk Density, lb/ft ³	76
Particle (Apparent) Density, lb/ft ³	140
Complete Fluidizing Velocity, ft/s	3.3
Fluidized Bed Density (At complete fluidization), lb/ft ³	65
Poured Angle of Repose	30 deg
Angle of Internal Friction	40 deg
Specific Heat, Btu/lb-°F	$0.18 + (7.78 \times 10^{-5})t; (t^{\circ}\text{F})$
Ash Content, %	85 - 95

Table D-22. INORGANIC ANALYSIS OF ASH AGGLOMERATES

Coke Type: Western Kentucky No. 9 Coal

<u>Mineral Analysis of Ash</u>	<u>Percent Weight Ignited Basis</u>
Silica, SiO_2	47.00
Alumina, Al_2O_3	22.21
Titania, TiO_2	1.03
Ferric Oxide, Fe_2O_3	19.75
Lime, CaO	3.00
Magnesia, MgO	1.02
Potassium Oxide, K_2O	2.55
Sodium Oxide, Na_2O	0.53
Sulfur Trioxide, SO_3	2.69
Phos. Pentoxide, P_2O_5	0.19
Strontium Oxide, SrO	0.00
Barium Oxide, BaO	0.01
Manganese Oxide, Mn_3O_4	0.00
Undetermined	<u>0.02</u>
Total	100.00
Alkalies as Na_2O , Dry Coal Basis	= 0.18
Silica Value	= 66.41
Base Acid Ratio	= 0.38

SCRUBBER WATER ANALYSIS

Trace Element Data

The trace elements present in the scrubber water were derived from the analysis of this water by EIA. These data have been compared with the trace element analysis performed on the feed, fines, and ash discharge. The primary purpose of this analysis is to indicate the tendency of the trace elements either to be volatile or to be retained in ash. The data presented in Table D-23 are completely based on the pilot plant data; the following assumptions are implicit in extrapolating the pilot plant data to the demonstration plant.

1. Fines and ash produced from run-of-mine coal and washed coal are similar.
2. The pilot plant material balance was adjusted to match the demo plant material balance.
3. Fines carryover in the demo plant was similar to the fines collected in the third-stage cyclone of the pilot plant.
4. Trace analysis of ash from the demo plant is assumed to be similar to that of the pilot plant ash regardless of their similarity in carbon/ash ratio.
5. Whenever the quantity of a particular trace element was analyzed to be greater in the ash than in the feed, it was assumed that the trace element did not volatilize.

Priority Pollutants Data

The priority pollutants expected to be found in the product gas emanating from the gasifier are based on the actual analysis performed on the scrubber water collected during the pilot plant tests. These data are given in Table D-24.

Minor Constituents

The data tabulated for minor constituents in Table D-25 are based on the analysis performed on the scrubber water collected during the pilot plant tests. The concentrations of these constituents were corrected for any of these constituents found in the water used for the scrubber.

Table D-23. TRACE ELEMENT DATA
(Based on Scrubber Water Analysis)

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	Western Kentucky No. 9 ROM Coal			Western Kentucky Washed Coal		
	Feed 1000 lb/hr	Ash 203 lb/hr	Fines 30 lb/hr	Feed 1000 lb/hr	Ash 154 lb/hr	Fines 30 lb/hr
	10 ⁻³ lb/hr					
Sb	--	--	--	0.4	--	--
As	6.6	0.5	0.3	3.8	0.3	0.17
B	20	9.8	2.1	40	19.6	4.2
Ba	9.9	16.1	0.5	2.8	2.8	0.1
Be	1.0	0.5	0.05	1.2	0.6	0.06
Cd	1.7	1.1	0.1	0.24	0.16	0.01
Cl	1900	30.3	--	1300	20.7	--
Cr	17.3	19.2	--	20	20	--
Co	6.1	6.4	0.3	2.5	2.5	0.2
Cu	11.0	13.4	1.0	11	11	1.0
F	120	3.2	--	47	1.3	--
Fe	19,000	22,288	457	12,000	12,000	288
Pb	5.4	2.6	0.5	7.8	3.8	0.7
Li	5.7	5.6	0.3	6.6	6.5	0.3
Mn	55.9	67.4	2.9	33	33	1.7
Mg	--	--	--	--	--	--
Mo	9.2	10.1	0.5	9.3	9.3	0.5
Ni	10.5	14.9	--	13	13	--
Se	1.7	0.5	0.03	0.1	0.03	0.2
Te	<0.1	<0.02	<0.003	6.0	<1	<0.2
Tl	<1.2	<8	0.5	0.88	0.88	--
Sn	--	--	--	7	--	--
V	57.8	53.9	3.4	76	70.9	4.5
Zn	61.7	97.8	4.9	27	27	--
Hg				0.064		

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Table D-24. PRIORITY POLLUTANTS DATA
 (Based on Scrubber Water Analysis;
 Basis: 1000 lb/hr Coal Feed)

<u>Pollutant</u>	<u>10⁻³ lb/hr</u>	<u>Pollutant</u>	<u>10⁻³ lb/hr</u>
Asbestos [Million Fibers per liter (MFL)]			
Phenanthrene	7.56	Phenols	3.1
Anthracene	0.6	Ni	0.2
Fluorene	0.2	Zn	6.51
Acenaphthylene	2.8	Be	<0.02
Acenaphthene	0.04	Tl	0.3
Naphthylene	23	Pb	2.5
Fluoranthene	11.2	Cd	0.09
Pyrene	12	As	0.08
Chrysene	0.2	Sb	0.05
Benzo(a)anthracene	0.2	Sc	0.3
Indeno Pyrene	0.2	Hg	<0.004
1,12 Benzoperylene	2.6	Ag	<0.04
Benzo(c)Phenanthrene	0.9	Total Cr	0.2
Biphenyl	0.04		
Benzo(a)pyrene	1.5		
Benzo(b)Fluoranthene	3.9		
Benzo(k)Fluoranthene	2.6		
Benzo(f)Fluoranthene	2.7		
Ethyl Hexyl Phthalate	ND		
Di-n-octyl Phthalate	ND		
Phenol	0.1		
2,4-Dimethylphenol	0.04		
Chloroform	1.3		
Benzene	27		
Toluene	0.7		

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Table D-25. MINOR CONSTITUENTS PRESENT IN THE SCRUBBER WATER
(Basis: 1000 lb/hr Coal Feed)

<u>Parameter</u>	<u>10⁻³ lb/hr</u>	<u>Parameter</u>	<u>10⁻³ lb/hr</u>
pH		Ca	460
Total Dissolved Solids	0	Total Fe	54
Total Suspended Solids	2950	Dissolved Fe	0.3
Total Acidity	400	Mg	0
Oil and Grease	40	K ⁺	23
Total Kjeldahl N	1300	Si	5.9
NO ₃ -N	1.45	Na ⁺	51
NO ₂ -N	0	B	66
NH ₃ -N	950	Bi	<20,000
Total Phosphorus	0	Cr ⁺⁶	0
Ortho Phosphate	0	Niobium	240
Total Dissolved Phosphate	0	Ga	0.9
BOD	118	Lanthanum	0.002
COD	272	Mn	0.5
TOC	296	Mo	0.07
Soluble Organic Carbon	70	Rb	0.1
Bicarbonate Alkalinity mg/l CaCO ₃	1000	Sc	1.1
Carbonate Alkalinity mg/l CaCO ₃	0	Sn	0.6
Cl ⁻	1187	Ti	<2.36
CN ⁻ (Total)	28.3	W	0.04
CN ⁻ (Free)	13	V	0.01
Iodide	0.05	U	0.01
F ⁻	14	Zr	2.4
SO ₄ ⁼	590	Carbonile	<0.1
Al	35	CS ₂	<0.1
Ba	1.0	H ₂ S	177
Bromide	1.9	HCN	15.4
Ge	<2.36	Total Sulfur	402
CO	0.02	Sulfide	225
		Sulfite	615
		SCN ⁻	47

SALTATION VELOCITY

The saltation velocity for feed, fines, bed material, and ash are given in Table D-26. Since the saltation velocity of the feed is close to 30 ft/s, it is recommended that the solid conveying velocity in excess of 30 ft/s be employed in the feed nozzles. (The saltation velocities were determined in the apparatus shown in Figure D-2.)

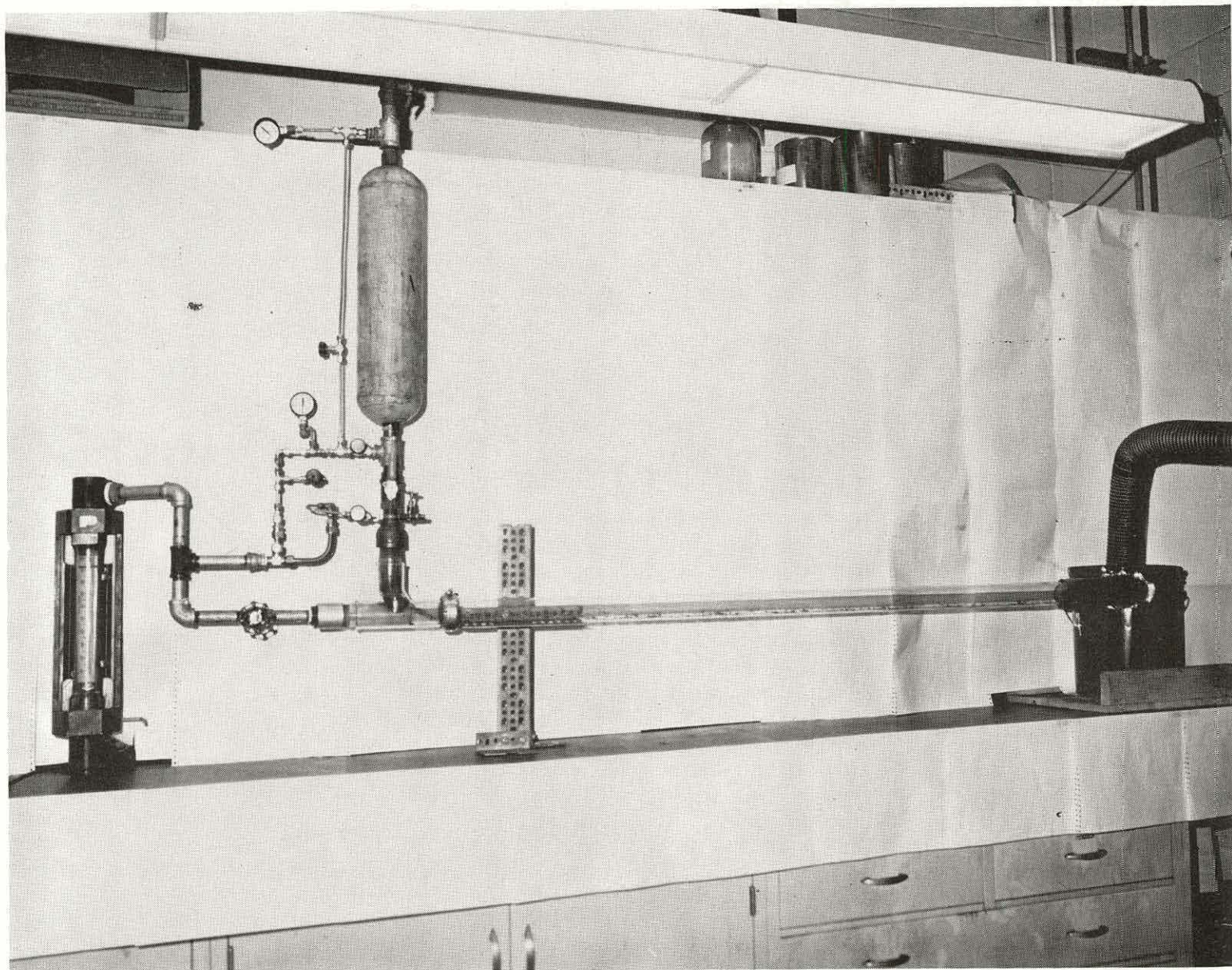


Figure D-2. EQUIPMENT FOR SALTATION VELOCITIES

Table L-26. DATA ON SALTATION VELOCITY

<u>Material</u>	<u>Saltation Velocity at Standard Temperature and Pressure in Air, ft/s</u>
Feed (ROM Western Kentucky Coal)	26
Feed (Washed Western Kentucky Coal)	28
Coke	
Bed Material	24
Ash Agglomerates	33
Second-Stage Fines	17
Third-Stage Fines	20

SYNOPSIS OF CALCULATIONS ON SCALE-UP OF MLGW/DOE DEMONSTRATION PLANT REACTOR

Introduction

The objective of this work is to develop a rational basis for scale-up of the reactor in the Institute of Gas Technology (IGT) Demonstration Plant. The bases of design of the demonstration plant are operations in the IGT Ash-Agglomerating Gasifier (AAG) Pilot Plant, which were conducted at a maximum pressure of 42 psig. The demonstration plant reactor is being designed to operate at 75 psig. In addition, the capacity of the demonstration plant reactor will differ from that of the pilot plant reactor. For these reasons, scale-up is required.

It is felt that the scale-up parameter of greatest importance is the mass fraction in the reactor of "agglomeratable" particles. Such particles are defined to be those that have been converted under such conditions that within the particle a temperature above some pseudo ash-melting temperature has been reached. Presumably, if there are equal concentrations of these (with the same dimensionless spatial distribution), the relative rate of agglomeration will be identical. Relative rate of agglomeration refers to the rate normalized by the feed ash rate. Pseudo ash-melting temperature is the temperature at which ash melts when the ash has a specified composition and has been exposed to the oxidizing atmosphere in the jet. These matters have been discussed fully in the Ash Chemistry in the U-GAS[®] Process special report.

The processes taking place in the vicinity of or in the central jet are of primary importance as approximately 50% of the total gas flow and almost all of the oxygen are admitted therein. The assumption has been made that the venturi jet will not contribute appreciably to the physico-chemical hydrodynamics in the vicinity of the central jet. The justification for this is as follows. The mass flow of gas in the venturi annulus is less than 1/3 that in the jet, and the velocity in the annulus is approximately 10% of the velocity in the central jet. The combined effect on the inertial forces, which are proportional to the mass times the velocity squared, results in a ratio of inertial forces of the central jet to inertial forces of the venturi of approximately 300:1.

To facilitate the analysis, three scenarios have been defined. These are referred to as Case I, Case II, and Case III.

- Case I - Pilot Plant, single venturi

- Case II - Demonstration Plant, multiple (3) venturis
- Case III - Demonstration Plant, single venturi

Further details of the specific process parameters that define the above cases appear in Table D-27. In certain instances, some of the parameters defining the cases are taken as variables; these instances are explained in the discussion.

Discussion

Jet Penetration

The jet penetration, or length of the cylindrical jet, was calculated according to the method of Merry.¹ An alternative correlation of jet lengths developed by Westinghouse² was compared with the Merry correlation at the specific conditions of Cases I through III and are presented in Table D-28. For Cases II and III (Demonstration Plant), the deviations between the two correlations were in the range of -1% to +23%, which is not significant for the purposes of this analysis. For the pilot-plant case (Case I), deviations of the order of 100% were found, with the Westinghouse correlation yielding higher jet penetrations. Figures D-3 and D-4 compare the correlations for jet penetration at three values of jet velocity and all orifice diameters. Deviations in Case II or Case III increase outside the range of orifice diameters of 2 to 4 inches. At this time, no recommendation as to which correlation should be adopted will be made. Experimental work should be performed on our specific materials to establish this point or to develop our own correlation. Because significant work was performed using the Merry correlation prior to the analysis conducted above, the latter will be adopted. The effect of the alternate assumption on the final results will be commented upon later.

Entrainment to Jet

The jet penetrations as computed above were required for further calculations of the quantity of emulsion phase entrained into the jet. The Merry³ correlation was used for this feature. This correlation was developed from observations of the behavior of lead-water systems, but it is believed to be applicable to jets in fluidized beds.

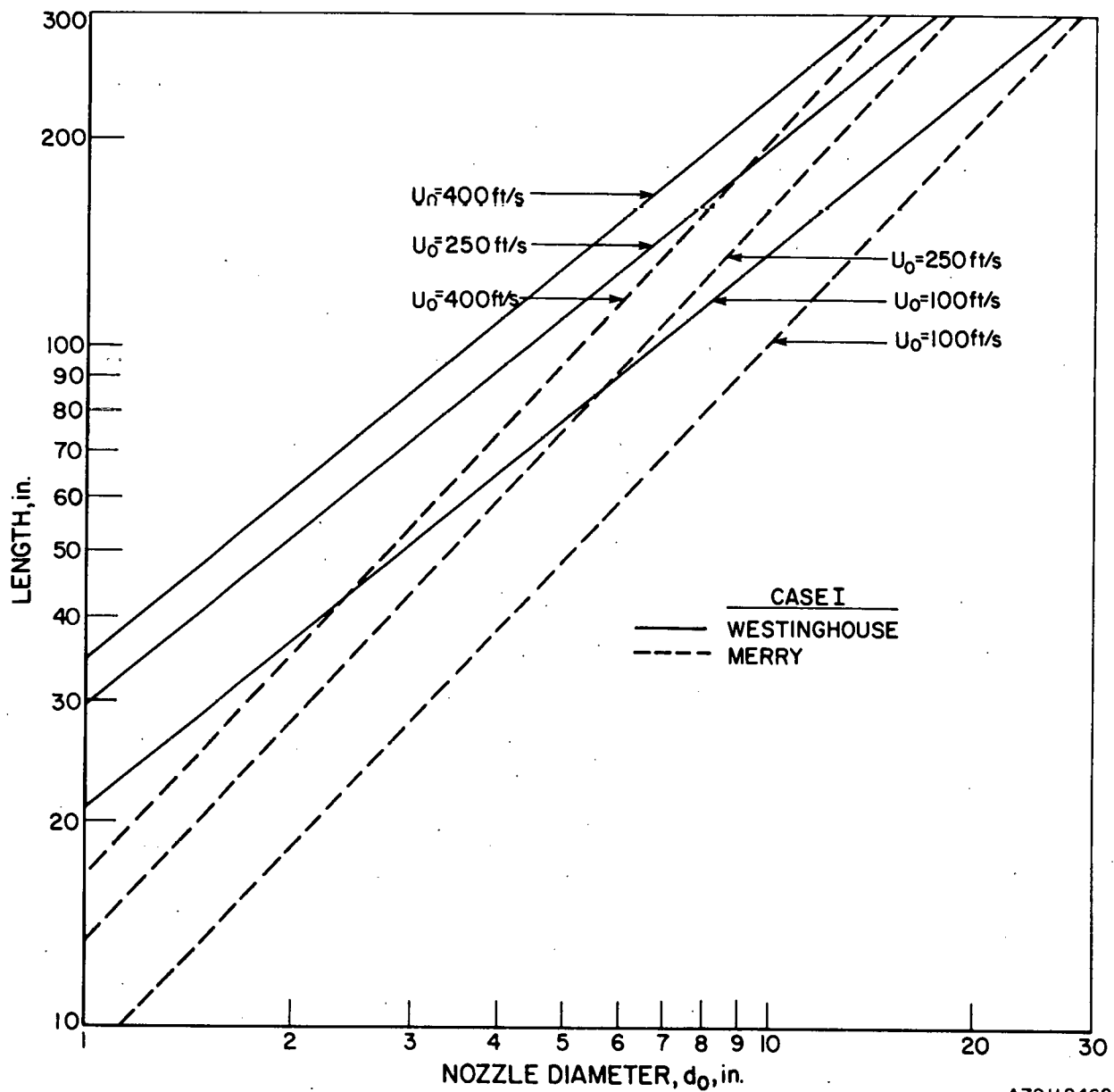
The additional assumptions required are the voidage of the emulsion, assumed at minimum fluidizing conditions, which was taken to be 0.60, and the minimum fluidizing velocity of the bed, which is known to be of the order of 1 ft/s. The angle of repose was taken to be 70° for the purposes of using the Merry correlation. The bed diameter is also needed, which was taken as 3 feet for Case I, 6 feet for Case II, and 13.5 feet for Case III. The bed diameter in Case II is an effective bed diameter for a multiple-venturi arrangement where the individual venturis are assumed to act independently of each

Table D-27. DEFINITION OF PARAMETERS FOR CASES I, II, AND III

<u>Parameter</u>	<u>Pilot Plant</u>	<u>Demonstration Plant</u>	
	<u>Case I</u>	<u>Case II</u>	<u>Case III</u>
Orifice Diameter, d_o , in.	1.5	3.55	6.07
Orifice Velocity, U_o , ft/s	400	400	400
Gas Density, ρ_g , lb/ft ³	0.134	0.265	0.265
Pressure, psia	57	90	90
Gas Temperature in Jet, °F	600	600	600
Particle Density, ρ_p , lb/ft ³	90	90	90
Average Particle Diameter, d_p , in. of Bed Material	0.09	0.09	0.09
Molar Flow Rate of Gas to Each Jet, lb-mol/hr	100	1117	3350
Oxygen Concentration in Jet, %	40	40	40
Total Gas Fed to Gas in Jet, %	50	50	50

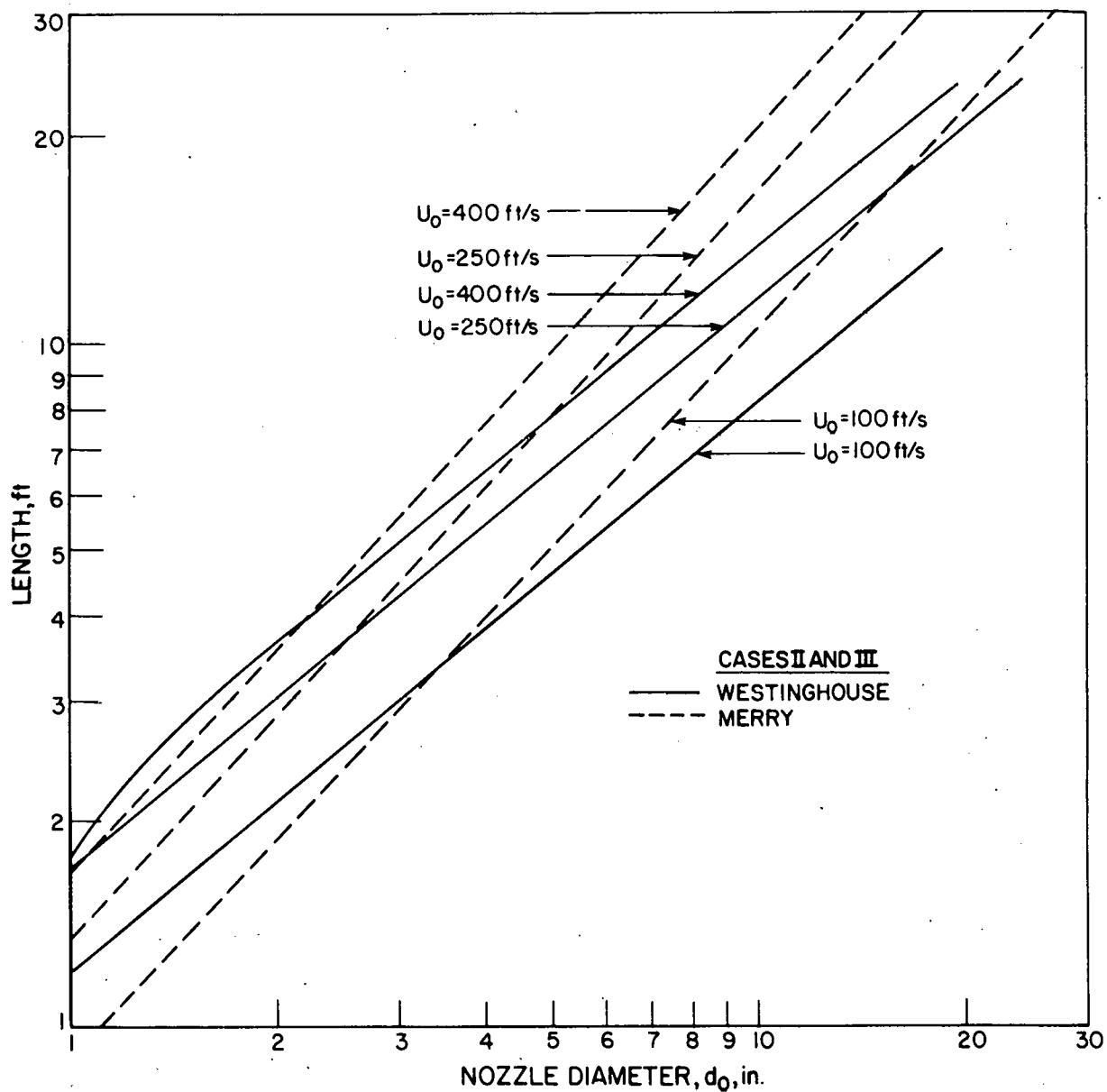
Table D-28. COMPARISON OF MERRY AND WESTINGHOUSE JET PENETRATION
CORRELATIONS FOR BASE CASE CONDITIONS

Orifice Velocity, U_o , ft/s	L (Merry)	L (Westinghouse)	$\frac{L_{\text{Merry}} - L_{\text{Westinghouse}}}{L_{\text{Merry}}} \times 100$
Case I ($d_o = 1.5$ in.)			
100	1.12	2.40	-114.3
250	1.71	3.37	-97.1
400	2.10	4.02	-91.4
Case II ($d_o = 3.55$ in.)			
100	3.41	3.45	-1.1
250	5.27	4.85	+8.0
400	6.53	5.79	+11.4
Case III ($d_o = 3.55$ in.)			
100	5.98	5.33	10.9
250	9.35	7.51	19.7
400	11.62	8.95	23.0



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Figure D-3. COMPARISON OF "MERRY" JET PENETRATION
AND WESTINGHOUSE CORRELATION (Case I)



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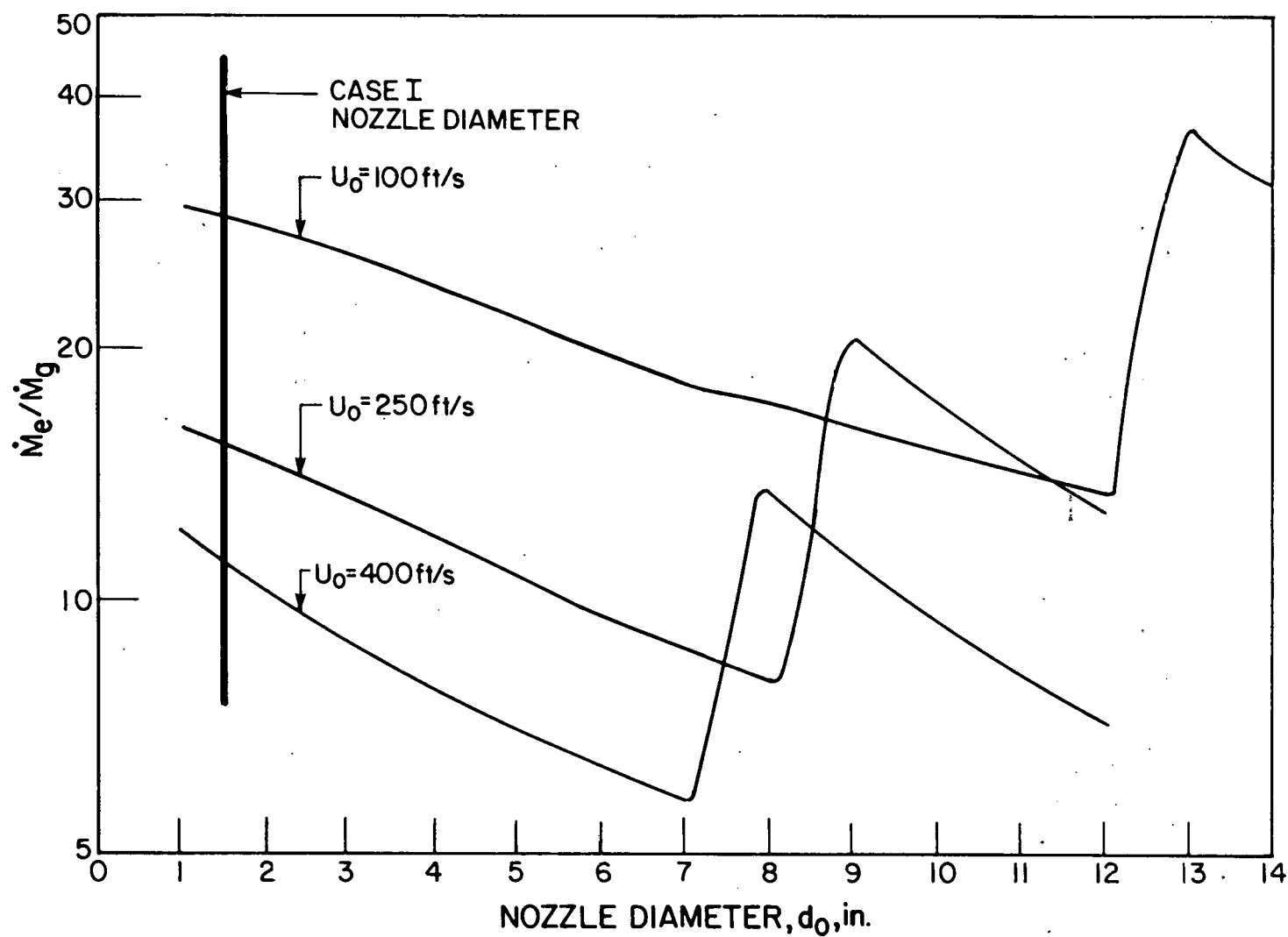
Figure D-4. COMPARISON OF "MERRY" JET PENETRATION AND WESTINGHOUSE CORRELATION (Cases II and III)

other. It is unknown whether this assumption or the alternative of using the actual bed diameter (13.5 feet) is appropriate. Figure D-5 gives the prediction of the Merry correlation of entrainment for Case I over all orifice diameters at three levels of orifice velocity. As the true orifice velocity used appears to be approximately 400 ft/s, the entrainment rate at the actual orifice diameter of 1.5 inches is 11.0 lb solids/lb gas. The correlation behaves in a peculiar fashion at large orifice diameters (about 7 inches at $U_o = 400$ ft/s), but, fortunately, this region is beyond the tried pilot plant conditions to date. Similar curves for Cases II and III have been developed and are presented in Figure D-6.

Because the curves in Figure D-6, which represent all values of d_o for Case II and Case III, differ only in bed diameter, this figure indirectly gives us a measure of the effect of bed diameter. Increasing the bed diameter increases the entrainment $\dot{M}_e/\dot{M}_g$, lb solids/lb gas in the jet. At the particular values chosen for orifice diameters in Cases II and III, 3.55 and 6.07 inches, respectively, the deviations of the assumed Case II bed diameter go through a minimum at 250 ft/s (Case II and Case III lines coincide at $d_o = 3.55$ inches). The maximum deviation of concern to us is at $U_o = 400$ ft/s, which is approximately 14% of the predicted value based on the smaller bed diameter that was chosen.

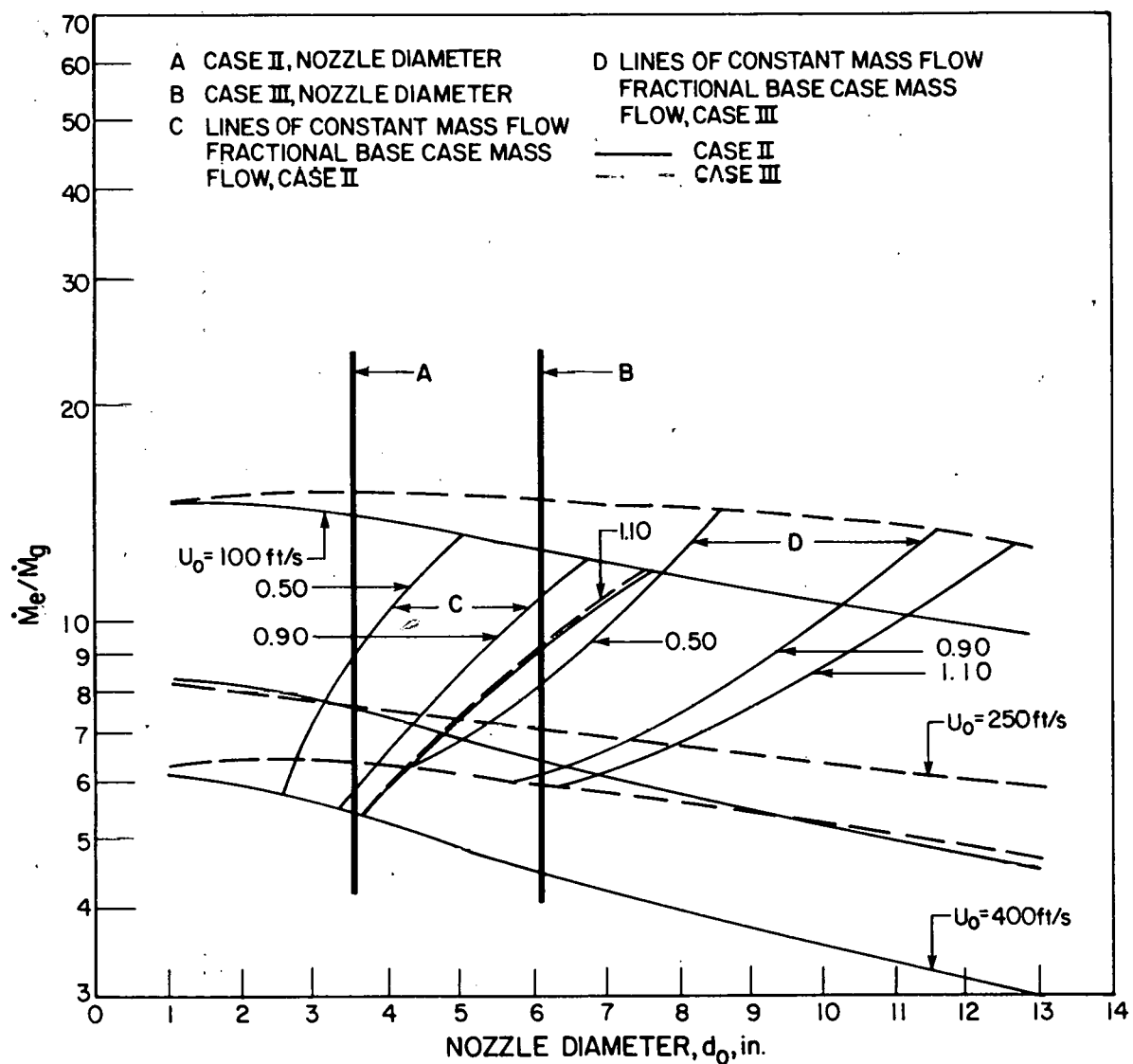
The sensitivity of the curves ($\frac{\partial [\dot{M}_e/\dot{M}_g]}{\partial d_o}$) is greatest for higher orifice velocities and for smaller bed diameters. The maximum sensitivity is present on the lowest curve in Figure D-6 and is approximately -0.28 lb solids/lb gas-inch. Even this degree of sensitivity doesn't seem appreciable. Thus, lowering the orifice size from approximately 3.6 inches to 2.6 inches would improve the total entrainment in the reactor by only 0.28 lb solids/lb gas, which is only 5% of the current value of 5.5 lb solids/lb gas.

We have been considering the effect of a change in d_o on the entrainment. Because $\dot{M} = \rho U_o \frac{\pi}{4} d_o^2$, and we are considering movement along a line of constant U_o , we must be considering changes in $\dot{M}$ (gas mass rate $\dot{M}$ to central jet). Gas displaced in this fashion must be utilized in the venturi annulus or grid. If, instead, we maintain our mass rate in the jet, then $U_o d_o^2$ will be constant. Lines corresponding to this equation are indicated in Figure D-6 at 50%, 90%, and 110% of the base case central jet flow. (See Table D-27.) Only lines representing the same cases can be connected in this manner. Higher $\dot{M}_e/\dot{M}_g$ can be obtained by reducing the gas velocity in the jet at constant mass rate $\dot{M}$ in the jet, which necessitates higher orifice sizes. Higher entrainment rates can be obtained by reducing the mass rate in the jet, which requires a reduced orifice diameter, at any value of velocity. These rate changes may also be made in concert.



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Figure D-5. GAS ENTRAINMENT RATE INTO JET/GAS RATE INTO JET FROM NOZZLE VS. NOZZLE DIAMETER



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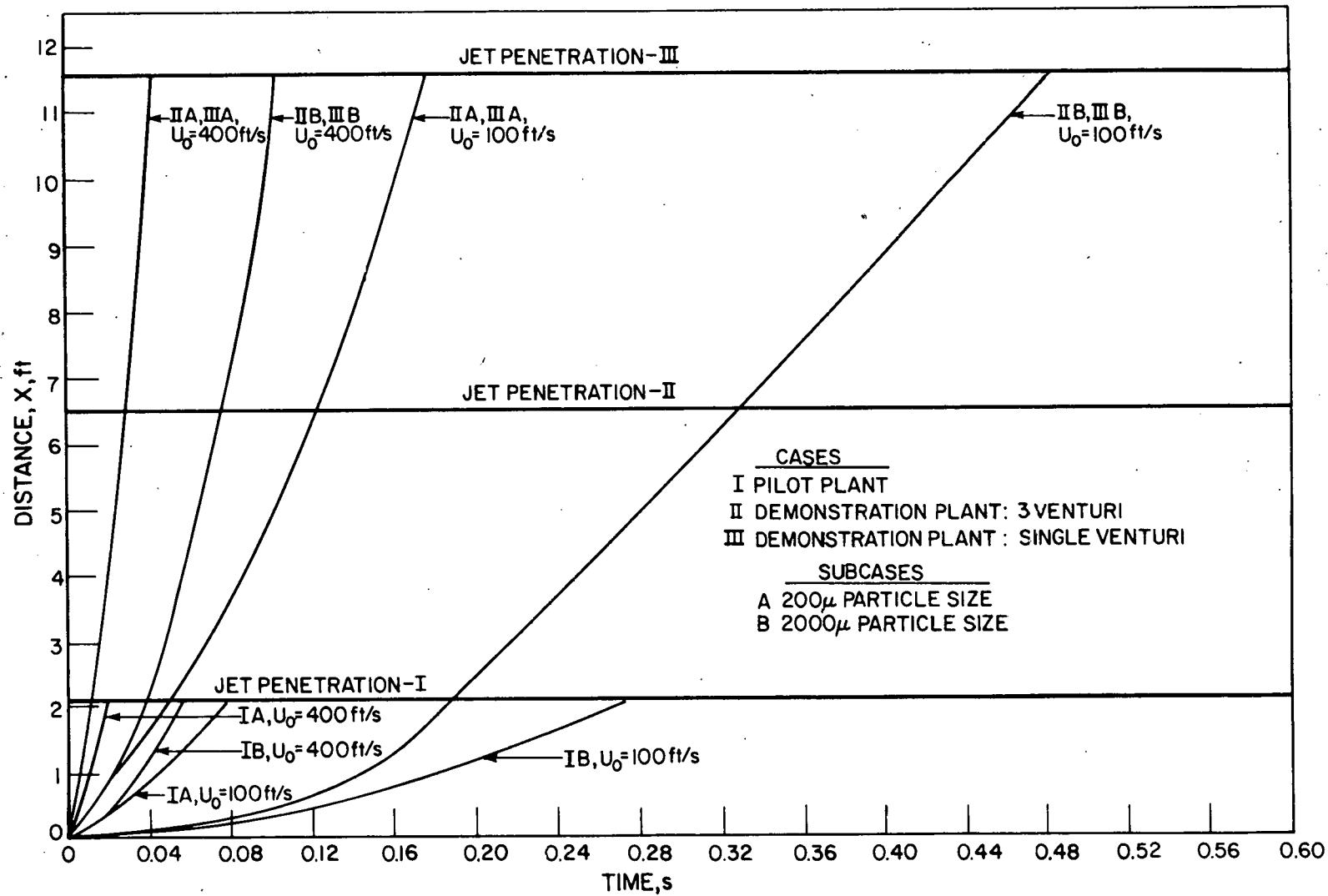
Figure D-6. GAS ENTRAINMENT RATE INTO JET/GAS RATE INTO JET FROM NOZZLE VS. NOZZLE DIAMETER

Ultimately, the maximum entrainment is at the extreme right-hand corner of the graph. Values of the entrainment there can be as high as those for the specific conditions of the pilot plant (11.0). In particular, a value of $U_o = 120$ ft/s (at the current Case II $\dot{M}$) or $U_o = 150$ ft/s (at 40% of the current Case II $\dot{M}$) would suffice equally well. Similar statements with respect to Case III can be made in which the results are approximately the same.

Particle Residence Time in the Jets

Considering only the entrainment into the jet is not satisfactory because the fate of the particles once entrained into the jet is also of concern. It is first necessary to calculate the time spent by a particle within the jet. To calculate this precisely we would need to know the initial trajectory of the particle into the jet, the velocity profile at all points in the jet, and the drag coefficient on an accelerating particle. In addition, if the conversion of the particle was appreciable in a single pass, its mass would change, and this would change its trajectory. This would depend on the conversions and trajectories of all the other particles in the jet as well. Such an analysis would be a monumental undertaking, and the likely improvement in the accuracy of the results is open to question. As a result, the following simplifications were made. Individual particles were assumed to be placed at a zone of constant jet velocity throughout their stay in the jet. Because the drag force was assumed constant at a value of 0.5, no acceleration effect on the drag force was assumed. In general, however, a higher value than would be expected for non-accelerating particles was taken, and this is in the right direction to account for the effect. No reduction in the mass of particles was permitted in the determination of particle trajectories. With these simplifications, we were able to determine the particle trajectories for various cases.

These results are presented in Figure D-7 as curves of distance travelled versus particle residence time in the jet. The particle size of the char in the bed material was considered to be of greatest importance; for this we used the information presented in the Appendix to the May 1979 Monthly Report, page 5. Part of that information is reproduced here as Figure D-8. Agglomerates were removed from Run 133 bed material, and a screen analysis was performed. The particle diameters corresponding to 10% and 90% of the cumulative size distribution of the char were selected as parameters for the remaining aspects of the study. These were 200μ and 2000μ size particles respectively. The particle trajectories are referred to as Subcases A (200μ) and B (2000μ) and are shown for Cases I, II, and III in Figure D-7. (Agglomerates are assumed to be entrained into the jet according to their proportion in the bed, but, because no carbon remains on them,



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Figure D-7. PARTICLE RESIDENCE TIMES IN THE JET

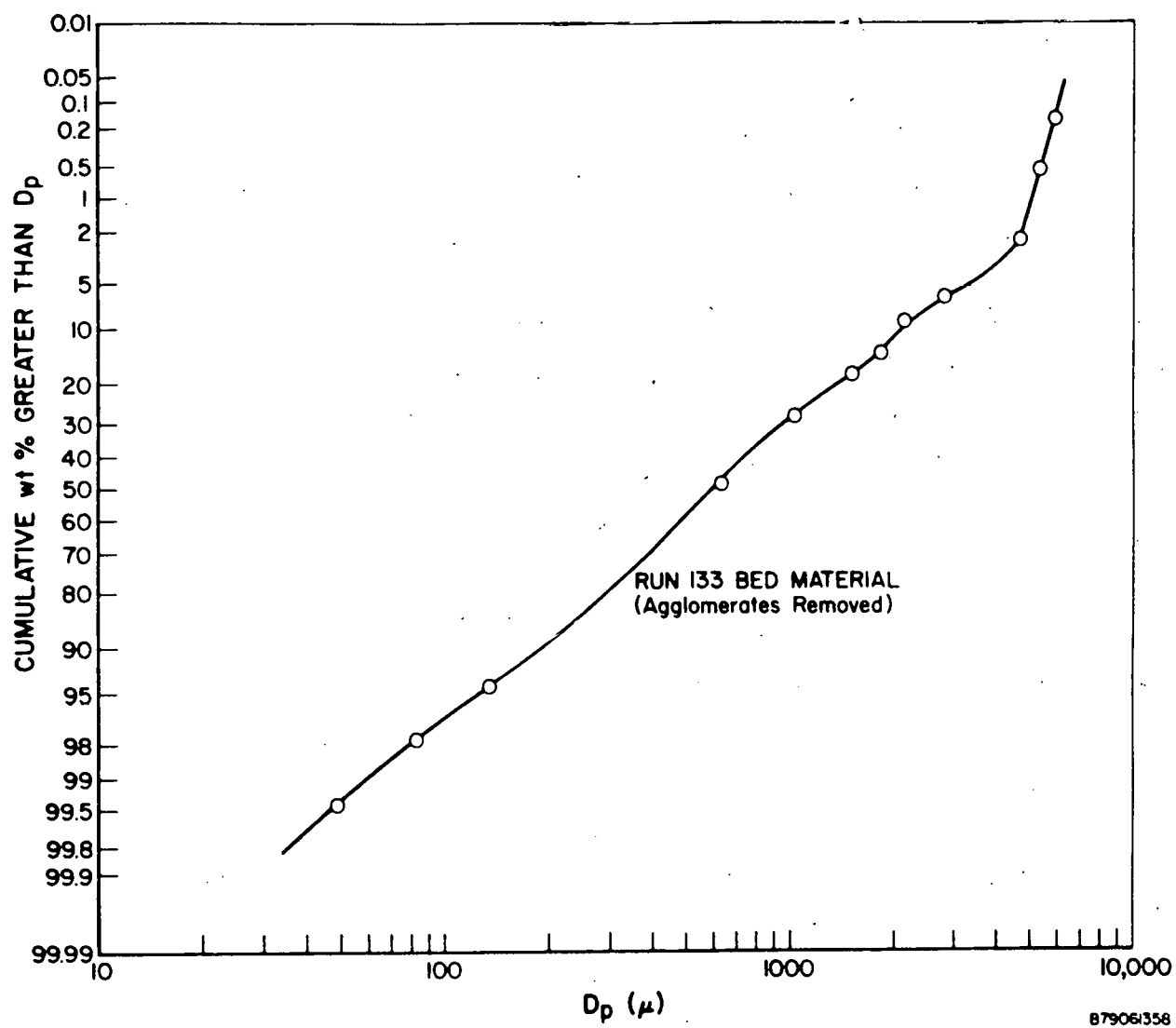


Figure D-8. CUMULATIVE SIZE DISTRIBUTION OF CHAR IN BED

they do not contribute to the conversion of oxygen or heat up in the jet.) Also shown in Figure D-7 are the jet penetrations (Merry formula) for Cases I through II for jet velocities of 100 and 400 ft/s. The appropriate intersections of the trajectory lines with the jet penetrations can be obtained from Figure D-7 and are presented in Table D-29. This intersection represents the total time spent by the particle in the jet under our assumptions.

Temperature Rise of the Particles

Following these calculations, the temperature rise of the individual particles was computed. Rehmat and Saxena have developed a model for multiple nonisothermal noncatalytic gas-solid reactions that was adapted for this purpose.^{4,5,6,7} In this model, the particle is converted by an unreacted shrinking core model. An ash layer develops that contributes to an intraparticle diffusion resistance of mass transfer of the reactants and of heat transfer. The simultaneous combustion and steam-carbon reactions are considered. The kinetic information on these reactions is taken from the open literature. The volume of ash remaining from a specific volume of char gasified is a parameter. The full unsteady-state solution to a problem of this nature is extremely difficult to obtain, but reasonable results can be obtained by making an assumption known as the pseudo steady-state assumption. The validity of this assumption for similar conditions was investigated by Wen⁸; he found that the assumption was valid except for situations involving extremely low solid reactant concentrations or very high pressures.

A computer program was developed to solve a partial differential equation with certain boundary conditions by a finite difference method. The input values for such parameters as the solid heat capacity, the gas viscosity, and the thermal conductivities are summarized in Table D-30. The results of the computer calculations consist of the extent of conversion of the particle versus time and the temperature of the core (assumed isothermal) versus time.

As shown in Figure D-9, higher pressure results in a higher core temperature for the same time of exposure to steam-oxygen mixtures of the same concentration. For particles with larger initial sizes, this effect of pressure is greater. As an extreme example, if a 2000 μ particle was exposed to the inlet concentration of O₂ for its entire residence time in the jet, we should have an approximately 1900°F particle temperature for the pilot plant case, a 2050°F particle temperature in the three-venturi demonstration plant case, and a 2100°F temperature in the single venturi demonstration-plant case. The greatest difference is the pressure effect between Case I and Cases II and III. The net effect of different residence times in the jet, caused by pressure and jet

Table D-29. RESIDENCE TIME OF PARTICLES IN THE JETS

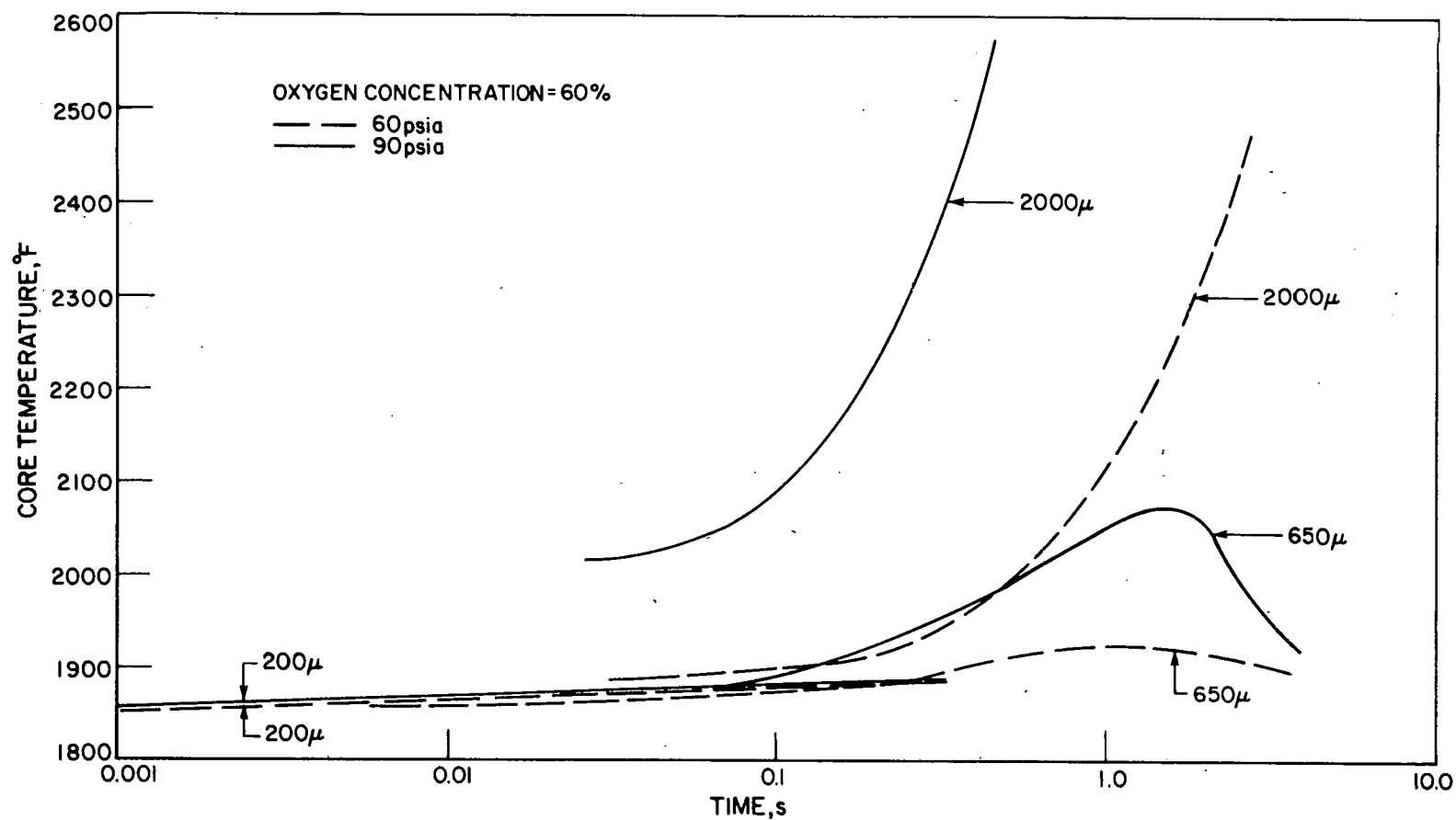
<u>Configuration</u>	<u>Particle Diameter, μ</u>	<u>Jet Velocity, ft/s</u>	<u>Residence Time, s</u>
Pilot Plant	200	100	0.08
Pilot Plant	2000	100	0.27
Pilot Plant	200	400	0.02
Pilot Plant	2000	400	0.05
Demo-3 jets	200	100	0.12
Demo-3 jets	2000	100	0.33
Demo-3 jets	200	400	0.03
Demo-3 jets	2000	400	0.07
Demo-1 jet	200	100	0.16
Demo-1 jet	2000	100	0.48
Demo-1 jet	200	400	0.03
Demo-1 jet	2000	400	0.10

Table D-30. INPUT DATA FOR COMPUTER CALCULATIONS

	Symbol	Total System Pressure, psia	
		60	90
Average Bed Temperature, °F	T_o	1850	1850
Mole Fraction of Oxygen	X_{O_2}	0.6	0.6
Particle Radius, cm	R	0.15	0.15
Specific Heat of Bulk Gas at Constant Pressure, cal/mol °K	C_p	0.39621	0.39621
Specific Heat of Coke, cal/mol °K	C_{pc}	0.368	0.368
Specific Heat of Ash, cal/cm ³ °K	C_{pe}	0.540	0.640
Molecular Diffusivity of Species i in the Bulk Gas Phase, cm ² /s	D_i	0.762	0.508
Effective Diffusivity of Species i in the Ash Layer, cm ² /s	D_{ei}	0.0762	0.0508
Activation Energy for Reaction $C + 1/2O_2 \rightarrow CO$, cal/mol	E_1	35,468	35,468
Activation Energy for Reaction $C + H_2O \rightarrow CO + H_2$, cal/mol	E_2	42,124	42,124
Heat of Reaction $C + 1/2O_2 \rightarrow CO$, cal/mol	ΔH_1^*	-25,996	-25,996
Heat of Reaction $C + H_2O \rightarrow CO + H_2$, cal/mol	ΔH_2^*	33,880	33,880
Thermal Conductivity of the Bulk Gas, cal/cm-s°K	k	2.30926×10^{-4}	2.30926×10^{-4}
Effective Thermal Conductivity of the Ash Layer, cal/cm-s°K	k_e	0.0032	0.0032
Rate of Reaction $C + 1/2O_2 \rightarrow CO$, g/cm ² -s-atm	k_A	2.108×10^{-2}	2.108×10^{-2}
Rate of Reaction $C + H_2O \rightarrow CO + H_2$, g/cm ² -s-atm	k_B	1.5397×10^{-6}	1.5397×10^{-6}
Partial Pressure of Oxygen, atm	P_A	2.45	3.67
Partial Pressure of Steam, atm	P_B	1.63	2.45
Density of the Bulk Gas, g/cm ³	ρ	1.024×10^{-3}	1.536×10^{-3}
Density of the Solid Reactant, g/cm ³	ρ_s	1.15	1.15
Radiative Heat Transfer Coefficient, cal/cm ² -s°K ⁴	C^{**}	1.35596×10^{-12}	1.35596×10^{-12}
Viscosity of Bulk Gas, g/cm-s	μ	0.053059	0.053059

* Positive for endothermic and negative for exothermic reactions.

** Black-body radiation assumed.



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Figure D-9. CORE TEMPERATURE VS. TIME IN JET AT CONSTANT OXYGEN-STEAM CONCENTRATION (60% O_2 -40% Steam) FOR VARIOUS PARTICLE SIZES

penetration length (also indirectly partially caused by pressure), is relatively minor in this instance. The results shown in Figure D-8 pertained to the case of the 400 ft/s jet velocity with the particle at such a position in the jet that the velocity is also 400 ft/s. (This is not the centerline because there is a velocity profile in the jet.) In similar analyses made for the other cases, larger particles and higher pressures result in higher particle temperatures, but the difference in the particle temperatures in the two demonstration plant alternatives are not too severe. (See Table D-31.)

Because the solids-to-gas mass ratio is higher in Case III than in Case II, one would expect the average oxygen concentration in the jet to be less because the temperatures at any given O_2 concentration in Case III are, on the average, at least as great as those in Case II. This indicates that the true situation is smaller temperature differences between Cases II and III. By the same token, the higher solids-to-gas ratio for the pilot plant Case I indicates even lower temperatures than those shown in Table D-31 on a comparative basis to Cases II and III. In addition to the effect that the higher solid-to-gas ratios has on average oxygen concentration in the jet, there would also be a greater sensible heat effect that would lower the initial temperature in the jet below that of the other cases. This would lower even further the average particle temperatures.

Table D-31 addresses the situation only for the case of initially unconverted particles at a fixed particle velocity and oxygen-steam composition in the jet of 60% O_2 /40% H_2O . The true situation would have to consider particles of a distribution of sizes, conversions, initial velocities, etc., entering the jet; the oxygen concentration varying according to the conversion of particles in the jet; and the infusion of emulsion gases and particle reaction product gases. Such an analysis would be prohibitively time consuming, and there is considerable doubt if meaningful values of this necessary input information could be obtained. We have, however, considered the case of a particle being exposed to a concentration profile in the jet. The concentration profile was arbitrarily assumed to be linear with the O_2 concentration equal to 0 at the terminus of the jet for the single nozzle (Case II). Because the solid-to-gas ratios are close in Cases II and III, the oxygen profile should be very similar with the exception that there should be some oxygen left at the terminus of the jet for Case III. The amount of oxygen assumed to be remaining was taken to be proportional to the height of the jets; this equals 32% in Case II. With this information, we were able to compute the core temperature versus time trajectory of the particles. This was done for Cases IIA, IIB, IIIA, and IIIB. The remaining assumptions of the simpler calculations (constant mass of particle local jet velocity, drag coefficient, bulk gas temperature, initial conversion, and initial velocity upward equal to 0) still hold for this calculation. These trajectories are shown in Figures D-10 through D-13.

Table D-31. MAXIMUM TEMPERATURE OF PARTICLES IN JET AT 60%
OXYGEN CONCENTRATION

<u>Jet Velocity, ft/s</u>	<u>Case A, 200 μ Particle</u>		
	<u>Case I</u>	<u>Case II</u>	<u>Case III</u>
	<u>Temperature, °F</u>		
100	1870	1880	1883
400	1863	1870	1876

<u>Jet Velocity, ft/s</u>	<u>Case B, 2000 μ Particle</u>		
	<u>Case I</u>	<u>Case II</u>	<u>Case III</u>
	<u>Temperature, °F</u>		
100	1935	2400	2550
400	1900	2050	2100

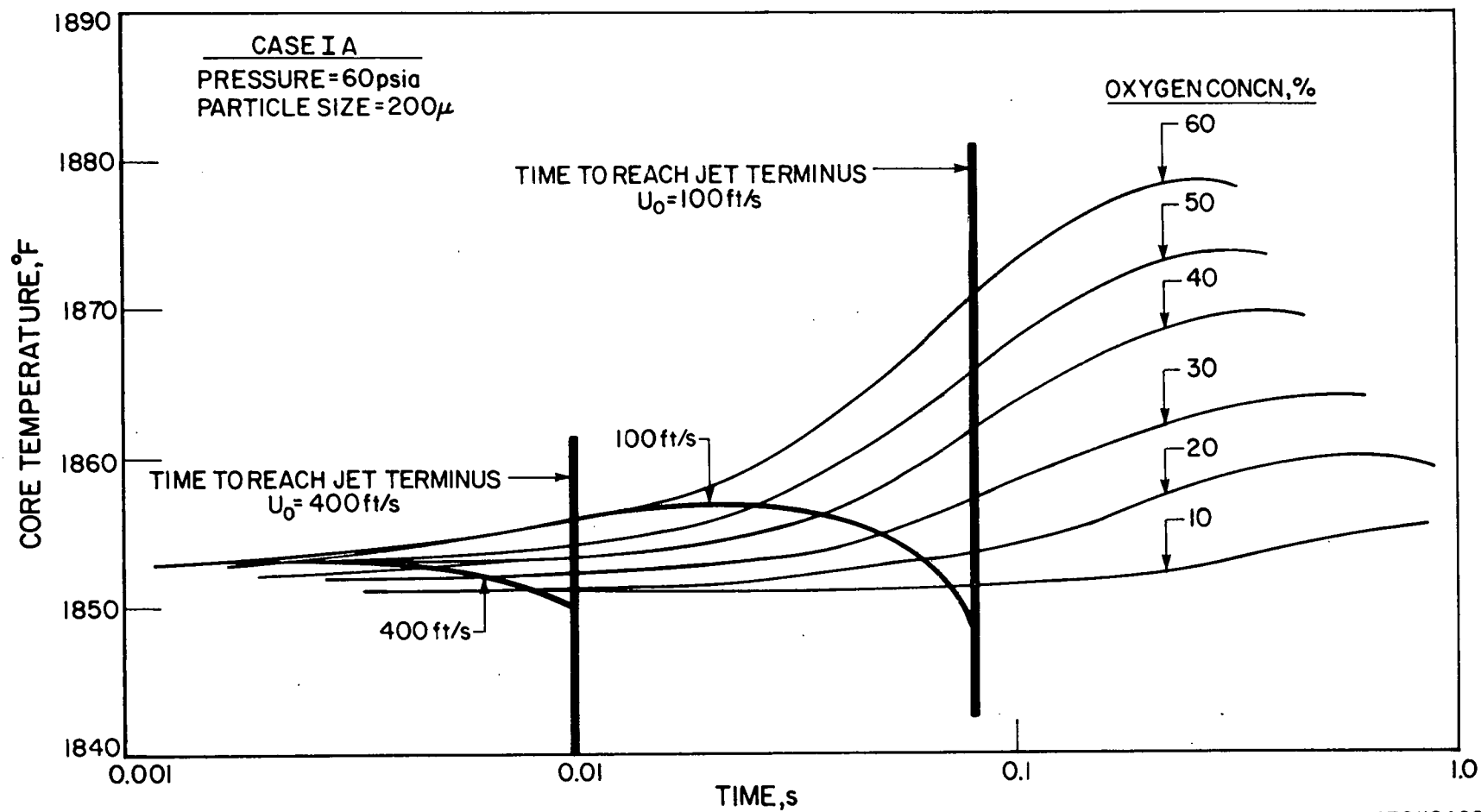


Figure D-10. CORE TEMPERATURE VS. TIME IN JET FOR LINEARLY VARYING JET OXYGEN CONCENTRATION PROFILES
 (100 and 400 ft/s Local Jet Velocity)

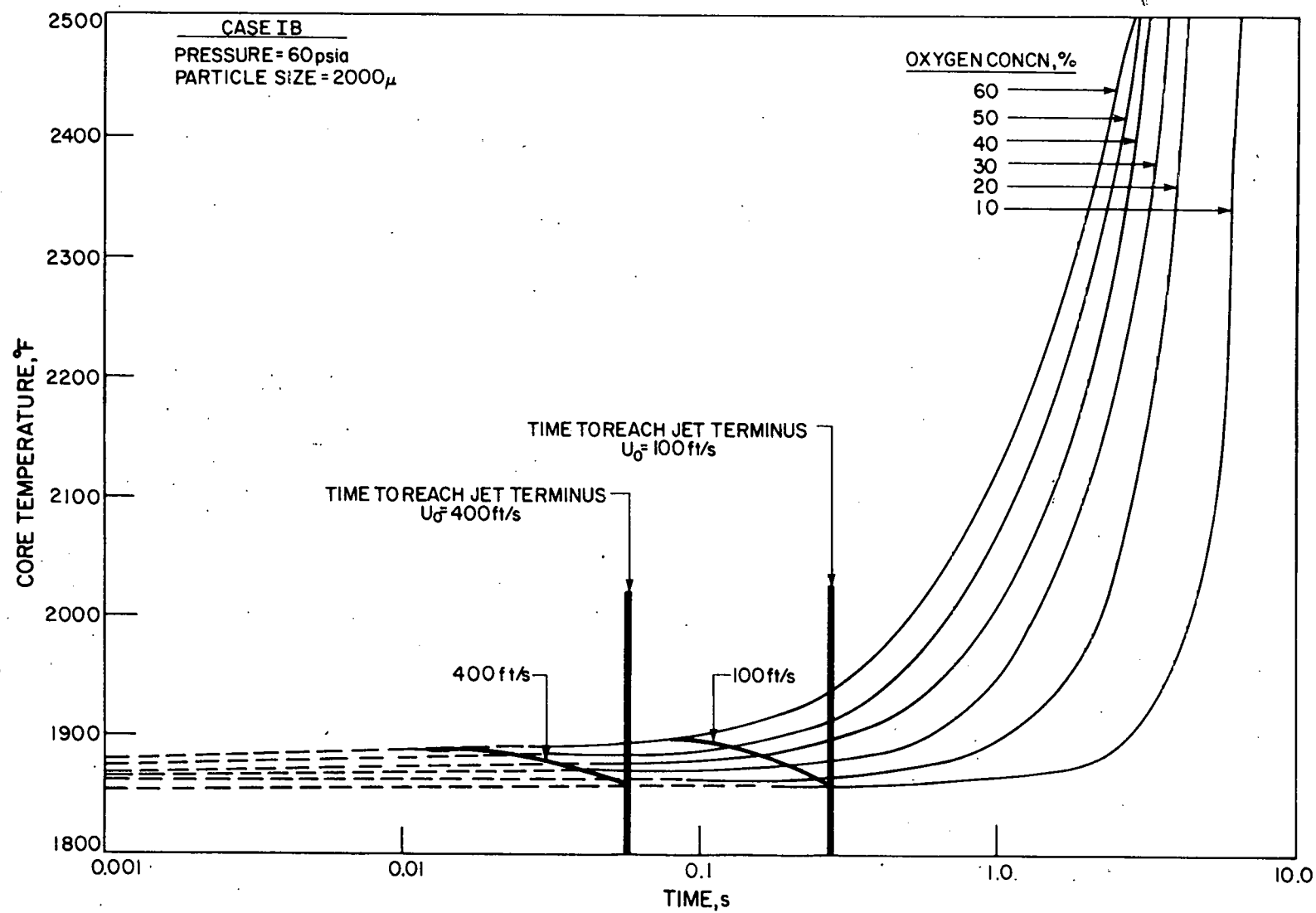
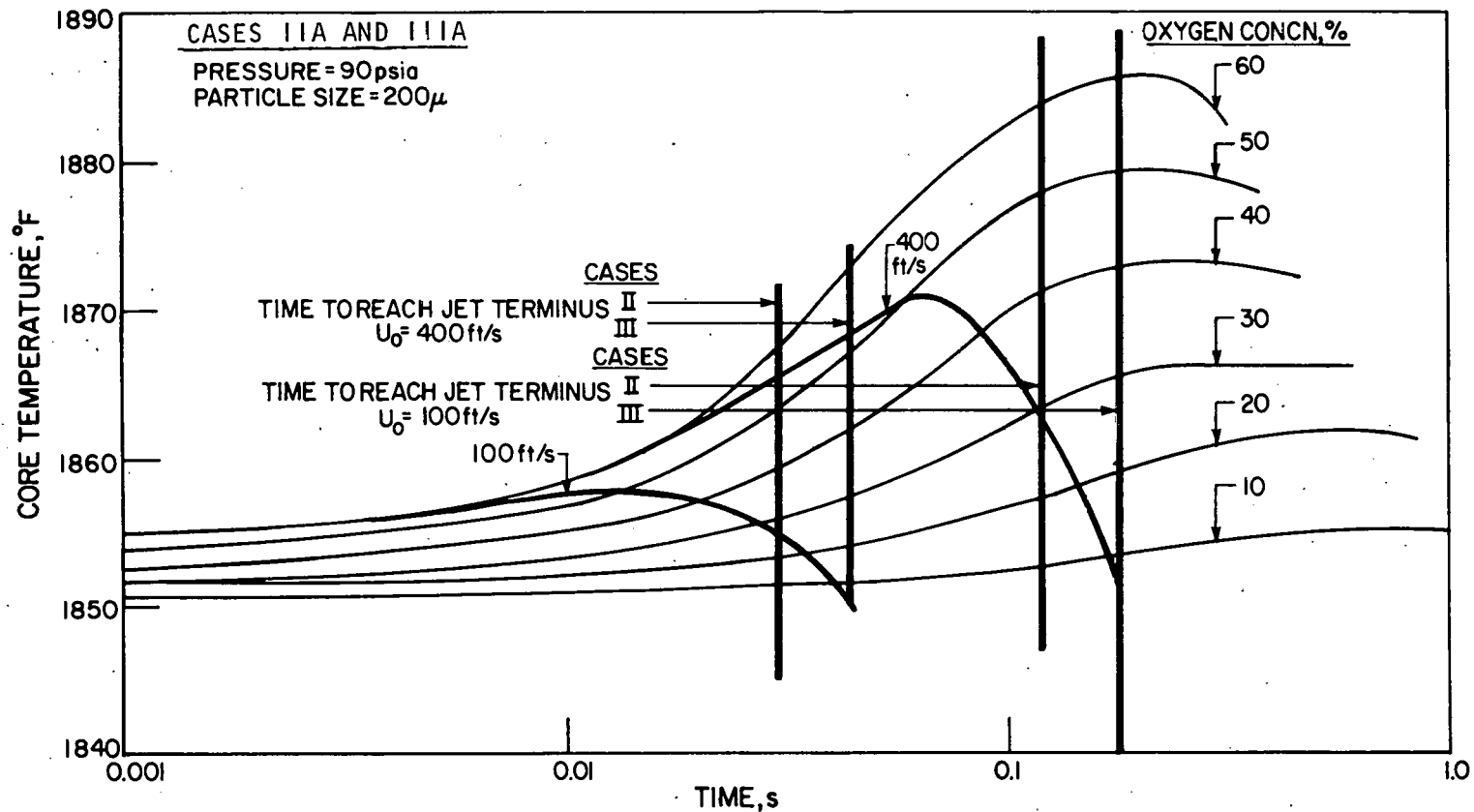
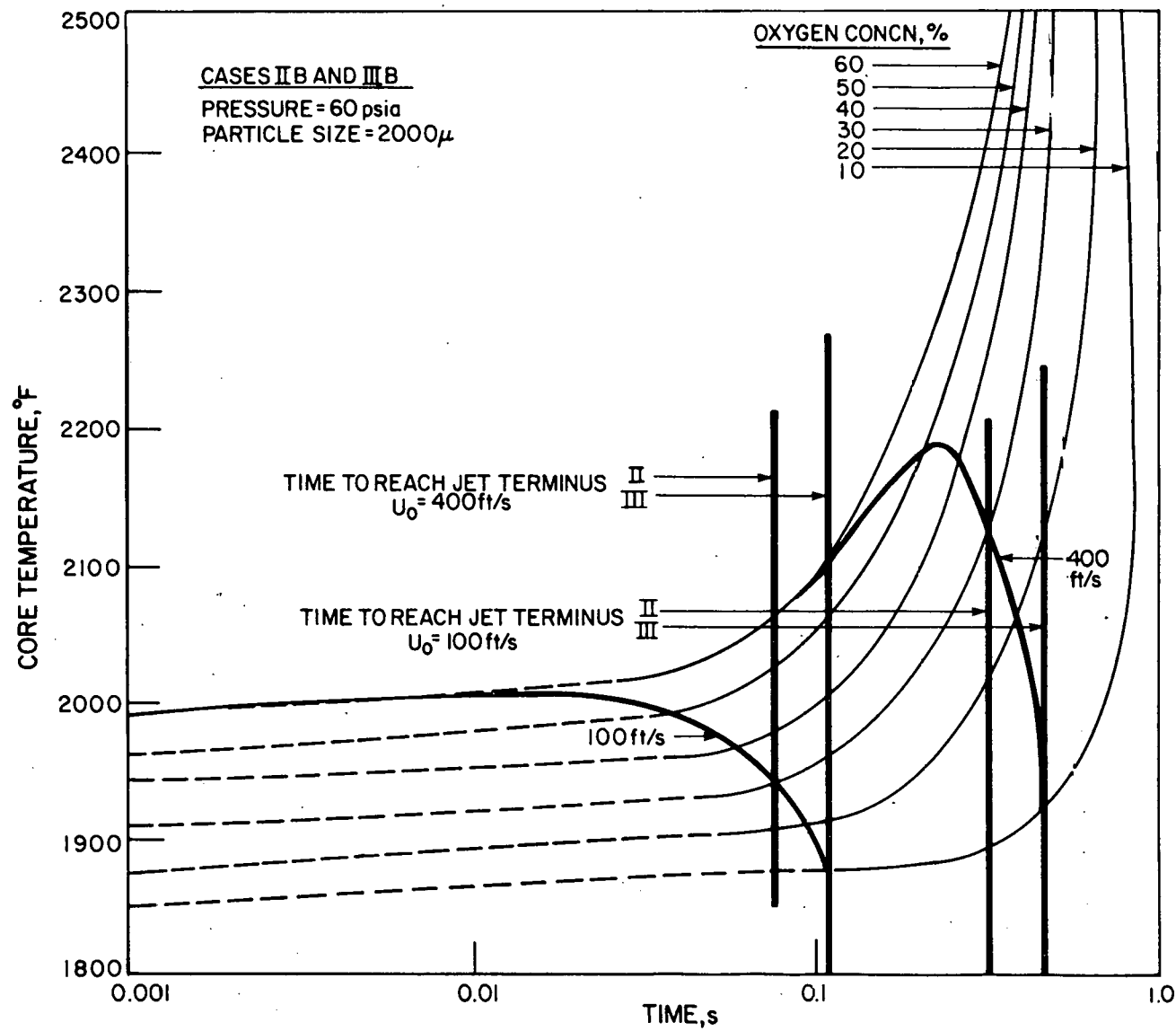


Figure D-11. CORE TEMPERATURE VS. TIME IN JET FOR LINEARLY VARYING JET OXYGEN CONCENTRATION PROFILES (100 and 400 ft/s Local Jet Velocity)



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Figure D-12. CORE TEMPERATURE VS. TIME IN JET FOR LINEARLY VARYING JET OXYGEN CONCENTRATION PROFILES (100 and 400 ft/s Local Jet Velocity)



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Figure D-13. CORE TEMPERATURE VS. TIME IN JET FOR LINEARLY VARYING JET OXYGEN CONCENTRATION PROFILES (100 and 400 ft/s Local Jet Velocity)

Conclusions

We can draw several conclusions from these calculations. The major conclusion is that the maximum temperature is reached while the particle is still in the zone of high oxygen concentration. The results of the previous analysis where the oxygen concentration was not varied yielded temperatures that are higher than those reached in the present case. In all cases, the maximum temperature reached is higher at lower local jet velocity (larger particle jet residence time). Thus, lowering the jet velocity can be expected to result in higher particle temperatures on the average. We can also conclude that smaller particles result in lower particle temperatures under otherwise identical conditions.

Although the oxygen concentration profile was assumed in all cases, if the assumptions are valid on a relative basis we would conclude that the particles entering the jet in Cases II and III reach at least as high a temperature as those in Case I. In fact, the temperature rise is generally considerably higher in Cases II and III. We can expect, then, that the fraction of particles entering the jet that reach 'agglomeratable' conditions of temperature is greater in these cases than in Case I. Thus, a lower solids-to-gas ratio in these cases may still result in adequate numbers of particles of this type being created.

Figures of Merit

In addition to the mechanistic model and analysis discussed in the previous sections, we have also calculated several figures of merit of interest in the scale-up of this process. These results are presented in Table D-32. (The figures in Table D-32 pertain to one reactor of three.) We determined the heat release assuming all oxygen fed to the jet is combusted to carbon dioxide. The solids circulation was calculated by determining the bubble volumetric rate or gas rate in excess of minimum fluidization and assuming that 20% of the bubble volume is wake (emulsion solids). The total gas above minimum fluidizing velocity is the same regardless of the number of venturis employed. The higher pressure of the demonstration plant results in less relative top-to-bottom circulation than in the pilot plant because the bubble volume is reduced at higher pressure. The ratio of demonstration plant circulation to pilot plant circulation is approximately 19:1. This ratio alone tells us little, but if we compute the ratios of heat release in the jet to solids circulation, we obtain an idea of the adequacy of the fluidized bed to dissipate the localized reaction heat. The demonstration plant figure on this basis is about three times larger than the pilot plant figure so there is a greater tendency to

Table D-32. FIGURES OF MERIT IN THE SCALE-UP OF THE
PILOT PLANT TO THE DEMONSTRATION PLANT

	Pilot Plant (Case I)	Demonstration Plant, 3 Venturis (Case II)	Demonstration Plant, 1 Venturi (Case III)
Heat Released Per Reactor, $\dot{Q}$, Btu/s	1880	73,200	73,200
Solids Circulation Rate Per Reactor, J, lb/s	85	1600	1600
$\dot{Q}/J$, Btu, lb	22	46	46
Jet Surface Area Per Jet, A_J , ft ²	3.8	23.0	63.8
Heat Release Per Unit Total Jet Surface Area, $\dot{Q}/nA_J$, Btu/ft ² ·s	495	1060	1147
Jet Volume, V_J , ft ³	0.54	7.0	30.3
Heat Release Per Unit Total Jet Volume, $\dot{Q}/nV_J$, Btu/ft ³ ·s	3481	3485	2415

avoid hot spots and nonisothermal operation in the pilot plant. Whether or not any appreciable hot spots or top-to-bottom temperature gradients would actually develop cannot be determined without more detailed analysis.

The ratios of heat released to jet surface area and jet volume were also calculated. In this case, there is about twice as much heat dissipated per unit jet area in either Case II or Case III than in Case I. This is, of course, only significant if all the oxygen is consumed within the jet. If convective heat transfer of gas and solids across the jet surface to dissipate heat is important, higher jet surface temperatures would result in the demonstration plant cases. The figures relating to the ratio of heat release in the jet to jet volume indicate a correspondence between Case I and Case II; Case III's figure is about 70% of the former cases. With the same caveats as apply to the surface-area-based heat release ratio, these figures indicate that higher particle and jet gas temperatures could be reached in the jet.

Some of the more conventional scale-up ratios are shown in Tables D-33 through D-35. Table D-33 discusses scale-up from the standpoint of coal throughputs. Data relating to actual pilot plant operation and proposed demonstration plant operation are presented as well as the mass velocity, both unadjusted and adjusted for pressure and velocity. The adjusted ratios represent a quite conservative scale-up.

Gasifier performance ratios are presented in Table D-34. These are primarily concerned with the extrapolation of steam and oxygen consumption ratios plus product gas quality. Again, the projected demonstration plant values are well within the range obtained in the pilot plant.

Finally, the grid-venturi design information is presented for the pilot plant and demonstration plant (three-nozzle basis). The ash mass velocity obtained in the pilot plant cannot be directly compared with that of the demonstration plant because in the pilot plant, the theoretical maximum rate obtainable with the venturi is limited by the ash feed and the agglomeration rate. Comparisons can be made between certain results obtained in the cold model and those projected for each nozzle of the three-venturi demonstration plant design, in which case the design rate is less than that already achieved.

References

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Table D-33. COMPARISON OF COAL THROUGHPUT

	Pilot Plant		Demonstration Plant
	<u>30 psia</u>	<u>55 psia</u>	<u>90 psia</u>
Coal Feed Rate lb/hr	1200	2000	76,000
Reactor Diameter, ft	3	3	13
Area of Reactor, ft ²	7	7	133
Mass Velocity, lb/ft ² -hr	172	286	573
Adjustment Due to Pressure, lb/ft ² ·P	5.7	5.2	6.4
Superficial Velocity, ft/s	4.0	3.5	4.0
Adjustment Due to Velocity, lb/ft ² ·Pv	1.42	1.49	1.6

Table D-34. GASIFIER PERFORMANCE

	<u>Pilot Plant</u>	<u>Demonstration Plant</u>
Coal	W. Ky*	W. Ky*
Temperature, °F	1825-1890	1875
Pressure, psia	25-60	90
Coal Utilization, %	86-94	95
Agglomerate Ash Content, %	80-95	85
Gas HHV, Btu/SCF	250-280	284
1b H ₂ O/1b coal (MAF)	1.4-2.0	1.4
1b O ₂ /1b coal (MAF)	0.6-0.9	0.75
SCF Gas (Dry)/1b coal (MAF), SCF/1b	32-43	42
CO + H ₂ /1b coal (MAF), SCF/1b	23-28	30

*Western Kentucky

Table D-35. DEMONSTRATION PLANT GRID DESIGN

	<u>Pilot Plant</u>	<u>Cold Model</u>	<u>Demonstration Plant</u>
Gasifier Diameter, ft	3	4	13
No. of Venturi	1	1	3
Venturi Diameter, in.	4-1/2	6	8
Jet Diameter, in.	1-1/2	--	3-1/2
Coal Feed, lb/hr	2000	--	76,000
Ash Discharge per Venturi, lb/hr	200*	3800	3900
Mass Velocity in Venturi, $\frac{\text{lb}}{(\text{hr})(\text{ft}^2)}$	--	19,400	12,600

*Limited by gasification rate.

2. Keairns, D. L. and Salvador, L. A., "Advanced Coal Gasification System for Electric Power Generation," Second Quarterly Progress Report - Fiscal Year 1978, (FE-1514-84) prepared for U.S. Department of Energy, Division of Coal Conversion, Chicago: Institute of Gas Technology, June 1978, p. 26.
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FLOWABILITY AND FLUIDIZATION TESTS OF ASH-WATER SYSTEMS

The objective of these studies was to aid in the design of the lower section of the demonstration plant gasifier. For flowability studies, the discharge lockhopper arrangement of the half-section atmospheric cold-flow model (Figure D-14) was used. For fluidization studies, a 2-inch Plexiglas column with the necessary flowmeters (Figure D-15) was set up. The tests were conducted with final bed material from Run 157, coke agglomerates, coal agglomerates (Runs 133 and 150), and coke breeze feed material. The results of the tests, presented in Tables D-36 through D-39, suggest that the char or coke particles do not settle as freely in water as coal or coke agglomerates.

Equipment Setup

Flowability Tests

For the flowability tests, the bottom section of the half-section atmospheric cold flow model (Figure D-14) was used. This setup consists of two vessels joined together with a 4-inch full-port ball valve. A second 4-inch ball valve connected to the bottom of the lower vessel discharges the slurry into a 55-gallon drum.

Fluidization Test

Figure D-15 shows a schematic diagram of the equipment setup used in the experiment. It mainly consists of two, 2-inch X 2-inch and 2-inch X 36-inch Plexiglas tubes joined together with a 2-inch X 2-inch Plexiglas tee, the branch of which connects a 2-inch rubber hose to drainage. A 2-inch male slip-on and female threaded special PVC fitting with a 10-mesh size screen plate resting on its projected inner edge provides the necessary base for the support of the column. Water flows to the column are controlled by two flowmeters in parallel. Materials are added from the open top end of the column. A 1/2-inch drain connection is provided at the bottom of the column to drain out the system at the end of the experiment.

Operation and Results

Flowability Tests

Three different types of tests were conducted on the solids mentioned in Table D-36 to determine the flowability characteristics of ash-water streams.

In the first series of tests, Vessel II (Figure D-14) was completely filled with water and solids were added slowly from the top. After 5 minutes, the lower ball valve was opened, and material was collected in a drum. The collected material was weighed, and

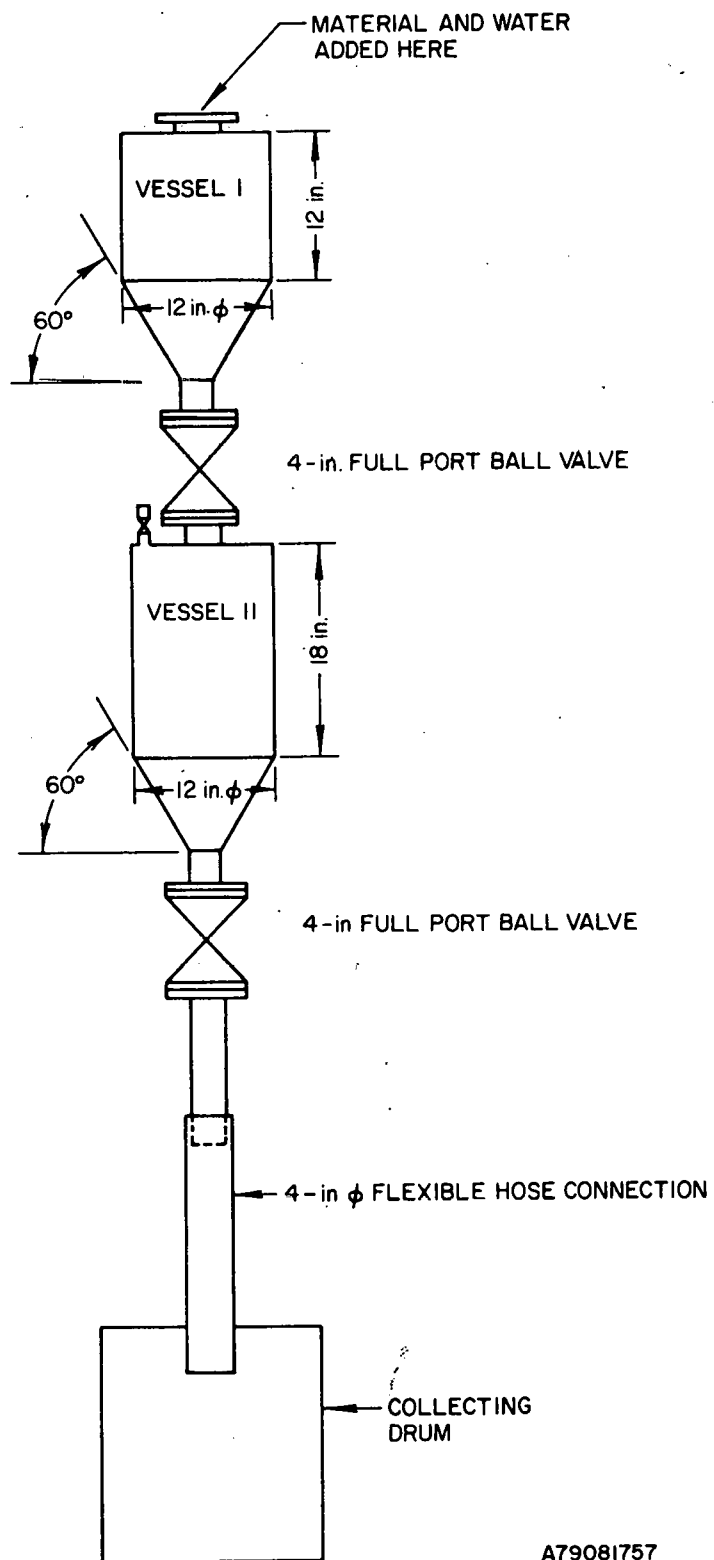


Figure D-14. EXPERIMENTAL SETUP FOR FLOWABILITY TEST FOR ASH-WATER SYSTEM

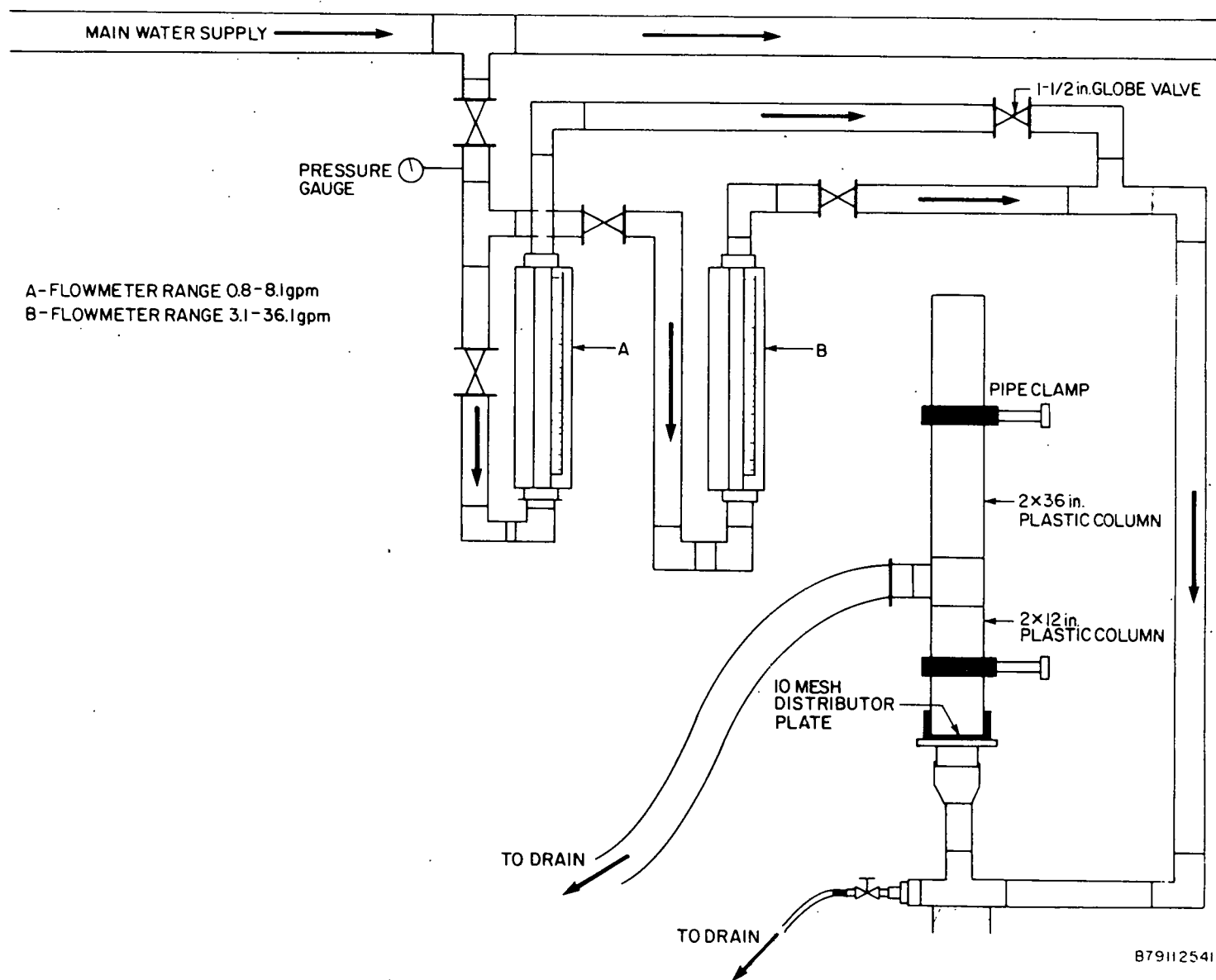


Figure D-15. FLOW SCHEMATIC OF ASH-WATER FLUIDIZATION TEST UNIT

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Table D-36. FLOWABILITY RESULTS OF SOLIDS-WATER SYSTEM WITH
UPPER VESSEL EMPTY AND LOWER VESSEL FILLED WITH WATER

<u>Solids Used</u>	<u>Material, lb</u>	<u>Water, lb</u>	<u>Observations</u>
Final Bed Material from AAG Run 151	35.8	88.7	First came clear water, then slurry, and finally water with suspended char.
Coke Agglomerates from AAG	32.9	89.2	First slurry for 4-5 seconds and then clear water, in- dicating that the solids will settle in water. However, about 2 lb of solids were found stuck up on the walls of lower and upper vessel.
Coal Agglomerates from Run 150	67.0	72.5	In the beginning, the mate- rial came out in form of thick slurry which grew thicker in the end; took 35 seconds for the whole thing to drop out.
Coal Agglomerates from Run 133	54.0	78.0	All material came out with water stream. The stream was thin in the beginning and became thick at the end. Took 25 seconds to drop out.
Coke Breeze	48.0	96.0	Had problems in pouring the solids in Vessel I. When bottom ball valve was opened, first came slurry, then big stream of water, then again slurry, and finally only water. A material balance at the end of test indicated that about 27.5 lb of material were still in system. 25 lb of material were taken out by rodding through the bottom. Rest will be taken out when system is dried.

Table D-37. FLOWABILITY RESULTS OF SOLID-WATER SYSTEM WITH UPPER VESSEL
PARTIALLY FILLED AND LOWER VESSEL COMPLETELY FILLED WITH WATER

<u>Solids Used</u>	<u>Weight of Water, lb</u>	<u>Initial Weight of Solids, lb</u>	<u>Final Weight of Solids</u>		<u>Remarks</u>
			<u>In Upper Vessel (Collected in Drum #2)</u>	<u>In Lower Vessel (Collected in Drum #1)</u>	
Coal Agglomerates From AAG Run 150	119.4	27.10	0.50	25.10	In Drum #1, first came slurry, then clear water. In Drum #2, it was all water. About 2 lb of material found stuck up on the walls of both vessels.
Coal Agglomerates From AAG Run 133	117.90	32.40	0.00	32.15	In Drum #1, first came slurry, then clear water. In Drum #2, it was all water. Practically no loss of material.
Coke Agglomerates From AAG	119.5	22.50	0.00	22.30	Same as above.
Coke Breeze	113.9	24.4	5.05	18.05	In Drum #1, first came slurry, then clear water with suspended coke particles. In Drum #2, it was all slurry. Some material was found stuck up on wall.
Final bed Material From AAG Run 151	123.00	26.30	5.20	17.3	In Drum #1, first came slurry, then a big stream of water, then a slurry again, and finally clear water. In Drum #2, it was all slurry. About 3 lb of material found stuck up on the walls of vessel.

Table D-38. FLOWABILITY RESULTS OF SOLIDS-WATER SYSTEM IN THE
CASE OF PROLONGED HOURS FOR SETTLING

<u>Solids Used</u>	<u>Weight of Material, lb</u>	<u>Weight of Water, lb</u>	<u>Time Allowed, hr</u>	<u>Observations</u>
Coal Agglomerates From AAG Run 150	35.1	90.2	66	A very thick slurry followed by a clear stream of water discharged freely from the system.
Final Bed Material From AAG Run 150	30.0	87.2	26	At first came thick slurry, then thin slurry, and finally a clear stream of water.

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Table D-39. WATER FLUIDIZATION CHARACTERISTICS OF CHAR, COKE BREEZE, COAL, AND COAL AGGLOMERATES

Solids Used	Weight of Solids, oz	Minimum Fluidization Velocity, ft/s	Elutriation* Velocity, ft/s	Observations
Agglomerates From AAG Run 133	10.0	0.15	0.61	Settled bed height of material was 5-3/4 in. in the tube which expanded to 6-1/4 in. at the U_{mf} . At U_{op} , about 90% of solids were carried away leaving behind a few big particles. Particles settled freely and were well-fluidized at U_{mf} .
Agglomerates From AAG Run 150	10.0	0.15	0.61	Settled bed height = 5-1/2 in. Expanded bed height at U_{mf} = 6-1/2 in. Solids settled freely and were well fluidized at U_{mf} .
Agglomerates From AAG Run 151	10.0	0.15	0.61	Settled bed height = 5 in. Expanded bed height at U_{mf} = 7 in. Solids settled freely and were well Fluidized at U_{mf} .
Coke Agglomerates From AAG Run 39	10.0	0.26	0.87	Settled height = 5.0 in. Expanded bed height at U_{mf} = 8.0 in. Solids settled freely and were well fluidized at U_{mf} .
Final Bed Material From AAG Run 151	10.0	0.05	0.51	Material did not drop freely into the bed of water. Whole solids settled to a 8-1/2 in. column.
Coke Breeze	10.0	0.05	0.12	Same as above except column height of settling mass was 7-1/2 in. in this case.

* This is the velocity at which 90% of material is carried away with the water stream within ~15 seconds.

the resulting weight was compared with that of the material fed. In the case of coke breeze and final bed, the slurry had to be filtered out and the moist material analyzed to determine the weight of char on coke breeze in the slurry. The results are summarized in Table D-36.

In the second series of tests, the lower vessel (Vessel II) was completely filled, and the upper vessel (Vessel I) was one-third filled with water. Solids were added to the upper vessel, keeping the connecting valve between the two vessels open and the lower valve closed. After 15 minutes, the connecting valve was closed, and the lower valve was opened. The vent valve on the top of the lower vessel was also opened to facilitate the discharge. The contents of the lower vessel were collected in a drum, weighed, and analyzed. Similarly, the contents of the upper vessel were collected in a separate drum, weighed, and analyzed. By comparison with predetermined weights of water and solids, the fractions of solids distributed in the two vessels over a period of 15 minutes were determined. In the case of coke breeze and final bed, the slurry had to be filtered to determine the weight of solids in each drum. The results are summarized in Table D-37.

In the third series of tests, the lower vessel was filled with water, and coke agglomerates from Run 150 were added to it. The solid was allowed to settle for a total period of approximately 66 hours. At the end of the test, when the bottom valve was opened, a very thick slurry followed by a clear stream of water discharged freely from the system. In a similar test, final bed from Run 150 was added to the upper vessel, and the material was allowed to settle for 26 hours. When the lower valve was opened, a thick slurry followed by a thin slurry and a clear water stream discharged from the system. The results are summarized in Table D-38.

Fluidization Tests

A small quantity of each one of the five materials tested (listed in Table D-39) was placed in the Plexiglas tube. Water flows were initiated and gradually increased until complete fluidization of the added material occurred. At this point the water flow rate was noted. The water velocity was increased further until a point was reached where all the material was swept out of the column, in at most 15 seconds. Although not a precise measure, this velocity indicates a flow sufficient to elutriate all materials from the apparatus in a short time with relative certainty. The two velocities so determined are reported in Table D-39 along with any significant observations pertaining to the tests.

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Conclusions

The characteristics of agglomerates appear to be such that no difficulty is to be expected in discharging them from a water-filled tank as proposed in the demonstration plant. Coke breeze feed and final bed material discharge can be expected to result in difficulties associated with hang up of material on the vessel walls. The fluidization velocities presented would allow us to design a discharge vessel and associated pumps and spargers that can obviate these potential problems.