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MULTICELL FLUIDIZED-BED BOILER DESIGN, CONSTRUCTION AND TEST PROGRAM

INTERIM REPORT FOR THE
PERIOD JULY 1977 - JUNE 1978

PREPARED BY
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16. Abstracts Design, construction and test program of a 300,000 lb/hr steam generating capacity multicell fluidized-bed boiler (MFB), as a pollution free method of burning high-sulfur or highly corrosive coals, is being carried out. The concept involves burning fuels such as coal, in a fluidized-bed of limestone particles that react with the sulfur compounds formed during combustion to reduce air pollution. Nitrogen oxide emissions are also reduced at the lower combustion temperatures. The $CaSO_4$ produced in the furnace is discharged with the ash or regenerated to CaO for reuse in the fluidized-bed.

This report presents information on continued operation of the Rivesville MFB steam generating plant in a commercial mode and for determining performance and emission characteristics; studies and tests on fly ash characterization and reinjection, fuel feed eductors and needles, air distributor, corrosion-erosion and sulfur capture; engineering studies to improve MFB performance and reliability.

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Agglomerates
Combustion Efficiency
Eductors
Fly ash
Nitrogen Oxides
Sulfur Capture
Sulfur Dioxide
Superficial Velocity

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Fluidized-Bed Module (FBM)
Multicell Fluidized-Bed Boiler (MFB)
Carbon Burn-up Cell (CBC)
Process Development Unit (PDU)

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ABSTRACT

The objective of this program is to design, construct and test a multicell fluidized-bed boiler, as a pollution-free method of burning high-sulfur or highly corrosive coals. The fluidized-bed boiler produces approximately 300,000 pounds of steam per hour. Steam pressure and temperature conditions were selected to meet requirements of the site at which the boiler was installed.

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INTRODUCTION

General

The Government, in order to implement research and development work on a multicell fluidized-bed boiler operating under utility electric power generation conditions, awarded DOE/ERDA/OCR Contract No. EX-76-C-01-1237 to Pope, Evans and Robbins Incorporated (PER) on October 5, 1972. The work under this contract is a follow-on to work previously performed by PER under OCR Contract No. 14-01-0001-478 as amended, and OCR Contract No. 14-02-0001-1229, which indicated that continued development would have a high probability of success.

Tasks and Phases

The objective of the program covered by DOE/ERDA/OCR Contract No. EX-76-C-01-1237 (E(49-18)-1237) is to test a multicell fluidized-bed boiler as a pollution-free method of burning high-sulfur coal or highly corrosive coals.

The objective is to be accomplished by designing, constructing and operating a multicell fluidized-bed boiler under utility electric power generation conditions in four technically distinct but chronologically overlapping phases:

Phase I	Boiler and Plant Design: Performance of Experiment to Optimize Certain Boiler Features.
Phase II	Fabrication, Installation and Component Testing
Phase III	Demonstration Operation
Phase IV	Preparation of Atmospheric Fluidized Bed (AFB) Boiler and conceptual design of packaged industrial size AFB boiler.

This Interim Report covers work performed under Phase III and IV of the Contract during the July 1977 through June 1978 time period.

1.1 Work Covered by This Report

This report includes description of work performed under Phases III and IV of the contract and includes:

a) Operation of the boiler and auxiliary systems in hot mode demonstration status to determine performance and emission characteristics, establish operational reliability, and commercial operation feasibility,

b) Preparation of Design Manual,

c) Preparation of conceptual designs for packaged industrial size unit,

d) Conduct approved laboratory test programs to support operation of the Rivesville MFB unit and to continue investigation of the atmospheric fluidized bed combustion process.

e) Administration and supervision of subcontracted work.

1.2 Review of Significant Events

The following significant events occurred during calendar years 1977 and 1978.

Event	Date	
Performed tests for bed material transfer between Cells D (CBC) and C of the Rivesville, W. Va. MFB	January	1977
Accomplished first light-off of Cell C and operation of two Cells (D & C) of the Rivesville, W. Va. MFB unit	February	1977
Initiated coal feeder test and evaluation program at Alexandria Laboratory	March	1977
Performed tests at Alexandria Laboratory to develop alternate light-off procedures for the Rivesville, W. Va. MFB	March	1977
Completed all Rivesville MFB plant auxiliary systems subcontract work	March	1977
Accomplished first light-off of Cell B and concurrent operation of three cells, Cells D, C and B of the Rivesville MFB unit	April	1977
Initiated modifications to MFB unit fuel feed systems to eliminate continuing operational problems	May	1977
Set safety valves on MFB unit steam drum and superheated steam outlet header	July	1977
Accomplished first light-off of Cell A and concurrent operation of all four cells of the Rivesville MFB unit	August	1977
Dedication ceremony held for Rivesville MFB Plant	August	1977
Operated MFB in commercial operating mode and exported steam to MPC for electric power generation	September	1977
Operated MFB in continuous commercial operating mode and exported steam to MPC for electric power generation for the longest duration to date; 9 hours	November	1977
Scheduled outage of MFB Plant for modifications and improvements	December	1977
Completion of Alexandria PDU construction and initial light-off	December	1977
Conducted FBM/PDU verification tests at the Alexandria Laboratory	April	1978
Initiated shakedown operation of the Rivesville MFB on completion of modifications and improvements	April	1978

Event	Date
Operated CBC continuously for 200 hours	May 1978
Started Comprehensive performance test plan	May 1978
Completed D Cell Performance tests	June 1978

1.3 Summary of Activities

1.3.1 General

Substantial progress was made during this reporting period in achieving overall program objectives. Most significant progress was made in on-line operation of the Rivesville MFB unit. The duration of operating was continuously increased resulting in increased reliability and higher levels of confidence in the unit performance to the extent that Monongahela Power Company permitted export of MFB generated steam to drive their turbine-generator unit for commercial production of electricity. Continuing problems were experienced with coal and limestone feed systems which were more pronounced when moist coal fines were present. Ongoing studies are being conducted to improve this condition.

Extensive test instrumentation was installed in the MFB unit during the outage resulting from a coal strike. Performance and emission testing was conducted when the unit was again operational. Gathered data was reduced and the results in general prove that large scale atmospheric fluidized bed combustion with a suitable sorbent material will produce emissions within environmentally acceptable levels.

Rivesville operations activities were supported by the Alexandria Laboratory, the New York Office engineering and design staff and various subcontractors.

1.3.2 Laboratory Work at Alexandria, Virginia

Work at the Alexandria fluidized-bed combustion laboratory, supported operation of the Rivesville multicell fluidized-bed boiler (MFB) and continued development of the atmospheric fluidized-bed combustion process. Following is a summary of the work performed at the laboratory, including experimental test results and process analysis results.

Test Results

A study of the effectiveness of simultaneously burning coal and fly ash at 1600°F in the Fluidized-Bed Module (FBM) was completed. Open cycle and closed cycle tests were conducted to measure the fly ash feed rate. Closed cycle tests were conducted, in which fly ash was recycled as it was produced. Open cycle tests were conducted in the previous reporting period using stored fly ash, to determine the once-through carbon burnup of the fly ash.

To provide new process information for the design and selection of dust collection equipment, chemical analysis of the fly ash as a function of particle size was determined at two different superficial velocities.

A test was conducted using the FBM to compare the effectiveness of an agglomerate of finely ground Greer limestone with the typical crushed Greer stone. Since the agglomerates did not remain in the bed, little significant difference in sulfur capture was observed. Improvement of resistance of the agglomerates to attrition by means of optimum drying or by use of binders is necessary.

Experimental runs were performed using eductors to feed coal to the FBM. Aerodynamic tests were conducted to optimize eductor designs for feeding coal up to 1/2 in particle size with minimum air supply pressure. A semiempirical mathematical model was developed to describe the results of the tests and to size eductors.

The FBM was replaced by the PDU in the second quarter of the reporting period. The PDU was successfully put into operation after four shake-down runs. Light-off and shut down procedures were established and routine testing initiated.

Corrosion/erosion test racks containing fifteen different metals which were installed in the bed and in the freeboard areas of the FBM were analyzed. Sections of the FBM waterwall tubes, the water door and the previously used horizontal tube bundle were removed and examined metallographically.

A series of tests were conducted in the PDU to evaluate air distributor protective materials suitable for installation over the air distributors of the Rivesville MFB.

Tests were conducted to optimize the design and location of four coal feed needle configurations of the PDU. Results show that the straight needle performed about as well as an in-bed "Y" feeder and is adequate for the nine square feet bed cross section. A flue gas sample probe traverse arrangement was used to determine a representative sampling point and the effects of coal feeder type and location.

Tests were conducted to compare the characteristics of the PDU and the FBM for the purpose of identifying process related effects of boiler configuration and to verify certain process characteristics.

Process Analysis Results

A standardized procedure to calculate the overall and outside heat transfer coefficients for the PDU in-bed surface was developed and applied to test data.

Data from MFB light-offs was used to develop a correlation which can be applied to cell light-off by hot bed transfer.

An analysis of the sulfur capture reaction in earlier CBC tests was carried out using thermodynamic calculations and test data. The analysis shows a potential for sulfur capture in the CBC.

1.3.3 Rivesville MFB Plant Operations

During this reporting period, Rivesville activities were concentrated on improving operating performance and increasing reliability of the MFB unit. In July 1977, all steam safety valves were set. In August, the unit was operated for the first time in four cell mode. In September, the unit was operated for the first time in a commercial mode, exporting steam to the Monongahela Power Company turbine generator for commercial production of electric power. Full design load operation was also achieved in September. In November, a series of test runs were made to obtain data for preliminary performance results; i.e. boiler efficiency, combustion efficiency, sulfur capture and gas emissions. A major outage occurred from December 1977 to March 1978 due to a strike by the United Mine Workers and during that period, extensive test instrumentation was installed. On start up of the unit after the outage, long duration continuous runs were achieved in April. In May and June, performance and emission test runs were performed to gather data for determination of boiler operating characteristics in various modes and varying load conditions.

1.3.4 New York Office Engineering and Design Work

Design and engineering work was performed in PER's New York Office in support of MFB operations. Various tasks were accomplished to improve performance, increase reliability and alleviate operational problems of the MFB unit.

Work included analysis of the existing structure to support additional loads imposed by installation of steam vent silencers, weigh belt feeders and modifications and additions to the boiler. An instrument compressed air system was developed to provide for additional capacity, higher quality air to improve operational reliability.

MFB operational problems were evaluated and corrective action initiated. These included:

- limestone unloading system vent filter,
- supplementary boiler feedwater supply,
- stack gas sampling platform,
- sample ducts for EPA Test Program,
- test coal bin for overbed feeder,
- flash tank vibration suppression,
- bed drain system improvement,
- fuel feed system improvement studies,
- access platforms and stairs to facilitate maintenance.

1.3.5 Subcontractor Activities

The specialized expertise of several organizations was utilized under subcontract arrangements to support MFB unit operation and maintenance activities and AFB technology development. These included:

- Foster Wheeler Energy Corporation - MFB unit operative assistance and improvement.
- Stone and Webster Engineering Corporation - plant advisory operations assistance.
- Zurn Industries - conceptual design of a package industrial fluidized bed boiler.

II. ALEXANDRIA, VA. LABORATORY HOT MODEL EXPERIMENTS AND DESIGN VERIFICATION TESTS

2.1 Introduction

The Alexandria, Virginia fluidized-bed combustion laboratory serves as the process development facility for the multicell fluidized-bed boiler design, construction and test program. Hot model experiments and design verification tests were conducted using two steam generators. The rectangular cross section FBM (fluidized bed module) which had been used for the past ten years was replaced by a square cross section PDU (process development unit) during the second quarter of this reporting period. Both boilers have nine square feet of bed area. Heat transfer surfaces of these boilers are capable of handling 10^7 Btu/hr heat release and are complemented by boiler auxiliary systems capable of feeding over 800 lb/hr of fuel and 400 lb/hr of sorbent for the reduction of sulfur dioxide emissions. To provide necessary data, the combustion systems are supplemented by a laboratory for chemical analysis of solid samples and on-line analysis of flue gas.

Material handling systems to process 3 1/2 tons per hour of coal and limestone are located adjacent to the laboratory building. Processed material is stored in 620 cubic feet capacity hoppers until automatically transferred to weigh bins on the laboratory building roof.

2.2 Laboratory Systems

2.2.1 Comparison of FBM and PDU

The PDU was designed for flexibility in testing and was the result of ten years of test experience with the FBM. Both the FBM and the PDU are atmospheric pressure fluidized-bed combustors of waterwall membrane construction with 350 psi design pressure. The same punched plate air distributor design was used in both combustors.

The PDU has a square air distributor cross section, measuring 3 ft x 3 ft, while the FBM is rectangular, measuring 1.5 ft x 6 ft. The PDU is equipped with horizontal in-bed superheater, economizer, and boiling bundles with the in-bed section of waterwall covered by refractory. The FBM, as operated during the report period, utilized only vertical wall in-bed surface. The PDU has been equipped with an air cooled steam condenser for total water recycle while the FBM was operated essentially in a once through blow-off mode. Air for the PDU is preheated by heat exchange with steam, while in the FBM the air was preheated by cooling the flue gas.

Table 1 compares the design parameters of the PDU and the FBM, while Figure 1 shows the dimensions of the two boilers. As can be seen, the PDU has fifty percent more freeboard than the FBM for the same operating bed depth.

For the FBM, samples of flue gas analysis were taken from the undiluted flue gas stream at the furnace exit. A 3 inch pipe, slotted at the bottom, extended 6 feet across the flue gas path along the top of the furnace. This averaging header was used to extract a sample stream with a fan which carried the hot flue gas through a cyclone collector to the chemistry laboratory, where samples were withdrawn from the loop. For the PDU, a sintered stainless steel probe installed above the upper tube bundle was used to traverse the freeboard. Figure 2 shows the probe traverse assembly for the PDU viewed from outside the top front door. Traverse sampling was conducted to determine the sensitivity of gas analysis to position of the sample in the freeboard. A schematic diagram of the FBM flue gas sampling system is shown in Figure 3.

For a more complete description of the FBM, refer to PER Interim Report No. 3 (August 1976) and No. 4 (August 1977).

2.2.2 Process Development Unit

In the PDU, the fluidized bed is contained in a square enclosure with refractory lined walls of vertical tubes, seal-welded to form a gas-tight enclosure.

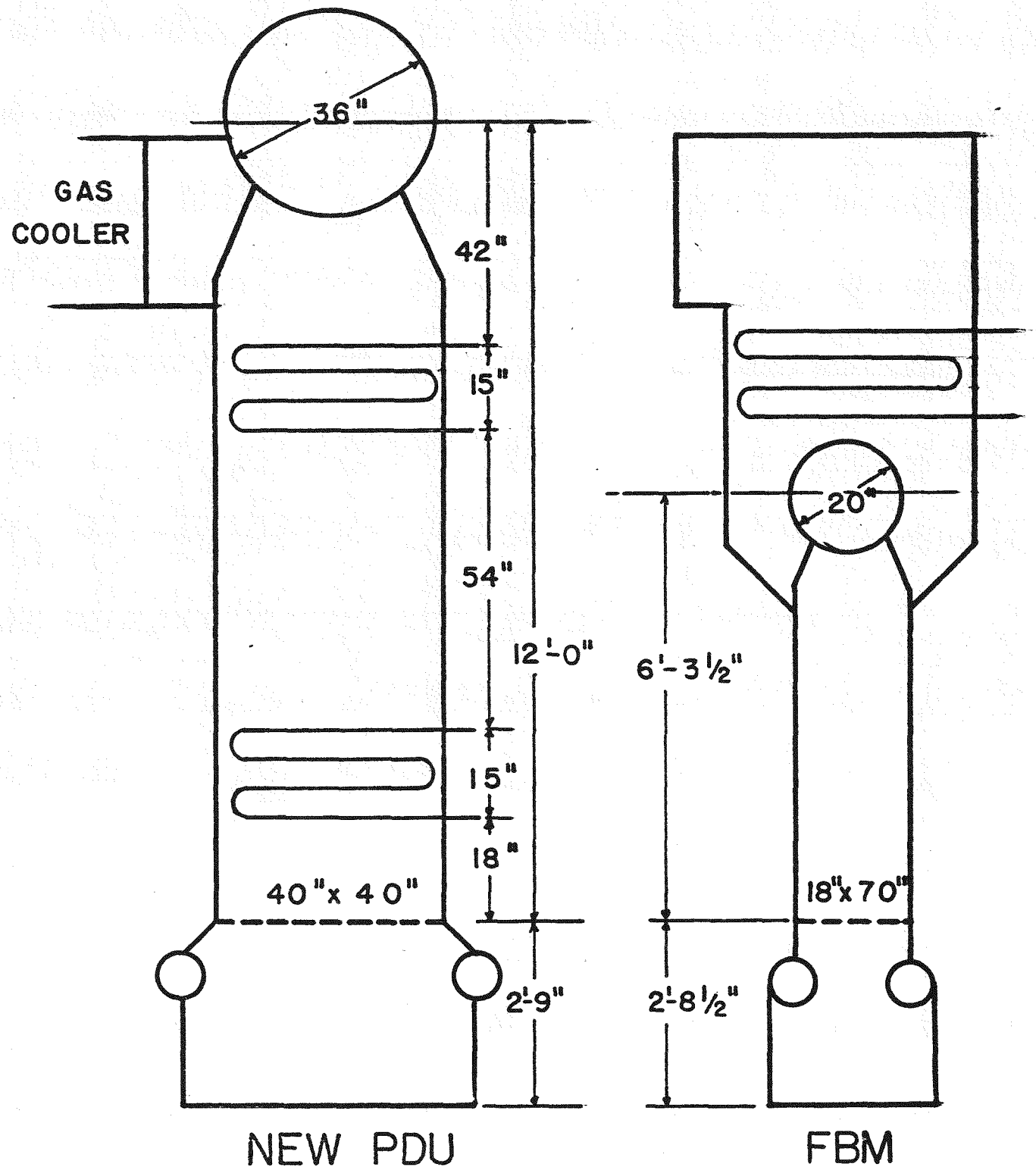
The PDU has a maximum air flow cross-sectional area of 11 sq ft but with a 2 in thick refractory lining, has a cross-sectional area of 9 sq ft. The bed is surrounded by vertical tubes which extend from four cross headers below the grid plate to the steam drum. The combustion space is accessible through six manways located on the front and back walls, below, between and above the horizontal tube bundles. An overbed light-off burner is installed in the lower rear manway and a 9 in diameter quartz window is installed in the lower front manway to permit viewing of the hot fluidized bed. Figure 4 is a photograph of the hot bed taken through the quartz window. A second burner is located in the plenum to preheat the incoming air during light-off.

Air entering the plenum at the base of the unit is directed upward through the air distributor to fluidize the bed. The static bed depth may be as high as 30 in. An automatic bed drain valve for controlling bed level and a bed material sampling valve are installed 1 3/4 in above the grid plate. Nine thermocouples are mounted in the combustion space. Five are located in the bed while four are located in the freeboard.

Table 1

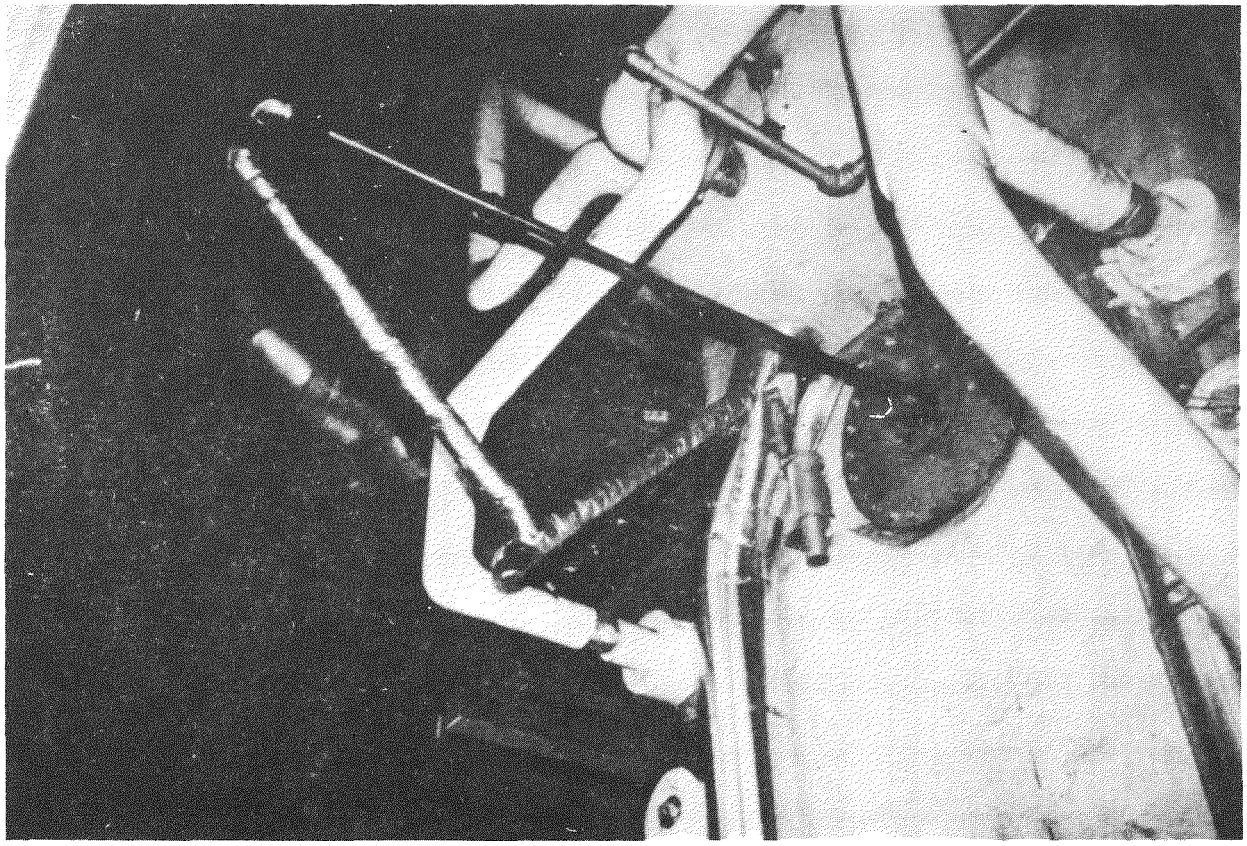
Comparison of the FBM and the PDU

	FBM	PDU
Bed Cross Section Area	9 sq ft	9 sq ft
Heat Release Rate	10^7 Btu/hr	10^7 Btu/hr
Grid Plate Design	punched plate	punched plate
Boiler Design Pressure	350 psi	350 psi
Combustor Geometry	1.5 ft x 6 ft	3 ft x 3 ft
Steam Drum Size	20 in I.D.	36 in I.D.
Freeboard Height	33.5 in	53.5 in
In-bed Bundle	none	superheater, economizer and evaporator
Waterwall Refractory	none	in-bed
Air Preheat Medium	flue gas	steam
Percentage Water Recycled	12%	100%



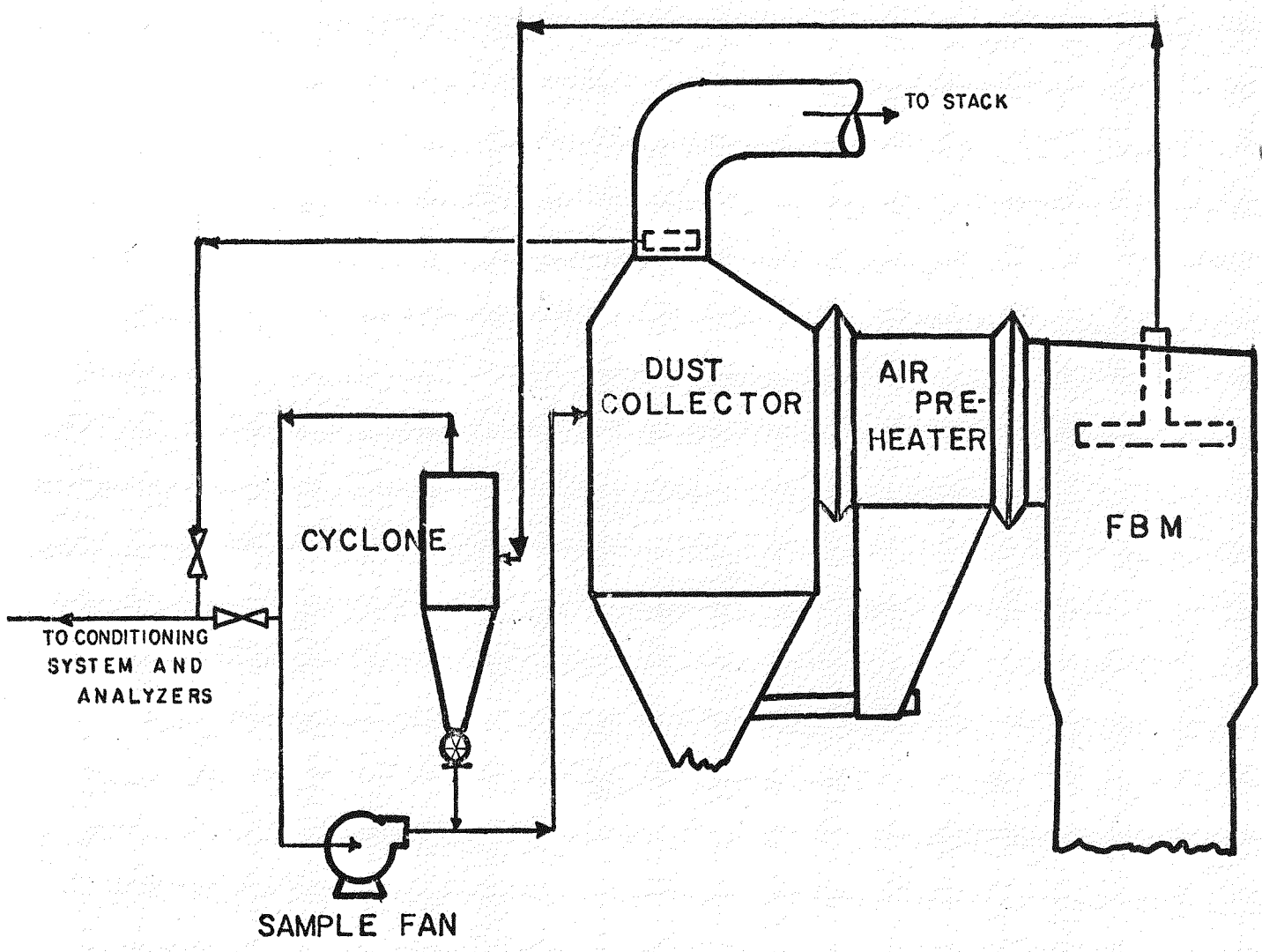
ALEXANDRIA FLUIDIZED-BED BOILERS

Figure 1



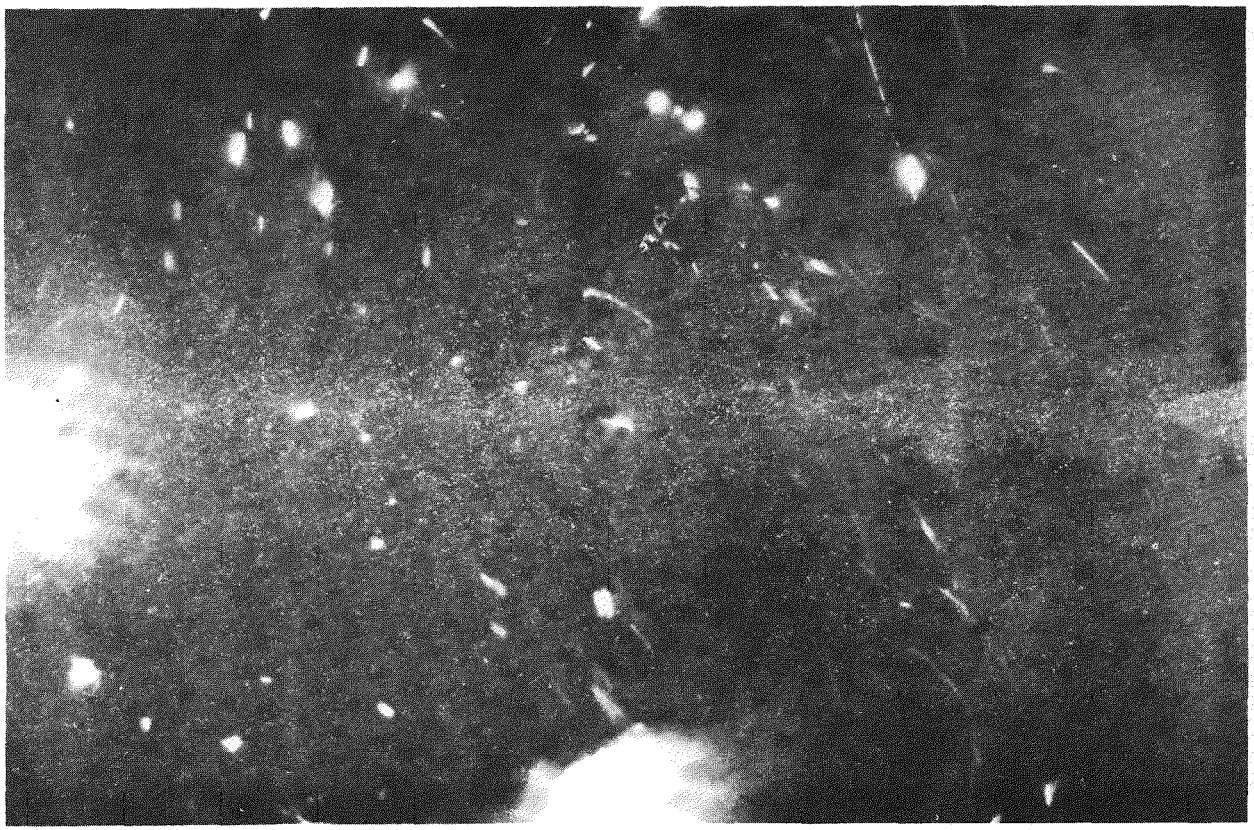
View of the front upper access door on the PDU showing the gas sample traverse assembly. At the time the photo was taken the probe was located inside the front wall of the unit.

Figure 2



FBM FLUE GAS SAMPLING SYSTEM

Figure 3



Photograph taken through 9" quartz window located in the fluidized bed. The small white spots are burning coal particles. The large white spots on the bottom and left side of the photo show devolatizing coal pieces burning. The dark background is the cooler bed material.

Figure 4

Figure 5 shows the PDU before installation of insulation and piping. Figure 6 and 7 show the PDU after installation of insulation and piping. Figure 8 shows the steam header for the superheater bundle, as well as one of the manways open. Dimensions of the PDU are shown in Figure 9. Figure 10 shows the cross section of the water-wall refractory bricks used in the PDU.

To light-off the boiler, the temperature of an eight inch static bed is raised to between 600° to 700°F with the overbed and the plenum propane fired burners. This temperature is maintained until a small amount of steam flow through the superheater bundle is obtained. A coal charge of 20 pounds is then added, the fluidizing air reduced, and when the bed temperature reaches 1000°F, continuous coal feed is initiated.

The PDU produces about 4,000 lb/hr of steam at 300 psig and 150°F of superheat. Figure 11 shows the schematic gas and solids flow diagram for the combustor and auxiliary systems utilized during the reporting period.

To cool the combustion gases leaving the boiler, a cooling water system is used. The cooling water system includes a gas to water heat exchanger, a circulation pump and an evaporative cooling tower.

2.2.3 Fuel and Sorbent Systems

High sulfur coal and fly ash were fired with Greer limestone as the sorbent bed material. All solid fuels and sorbents are processed, stored, and delivered by a materials preparation and handling system which is described in detail in PER Interim Report No. 3.

Coal is dried in a fluid bed dryer, crushed in a roll crusher, screened on a vibrating screen with the + 1/4 in size recycled. Coal under 1/4 in is stored in a silo and pneumatically conveyed to a weigh hopper located on the roof of the combustion laboratory. The coal feed rate from the weigh hopper is controlled by a variable speed rotary valve.

The limestone sorbent is screened through a #8 mesh screen and stored in a silo. From the silo, the limestone is pneumatically conveyed to a second weigh hopper on the roof of the combustion laboratory. The limestone feed rate is also controlled by a variable speed rotary valve. The limestone is mixed with the coal and fed pneumatically into the boiler.

SUPPLY AIR

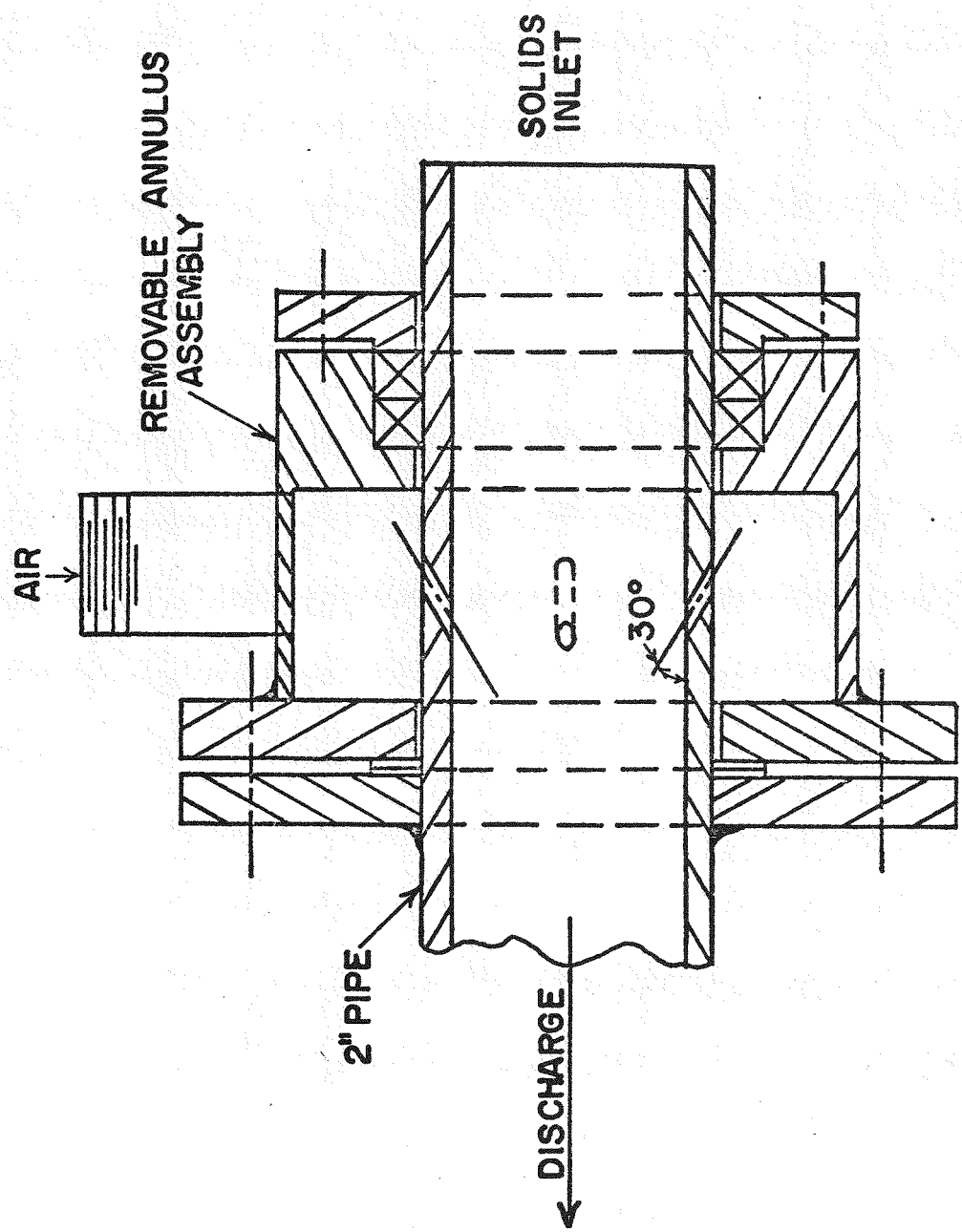
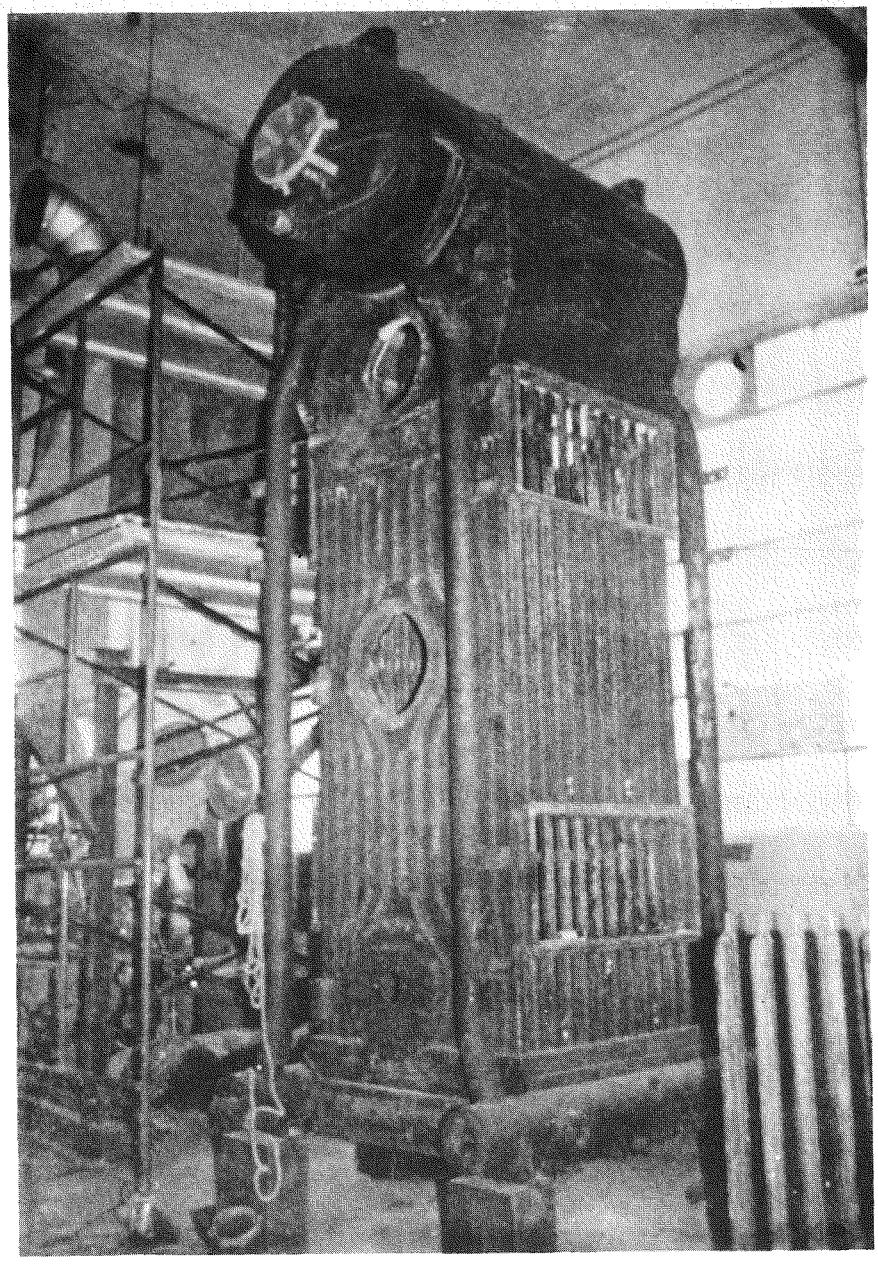
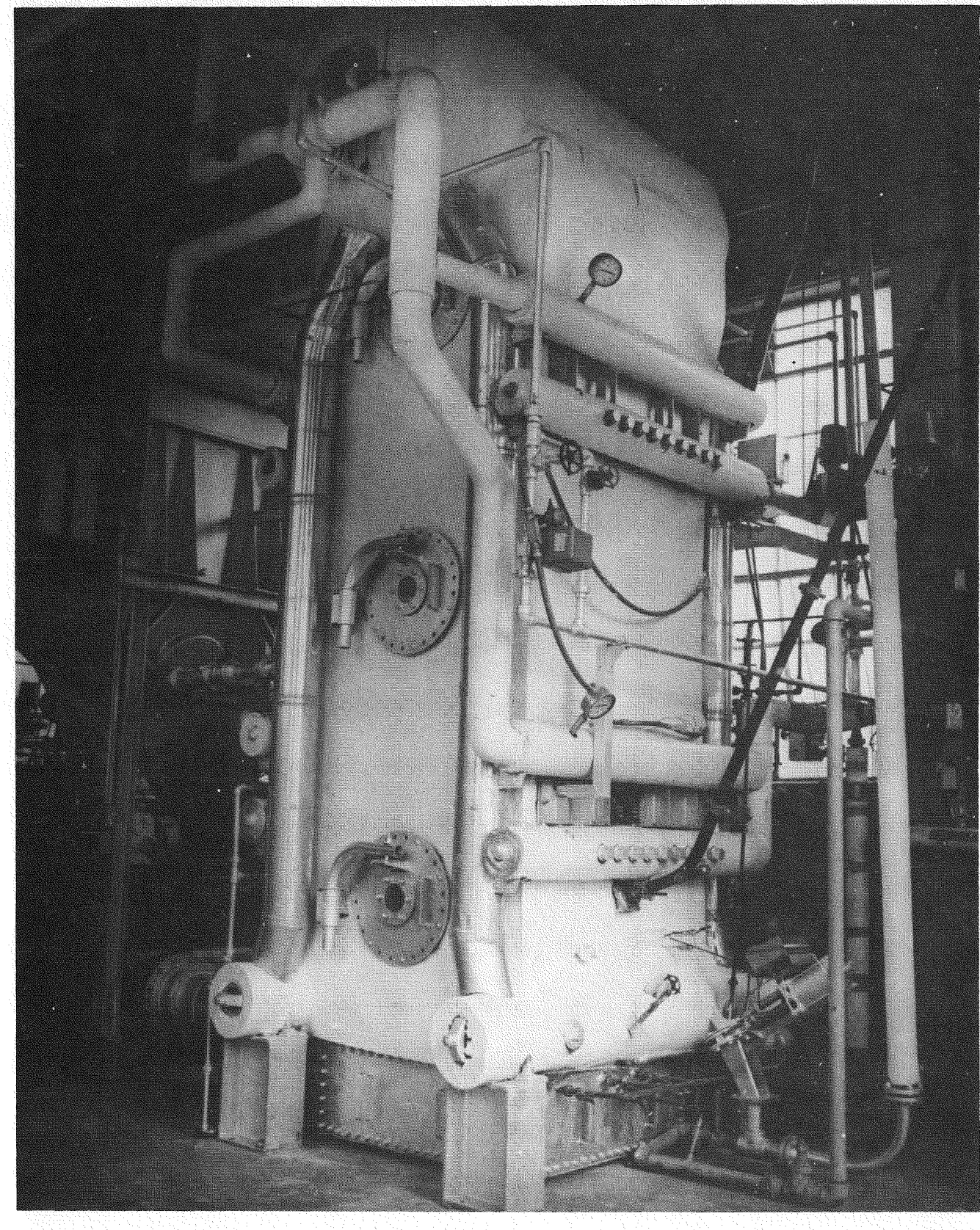

SOLIDS INLET
REMOVABLE ANNULUS ASSEMBLY
AIR
30°
2" PIPE
DISCHARGE

Figure 5



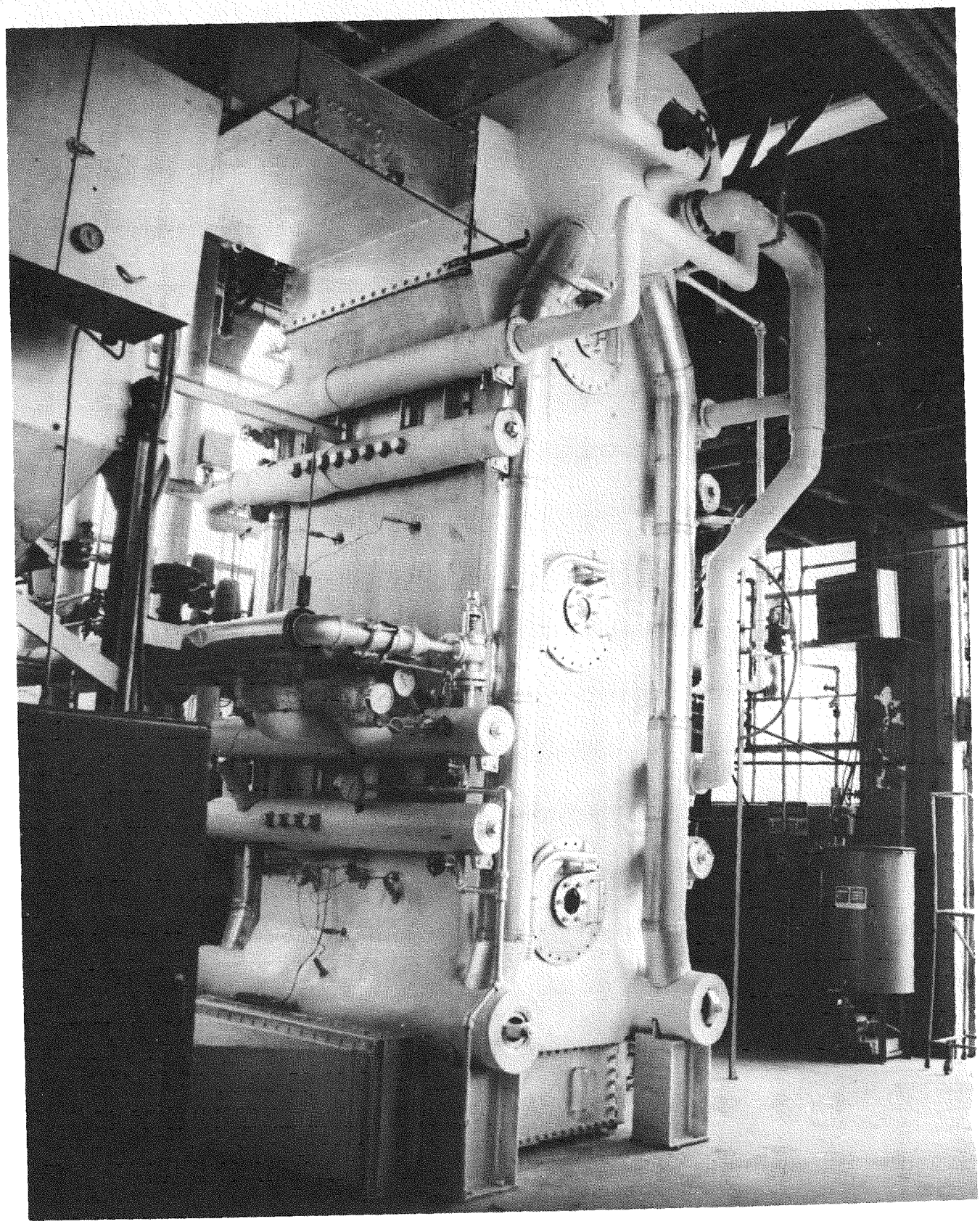
PDU BEFORE INSULATION AND PIPING

Figure 5



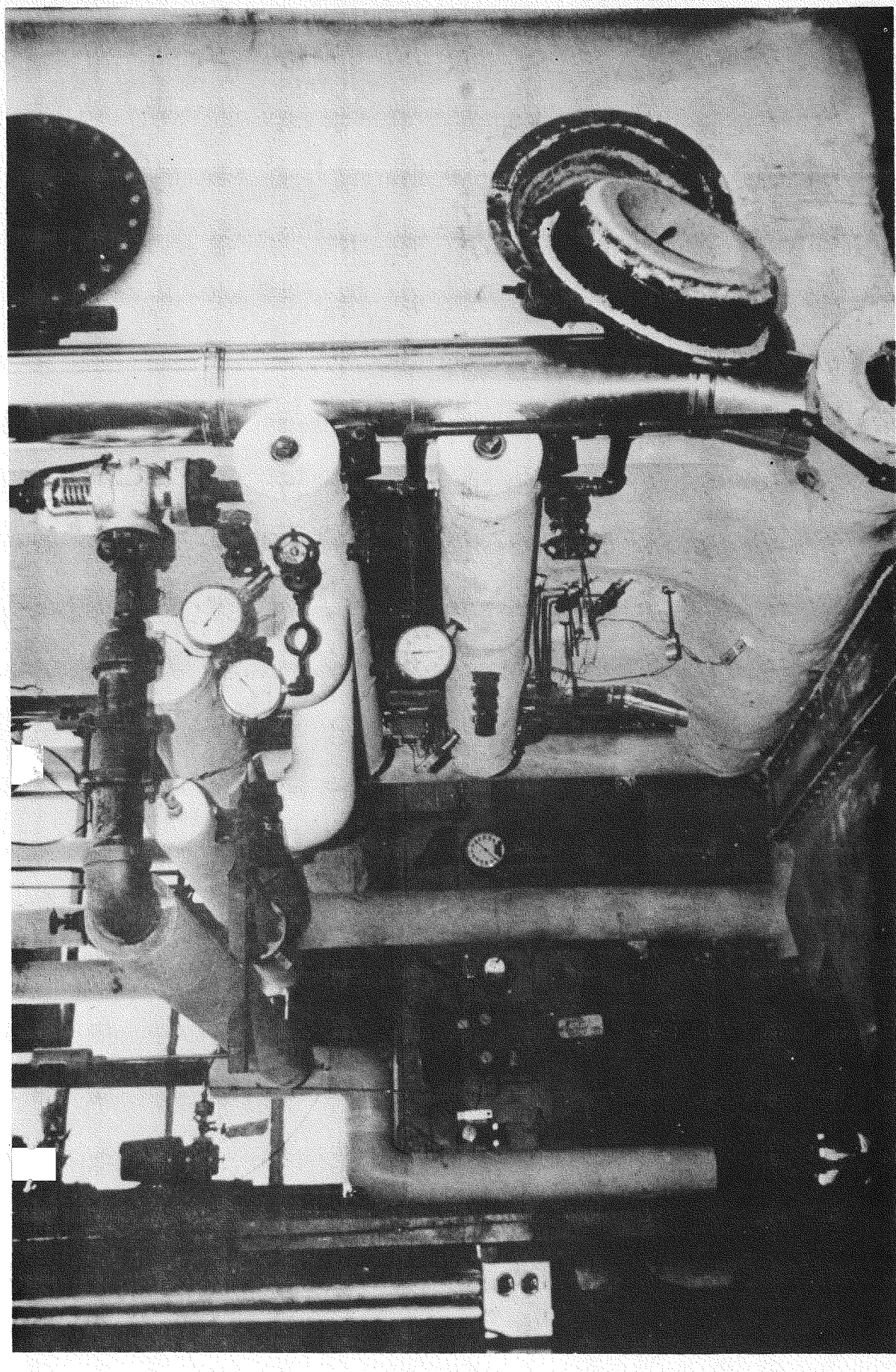
VIEW OF PDU COAL FEEDER AND BED DRAIN SIDE

Figure 6



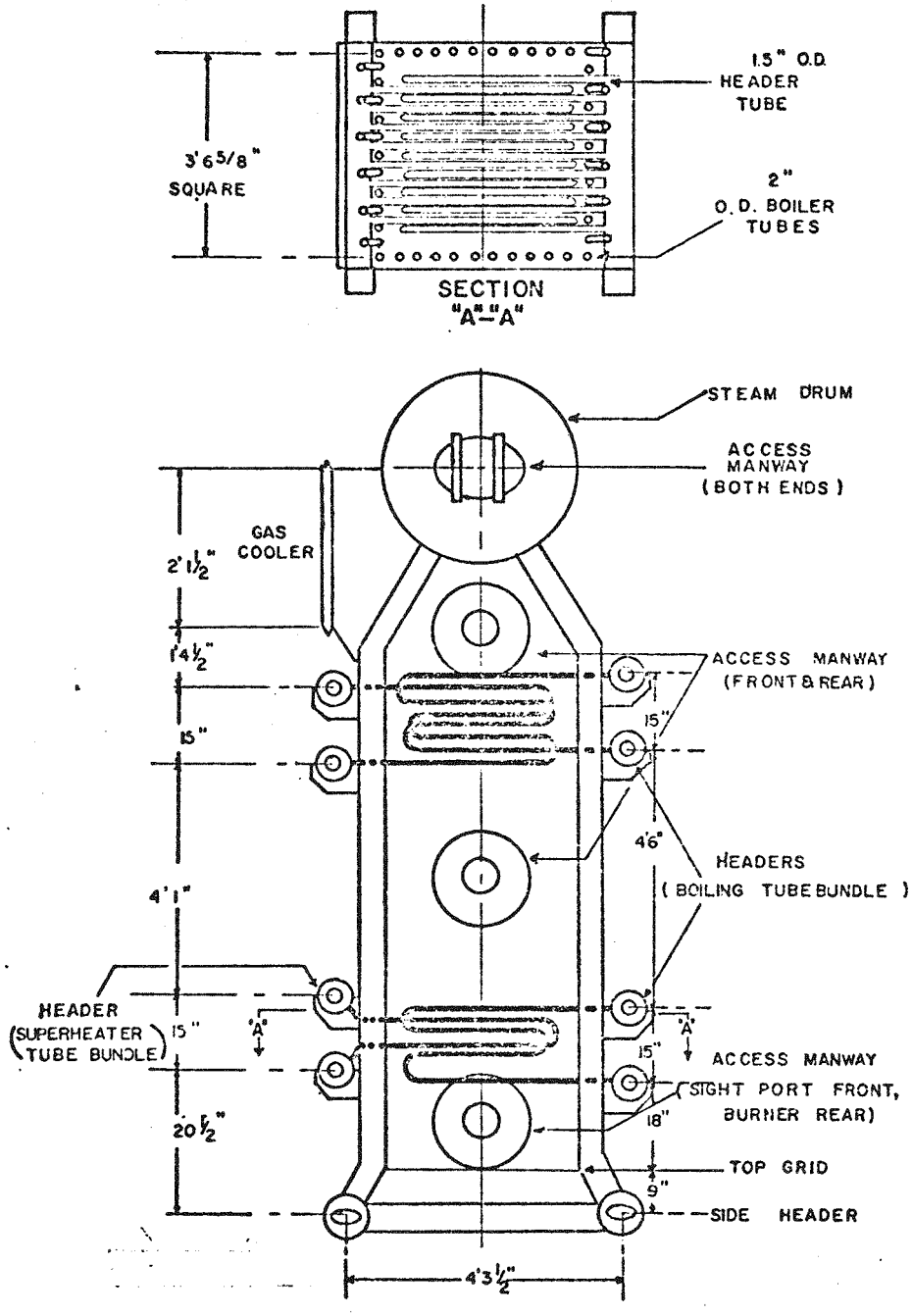
VIEW OF PDU AIR INLET AND GAS OUTLET SIDE

Figure 7



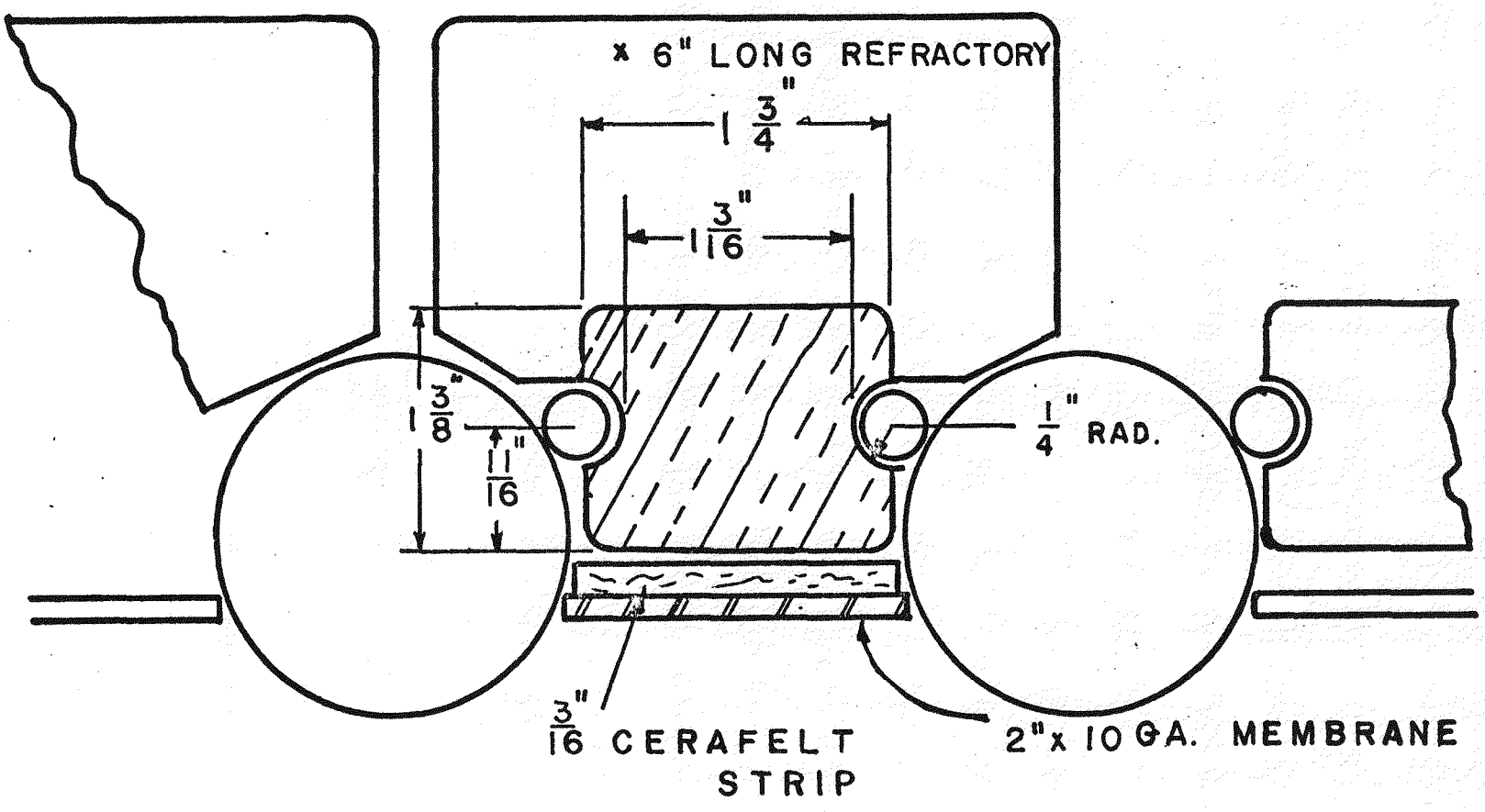
SUPERHEATER BUNDLE IN PDU

Figure 8



DIMENSION DIAGRAM OF THE PDU

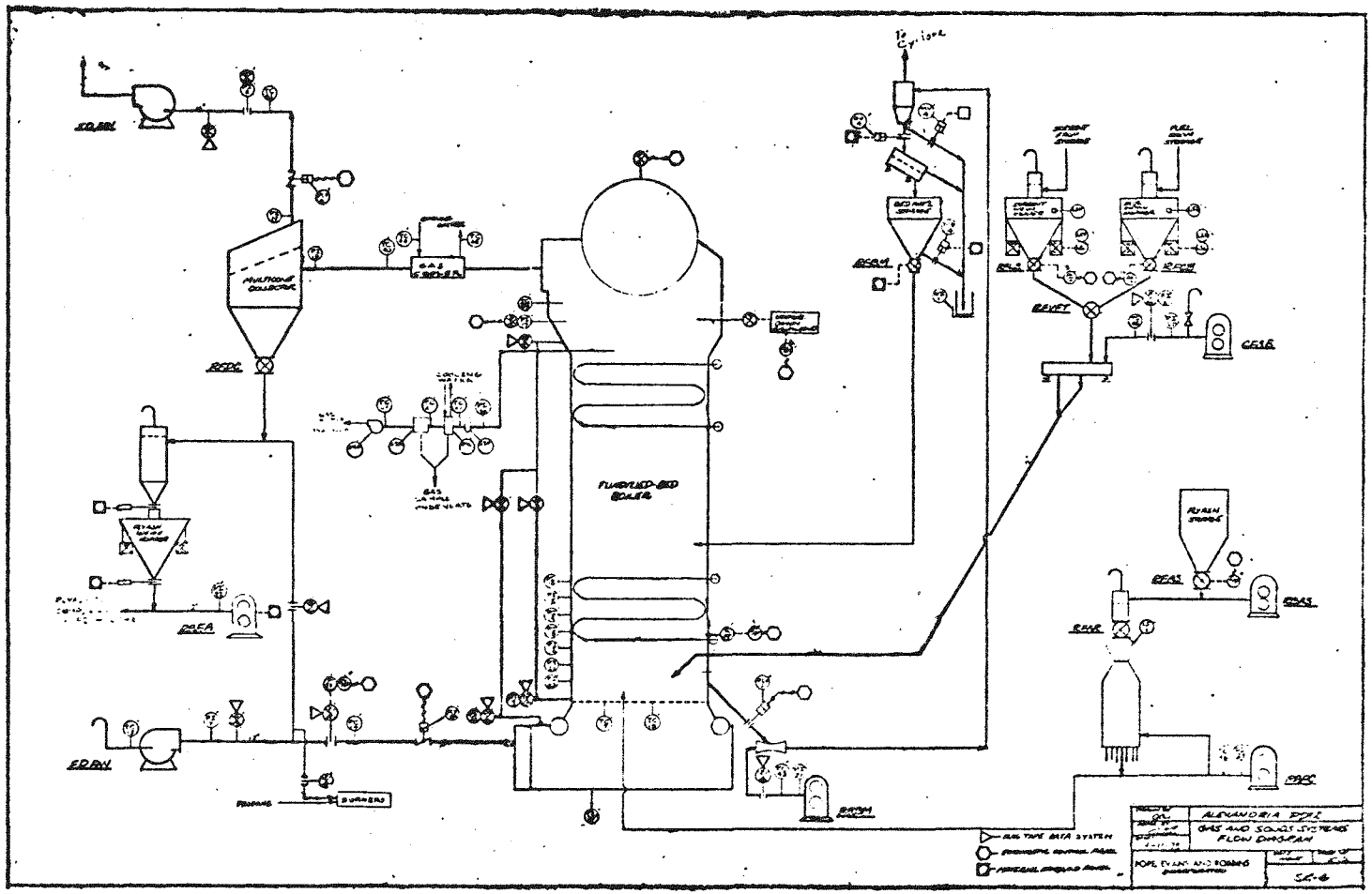
Figure 9



CROSS SECTION OF WATERWALL

REFRACTORY BRICKS USED IN PDU

Figure 10



GAS AND SOLIDS SYSTEMS FLOW DIAGRAM

Figure 11

Fly ash is collected in dust collectors and pneumatically transported to a weigh hopper. From there it can be pneumatically transported to a fly ash storage silo located in the materials preparation system area. A variable speed rotary valve controls the fly ash feed rate out of the storage silo. From the silo, the fly ash is transported to a fly ash injection system which is described later.

2.2.4 Air Supply and Control System

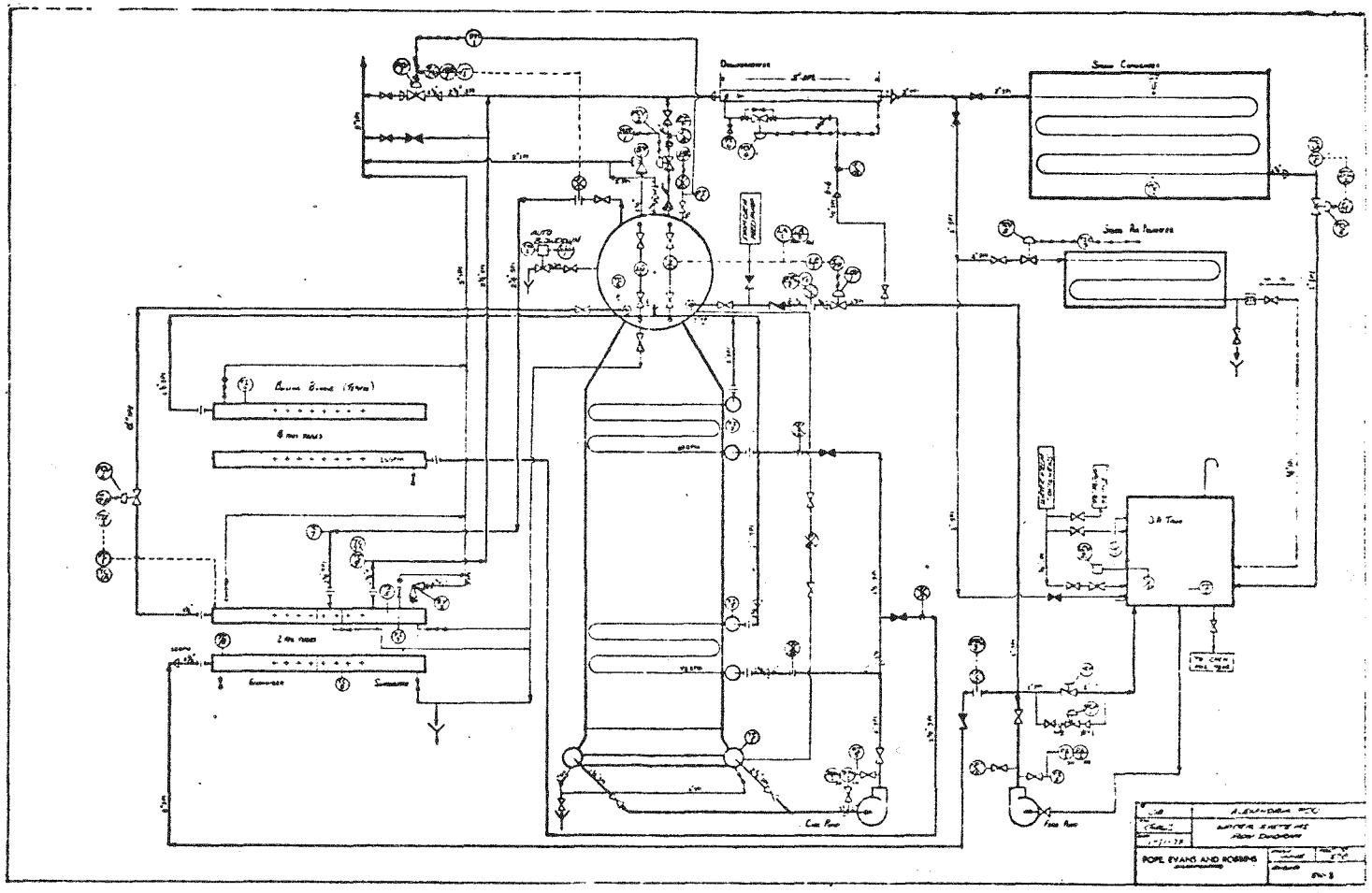
A high pressure fan capable of providing 12,000 pounds of air per hour at 50" H_2O pressure is used as the forced draft (F.D.) fan to provide combustion/fluidizing air. The air flow rate is set by a pneumatically operated damper. The air is preheated by a steam coil and metered by the main air orifice. The freeboard space of the PDU is operated slightly below atmospheric pressure in accordance with conventional boiler practice, by means of an induced draft (I.D.) fan. A pneumatically operated damper is used to maintain the draft in the furnace.

The position of both the F.D. and the I.D. fan dampers can be controlled from the automatic combustion control panel. The I.D. fan damper can either be set manually or automatically to maintain a set furnace pressure. The F.D. fan damper can either be set manually to deliver a constant air flow, or automatically in response to system variables, through the automatic combustion control system which is described elsewhere.

2.2.5 Feed Water and Steam Systems

Figure 12 shows the water system schematic for the new PDU with the control valves indicated. Some boiler feed water passes through the in-bed economizer bundle before going into the steam drum. This flow rate is controlled by a temperature controller on the economizer outlet to prevent boiling. The in-bed economizer bundle is instrumented to measure in-bed heat transfer coefficients. The remainder of the feedwater passes through the liquid level control system and into the steam drum. Water for the three boiling bundles is withdrawn from the main lower headers by a recirculation pump. The steam produced passes through the in-bed superheater bundle. The superheater bundle is also instrumented for heat transfer coefficient measurements.

Flow through the superheater can be controlled by three methods. First, a constant flow controller can be used to maintain a constant flow of steam regardless of changes in boiler pressure. A back pressure regulator can be used



WATER SYSTEMS FLOW DIAGRAM

Figure 12

to maintain a constant pressure in the steam drum, thereby allowing the flow to fluctuate. The third method of steam flow control is through the steam condenser. The steam condenser can be used to vary the load to the boiler by changing the air flow through the condenser coil. Prior to entering the steam condenser, the superheated steam passes through a desuperheater to lower the temperature to near saturation.

Some of the desuperheated steam passes through a Trane air heater coil to preheat the main air to 300°F. Steam flow to the condenser is controlled by a temperature controller to maintain the plenum temperature. A steam trap prevents steam from leaving the coil with the condensed water.

The PDU uses water of a quality commensurate with the normal operating pressure of 300 psig. Scale forming minerals are removed from the water by a Zeolite softener and the water is raised to 200°F in a open tank which acts as a deaeration feedwater heater. The feed water is controlled by a proportional-integral (P-I) controller that maintains the boiler water level in the steam drum at a constant set point. Steam produced in the PDU goes to a steam condenser and is returned to the deaeration tank. A back pressure regulator is also used to maintain steam pressure in the boiler by venting a small amount of the steam. Steam generation takes place in the waterwalls by natural convection and by forced circulation in boiling tube bundles located inside the bed and above the bed. Two other smaller tube bundles in the bed act as a superheater and an economizer. The constant steam flow controller operates by utilizing a flow signal from the main steam flow orifice and vents steam to the atmosphere.

2.2.6 Fly Ash Collection System

The products of combustion are separated from the gas by a multi-cyclone dust collector. A heavy duty vibrator is installed on the hopper under the cyclone collector to prevent fly ash from accumulating on the hopper walls. The ash is conveyed from the cyclone hopper to a baghouse and weigh station outside the laboratory.

2.2.7 Fly Ash Reinjection System

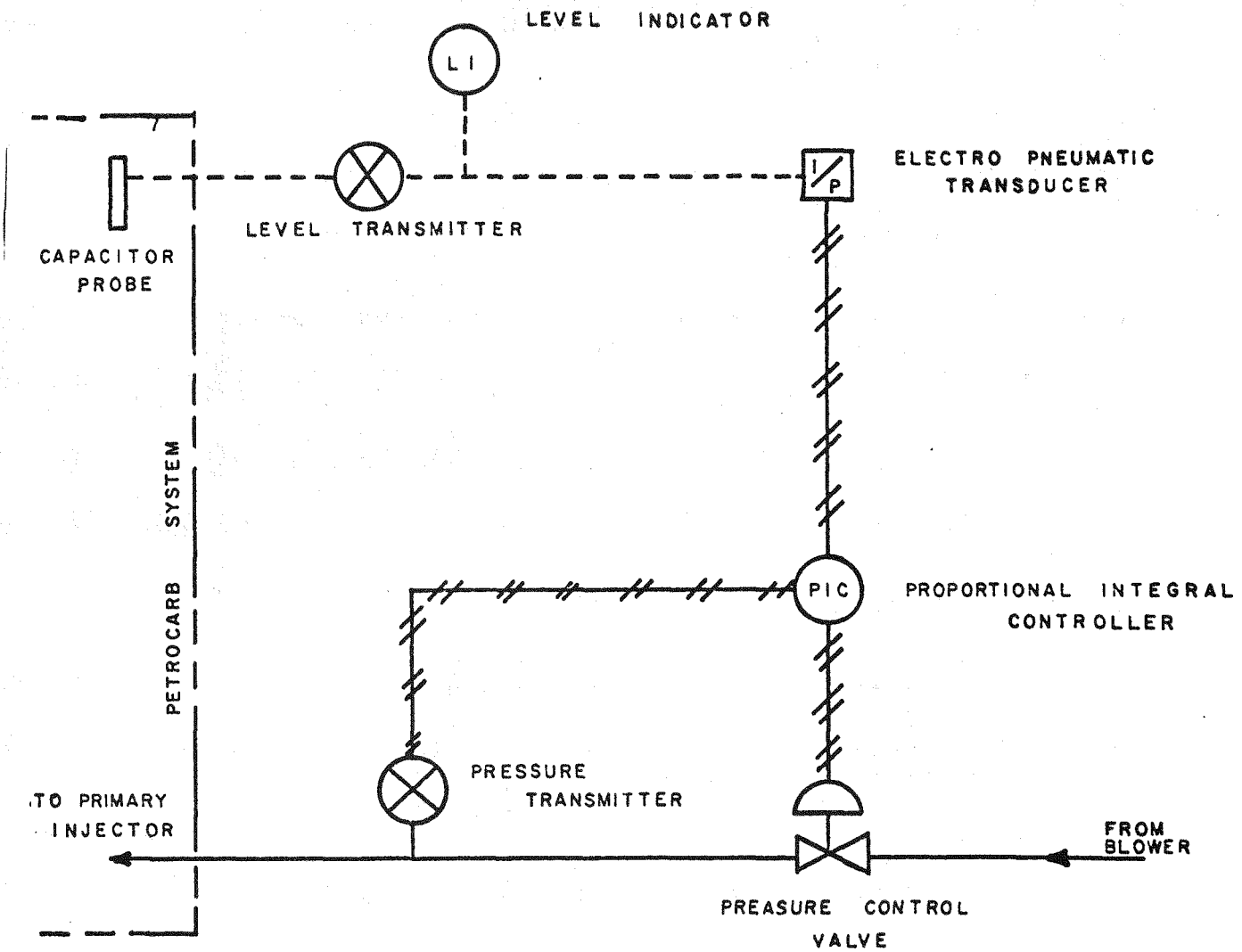
A system to reinject fly ash into the FBM was evaluated for use in the Rivesville MFB. The injector is a proprietary design of Petrocarb Incorporated. Since its installation, the injector has been used to feed both fly ash and 1/8 in coal to the FBM in various test programs.

Originally the Petrocarb vessel consisted of three separate sections; the primary injector, the seal leg, and the surge vessel. The primary injector is the heart of the Petrocarb system and in general, is similar for all their units. This chamber splits the fly ash into the number of required streams. The unit at Alexandria is capable of splitting the fly ash into six streams. The seal leg is a 24 ft long, 4 in I.D. pipe located on the top of the primary injector which remains full of fly ash during the test. The purpose of the seal leg is to create a pressure differential between the bottom and the top of the leg enabling pressures of up to 10 psi to be developed in the primary injector. During the testing programs, it was discovered that injector pressures less than 4 psi in the primary injector were sufficient to achieve the desired feed rates of coal or fly ash. As a result of this fact and because hold up of fly ash or coal in the seal leg was a constant operational problem, the seal leg has been eliminated from the system. The surge vessel, which was located at the top of the seal leg, is now mounted on top of the primary injector. This vessel is essentially an expanded section of the seal leg. A capacitance level probe located in the surge vessel monitors the level of fly ash in the vessel for control purposes and a rotary air lock seals the injector pressure.

Fly ash is fed to the Petrocarb injector through the rotary feeder located on the top of the surge vessel. The fly ash then passes down into the primary injector. In the primary injector, the fly ash is split into multiple streams and pneumatically transported with air. Valves between the primary injector and the transport air permit operation of any number of feeders. Flow through the various injector lines can be balanced by adjusting the flow of transport air.

The Petrocarb fly ash injector was equipped with an automatic control system. This system was designed so that the unit would deliver as much fly ash as was fed to it. Figure 13 shows a schematic of the automatic control system. The capacitance level probe produces a signal which acts as the set point for the automatic controller. The controller maintains the pressure in the primary injector at the set point. In this way the level in the surge vessel reaches a stable level which gives the proper primary injector pressure for the fly ash feed rate.

A considerable amount of time was expended calibrating the fly ash injector and solving operational problems. Problems with hold up of fly ash or coal in the seal leg and with the automatic controls prevented normal operation of the unit during most of the testing programs. Operation



FLY ASH CONTROL SYSTEM

Figure 13

of the unit during these tests is discussed in the test results section. However, with the elimination of the seal leg, the Petrocarb injector system appears to be operating satisfactorily.

2.2.8 Automatic Control System

The control system was designed to operate the PDU based on the control logic used in the Rivesville MFB. The purpose of the control system was to permit longer test runs of the Alexandria unit and to verify the control logic of the MFB system.

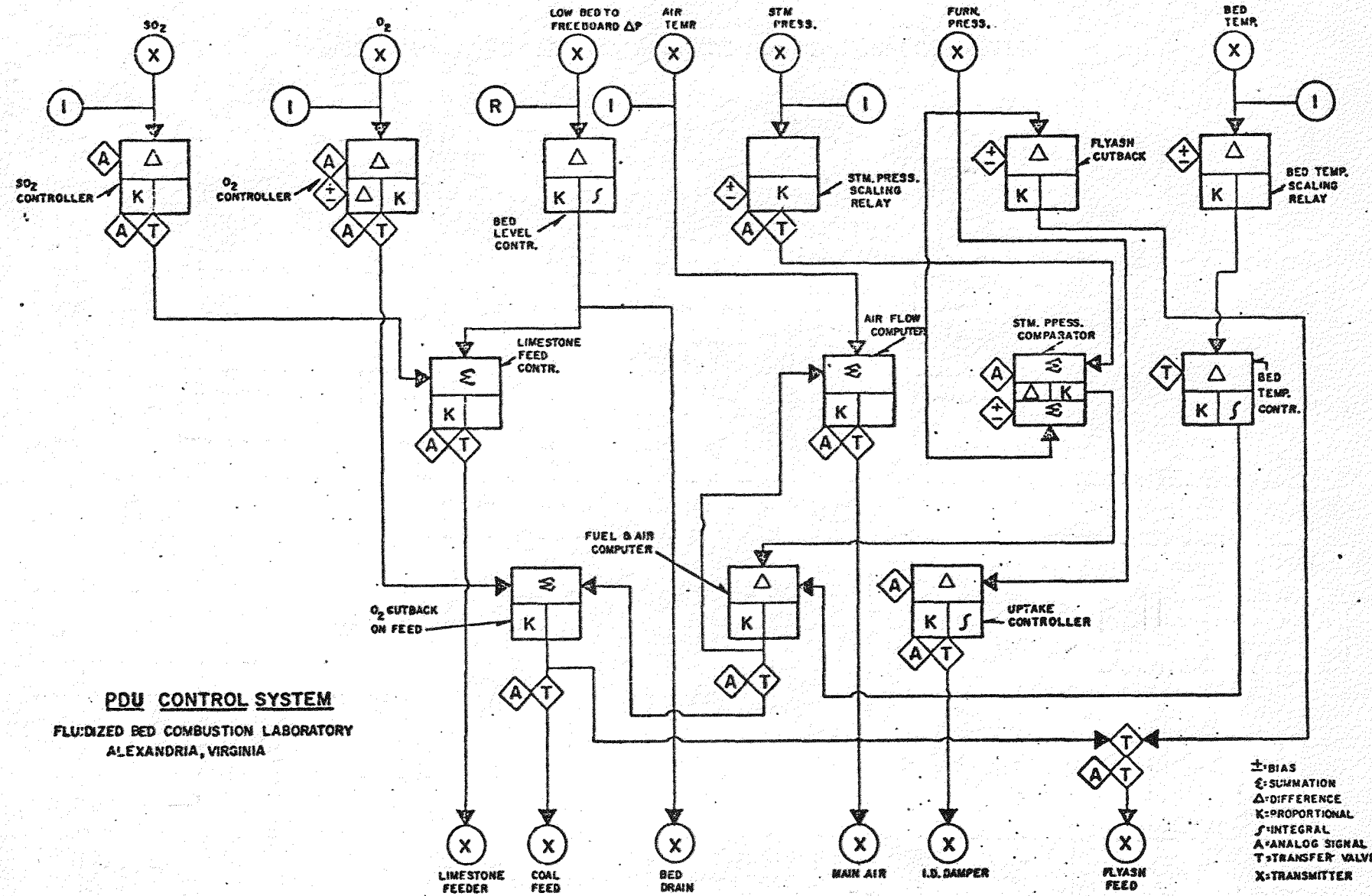
Figure 14 shows a scheamtic of the PDU control system. The control logic consists of three separate, but not independent control loops. Loop one, the uptake controller, is a simple P-I controller which positions the I.D. fan damper to maintain a constant furnace pressure. Loop two, the bed temperature controller, consists of a P-I controller to maintain bed temperature and eight proportional controllers to adjust the coal and combustion air feed rates to compensate for fluctuations in steam pressure, air temperature, furnace pressure, and percent O_2 in the stack. The third control loop is the bed level and SO_2 control loop. This control loop uses a P-I controller to maintain stone feed and bed removal rates to maintain a set SO_2 concentration in the stack. Figure 15 shows the interface schematic between the FBM and the control system.

2.2.9 Chemical Analysis Systems

The fluidized-bed combustion laboratory includes a chemical laboratory which has the capability of handling nearly all of the required gas and solid chemical analyses. On-line gas analyses are performed and recorded during the tests by automatic flue gas monitoring equipment. Chemical analyses of solid samples taken periodically during the test runs, permit total material balances to be conducted on the combustion system. From these analyses combustion efficiencies and other performance parameters can be determined.

Solid Sampling System

Coal and limestone samples are taken at the base of the coal and limestone silos. A hole has been cut and a nipple welded into the tee just under the rotary feeder on the coal and limestone silos. A cap has been placed on the end of the nipple with a handle welded-on for easy removal. To collect a sample, the blower is turned off, the cap on the nipple removed and a sample container



PDU CONTROL SYSTEM

FLUIDIZED BED COMBUSTION LABORATORY
ALEXANDRIA, VIRGINIA

Figure 14

SCHEMATIC OF PDU CONTROL SYSTEM.

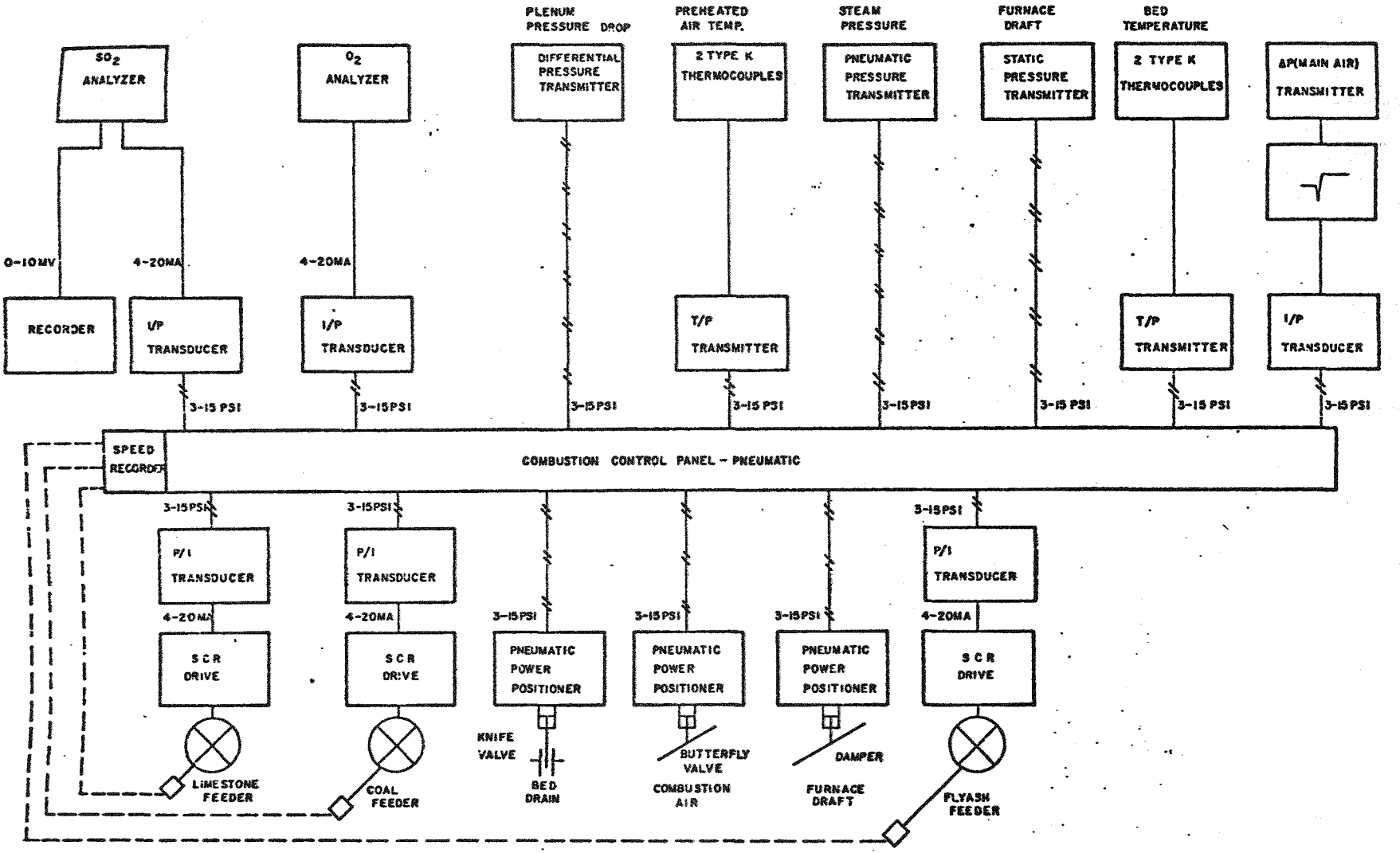


Figure 15

PDU CONTROL PANEL TO PROCESS INTERFACE SCHEMATIC

placed under the nipple. The rotary feeder jog button is momentarily pushed, allowing a sample to fall into the sample container. Samples of coal and limestone are taken every hour during a test and combined for analysis.

Bed material samples are withdrawn via a tube which penetrates the furnace wall $1\frac{3}{4}$ in above the grid plate. A nitrogen purged stainless steel bed sampling tube is used in order to preserve the carbon content and eliminate moisture deterioration of the sample.

Fly ash samples are taken from the penumatic transport line between the cyclone dust collector and the fly ash bag house. A Y has been placed in the line with gate valves in each leg as shown in Figure 16. To collect a sample, a nipple with a bag filter is screwed into the open leg of the Y. The flow of fly ash to the bag house is then diverted to the collection bag by opening the sample valve and closing the other valve.

Dust samples from the stack are collected using a Rader high volume isokinetic sampler. The sampling port is located between the cyclone dust collector and the I.D. fan. A pitot tube is used to obtain an average stack gas velocity before the sample is taken. The pitot reading is used to set isokinetic conditions during sampling.

Flue Gas Sampling System

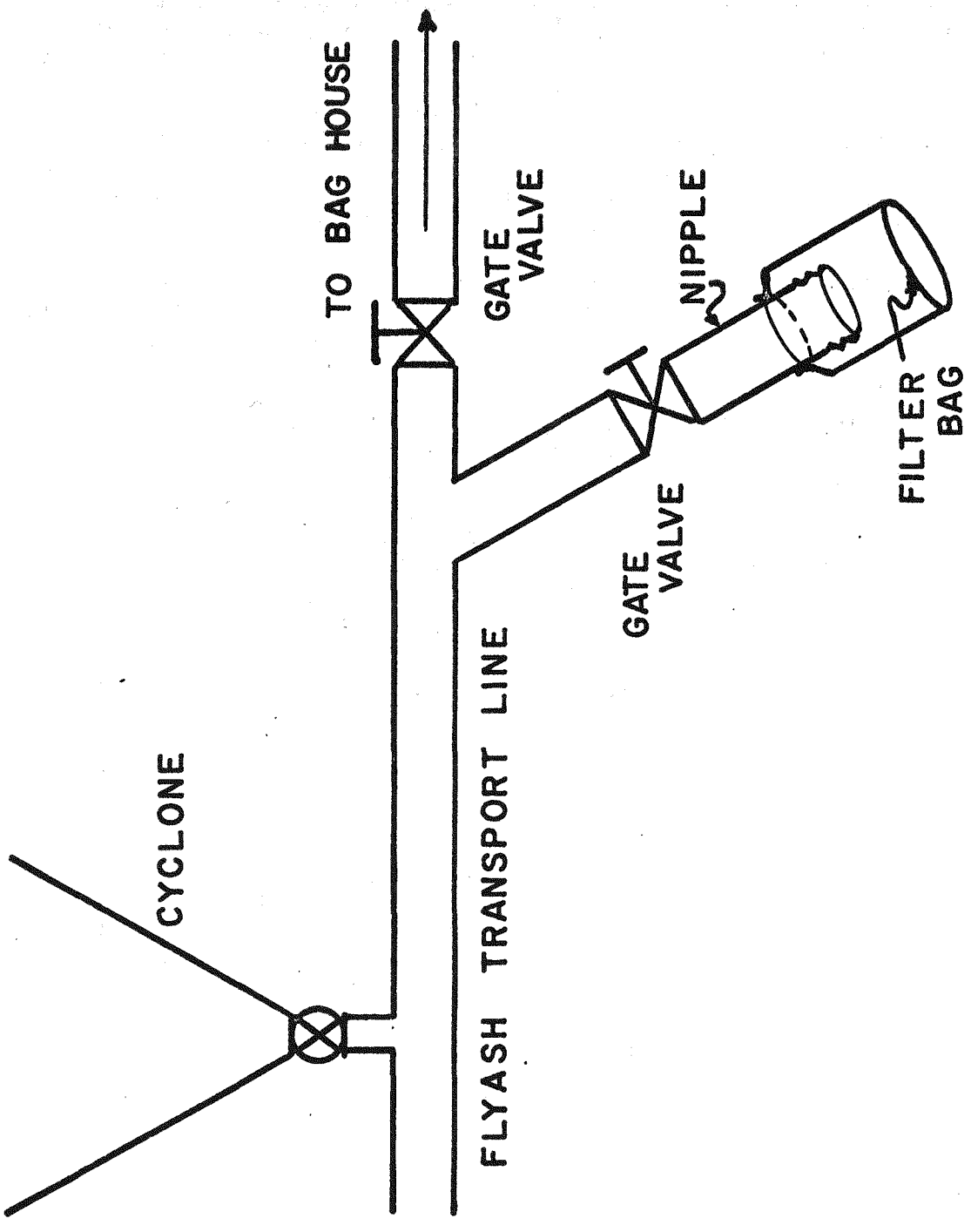
Samples of flue gas are continuously taken from the freeboard region of the boiler. A probe consisting of a sintered stainless steel filter and 1/2 in stainless steel tubing enters the boiler through the top front door. The probe penetrates the freeboard to a depth of 18 in. A vacuum pump is used to withdraw the flue gas from the boiler to a gas conditioning system and then to the gas analysers.

Flue Gas Sample Conditioning System

The flue gas sample conditioning system removes dust and water from the gas sytem and is shown in Figure 17.

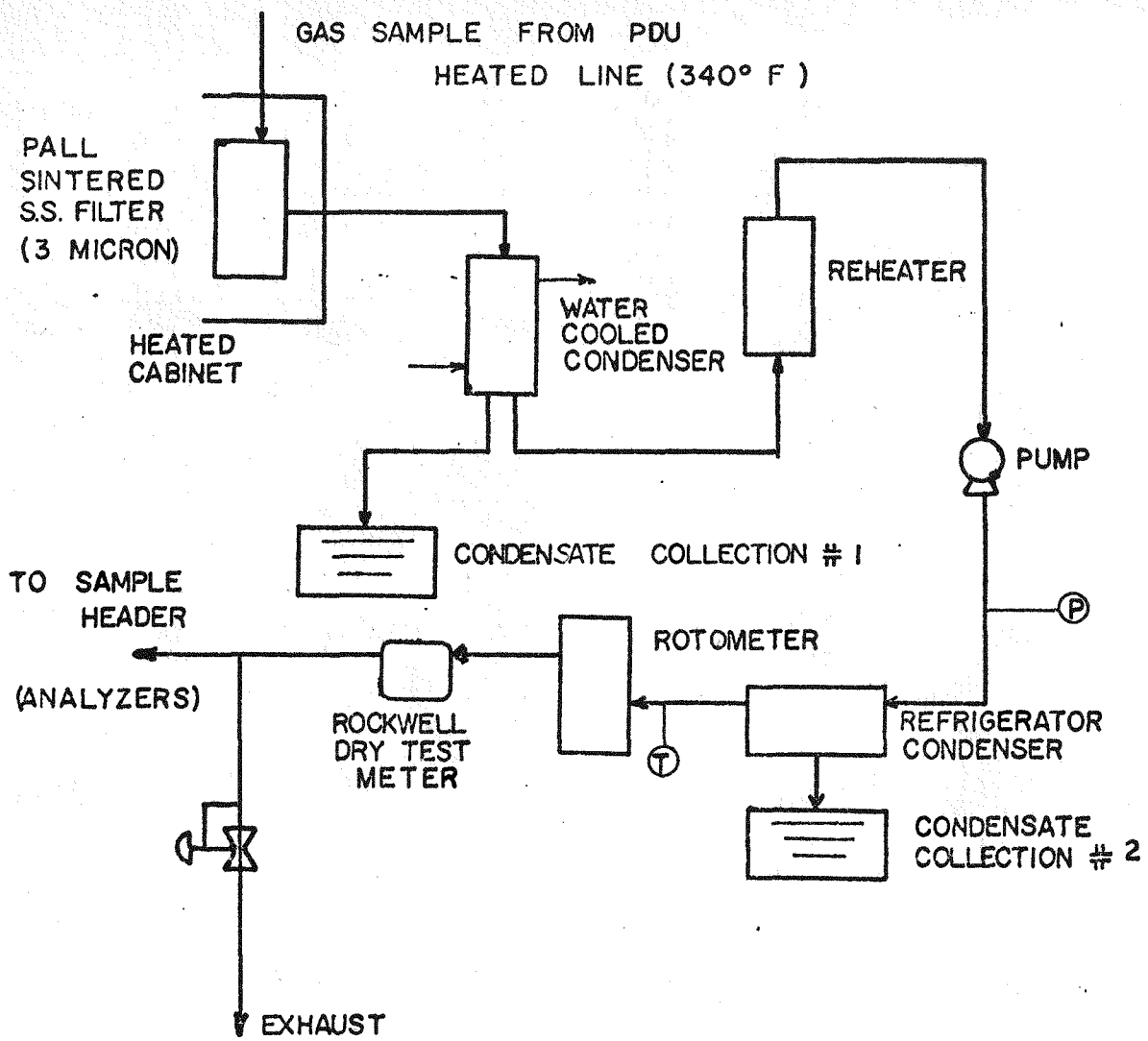
The system consists of two parallel heated filters, a water cooled condenser, a "warm" pipe leading to the sample pump and a refrigerated cooler. The sample handling and preparation system has a response time of less than one-half minute. The components consist of:

(a) Dust filters which are housed in an electrically heated, insulated enclosure. The enclosure temperature is maintained at approximately 350°F.



FLY ASH SAMPLING ARRANGEMENT

Figure 16



Notes

1 - Total condensate collected is #1 + #2

2 - Assume gas to be saturated with H_2O leaving refrigerated condenser

3 - Total volume of gas samples measured by Rockwell Dry Gas Meter

SAMPLE CONDITIONING AND MOISTURE DETERMINATION SYSTEMS

Figure 17

(b) The water cooled condenser is of the Friedrichs type. There is always some concern that a gas sample containing small quantities of SO_2 will lose the SO_2 to the condensate. Condensate is therefore stored away from the gas path. Analysis of the condensate shows that at the SO_2 levels encountered, the loss into the condensate is negligible in an overall sulfur balance.

(c) A heated section of pipe connects the water cooled condenser to the sample pump. This is to ensure that no water condenses within the sample pump.

(d) Sample pumps have been a source of problems until Metal Bellows Company furnished their Model MB-235, which has given noiseless, trouble-free service.

(e) A Freon cycle refrigerator is used to chill the sample and remove most remaining water vapor.

(f) A Perma Pure dryer is used to remove the last traces of moisture from the refrigerated sample to the infrared instruments. The dryer resembles a shell and tube heat exchanger with a 1/2 in stainless steel tube as the shell. The tubes are made of material which absorb water. Dry gas is passed through the shell to remove the moisture.

Methods of Analysis

Table 2 lists the currently utilized gas analysis equipment as well as methods used to analyze solid samples. Flue gas is continuously analyzed for carbon dioxide, carbon monoxide, oxygen, nitric oxide, sulfur dioxide, and total hydrocarbons. Routine analyses are performed on samples of coal, limestone, bed material, fly ash, dust, flue gas, and boiler water. The samples are collected according to the following methods:

ASTM Designation D2234-72, Standard Methods for Collection of a Gross Sample of Coal.

ASTM Designation C50-57, Standard Methods of Sampling, Inspection, Packing, and Marketing of Lime and Limestone Products.

ASTM Designation D29828-71, Standard Method of Sampling Stack for Particulate Matter.

ASTM Designation D860-54 (1971) Standard Method for Sampling Water from Boilers.

Table 2

Methods of Analysis

Analysis	Sample Type	Method
Alkalinity	Boiler H_2O	Hack Model AL-AP Alkalinity Test Kit.
Ash Loss on Ignition (LOI)	Coal Limestone Bed material Fly ash Dust	ASTM Designation D3174-73, Standard Method of Test for the Analysis Sample of Coal and Coke.
Bulk density	Coal Limestone Bed material	Volume and mass unpacked sample is measured in a graduated cylinder.
Calcium	Coal Limestone Bed material Fly ash Dust	ASTM Designation D2795-69 (Reapproved 1974), Standard Method of Analysis of Coal Ash and Coke Ash.
Carbon Hydrogen	Coal Limestone Bed material Fly ash Dust	Coleman Model 33 Carbon-Hydrogen Analyzer.
Carbon dioxide Carbon monoxide	Flue gas	Beckman NDIR
Chloride	Boiler Feed H_2O	Hack Model 7-P, Chloride Test Kit.
Conductivity	Boiler H_2O	Hack Portable Conductivity Meter Model 16300.
Hardness	Boiler Feed H_2O	Hack Model HA-71A, Total Hardness Test Kit.
Heat of Combustion	Coal Fly ash	Plain Jacket Parr Oxygen Bomb Calorimeter.
Hydrocarbons	Flue gas	Beckman ID8A Hydrocarbon Analyzer.

Table 2

Methods of Analysis (continued)

Analysis	Sample Type	Method
Hydrazine	Boiler H_2O	Hack Model HY-2, Hydrazine Test Kits
Iron	Coal Limestone Bed material Fly ash Dust	ASTM Designation D 1068-68
Moisture	Coal Limestone	ASTM Designation D3173-73, Standard Method of Test for Moisture in the Analysis Sample of Coal and Coke.
Nitric oxide and NO_x	Flue gas	Thermo Electron Corp. Chemiluminescent NO-NO_x Gas Analyzer.
Organic Phosphate	Cooling Tower H_2O	Olin Water Service O-P Test Kit.
Oxygen	Flue gas	Taylor Servomax Oxygen Analyzer.
Particle density	Coal Limestone Bed material	Mass is determined by weighing in a graduated cylinder. Liquid displacement is used for volume measurement.
pH	Boiler water	Orion Model 801 digital pH/MV meter.
Phosphate	Boiler H_2O	Hack Model PO-19/PO-19A, Orthophosphate Test Kit.
Silica	Coal Limestone Bed material Fly ash Dust	ASTM Designation D859-68, Standard Method of Test for Silica, Gravimetric Technique.

Table 2

Methods of Analysis (continued)

Analysis	Sample Type	Method
Size distribution	Coal Limestone Bed material	ASTM Designation D410-38, (Reapproved 1969), Standard Method of Sieve Analysis of Coal.
Sulfite	Boiler H_2O	Hack Model SU-5, Sulfite Test Kit.
Sulfur	Coal Limestone Bed material Fly ash Dust	LECO Sulfur Determinator Model 517.
Sulfur dioxide	Flue gas	Thermo Electron Corp. SO_2 Pulsed Fluorescent Analyzer.

Additional analyses are performed as required. Samples are also sent to commercial laboratories for analysis which cannot be performed by our laboratory.

2.2.10 Data Acquisition and Reduction Systems

Instrumentation and associated data acquisition equipment includes: thermocouples with multipoint strip chart recording and digital data logging for main air, steam and tube bundle water flow orifices: Westinghouse, Taylor and Thermox oxygen analyzers with recorders; and flue gas analyzers listed previously with strip chart recorders. Solids feed rates are determined from load cell readings recorded on a digital data logger. The auxiliary air and gas flows and operating pressures are observed on manometers and logged on data sheets by laboratory personnel as required. A time sharing computer and an HP-97 programmable calculator have been used to reduce the time required for data handling and reduction after the data has been collected.

2.3 Test Results

Test programs included fly ash reinjection tests, fly ash characterization tests, limestone tests, coal feeder tests, PDU performance characterization tests, air distributor tests, and corrosion/erosion studies. No hot tests were conducted from August 27, 1977 to December 28, 1977 due to the replacement of FBM by PDU. Hot tests conducted during the reporting period were as follows:

Test No.	Date	Description	Duration/Hrs.
650-12	July 5, 1977	Simultaneous coal and fly ash test, closed loop	6
650-13	July 11, 1977	Same as above	5
650-14	July 14, 1977	Same as above	6
651-1	July 15, 1977	Transient response test	6
651-2	July 19, 1977	Same as above	9
651-3	July 29, 1977	Same as above	9
652-1	July 21, 1977	Fly ash characterization test	6
652-2	July 22, 1977	Same as above	4
652-3	July 25, 1977	Same as above	6
653-1	Aug. 2, 1977	Continuous wet miniball test	25 1/2
653-2	Aug. 11, 1977	Batch dried miniball test	2 1/2
653-3	Aug. 12, 1977	Continuous feed dried miniball test	9
654-0	Aug. 16, 1977	Eductor test	1
654-1	Aus. 17, 1977	Same as above	7
654-2	--	Number not used	--
654-3	Sept. 21, 1977	Eductor test	3
654-4	Sept. 22, 1977	Same as above	4
654-5	Sept. 27, 1977	Eductor test using air and steam	4
654-6	Sept. 28, 1977	Same as above	5

Test No.	Date	Description	Duration/Hrs
656-1	Aug. 26, 1977	Eductor test with 1/2 inch down coal	6
PDU-1	Dec. 29, 1977	Shakedown	6
PDU-2	Jan. 10, 1978	Same as above	5 1/2
PDU-3	Jan. 23, 1978	Same as above	5
PDU-4	Jan. 31, 1978	Same as above	5 1/2
700-1	Feb. 22, 1978	Air distributor test	4 1/2
700-2	Feb. 27, 1978	Same as above	5
700-3	Fe. 28, 1978	Same as above	6 1/2
700-4	Mar. 15, 1978	Air distributor test with rocks and coal feeder test	7
700-5	Mar. 21, 1978	Air distributor base line and coal feeder test	6
700-6	Mar. 22, 1978	Sulfur capture test	7
700-7	Mar. 28, 1978	Coal feeder test with gas sample traverse	8
700-8	Mar. 30, 1978	Same as above	8
700-9	Apr. 6, 1978	Same as above	8 1/2
700-10	Apr. 18, 1978	Sulfur capture test with gas sample traverse	17
700-11	Apr. 20, 1978	Coal feed test with gas sample traverse	8
700-12	Apr. 27, 1978	Air distributor test with porous refractory plates	8
700-13	May 2, 1978	Same as above	7
700-14	May 4, 1978	Same as above	7 1/2
700-15	May 16, 1978	Same as above	5
700-16	May 18, 1978	Same as above	5 1/2
700-17	May 23, 1978	Load following test, no air distributor protection	7
700-18	May 31, 1978	Air distributor test with alumina stones	6
700-19	June 8, 1978	Material and energy balance	16
700-20	June 13, 1978	Turndown	7
700-21	June 22, 1978	Air distributor test with 3/4 inch alumina spheres	5
700-22	June 27, 1978	Same as above	5

2.3.1 Fly Ash Reinjection Tests

The test program for fly ash reinjection was continued to study the effectiveness of burning coal and fly ash simultaneously at the lower end of the effective CBC operating temperature range, where sulfur capture is near optimum. Test results would also be applicable to industrial FBC technology is which elimination of the CBC would have economic advantages.

Both open cycle and closed cycle tests were conducted with the closed cycle tests being conducted during the reporting period. The open cycle tests were conducted to

determine the once-through carbon burnup efficiency at 1600°F bed temperature. Fly ash used in these tests was obtained from previous FBM operation. In Test No. 650-4, limestone feed and bed removal rates were zero as opposed to Test No. 650-2 in which the limestone feed and bed removal rates were both normal. In both tests, data was collected for steady state periods when coal alone was fed and when fly ash and coal were fed simultaneously. The data obtained from these tests is summarized in Table 3.

The objective of these tests was to determine the percent carbon burnup of the fly ash being fed. In order to make this determination, chemical analyses, which show the percent carbon in the various materials, were performed. The results of these analyses are shown in Tables 4 and 5. Table 6 shows burnup of the fly ash, based on carbon analysis in each test and the formula used to calculate this quantity. In making the calculations, it was assumed that the amount of fly ash produced from the coal is the same whether or not coal or coal plus fly ash is burned. Since the coal feed rates were slightly less when coal and fly ash were fed than when coal alone was fed, the fly ash produced from the coal when coal and fly ash were fed together was calculated by multiplying the reference fly ash produced from burning coal only by a coal feed rate ratio. This feed rate ratio is the coal feed rate when coal and fly ash are burned, divided by the reference coal feed rate when coal alone was burned. Each steady state value of the fly ash produced by burning coal alone was taken as the reference. In Test No. 650-2, two reference points and two combined coal and fly ash points were taken for each test, thus all four combinations of reference points and combined points were used to calculate the percent burnup. In Test No. 650-4, only one steady state point was taken during the fly ash feeding and two reference points were compared.

The results show that in Test Nos. 650-2 and 650-4, the values of the average percent burnup based on carbon analysis data were 60.9 percent and 60.1 percent respectively. These results indicate that if the fly ash were recycled to the main cell, three passes would be required to achieve 95 percent carbon burnup of the fly ash.

Test Nos. 650-6 through 650-8 were closed cycle tests where the fly ash produced was recycled to the FBM. In Test Nos. 650-6 and 650-8, all of the fly ash collected in the cyclone was recycled to the FBM. In Test No. 650-7, some of the fly ash was bled off. Table 7 summarizes operating conditions for these tests. Chemical analyses

Table 3

Test Data for Tests 650-2 and 650-4

Test No.	650-2				650-4		
Test Date	5-23-77				5-26-77		
Time	11:30	12:30	13:30	14:00	11:30	13:30	14:15
Coal Feed Rate, lb/hr	593	591	579	577	546	535	530
Limestone Feed Rate, lb/hr	300	293	293	293	0	0	0
High Carbon Fly Ash Feed Rate, lb/hr	0	0	427	427	0	318	332
Low Carbon Fly Ash Collection Rate, lb/hr	170	165	320	340	160	300	300
Bed Removal Rate, lb/hr	175	180	215	240	0	0	0
Dust Loading, lb/hr	19.0	20.9	21.1	30.0	16.93	26.32	18.19
Bed Temp., °F	1620	1610	1570	1570	1630	1625	1570
Superficial Velocity, ft/sec	11.0	10.3	10.0	9.9	9.5	9.0	8.7
Bed Depth (static) in.	19.9	19.7	19.7	19.5	19.6	20.1	20.2
% Excess O_2	1.6	2.2	2.1	2.0	3.2	1.0	1.3

Table 4

Chemical Analysis of Solid Streams
for Test 650-2

MATERIAL	TIME Hr.	C %	H %	LOI %	ASH %
Coal		67.58	5.43	85.12	14.88
Bed	11:30	0.33	0.14	0.69	99.31
	12:30	0.49	0.07	0.74	99.26
	13:30	0.82	0.23	1.28	98.74
	14:00	1.67	0.13	0.98	99.03
Fly Ash In	13:30	33.84	0.93	36.62	63.38
	14.00	35.08	1.01	38.62	61.39
Fly Ash Out	11:30	46.51	0.89	50.82	49.18
	12:30	46.45	0.51	48.94	51.06
	13:30	40.56	0.66	44.23	55.78
	14:00	37.53	0.81	42.38	57.62
Dust	11:30	16.80	0.58	20.38	79.63
	12:30	15:13	0.44	19.48	80.53
	13:30	15:66	0.98	20.29	79.71
	14:00	22.32	0.63	24.80	75.21

Table 5

Chemical Analysis of Solid Streams
for Test 650-4

MATERIAL	TIME HR.	C %	H %	LOI %	ASH %
Coal		64.34	4.36	80.48	19.52
Bed	11:30	0.2662	0.3616	--	>100
	13:30	0.4912	0.3916	0.81	99.19
	14:15	1.1810	0.1669		
				0.68	99.32
Fly Ash In	13:30	41.89	0.2037	47.03	52.97
	14:15	37.00	2.0340	40.85	59.15
Fly Ash Out	11:30	52.16	0.9464	58.75	41.25
	13:30	43.34	1.1260	46.24	53.78
	14:15	44.35	0.7586	48.96	51.04
Dust	11:30	20.13	0.5160	22.79	77.21
	13:30	20.73	0.5480	23.27	76.73
	14:15	16.66	0.6223	18.99	81.01

Table 6

Calculated Fly Ash Burnup

Test No.	650-2				650-4	
Coal Reference Time	11:30	11:30	12:30	12:30	11:30	11:30
Coal and Fly Ash Reference Time	14:00	13:30	14:00	13:30	14:15	13:30
% Burnup Based on Carbon Analysis	61.3	61.4	60.0	60.7	57.8	62.33

$$\text{Burnup of Fly Ash \%} = \Big[W_{FAin}C_{FAin}(W_{Fao}-W_{FAcoal}C_{FAcoal})-(W_{Do}C_{Do} -W_{Dcoal}C_{Dcoal})-(W_{Bo}C_{Bo}-W_{Bcoal}C_{Bcoal})\Big] \Big/ W_{FAin}C_{FAin}$$

where:

W_{FAin} is the fly ash feed rate.

C_{FAin} is the weight fraction of carbon of fly ash feed.

W_{FAo} is the fly ash collection rate from burning coal and fly ash.

C_{FAo} is the fly weight fraction of carbon of fly ash collected from burning coal and fly ash.

W_{FAcoal} is the fly ash produced from coal when burning coal and fly ash.*

C_{FAcoal} is the weight fraction of carbon of fly ash collected by burning coal alone.

W_{Do} is the dust collection rate from burning coal and fly ash.

C_{Do} is the weight fraction of carbon of the dust when burning coal and fly ash.

W_{Dcoal} is the dust collected from coal when burning coal and fly ash.*

C_{Ccoal} is the weight fraction of carbon of the dust from burning coal alone.

W_{Bo} is the bed removal rate when burning coal and fly ash.

C_{Bo} is the weight fraction of carbon of the bed when burning coal and fly ash.

W_{Bcoal} is the bed removal rate of coal when burning coal and fly ash.*

C_{Bcoal} is the weight fraction of carbon of the bed when burning coal alone.

* The feed removal rate of a substance time the ratio of the coal feed rate when burning coal and fly ash to the coal feed rate method and results of burning of coal alone.

Table 7

Summary of Closed Cycle Fly Ash Reinjection Tests

Test No.	650-6		650-7				650-8
Time, Hr.	1415	1500	0930	1115	1230	1400	1115
Bed Temp., ^{o}F	1610	1625	1595	1580	1570	1550	1635
O_2 in flue gas, %	2.3	1.7	1.3	0.5	1.0	1.5	2.8
Coal feed rate, lb/hr	499	486	520	503	488	503	473
Fly ash collection rate, lb/hr	0	0	143	113	90	90	0
Limestone feed rate, lb/hr	0	0	288	288	173	173	266
Stack dust rate, lb/hr	27.1	28.1	19.9	32.8	31.3	28.7	37
Main air flow rate, lb/hr	4358	4350	4464	4430	4415	4415	4369
Total air flow rate, lb/hr	4893	4887	5012	4918	4956	4957	4916
Superficial velocity, ft/sec	8.4	8.5	8.6	8.5	8.4	8.3	8.6
Static bed depth, inches	18.9	18.9	19.4	21.9	21.5	21.5	15.6
EOBH (effective operating bed ht.) inches	20	20	21.5	22.5	21.0	21.0	20
SO_2, ppm	2545	3595	580	440	620	630	1285
Combustion Efficiency, %	95.7	0	79.3	82.0	83.9	83.0	91.8

of these tests are shown in Tables 8, 9 and 10. It is seen from the data for Test No. 650-6 that the combustion efficiency was greater than 95 percent.

After Test No. 650-6, it was noticed that fly ash had accumulated in the dropout hopper located below the air preheater. Normally, a screw feeder located at the bottom of this dropout hopper transports this fly ash to the cyclone where it can be sent to the fly ash weigh hopper.

Apparently, with increased fly ash loading that occurs during fly ash recycle, more fly ash was collecting in the preheater dropout hopper than could be sent to the cyclone by the transport screw. To overcome this problem, the speed of the transport screw was increased, thus increasing the fly ash transport rate out of the dropout hopper.

During Test Nos. 650-6 and 650-8, it was observed that the feed rate of recycled fly ash (as indicated by the pressure in the primary injection vessel) reached an equilibrium value. This indicated that in a total recycle situation, a steady state could be achieved where the amount of recycled fly ash burned equals the amount of fly ash produced.

In order to examine this observation more closely, it was first necessary to calibrate fly ash delivery of the Petrocarb system. The next series of tests, Test No. 650-9 through 650-11, were run in the open cycle so that the Petrocarb system's operational parameters could be calibrated. Figure 18 shows the result of this calibration.

During Test Nos. 650-9 and 650-11, it was noticed that fly ash continued to collect in the preheater dropout hopper when the fly ash feed rates were high. The speed of the dropout hopper screw which removes the fly ash from the hopper could not be increased any further. In order to prevent further accumulation of fly ash in the hopper when steady state operating data was being collected, it was decided to let the dropout hopper fill up with fly ash. During Test No. 650-11, the dropout hopper screw was turned off and the fly ash was allowed to collect in the hopper.

Three additional tests, Nos. 650-12, 650-13 and 650-14 were conducted in the closed loop mode. The purpose of these tests was to examine the combustion efficiencies at 1550°F, 1600°F and 1650°F with complete fly ash recycle.

Table 8

Chemical Analysis of Solid Samples

Test 650-6

MATERIAL	TIME Hr.	C %	H %	LOI* %	ASH %	Ca %	S %
Bed	14:15	.6064	.1554	.83	99.17	5.148	2.526
Dust	14:10	26.33	.4046	31.27	68.73	31.41	11.92
Fly Ash Out	14:15	52.98	.5745	57.80	42.20	6.535	2.476

*LOI = Loss on ignition

Table 9

Chemical Analysis of Solid Samples

Test 650-7

Fly Ash Reinjection Test

MATERIAL	TIME Hr.	C %	H %	LOI %	ASH %	Ca %	S %
Bed	09:30	.9233	.2608	1.37	98.63	7.8	6.382
	11:15	1.168	.6562	1.56	98.44	12.8	6.362
	12:00	.6802	.0836	2.47	97.53	12.85	6.812
	14:00	1.709	.6464	3.43	96.57	16.90	7.066
Dust	09:30	17.15	.4364	24.85	75.15	9.70	2.556
	11:15	24.45	.6117	30.34	69.66	7.55	2.611
	12:30	20.56	.6806	33.44	66.56	4.70	2.100
	14:00	28.25	.9251	31.76	68.64	6.65	1.948
Fly Ash	09:30	43.46	.9805	43.16	56.84	16.4	2.393
	11:15	39.56	.3531	43.31	56.69	11.85	2.269
	12:30	48.16	.8816	46.24	53.76	7.0	2.485
	14:00	44.06	.6417	49.24	50.76	5.25	2.395

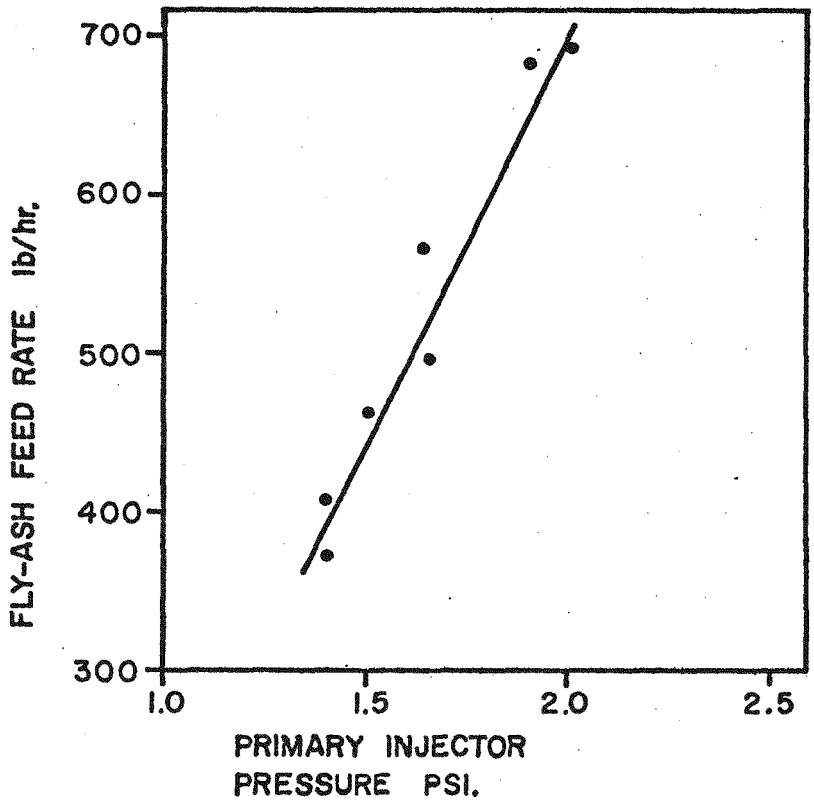
Table 10

Chemical Analysis of Solid Samples

Test 650-8

MATERIAL	TIME Hr.	C %	H %	LOI %	ASH %	Ca %	S %
Bed	11:15	.9195	.4990	2.48	97.52	36.22	8.299
Dust	11:15	28.85	4.587	29.90	70.10	14.43	2.436
Fly Ash	11:15	43.19	1.536	48.56	51.44	2.881	6.436

CALIBRATION OF PETROCARB FLY-ASH INJECTOR



NOTE
Feeders 1 and 2 to FBM
Feeders 3 through 6 off

Figure 18

Test No. 650-12 - Bed temperatures for this test ranged from 1630°F to 1680°F. Table 11 lists operating conditions for the four steady state analyses periods studied during this test. Flue gas analysis and chemical analysis of this test are shown in Tables 12 and 13 respectively. The most noticeable difference in the four time periods is the decrease in the amount of CO and hydrocarbons in the off-gas at 14:30 hrs. During this time interval, the coal feed rate was decreased resulting in higher stack oxygen and a lower bed temperature. Combustion efficiencies greater than 95 percent were achieved.

Test No. 650-13 - Bed temperatures for this test ranged from 1545°F to 1570°F which represents normal FBM operating conditions. During this test, as in the two others, the rate of fly ash reinjected reached a steady state level as indicated by the primary injection pressure of the Petrocarb system. However, the primary injection pressure never rose above 1 psi, which is below the range of calibration. The operating conditions for this test are listed in Table 14. Flue gas analysis and chemical analysis for this test are shown in Tables 15 and 16 respectively. The most notable fact from this test is the low amount of hydrocarbons and CO in the flue gas despite the relatively shallow bed depths. This is probably due to the high excess oxygen level. Again, combustion efficiencies greater than 95 percent were attained.

Test No. 650-14 - Temperatures for this test ranged from 1610°F to 1620°F. Table 17 shows the operating conditions for the three steady state periods and the calculated combustion efficiencies obtained in this test. Table 18 lists the flue gas analysis and Table 19 lists the solid sample analysis. As can be seen in Table 17, combustion efficiencies as high as 95 percent were again obtained with fly ash recycling.

In order to confirm the existence of the steady state recycle rate of fly ash, samples of the fly ash were taken and analyzed every half hour during this test. The analyses of these fly ash samples is also shown in Table 19. The analyses show no trend in either the carbon content or the ash content of the fly ash. This indicates that a steady state was achieved rather quickly during the test.

The results of the fly ash reinjection tests show that high combustion efficiencies can be achieved with fly ash recycle. Combustion efficiencies of 96 percent appear to be possible. The steady state fly ash recirculation rate achieved during the tests implies that some fly ash remains in the bed. This could be possible if during the combustion process the fly ash particles heat-up to above the ash

Table 11

Test 650-12

Fly Ash Reinjection Test

Operating Conditions

Time; Hr.	10:30	11:30	13:00	14:30
Bed. Temp., °F	1670	1675	1680	1630
O_2 in flue gas, %	3.0	2.6	2.8	4.2
Coal feed rate, lb/hr	527	513	520	460
Fly ash collection rate, lb/hr	0	0	0	0
Limestone feed rate, lb/hr	282	282	282	260
Stack dust rate, lb/hr	15.7	15.7	29.2	25.4
Main air flow rate, lb/hr	4991	4879	4925	4964
Total air flow rate, lb/hr	5554	5445	5491	5527
Superficial velocity, ft/sec	9.8	9.7	9.8	9.6
Static bed depth, inches	13.8	13.6	13.6	13.7
EOBH (effective operating bed ht.) inches	20	20	20	20
SO_2, ppm	1703	1188	1628	778
Bed drain rate, lb/hr	70	118	118	118
Combustion efficiency, %	96.1	95.9	95.8	97.4

Table 12

Test 650-12

Flue Gas Analysis

Time; Hr.	10:30	11:30	13:00	14:30
CO, %	0.3	0.26	0.27	0.08
HC, ppm	1928	1538	1448	428
CO_2, %	13.7	14.0	13.9	13.2
NO, ppm	← Data Unavailable →			
NO_x, ppm	295	340	325	350
SO_2, ppm	1703	1188	1628	778
O_2, %	3.0	2.6	2.8	4.2
Moisture, wt %	10.86	10.86	10.9	10.95

Table 13

Chemical Analysis of Solids Samples
Test 650-12
Fly Ash Reinjection Test

Material	Time, hr	C, %	H, %	LOI*, %	Ash, %	Ca, %	S, %
Coal	10:17	69.7	4.96	85.5	14.5	6.2	3.5
Bed	10:30	0.4	0.23	2.0	98.0	35.0	8.3
	11:30	1.3	0.51	2.5	97.5	36.9	7.8
	12:30	0.0	0.53	2.9	97.1	39.3	7.0
	14:30	0.3	0.34	2.3	97.7	40.8	6.9
Dust	10:32	13.7	0.70	17.1	82.9	24.5	2.8
	11:30	14.7	0.71	17.8	82.2	17.5	2.6
	13:00	13.7	1.04	18.0	82.0	18.7	2.8
	14:35	15.6	1.16	21.2	78.8	19.6	2.3
Fly Ash Recirculated	10:00	50.3	1.31	54.0	46.0	10.4	2.5
	10:30	50.0	1.08	55.3	44.7	8.1	3.6
	11:30	49.2	2.01	49.3	50.7	16.0	2.8
	12:00	44.0	1.02	47.1	52.9	22.4	2.6
	12:30	43.1	0.31	47.4	52.6	13.1	2.3
	13:00	41.8	0.72	45.9	54.1	20.2	2.5
	13:30	40.6	0.96	44.9	55.1	--	2.8
	14:00	41.6	1.09	45.7	54.3	19.5	2.8
	14:30	44.7	1.12	48.4	51.6	11.8	2.6

* Loss on ignition.

Table 14

Test 650-13

Fly Ash Reinjection Test
Operating Conditions

Time, hr	10:00	13:00
Bed Temp., °F	1545	1570
O_2 in flue gas, %	7.2	6.7
Coal feed rate, lb/hr	443	444
Fly ash collection rate, lb/hr	0	0
Limestone feed rate, lb/hr	291	284
Stack dust rate, lb/hr	20.7	28.4
Main air flow rate, lb/hr	5009	4989
Total air flow rate, lb/hr	5566	5548
Superficial velocity, ft/sec	9.4	9.3
Static bed depth, inches	12.7	12.6
EOBH (effective operating bed ht.) inches	14.5	14.0
SO_2, ppm	420	490
Bed drain rate, lb/hr	174	169
Combustion efficiency, %	96.4	95.8

Table 15

Test 650-13

Flue Gas Analysis

Time; Hr.	10:00	13:00
CO, %	0.142	0.112
HC, ppm	143	63
CO_2, %	11.1	11.6
NO, ppm	350	360
NO_x, ppm	318	350
SO_2, ppm	420	490
O_2, %	7.2	6.7
Moisture, wt %	9.35	10.11

Table 16

Chemical Analysis of Solid Samples
Test 650-13
Fly Ash Reinjection Test

Material	Time, hr	C, %	H, %	LOI*, %	Ash, %	Ca, %	S, %
Coal	08:15	71.0	5.36	83.9	16.1	3.2	4.8
	11:30	69.1	5.08	83.1	17.0	17.3	4.3
Bed	08:45	0.6	0.20	0.0	>100	47.6	7.4
	11:00	0.9	0.34	0.0	>100	38.9	8.3
	11:30	0.3	0.10	0.0	>100	46.5	8.6
	12:30	1.2	0.11	0.0	>100	67.5	10.3
Dust	08:47	12.9	0.20	13.0	87.0	19.0	3.3
	11:00	22.1	0.46	24.4	75.6	16.4	2.8
	11:30	23.4	0.65	26.2	73.9	28.1	3.0
	12.35	22.7	0.55	25.7	74.3	18.8	2.6
Fly Ash Recirculated	08:00	44.0	0.96	47.5	52.5	--	3.0
	08:30	45.4	0.83	50.2	49.8	--	2.9
	08:45	46.9	1.81	49.8	50.2	13.5	2.6
	09:00	43.6	1.09	44.0	56.0	--	2.8
	09:30	40.0	0.85	40.6	59.5	--	2.4
	10.00	42.6	0.90	45.0	55.0	--	2.9
	10:30	43.8	1.40	44.4	55.6	--	3.0
	11:00	44.2	0.97	45.2	54.8	13.3	2.9
	11:30	41.1	0.87	47.0	53.0	13.6	2.9
	12:00	41.4	1.09	44.5	55.5	--	3.2
	12:30	41.2	0.74	46.8	53.2	14.5	2.9
	13:00	38.0	0.16	41.7	58.3	--	4.0
	13:30	40.3	0.88	44.6	55.4	--	3.3

* Loss on ignition.

Table 17

Test 650-14

Operating Conditions

Time; Hr.	11:00	11:30	12:30
Bed Temp., °F	1610	1615	1620
O_2 in flue gas, %	2.5	1.9	1.8
Coal feed rate, lb/hr	578	578	550
Fly ash collection rate, lb/hr	0	0	0
Limestone feed rate, lb/hr	287	287	285
Stack dust rate, lb/hr	35.4	37.3	40.3
Main air flow rate, lb/hr	4858	4871	4961
Total air flow rate, lb/hr	5405	5412	5505
Superficial velocity, ft/sec	9.3	9.3	9.5
Static bed depth, inches	18.4	18.7	18.5
EOBH (effective operating bed ht.), inches	20	20.5	20.5
SO_2, ppm	668	1045	1050
Bed drain rate	165	150	150
Combustion efficiency, %	95.2	95.1	93.8

Table 18

Test 650-14

Flue Gas Analysis

Time ; Hr.	11:00	11:30	12:30
CO, %	0.337	0.417	0.397
HC, ppm	1350	1425	2107
CO_2, %	13.7	14.2	14.3
NO, ppm	280	280	290
NO_x, ppm	135	150	155
SO_2, ppm	668	1045	1050
O_2, %	2.5	1.9	1.8
Moisture, wt %	10.3	10.2	10.2

Chemical Analysis of Solid Streams
For Test 650-14

Material	Time Hr.	C %	H %	L.O.I* %	Ash %	Ca %	S %
Coal	08:15	70.96	5.36	83.94	16.06	3.15	4.84
	11:30	69.12	5.08	83.05	16.95		4.31
Bed	08.45	0.648	0.198	--	∿100.0	47.60	7.38
	11:00	0.855	0.343	--	∿100.0	38.92	8.35
	11:30	0.262	0.102	--	∿100.0	46.5	8.61
	12:30	1.218	0.111	--	∿100.0		10.32
Dust	08.45	12.90	.198	13.04	86.96	19.00	3.27
	11:00	22.06	.455	24.38	75.62	16.37	2.82
	11:30	23.37	.651	26.15	73.85		3.01
	12:35	22.72	.550	25.66	74.34	18.80	2.65
Fly Ash Recirculated	08:00	44.02	0.96	47.48	52.52		3.02
	08:30	45.40	0.83	50.21	49.79		2.85
	08:45	46.93	1.81	49.78	50.22	13.50	2.57
	09:00	43.63	1.09	44.00	56.00		2.79
	09:30	40.00	0.85	40.55	59.45		2.38
	10:00	42.62	0.90	45.00	55.00		2.93
	10:30	43.83	1.40	44.37	55.63		2.98
	11:00	44.22	0.97	45.16	54.84	13.27	2.91
	11:30	41.07	0.87	46.97	53.03	13.56	2.94
	12:00	41.43	1.09	44.50	55.50		3.22
	12:30	41.20	0.74	46.82	53.18	14.53	2.87
	13:00	38.00	0.16	41.73	58.27		4.02
	13:30	40.29	0.87	44.60	55.40		3.31

*Loss on Ignition

fusion point and became sticky. The sticky particles could then adhere to the cooler limestone particles in the bed. With the high combustion efficiencies achieved, it appears that fly ash recycle could be a way of eliminating the Carbon-Burnup Cell for small industrial boilers. However, the combustion efficiencies currently achieved with fly ash recycle are not sufficiently high to eliminate the CBC in utility size boilers.

2.3.2 Fly Ash Characterization Tests

The purpose of the test program was to provide information on the chemical analysis of particles entrained in the flue gas as a function of particle size. Since future AFB units will be equipped with dust collecting equipment having collection efficiencies different from the FBM cyclone, the size distribution of the fly ash collected for reinjection in these units could differ from the typical FBM fly ash. A difference in chemical composition of fly ash with respect to particle size will lead to different bulk analyses in fly ash of different size distributions. Therefore, the information generated from this test program will be useful in predicting the bulk chemical analysis of fly ash collected at different collection efficiencies.

Three fly ash characterization tests were conducted. Each test was conducted at the same normal operating conditions for high sulfur coal except for variations in the superficial velocity and bed depth. Test Nos. 651-1, 652-2 and 652-3 were conducted at approximately 10, 8 and 12 ft/sec respectively. Bed depths were adjusted for different superficial velocities in order to maintain a bed temperature of about 1500°F for each test with excess oxygen in the range of 3 to 5 percent. For each test, Sewickely coal of 4.5 percent sulfur content was the fuel and Greer limestone was the sulfur sorbent. The tests were conducted with a calcium to sulfur molar feed ratio of from 3.0 to 3.5 in order to maintain SO_2 emission at or below EPA regulation levels. A summary of test conditions is shown in Table 20.

For each test, fly ash samples were screened using Nos. 16, 30, 40, 70, 100, 140, 200 and 400 screen sizes and the dust samples were screened using Nos. 200, 325 and 400 screen sizes. The size distribution for coal, limestone, bed, fly ash and dust for Test Nos. 652-1, 652-2 and 652-3 are shown in Tables 21, 22 and 23. Each sample was size separated prior to chemical analysis and each size cut was individually analyzed for carbon, hydrogen, calcium, sulfur and loss on ignition. The analyses are presented on weight percent basis and the "as received" condition. These analyses are shown in Tables 24, 25 and 26.

Table 20

Operating Conditions of the Fly Ash Characterization Tests

Test No.	652-1		652-2		652-3	
Time	10:30	11:30	10:15	11:45	10:30	11:15
Bed Temp., °F	1480	1500	1500	1540	1515	1560
O_2 in flue gas, %	4.9	4.0	3.2	2.9	4.7	4.0
Coal feed rate, lb/hr	595	555	553	553	705	660
Fly ash collection rate, lb/hr	155	161	130	130	231	231
Limestone feed rate, lb/hr	333	333	273	273	444	444
Stack dust rate, lb/hr	25.1	25.6	39.7	39.7	38.9	41.1
Main air flow rate, lb/hr	5209	5194	4222	4209	6744	6744
Total air flow rate, lb/hr	6049	6032	5059	5939	7572	7577
Superficial velocity, ft/sec	9.3	9.4	7.8	7.9	11.2	10.3
Static bed depth, inches	23.5	23.0	19.5	19.5	24.0	24.0
SO_2, PPM	145	205	450	550	275	290
Ca/S	2.52	2.70	2.22	2.22	2.84	3.03
Combustion Eff.	76.2	73.3	79.8	79.6	75.8	74.1
lb $SO_2/10^6$ BTU	0.295	0.386	0.828	1.008	0.588	0.660

Table 21

Size Distribution of Solid Samples, Test 652-1
Fly Ash Characterization Test

% Smaller than stated size.

Test 652-1 @ 11:30 hr.

No.**	Coal	Limestone	Bed	Fly Ash	Dust
1/4"	100		100		
4	97.49	100	98.95		
8	86.97	98.16	95.87		
14	75.84	50.06	63.14		
16				100.00	
18	65.30	12.47	25.73		
25	55.77	8.14	14.00		
30				99.90	
40	41.82	6.39	4.47	99.70	
50	33.69	5.72	.78		
70	26.6	5.16	.21	84.83	
100				61.08	100.00
140				46.45	
200				32.57	99.83
325					97.63
400				10.67	96.13

** U.S. Sieve Series, except when marked otherwise.

Table 22

Size Distribution of Solid Samples, Test 652-2
Fly Ash Characterization Test

% Smaller than stated size.

Test 652-2 @ 10:15 hr.

No.**	Bed	Fly Ash	Dust
1/4"	100		
4	99.46		
8	96.52		
14	67.60	100	
16		98.88	
18	31.19		
25	18.58		
30		97.83	
40	7.43	97.22	
50	2.24		
70	0.54	77.11	
100		51.75	100
140		38.85	
200		25.64	97.96
325			64.58
400		8.22	28.25

** U. S. Sieve Series, except when marked otherwise.

Table 23

Size Distribution of Solid Samples, Test 652-3

% Smaller than stated size.

Test 652-3 @ 10:30 hr.

No.**	Bed	Fly Ash	Dust
1/4"	100.00		
4	99.30		
8	96.21		
14	66.22	100.00	
16		99.96	
18	27.02		
25	13.82		
30		99.62	
40	2.92	97.52	
50	.25		
70	.08	70.67	
100		46.04	100.00
140		35.49	99.80
200		25.31	96.25
325			90.33
400		9.66	59.86

** U.S. Sieve Series, except when marked otherwise.

Table 24

Chemical Analysis of Solid Samples for Test 652-1
Weight Percent of Component in Each Size Cut

Material	C,%	H,%	Ca,%	S,%	Ash,%	LOI,%
Bed						
+4	4.9	0.88	19.7	1.2	84.1	15.9
-4+8	2.2	0.92	15.6	2.5	94.4	5.6
-8+14	1.1	0.57	28.9	6.4	93.7	6.3
-14+18	3.0	0.79	35.9	8.3	95.6	4.4
-18+25	1.0	0.71	33.1	9.6	94.5	5.5
-25+40	1.4	0.57	33.3	10.3	96.6	3.4
-40+50	0.5	0.24	30.4	9.8	95.8	4.2
-50+70	4.3	0.70	28.8	10.1	92.8	7.2
-70	16.1	1.07	22.3	6.6	76.6	23.4
Fly Ash						
-16+30	20.7	1.51	20.8	5.0	--	--
-30+40	29.9	1.09	7.1	1.5	60.6	39.4
-40+70	44.9	1.20	10.2	2.9	48.8	51.3
-70+100	51.0	1.30	11.2	2.7	41.4	58.6
-100+140	53.2	3.50	9.0	2.6	43.0	57.0
-140+200	47.3	1.62	11.7	3.0	45.2	54.8
-200+400	41.5	1.35	8.7	3.1	51.0	49.0
-400	27.1	0.86	15.7	3.5	68.1	31.9
Dust						
+200	18.8	0.83	15.9	4.4	--	--
-200+325	48.6	1.87	5.9	3.0	38.8	61.2
-325+400	42.9	1.69	9.9	2.5	47.6	52.4
-400	22.0	0.88	16.6	8.8	74.1	25.9

Table 25

Chemical Analysis of Solid Samples for Test 652-2
Weight Percent of Component in Each Size Cut

Material	C%	H%	Ca%	S%	Ash%	LOI%
Bed						
+ 4	4.35	0.45	8.8	2.1	92.4	7.6
- 4+ 8	4.55	0.56	10.2	4.8	95.1	4.9
- 8+14	0.88	0.37	33.4	6.8	93.3	6.7
-14+18	0.39	0.50	34.8	8.4	94.3	5.7
-18+25	0.45	0.95	31.4	9.6	98.4	1.6
-25+40	0.56	0.80	30.1	10.0	95.8	4.2
-40+50	0.49	0.46	27.7	10.5	96.3	3.7
-50+70	1.45	0.48	30.7	10.7	94.5	5.5
-70	15.12	0.78	21.3	3.8	71.3	28.7
Fly Ash						
+16	1.6	0.55	24.2	6.6	93.2	6.8
-16+30	4.7	0.76	29.7	7.4	90.7	9.3
-30+40	30.5	1.28	11.8	3.7	62.8	37.2
-40+70	52.4	1.70	6.6	2.2	40.9	59.1
-70+100	50.8	1.26	7.6	2.5	41.0	59.0
-100+140	47.4	1.30	12.4	2.8	44.2	55.8
-140+200	45.9	1.37	8.2	2.8	47.2	52.8
-200+400	37.0	1.02	7.8	2.9	56.1	43.9
-400	22.8	0.94	12.5	3.2	70.7	29.3
Dust						
+200	26.7	1.43	12.9	2.0	68.9	31.1
-200+325	16.4	0.41	12.6	2.6	79.8	20.2
-325+400	16.7	0.51	12.93	2.4	79.5	20.5
-400	21.7	0.83	14.7	2.5	73.4	26.6

Table 26

Chemical Analysis of Solid Samples for Test 652-3

Weight Percent of Component in Total Sample for Each Size Cut

Material	C %	H %	Ca %	S %	Ash %	LOI %
Bed						
-¼"+4	8.32	0.54	8.998	3.42	90.69	9.31
-4+8	3.67	0.51	14.21	2.66	94.25	5.75
-8+14	0.81	0.64	31.05	6.05	90.69	9.31
-14+18	0.48	0.62	34.90	7.99	92.49	7.51
-18+25	0.22	0.47	33.47	10.37	94.87	5.13
-25+40	0.72	0.23	23.73	9.20	95.06	4.94
-40+50	0.71	0.29	23.40	10.23	95.35	4.65
-50+70	13.83	0.84	24.68	5.60	75.79	24.21
-70	5.80	0.75	29.05	8.48	88.61	11.39
Fly Ash						
-16+30	23.10	0.84	4.989	2.85	70.57	29.43
-30+40	40.71	0.59	15.32	2.91	54.40	45.60
-40+70	42.83	1.14	12.14	3.56	52.20	47.80
-70+100	48.41	0.75	9.736	3.53	44.86	55.14
-100+140	48.20	1.62	9.235	3.31	46.17	53.83
-140+200	44.86	1.35	10.29	3.74	49.34	50.66
-200+400	36.50	1.92	8.298	4.42	58.80	41.20
-400	20.25	0.95	17.32	4.546	76.62	23.38
Dust						
+200	30.16	1.67	9.702	3.65	61.52	38.48
-200+325	18.19	0.10	16.01	2.88	74.23	25.77
-325+400	11.95	0.00	17.18	2.33	80.31	19.69
-400	15.59	1.34	14.85	3.93	80.29	19.71

For the bed material samples, the analyses seem reasonable in most of the size cuts with the exception of the +14 mesh and -70 mesh size cuts in Test Nos. 652-1 and 652-2. In Test No. 652-3, the preceeding holds not only for the +4 mesh and -70 mesh, but in addition for the +70 mesh of the bed sample. Both of these size cuts showed an unusually high loss on ignition and the -70 mesh size cut had an unusually high carbon content. However, both of these cuts represent a very small fraction of the bed material sample and therefore these anomalies are insignificant.

The data in Tables 24, 25 and 26 were further analyzed by weighing the chemical analyses of each size cut by the fraction of material in that size cut. For example, in Table 24 for Test No. 652-1, the amount of carbon contained in the +4 mesh bed material fraction is only 0.05 percent by weight of the total bed sample. These results, shown in Tables 27, 28 and 29 represent the weight percent of the component in the total sample for each size cut. The composite analyses are the sum of the samples for each size cut and represent the bulk sample analysis.

Figure 19 is a plot of the amount of carbon in each screen cut as a percent of the total fly ash for Test Nos. 652-1, 652-2 and 652-3. All three tests show similar trends with the only major differences occurring at 55 and 315 microns. In general, these curves show two peaks, one at 55 microns and the other larger peak at 180 microns. From Figure 19 it can be concluded that in the range of 8 to 11 ft/sec, the superficial velocity has very little effect on the composition of the fly ash produced below 200 microns with the greatest effect being seen in the 300 micron size.

2.3.3 Limestone Tests

Improving Limestone Reactivity

Improved limestone utilization in the AFB process represents one of the primary areas of potential operating cost reductions. Present limestone requirements are between two and four times the stoichiometric calcium requirements. Furthermore, different limestones exhibit wide variations in reactivity, thus rendering some stones unacceptable for use in the FBC process.

The reactivity of any limestone can be improved significantly by pulverizing the stone. This increases the available surface area of the stone thereby increasing the amount of sulfur that can be captured. However, the superficial velocities encountered in the FBC process limit the size of the limestone to the range of 1/8 in by 40 mesh.

Table 27

Chemical Analysis of Solid Samples
Weight Percent of Component in Total Sample for Each Size Cut
Test 652-1

Material	C,%	H,%	Ca,%	S,%	Ash,%	LOI,%
Bed						
+4	.05	.01	.21	.01	.88	.17
-4+8	.07	.03	.48	.08	2.91	.17
-8+14	.36	.19	9.46	2.09	30.67	2.06
-14+18	1.12	.30	13.43	3.11	35.76	1.65
-18+25	.12	.08	3.88	1.13	11.08	.65
-25+40	.13	.05	3.17	.98	9.21	.32
-40+50	.02	.01	1.12	.36	3.54	.15
-50+70	.02	.004	.16	.06	.53	.04
-70	.03	.002	.05	.01	.16	.05
Composite	1.92	.676	31.96	7.83	94.74	5.26
Fly Ash						
-16+30	.02	.002	.02	.005	.05	.05
-30+40	.10	.002	.01	.003	.121	.08
-40+70	6.68	.178	1.52	.43	7.26	7.63
-70+100	12.11	.309	2.66	.64	9.86	13.92
-100+140	7.78	.512	1.32	.38	6.31	8.34
-140+200	6.57	.225	1.62	.42	6.29	7.61
-200+400	9.09	.296	1.91	.68	11.17	10.73
-400	2.89	.092	1.68	.37	7.27	3.40
Composite	45.24	1.62	10.74	2.92	48.33	51.76
Dust						
+200	.10	.004	.09	.02	0.41	.14
-200+325	.89	.034	.11	.05	0.71	1.12
-325+400	.64	.025	.15	.04	0.71	.79
-400	21.15	.846	15.95	8.46	71.23	24.89
Composite	22.78	.909	16.30	8.58	73.06	26.94

Table 28

Chemical Analysis of Solid Samples

Weight Percent of Component in Total Sample For Each Size Cut

Test 652-2

Material	C%	H%	Ca%	S%	Ash%	LOI%
Bed						
+4	.023	.002	.05	.01	.50	.04
-4+8	.134	.016	.30	.14	2.80	.14
-8+14	.254	.107	9.66	1.97	26.98	1.94
-14+18	.142	.182	12.67	3.06	34.33	2.08
-18+25	.057	.120	3.96	1.21	12.40	.20
-25+40	.062	.089	3.36	1.12	10.68	.47
-40+50	.025	.024	1.44	.55	5.00	.19
-50+70	.025	.008	.52	.18	1.61	.09
-70	.082	.004	.12	.02	.39	.16
Composite	.805	.553	32.07	8.25	94.69	5.31
Fly Ash						
+16	.02	.006	.27	.07	1.04	.076
-16+30	.05	.008	.31	.08	.95	.098
-30+40	.19	.008	.07	.02	.38	.227
-40+70	10.54	.342	1.33	.44	8.23	11.89
-70+100	12.87	.320	1.93	.63	10.40	14.96
-100+140	6.12	.168	1.60	.36	5.70	7.20
-140+200	6.06	.181	1.08	.37	6.24	6.98
-200+400	6.45	.178	1.36	.51	9.77	7.65
-400	1.87	.077	1.03	.26	5.81	2.41
Composite	44.16	1.287	8.98	2.75	48.52	51.48
Dust						
+200	.54	.029	.26	.04	1.39	.63
-200+325	5.48	.137	4.21	.87	26.64	6.75
-325+400	6.07	.185	4.70	.87	28.88	7.45
-400	6.13	.234	4.15	.71	20.74	7.51
Composite	18.22	.586	13.32	2.49	77.65	22.35

Table 29

Chemical Analysis of Solid Samples
Weight Percent of Component in Total Sample for Each Size Cut

Test 652-3

Material	C %	H %	Ca %	S %	Ash %	LOI %
Bed						
-¼"+4	.06	0	.06	.02	.63	.07
-4+8	.11	.02	.44	.08	2.91	.18
-8+14	.24	.19	9.31	1.81	27.20	2.79
-14+18	.19	.24	13.68	3.13	36.26	2.94
-18+25	.03	.06	4.42	1.37	12.52	.68
-25+40	.08	.03	2.59	1.00	10.36	.54
-40+50	.02	.01	.62	.27	2.55	.12
-50+70	.02	0	.04	.01	.13	.04
-70	0	0	.02	0	0	0
Composite	.75	.54	31.18	7.69	92.56	7.36
Fly Ash						
-16+30	.08	0	.02	.01	.24	.10
-30+40	.85	.01	.32	.06	1.14	.96
-40+70	11.50	.31	3.26	.96	14.02	12.83
-70+100	11.92	.18	2.40	.87	11.05	13.58
-100+140	5.09	.17	.97	.35	4.87	5.68
-140+200	4.57	.14	1.05	.38	5.02	5.16
-200+400	9.24	.49	2.10	1.12	14.88	10.43
-400	1.96	.09	1.67	.44	7.40	2.26
Composite	45.21	1.39	11.79	1.94	58.62	50.77
Dust						
+200	1.13	.06	.36	.14	2.31	1.44
-200+325	1.08	.01	.95	.17	4.39	1.53
-325+400	3.64	0	5.23	.71	24.47	6.00
-400	4.75	.41	4.52	1.20	24.46	6.01
Composite	10.60	.48	11.06	2.28	55.63	14.98

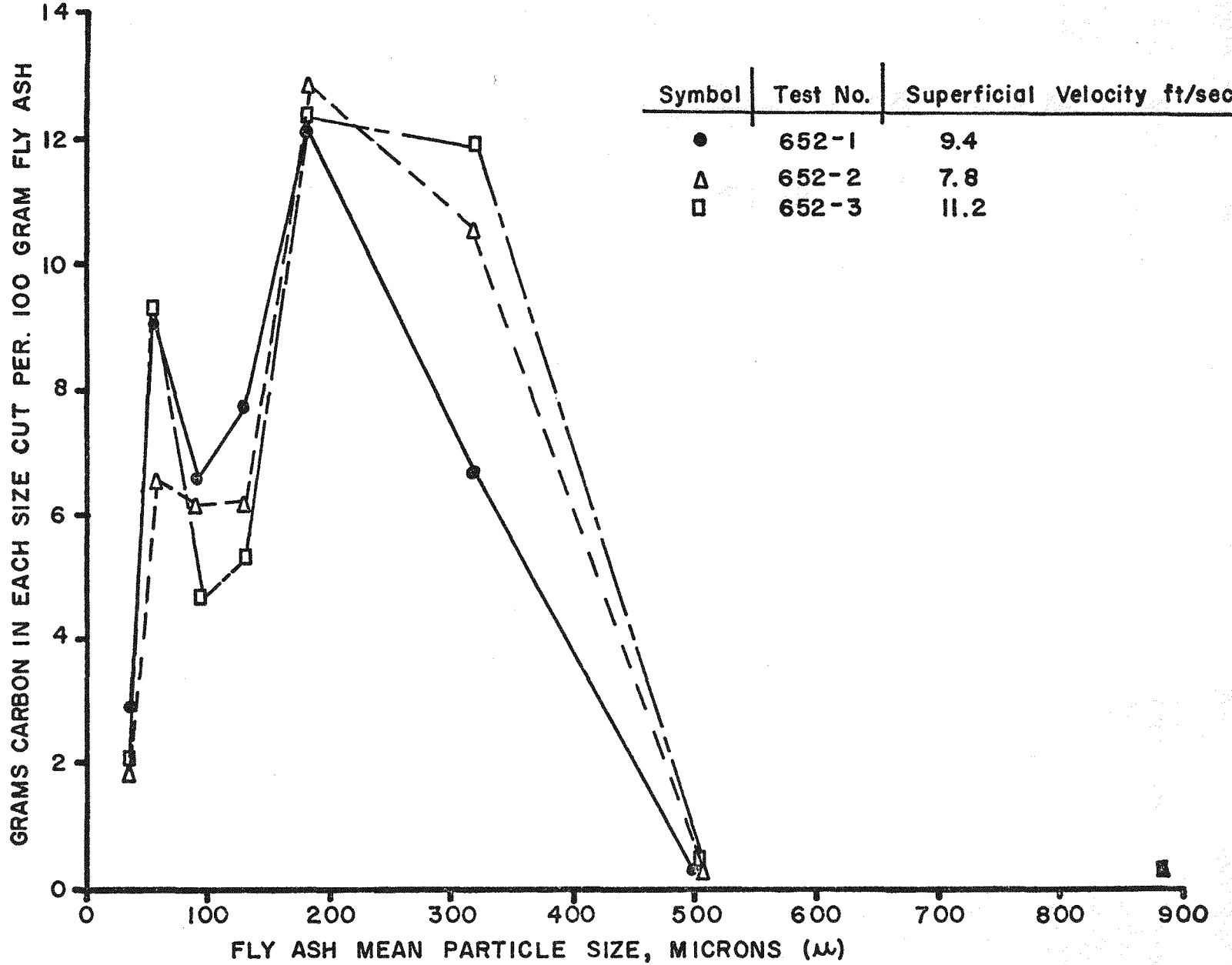


Figure 19

AMOUNT OF CARBON IN EACH SIZE CUT AS A PERCENTAGE OF THE TOTAL FLY ASH VERSUS MEAN PARTICLE SIZE OF EACH SIZE CUT FOR TESTS 652-1, 652-2 AND 652-3.

One way of overcoming these difficulties is to agglomerate the pulverized limestone with a binder to form pellets in the size range required by the FBC process. The agglomerates formed possess a highly porous structure which allow the agglomerates to have a reactivity near that of pulverized stone. Another advantage of agglomeration is the ability to utilize some limestone that could not be used normally due to their poor reactivity.

Description of the Limestone Agglomerates

In order to test the agglomeration method, approximately three tons of agglomerates were produced by the Hanna Mining Company and shipped to the Alexandria test facility. These agglomerates were produced by mixing smaller than 325 mesh Greer limestone with aporoximately 9 percent calcined lime in the agglomeration apparatus. The agglomerates were shipped to Alexandria wet. Analysis showed they contained 3.5 percent moisture.

The agglomerates were segregated prior to shipment into two size ranges. Two tons were in the size range of -6 +12 mesh (hereafter referred to as -6 mesh) and one ton was as +6 mesh (hereafter referred to as +6 mesh). The minimum fluidization velocity of the -6 mesh agglomerates was measured to be 2.8 ft/sec. The +6 mesh agglomerates had a minimum fluidization velocity greater than 4.5 ft/sec (the highest that could be measured by the small scale laboratory setup).

Test Objectives

Three tests were conducted to examine the operational characteristics of the agglomerates. The first test, No. 653-1, was a long duration test feeding the wet -6 mesh agglomerates to a bed of sulfated limestone. This test was conducted to determine the effectiveness of the moist agglomerates as a sulfur sorbent. The second test, No. 653-2, was a batch test using a fresh bed of a mixture of $^{1}/_{3}$ -6 mesh, $^{2}/_{3}$ +6 mesh dried agglomerates. The purpose of this test was to examine the attrition characteristics of the dried agglomerates. In the final test, No. 653-3, the dried mixture of agglomerates was fed to a sulfated limestone bed. The purpose of this test was to compare the performance of the moist agglomerates to the dried.

Test Results

Test No. 653-1 This test was a long duration test lasting approximately 26 hours. The objective of this test was to operate at steady state with both limestone and the -6 mesh agglomerates to compare the effect of the agglomerates on sulfur capture. The test started with a fresh bed of limestone. After what appeared to be a steady state with limestone feeding, feeding of the agglomerates was begun. Both limestone and agglomerate feed rates were controlled by the automatic combustion control system to maintain 500 ppm concentration of SO_2 in the flue gas. The bed drain rate and the coal feed rate were also automatically controlled. Table 30 lists the operating conditions during both steady state periods. Limestone was fed during the periods designated as 1500 hrs and 1600 hrs. The agglomerates were fed at 0900 hrs and 1000 hrs. Agglomerates feeding lasted for approximately 15 hours prior to sample taking.

Table 31 contains the chemical analysis of the solid samples taken during the test. Table 32 contains the chemical analysis of the bed material samples taken every hour during the test run. The bed material analyses prior to 1630 hrs on the first test day, August 2nd, tend to verify steady state operation on limestone.

Figure 20 shows a comparison of some previous limestone tests with Test No. 653-1. As can be seen in Figure 20, the limestone utilization for the Greer limestone was near that achieved by the previous tests. Furthermore, the agglomerates performed only marginally better than the Greer limestone.

A significant observation made during the test was that the agglomerates did not remain in the bed. This was shown by the increase in the fly ash production rate and by the fact that the bed drain rate decreased to zero. Further testing showed that when the moist agglomerates were placed in contact with hot bed material they exploded. It is believed that the moisture content of the agglomerates was responsible for the exploding particles. In order to determin the effect the moisture in the agglomerates had on their performance, the next two tests were conducted after drying the agglomerates.

Agglomerate Drying

Prior to conducting Test Nos. 653-2 and 653-3, the remaining agglomerates were dried in order to reduce their tendency to break apart in a hot, fluidized bed. Approximately 1,000 lbs of the -6 mesh and 2,000 lbs of the +6 mesh agglomerates remained. These agglomerates were mixed together and dried in batches of approximately 1,000 lbs.

Table 30

Test 653-1
Agglomerate Test Operating Conditions

Time	1500	1600	900	1000
Bed Temp.,°F	1540	1535	1495	1475
O_2 in flue gas, %	4.4	5.0	6.0	6.6
Coal feed rate, lb/hr	570	578	571	548
Fly ash collection rate, lb/hr	156	153	238	260
Limestone feed rate, lb/hr	256	256	228	228
Stack dust rate, lb/hr	38.8	39.8	39.7	39.6
Main air flow rate, lb/hr	5391	5227	5862	5595
Total air flow rate, lb/hr	5776	5610	6255	5995
Superficial velocity, ft/sec	9.6	9.3	10.2	9.7
Static bed depth, inches	22.2	21.9	19.1	19.3
EOBH (effective operating bed ht.)inches	21.5	21.5	19.0	19.0
SO_2, ppm	544	539	549	676
Ca/S feed ratio	2.87	2.83	2.56	2.66
Bed drain rate, lb/hr	139	139	0	0
% Sulfur capture	82.6	83.4	83.7	80.0
lb $SO_2/10^6$Btu input	0.98	0.94	1.09	1.34
% Combustion efficiency	82.4	83.7	80.2	77.7

Table 31

Chemical Analysis of Solid Samples Taken During Test 653-1

Material	Time Hr	C %	H %	Ash %	Ca %	S %
Coal						
composite 8/2*		72.64	5.00	11.46	1.40	3.78
composite 8/3**		70.20	4.82	13.65	1.49	4.36
Limestone						
composite*		10.84	0.64	61.32	30.18	0.32
Agglomerates						
composite**		10.03	0.35	64.84	34.88	0.29
Fly ash	15:00*	45.83	0.79	45.21	11.45	2.77
	16:00*	41.62	0.98	46.71	8.77	2.53
	09:00**	31.61	1.20	60.63	17.07	2.14
	10:00**	----No sample taken ----------				
Dust	15:00*	15.69	0.10	76.32	13.51	2.49
	16:00*	16.77	1.27	74.16	11.81	2.37
	09:00**	14.81	0.51	73.68	19.18	5.86
	10:00**	15.79	0.82	72.17	18.77	5.07
Bed	15:00*	1.05	0.30	94.37	36.34	9.57
	16:00*	0.72	0.68	95.83	34.72	10.79
	09:00**	-	-	-	31.04	-
	10:00**	-	-	-	27.66	-

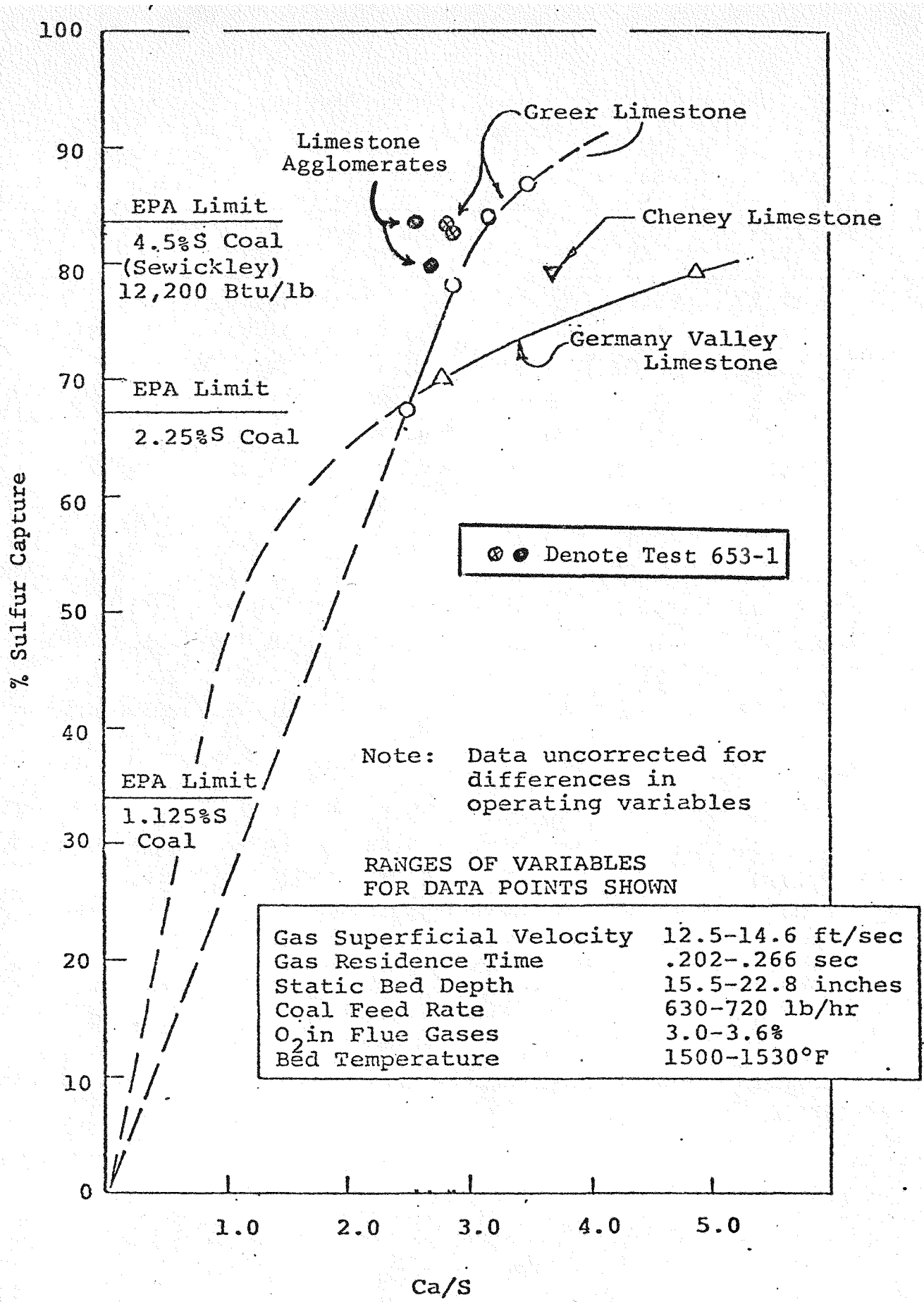
* Limestone feeding
** Agglomerate feeding

Table 32

Chemical Analysis of Bed Material Samples Taken During Test 653-1

Date	Time Hr	C %	H %	Ash %	Ca %	S %
8/2/77	11:30	0.86	0.51	94.18	32.88	10.00
	12:30	0.54	0.37	96.70	31.66	10.98
	13:30	0.22	0.30	96.57	31.76	9.08
	14:30	1.06	0.64	93.72	33.68	9.59
	15:00	1.05	0.30	94.37	36.34	9.75
	16:00	0.72	0.69	95.83	34.72	10.79
	17:00*	0.62	0.48	90.98	30.69	10.42
	18:00	0.41	0.47	97.65	28.47	11.41
	19:00	0.57	0.22	97.08	35.61	12.27
	20:00	0.31	0.32	96.09	33.02	12.60
	21:00	1.50	0.49	96.78	28.05	12.83
	22:00	0.59	0.38	97.92	31.42	12.84
	23:00	0.25	0.33	97.61	31.43	13.24
	24:00	0.40	0.28	97.10	34.94	13.66
8/3/77	01:00	0.33	0.26	96.91	30.72	12.34
	02:00	0.51	0.34	96.96	29.02	13.13
	03:00	0.40	0.21	97.41	29.06	11.89
	04:00	0.25	0.18	96.42	31.22	13.08
	05:00	0.18	0.32	97.87	29.62	11.88
	06:00	0.38	0.26	97.70	28.62	11.65
	07:00	0.35	0.35	97.50	32.41	12.65
	08:00	0.69	0.21	97.31	30.21	12.65

* Agglomerate feeding begun at 1630 hours on 8/2/77.



COMPARISON OF PREVIOUS LIMESTONE TESTS WITH TEST NO. 653-1

Figure 20

The drying was conducted in the FBM rather than in the materials processing system dryer in order to minimize particle breakage due to the handling system. It was felt that the drag conveyor used to transport solids to the dryer would promote particle breakage whereas, the agglomerates could be poured by hand into the FBM thus minimizing breakage.

The agglomerate drying was conducted with a superficial velocity in the FBM of 4 to 5 ft/sec and an inlet air temperature of approximately 500°F. The propane burner located in the air inlet duct was used to heat the incoming air. The equilibrium bed temperature was between 300°F and 400°F. Each batch of agglomerates was dried for a period of two hours.

After being dried, the agglomerates were analyzed for moisture content. During the analyses, the samples actually showed a gain in weight due to the CaO binder absorbing moisture from the atmosphere. This indicates that the agglomerates were dried to a negligible water vapor pressure which equalized with the ambient humidity after being exposed to the atmosphere. The equalized moisture content would analyze as 0 percent moisture. Prior to drying in the FBM, the agglomerates had an as-received moisture content of approximately 3.5 percent.

Test 653-2 - The second test was conducted to determine the attrition rate of the dried agglomerates in the hot fluidized bed environment. The test procedure consisted of loading the FBM with approximately 850 pounds of agglomerates. This gave an initial static bed depth of 14 in. A routine light-off procedure was followed in which the overbed burner was used to raise the front of the bed up to the ignition temperature of coal. When this temperature was reached, coal feeding was begun and the entire bed was brought up to 1500°F in a fully fluidized condition. Coal feeding was varied throughout the test to maintain bed temperatures between 1500°F and 1600°F. No agglomerate feed was used throughout the test. The test was terminated after the bed (static) depth had decreased to 4 inches due to attrition of the agglomerates.

The total test time period was approximately three hours. The light-off period lasted approximately two hours during which time approximately 300 lbs of the agglomerates elutriated from the bed.

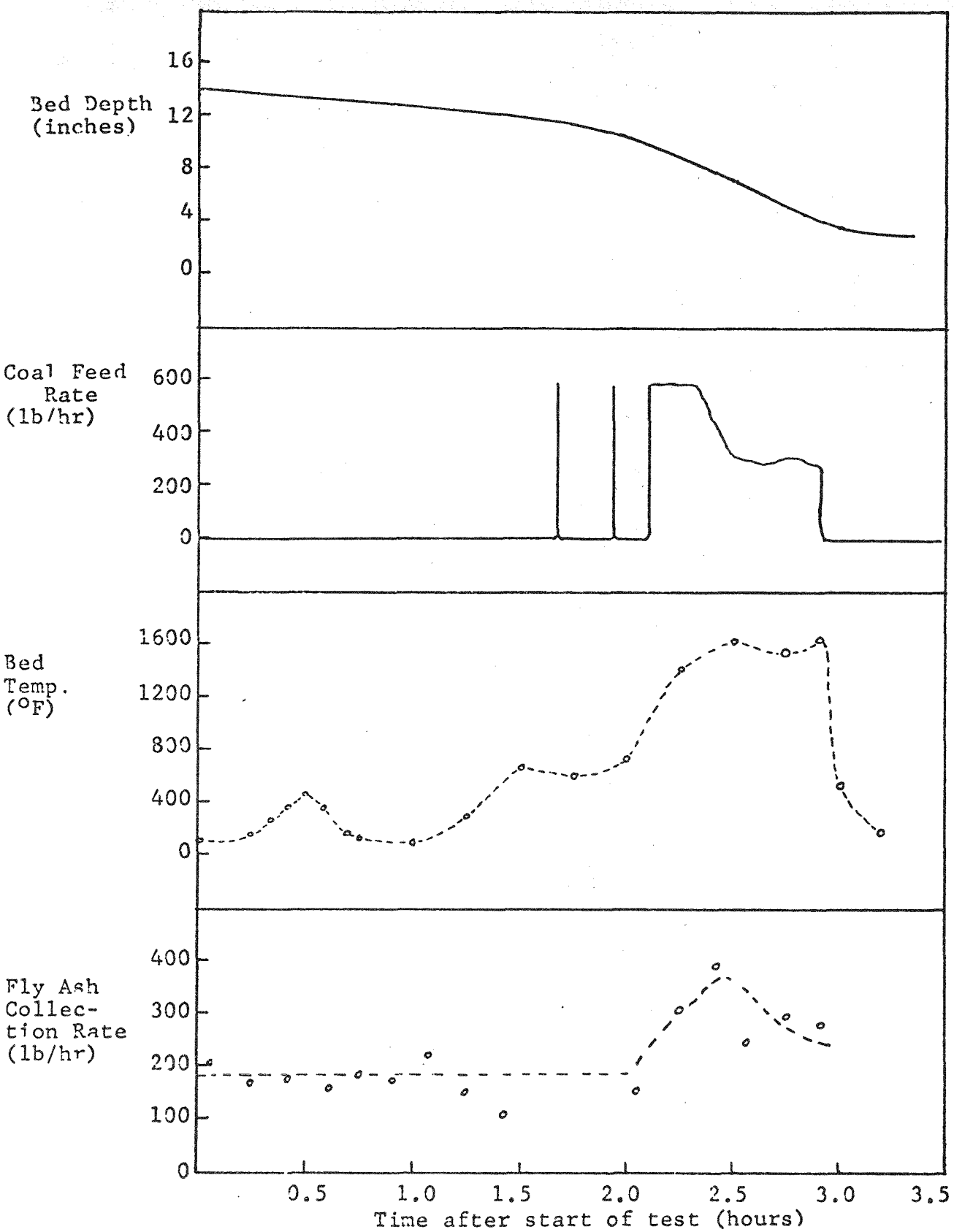
The period of operation at normal bed temperatures of 1450°F to 1600°F lasted approximately 40 min. during which time approximately 300 lbs more of agglomerates were elutriated from the bed. This corresponds to an agglomerate loss

rate of 450 lb/hr. Figure 21 shows test data for several parameters recorded during Test No. 653-2. Included in the figure are plots of bed depth, coal feed rate, bed temperature and fly ash collection rate versus time. From this figure it can be seen that the fly ash collection rate during the period of operation at 1500°F to 1600°F was approximately 380 lb/hr. This combined with the dust rate agrees fairly well with the bed depth decrease rate data which indicates a bed loss rate of 450 lb/hr. The dust rate, assuming an 85 percent cyclone collection efficiency, is approximately 60 lb/hr. An isokinetic dust sample was not collected during this time period to verify this rate due to the short period of hot operation. However, a dust rate of 60 lb/hr is very reasonable considering the unusually high fly ash collection rate.

Test No. 653-3 - During this test the dried agglomerate mixture was fed to a bed of spent limestone and coal ash. The procedure for this test was similar to the procedure for Test No. 653-2 with the exception that prior to agglomerate feeding, there was no limestone feed and the SO_2 level was above 1300 ppm. For this test, only 1,000 lb of agglomerates remained, permitting only a little over three and one-half hours of operation. Due to the limited amount of sorbent, it is believed that the test was not of sufficient duration to achieve a steady state feed rate of agglomerates. Results of the test are summarized in Table 33. Table 34 shows the chemical analysis of the solid samples taken during the test.

During the agglomerate feeding, the feed rate was controlled by the automatic control system. The controller was a proportional controller which attempted to maintain 500 ppm SO_2 in the flue gas. Due to the high SO_2 level prior to agglomerate feeding, the initial feed rate of the agglomerates was high. As the SO_2 level decreased the controller reduced the agglomerate feed rate. However, the system's response to the agglomerates was slow so that the SO_2 level was not able to reach 500 ppm before all agglomerates has been used. For this reason comparison of this test with Test No. 653-1 is not possible.

As in the previous tests, the agglomerates did not remain in the fluidized bed. This is evidenced by the increased fly ash production rate, the increase in the fly ash calcium content and the decrease in the fly ash carbon content. However, they did remain in the bed long enough to at least partially sulfate. This is evidenced by increased sulfur capture and sulfur content in the fly ash.



TEST 653-2 DATA VS. TIME

Figure 21

Table 33

Test 653-3
Agglomerate Test Operating Conditions

Time	930*	1315**	1400**
Bed. Temp., °F	1520	1515	1500
O_2 in flue gas, %	5.3	4.5	4.8
Coal feed rate, lb/hr	615	629	596
Fly ash collection rate, lb/hr	155	272	246
Limestone feed rate, lb/hr	0	353	264
Stack dust rate, lb/hr	39.2	39.2	39.2
Main air flow rate, lb/hr	5307	5614	5641
Total air flow rate, lb/hr	5707	6008	6040
Superficial velocity, ft/sec	9.4	9.9	9.9
Static bed depth, inches	22.4	21.3	21.5
SO_2, ppm	1325	679	898
Ca/S Molar feed ratio	0	4.01	3.17
Bed drain rate, lb/hr	85	85	85
lb $SO_2/10^6$Btu input	2.20	1.18	1.64
% Sulfur capture	60.7	78.9	70.7
% Combustion efficiency	84.2	84.3	82.6

* No limestone feed
** Agglomerate feed

Table 34

Chemical Analysis of Solid Samples Taken During Test 653-3

Material	Time Hr	C %	H %	Ash %	Ca %	S %
Coal composite		71.70	4.98	11.47	0.85	3.71
Agglomerates composite		9.74	0.42	63.96	33.51	0.00
Bed	09:30	0.40	0.15	98.25	29.60	12.70
	13:15	0.51	0.17	97.83	31.35	11.68
	14:00	0.45	0.16	98.32	31.41	10.37
Fly Ash	09:30	43.86	1.10	48.42	8.69	2.35
	13:15	26.57	0.54	68.01	18.65	5.10
	14:00	29.11	1.04	65.50	16.31	4.88
Dust	09:30	22.10	1.12	70.37	12.24	2.63
	13:15	9.68	0.43	80.48	24.55	6.67
	14:00	11.28	0.65	79.14	21.71	6.82

Discussion of Results

The wet limestone agglomerates tested showed only a slightly better sulfur capture than the Greer limestone. The residence time of the wet agglomerates in the bed was short. Nevertheless, the amount of sulfur capture was significant. It is not known at this time whether the agglomerates performed any differently than -325 mesh limestone would have under similar test conditions. Furthermore, it cannot be determined at this time whether an indigent quality limestone would have been as effective.

The dried agglomerates appeared to have a longer residence time in the bed. The batch Test No. 653-2 indicates particle residence times on the order of one hour. However, due to the limited amount of sorbent, a steady state evaluation of the sorbent capability was not possible.

Recommendations

Further testing of the limestone agglomerates is recommended. The first preliminary tests showed that even with short residence times in the bed, sulfur capture was still possible. Large scale testing in the Alexandria unit, however, should not be pursued at this time due to the expense and materials required.

Considerable laboratory testing is needed to optimize the agglomerate before large scale testing is recommended. Alternate binders (other than CaO) should be tried in an attempt to produce a porous abrasion resistant pellet that can survive in the FBM environment for several hours. Once an optimum pellet has been determined, based on laboratory tests, a sufficient amount of agglomerates should be produced for testing at Alexandria. Approximately four tons of agglomerates would be required for 20 hours of testing. A 100 hour test is recommended to fully determine the effectiveness of the agglomerates. If the results using Greer limestone appear favorable, other indigent quality limestone should be tested.

Lowellville Limestone

Test No. 654-6, the final eductor test, was also used as a limestone evaluation test. During the first half of this test, the attrition rate of limestone obtained from Lowellville, Ohio was studied to determine whether or not this limestone could be used in the Rivesville MFB.

In order to compare the Lowellville limestone to Greer limestone, Test No. 654-6 was compared to the fly ash characterization tests. Table 35 shows the operating conditions of Lowellville limestone test, No. 654-6 and the fly ash characterization tests, Nos. 652-1, 652-2 and 652-3. The calcium to sulfur molar feed ratio was 3.90 and the bed temperature was 1710°F for the Lowellville limestone test. For the characterization tests, the calcium to sulfur molar feed ratio ranged from 2.22 to 3.03 (lower than in Lowellville limestone test) and the bed temperature was around 1500°F (again lower than in the Lowellville limestone test). It can be seen from Table 35 that the rate of stack dust production when using the Lowellville limestone is about the same as that for Greer limestone. The Lowellville limestone was also subjected to higher operating temperatures which could be more favorable to attrition. Therefore, on the basis of these short run tests, it appears that Lowellville limestone can be used in the Rivesville MFB without formation of excessive amounts of stack dust. Due to the small quantity of the limestone tested, no prediction on the sulfur capture ability could be made at this time.

Table 36 shows analyses of bed material, fly ash, dust, coal and as-received limestone for Test No. 654-6. It should be noted that the chemical composition of the bed material obtained in this test is typical of the compositions obtained during tests feeding Greer limestone. However, the bed sulfur content is slightly lower due to the short duration of this test. Table 37 shows the particle size analysis of the Lowellville limestone.

2.3.4 Eductor Tests

A study of the potential use of single and multiple nozzle eductors in the coal feed systems of fluidized-bed boilers was performed. Experimental runs were performed over a wide-range of test conditions for several different eductor designs. The purpose of these tests was to determine an eductor design with no restriction to 1/2 in coal particles for which: (1) the coal inlet pressure would be at, or slightly below, atmospheric, with back pressures between 15 in and 30 in of water, (2) air flow rate would be 300-400 lb_m/hr, (3) required air supply pressure would be minimized. Although no eductor design was able to attain the desired conditions at the 3 psig supply pressure available at Rivesville, the test results indicate that single nozzle types come closer to meeting these conditions than any other type tested.

Table 35

Operating Conditions of the Fly Ash Characterization Tests

Test No.	652-1		652-2		652-3		654-6
Time	10:30	11:30	10:15	11:45	10:30	11:15	12:30
Bed Temp., °F	1480	1500	1500	1540	1515	1560	1710
O_2 in flue gas, %	4.9	4.0	3.2	2.9	4.7	4.0	5.2
Coal feed rate, lb/hr	595	555	553	553	705	660	413
Fly ash collection rate, lb/hr	155	161	130	130	231	231	129
Limestone feed rate, lb/hr	333	333	273	273	444	444	244
Stack dust rate, lb/hr	25.1	25.6	39.7	39.7	38.9	41.4	20.0
Main air flow rate, lb/hr	5209	5194	4222	4209	6744	6744	4415
Total air flow rate, lb/hr	6049	6032	5059	5939	7572	7577	6927
Superficial velocity, ft/sec	9.3	9.4	7.8	7.9	11.2	10.3	12.9
Static bed depth, inches	23.5	23.0	19.5	19.5	24.0	24.0	13.0
SO_2, PPM	145	205	450	550	275	290	194
Ca/S	2.52	2.70	2.22	2.22	2.84	3.03	3.90
Combustion Eff.	76.2	73.3	79.8	79.6	75.8	74.1	83.7
lb $SO_2/10^6$ BTU	0.295	0.386	0.828	1.008	0.588	0.660	0.570

Table 36

Test 654-6 Chemical Analyses

	C	H	Ca	S	Ash	L.O.I.
Bed Material						
11:00	1.45	0.26	43.66	6.69	98.04	1.96
11:30	0.20	0.16	41.86	7.12	99.10	0.90
12:00	1.13	0.33	42.18	6.83	98.56	1.44
12:30	0.64	0.18	41.57	6.00	98.15	1.85
Fly Ash 12:30	37.13	0.73	19.48	2.38	59.14	40.86
Dust 12:30	10.01	0.60	23.60	3.06	84.00	16.00
Coal 12:30	71.74	4.85	1.34	4.31	11.40	88.60
Limestone 12:30	11.58	0.00	35.60	0.00	58.92	41.08
Limestone Assay						
$CaCO_3$	88.90%					
$MgCO_3$	6.94%					
others	4.16%					

Table 37

Size Distribution of Lowellville Limestone

No.*	Percent smaller than stated size
1/4"	100.0
6	99.8
8	77.6
12	37.8
16	14.6
20	10.0
30	8.2
40	7.1

* U.S. Sieve Series

Introduction

In order to feed coal and limestone into a fluidized bed, the back pressure of the bed must be overcome by the solids feeding system. Although screws, rotary feeders, and seal legs are commonly used to meet this requirement, reliability of these devices has been less than desired, particularly when handling widely varying sizes of solids. A program was undertaken at the Alexandria laboratory to develop an eductor to reliably feed 1/2 in x 0 particle size coal into atmospheric fluidized-bed boilers at minimum supply pressure requirements.

Testing

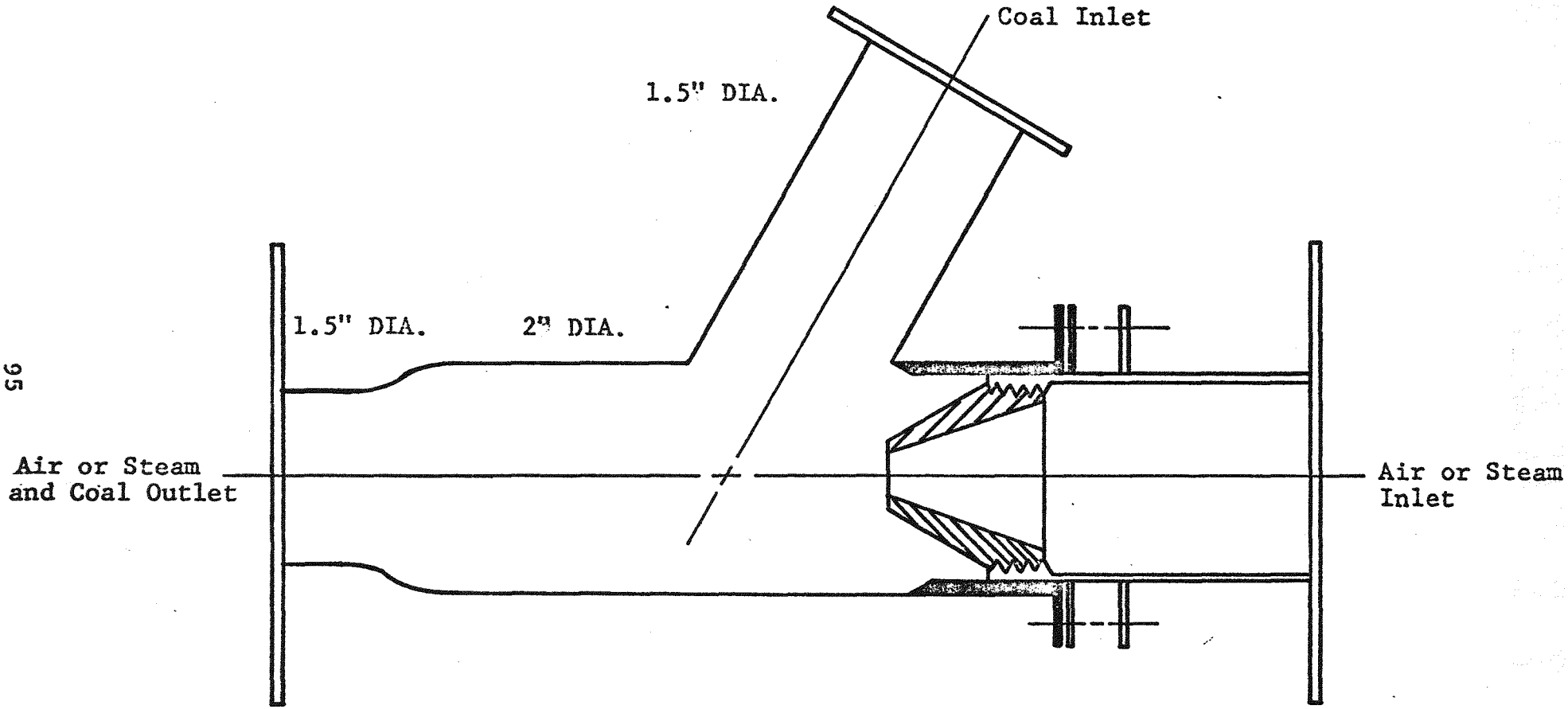
Preliminary testing indicated that eductor performance for feeding coal and sorbent into a fluidized bed could be satisfactorily estimated by testing without solids flow. Eductor tests were conducted using the coal feed system of the FBM for hot tests and a pneumatic test stand for the cold testing. The arrangement for the eductor tests using the coal feed system is shown in Figure 22. The pneumatic test stand setup was similar to the coal feed system setup except that the inlet and outlet piping sizes of the eductor were different and the back pressure was maintained by a valve instead of the fluidized bed.

When the coal feed system was used, a custom fabricated 2 in pipe size eductor body with 1 1/2 in inlet and outlet piping was used. This eductor assembly is shown in Figure 23. When the test stand setup was used, single nozzle types, multiple nozzle types, and commercial design booster fittings were tested. The single nozzle types tested on the test stand were a 2 1/2 in pipe tee with a 2 in inlet and a 2 1/2 in outlet, and a 2 in pipe Y with 1 1/2 in inlet and 2 in outlet, both shown in Figure 24 The single nozzle type of eductors were tested using varying nozzle sizes and were positioned such that the nozzle did not enter the solids flow path. The multiple nozzle eductor bodies were of two varieties. These were a design fabricated at Rivesville using four tubes as nozzles and a type fabricated at Alexandria by drilling holes around the circumference of a pipe at 30° to the pipe axis.

The Rivesville fabricated eductor body is shown in Figure 25. Four 3/8 in OD copper tubes connect the manifold to the pipe at 90° intervals around the circumference of the pipe and at a 30° angle with the axis of the pipe.

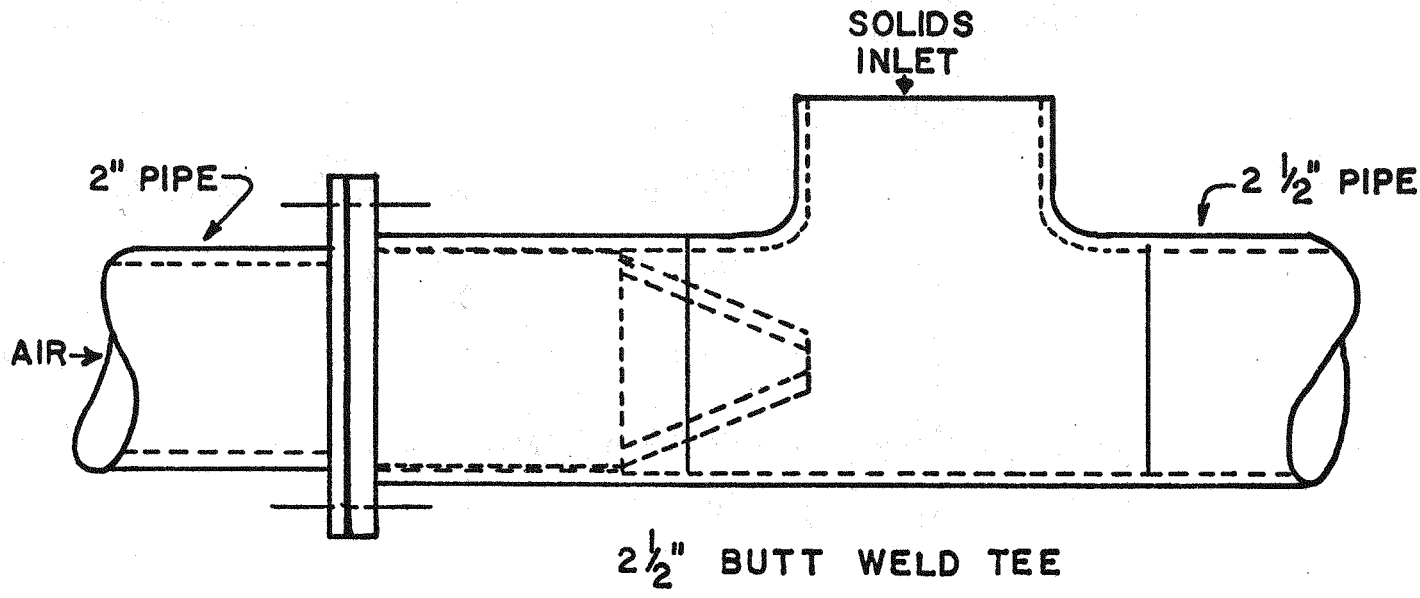
Figure 22

EDUCTOR TEST ARRANGEMENT

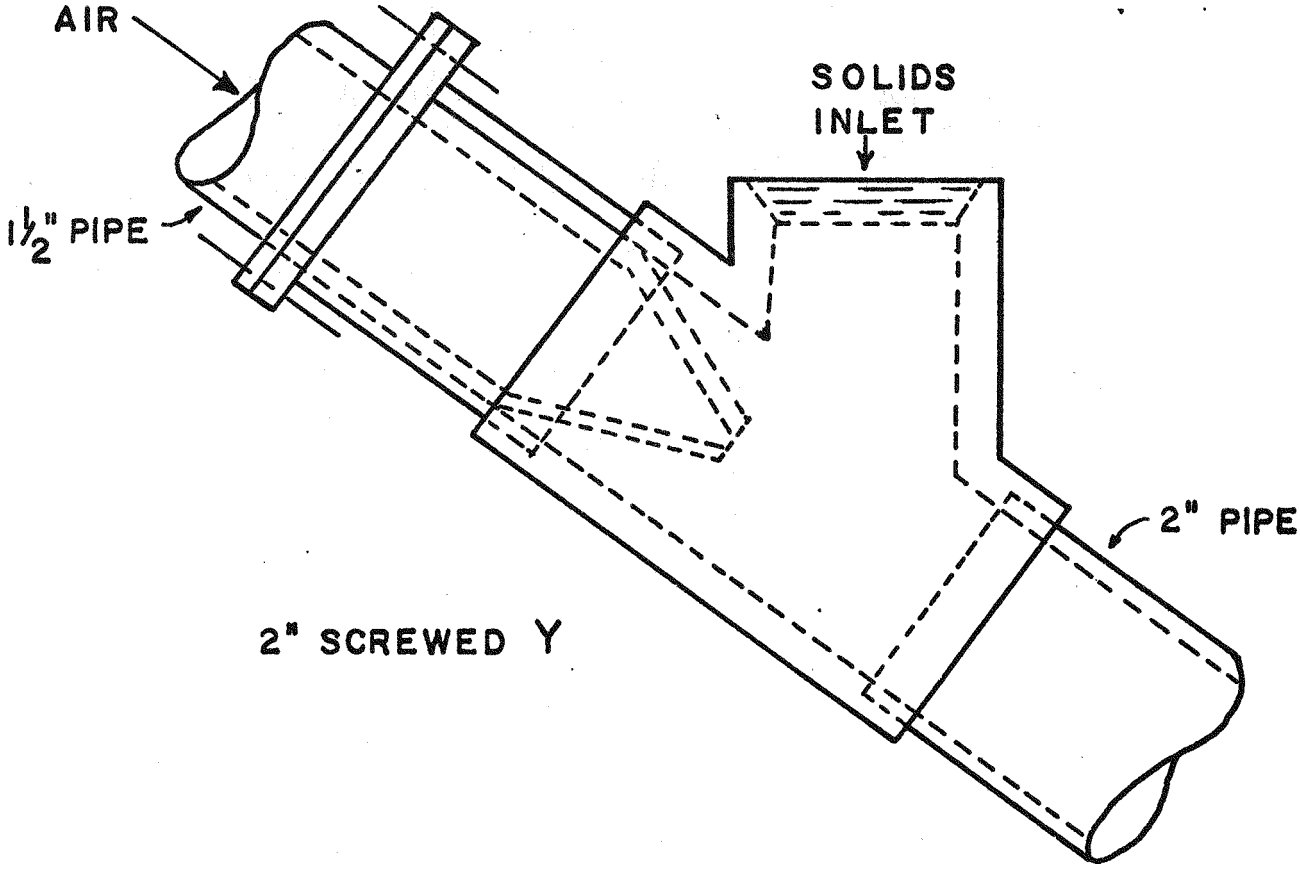


SKETCH OF EDUCTOR APPARATUS SHOWING A CONVERGING NOZZLE (HATCHED) IN PLACE.

Figure 23



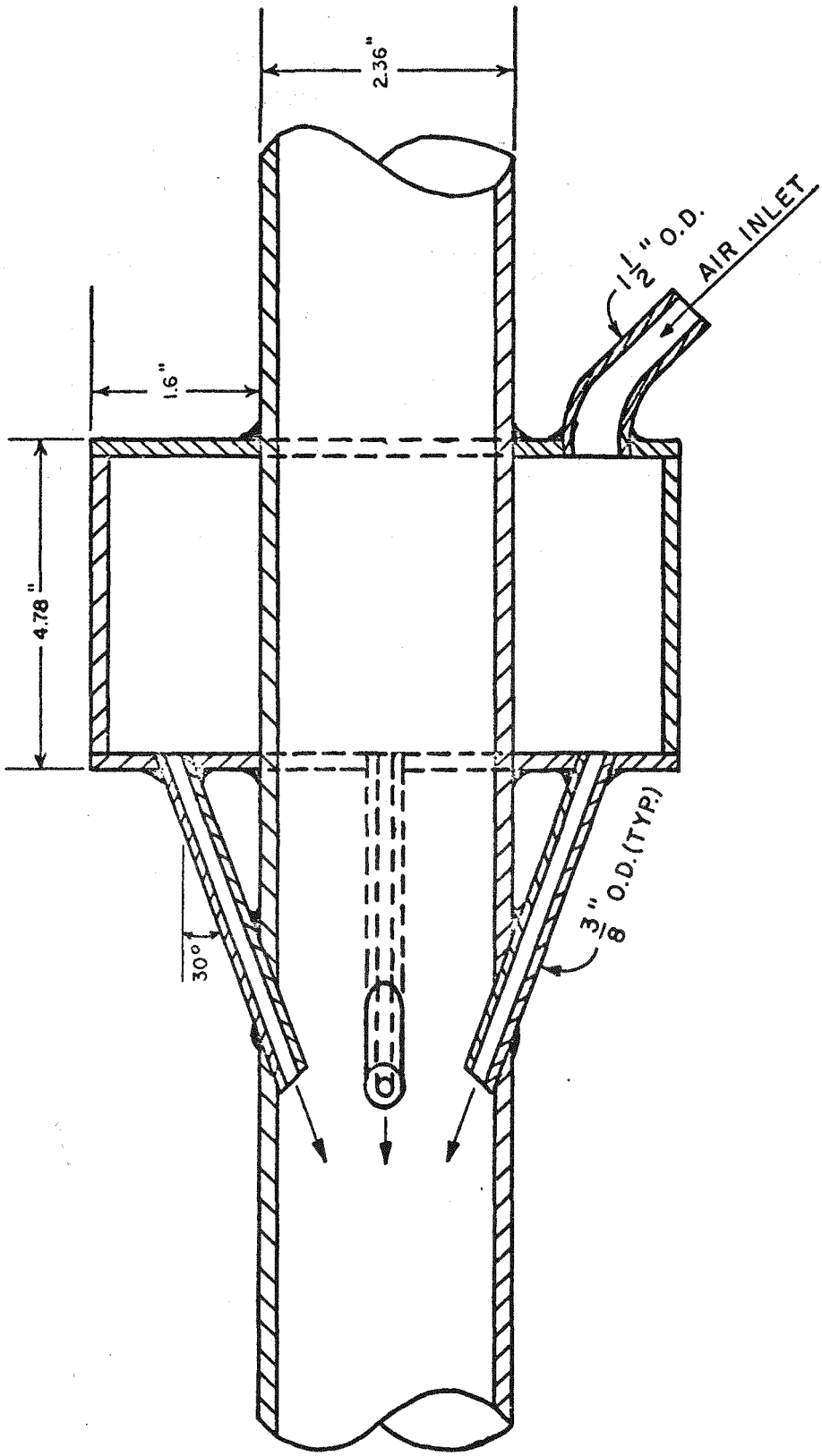
2½" BUTT WELD TEE



2" SCREWED Y

PIPE FITTING EDUCTOR BODIES

Figure 24



CROSS SECTION VIEW OF MULTIPLE NOZZLE EDUCTOR
FABRICATED AT RIVESVILLE

Figure 25

The drilled hole type eductor fabricated at Alexandria is shown in Figure 26. Holes were drilled in a pipe at intervals around the circumference. Air is supplied to these holes by an annulas manifold. The size and number of holes can be changed by drilling. The designs tested for this type of eductor body were four 3/16 in holes, four 1/4 in holes, four 5/16 in holes, and eight 3/16 in holes.

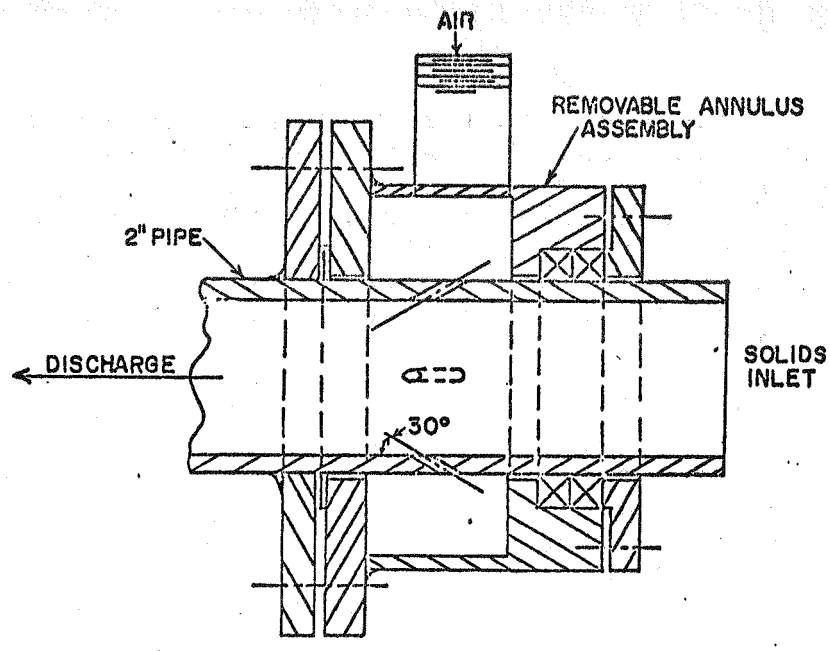
Two commercial design pneumatic transport boosters were also tested. A Dynachek and a Dynajet Booster were purchased from Dynamic Air Incorporated. The commerical design of the Dynachek is shown in Figure 26. In the Dynachek booster, a rubber gasket placed between the two cast iron pieces forms an air manifold. When the pressure in the conveying line is less than the supply pressure, the rubber gasket flexes, allowing air to enter around the entire perimeter of the conveying line. When the air supply pressure is less than the conveying line pressure, the rubber gasket contracts and seals the booster fitting. Thus solids cannot enter the booster and plug it. The Dynajet design is similar to the Dynachek except that the rubber gasket is removed and clearances adjusted.

Results and Discussion

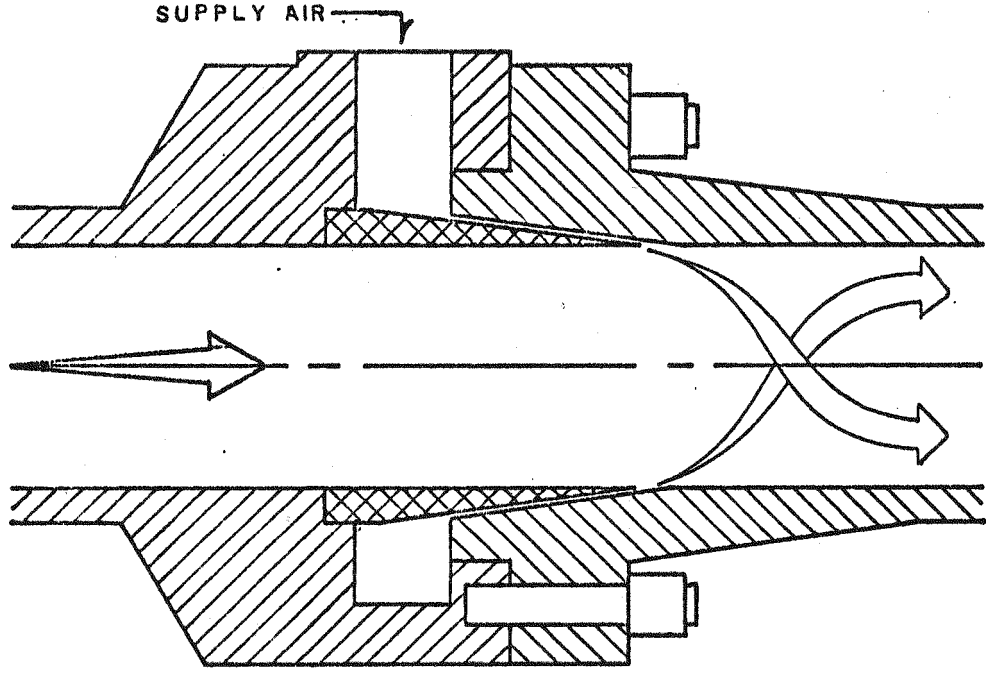
Tables 38 through 40 are summaries of the results. Table 38 compares selected data for single nozzle eductors, multiple nozzle eductors, and the commerical boosters at 30 in of water back pressure. This is about the maximum expected back pressure if the coal feed needles at Rivesville were extended into the beds. The data shows that the single nozzle design requires less supply pressure to meet the desired criteria for solids inlet pressure and mass flow than the multiple nozzle designs and boosters.

The "best" multiple nozzle design had eight .188 in diameter nozzles. This design is slightly better than design with four .25 in diameter nozzles because the required supply pressure is slightly lower for the eight nozzle design than for the four .25 in diameter nozzle design.

The Rivesville fabricated eductor did not perform well at the 30 in back pressure because of the high pressure losses through the 3/8 in OD copper tubes used as nozzles.



DRILLED PIPE EDUCTOR



DYNA-CHEK BOOSTER FITTING

Figure 26

Table 38

Test Results Summary at 30" H_2O Back Pressure

Description	Nozzle Diameter (in.)	Supply Air (psig)	Air Flow (lb/hr)	Back Pressure (in H_2O guage)	Solids Inlet Pressure (in H_2O guage)
Eductor Type					
Pipe Y'2"	0.5	10	347	30.0	1.2
"	0.5	5	247	30.0	14.6
Multiple Nozzle					
Drilled pipe 8 holes	.188	16	413	30.0	2.5
Drilled pipe 4 holes	.25	20	419	30.0	3.4
	.25	10	267	30.0	18.0
Rivesville Fabricated 4 tubes	.375 OD	20	306	30.2	11.0
	.375 OD	10	214	29.8	20.7
Drilled pipe 4 holes	.188	20	253	30.0	11.7
	.313	10	446	30.2	14.3
Commercial					
"Dynachek"	--	20	212	29.8	24.6
"Dynajet"	--	20	218	30.3	30

Table 39

Test Results Summary at 15" H_2O Back Pressure

Description	Nozzle Diameter (in.)	Supply Air (psig)	Air Flow (lb/hr)	Back Pressure (in H_2O guage)	Solids Inlet Pressure (in H_2O guage)
Eductor Type					
Single Nozzle					
Pipe Y-2"	.5	7	303	15.2	- 7.6
"	.5	5	258	15.2	- 1.9
"	.5	4	230	15.0	+ 1.4
Fabricated eductor on coal feed system with tapered nozzle	.47	10.2	289	18.0	-15.0
	.47	3.7	184	14.0	0
Multiple Nozzle					
Drilled pipe 8 holes	.188	16	406	15.0	-12.5
	.188	10	312	15.0	- 1.9
	.188	8	264	15.0	+ 2.5
Drilled pipe 4 holes	.313	9	436	15.0	+ 0.1
	.25	12	304	15.0	- 2.4
	.25	10	273	15.0	+ 0.6
	.188	20	254	15.0	- 4.0
Rivesville Fabricated 4 tubes	.375 OD	20	308	15.0	- 4.8
	.375 OD	16	263	15.0	- 1.2
"	.375 OD	14	241	15.0	+ 1.2
Commercial					
"Dynachek"	--	20	212	14.8	+ 9.4
"Dynajet"	--	20	225	14.8	+14.3

Table 40

Test Results Summary for Low Pressure Air Supply

Description	Nozzle Diameter (in.)	Supply Air (psig)	Air Flow (lb/hr)	Back Pressure (in H_2O guage)	Solids Inlet Pressure (in H_2O guage)
Eductor Type					
Single Nozzle					
Pipe Y-2"	.5	3.0	181	30.0	22.4
"	.5	3.0	199	14.7	14.6
"	1.8	2.4	527	30.0	17.0
"	1.0	2.0	518	15.0	2.0
Fabricated eductor on coal feed system	.47	2.8	159	13.0	5.0
Multiple Nozzle					
Drilled pipe 8 holes	.188	4.0	141	30.0	26.3
	.188	4.0	161	15.0	10.3
Drilled pipe 4 holes	.25	4.0	132	30.0	26.0
	.25	4.0	141	15.0	10.6
	.313	3.0	178	15.0	12.3
Rivesville Fabricated 4 tubes	.375 OD	3.0	106	30	28.5
Commercial					
"Dynachek"	--	4	62	15.0	14.4
"Dynajet"	--	4	82	15.0	15.4

Tests on multiple nozzle designs with larger and smaller than .25 in holes show that this size is near optimum for the design conditions. Compared to the other designs tested, the performance of the boosters was poor. This is not surprising since the boosters were designed for assisting the movement of solids already in transport and not for use as eductors.

Table 39 is a summary of the data for the single nozzle eductors, multiple nozzle eductors, and boosters at 15 in of water back pressure. This is about the back pressure expected for coal feed needles terminating 12 in above the air distributor with a 24 in static bed depth. At the lower back pressure, atmospheric and slightly negative solids inlet pressures were obtained for most designs. Again, the data shows that for the 15 in back pressure, the single nozzle designs require less supply pressure to meet the desired conditions than did the multiple nozzle and the booster designs. The tapered nozzle with a .5 in throat is slightly too small for the desired mass flow at low supply pressures.

The multiple nozzle designs all required a much higher supply pressure than the single nozzle designs to attain the desired conditions at the solids inlet. The .188 in diameter eight nozzle design and the .313 in diameter four nozzle design showed the "best" performance of the multiple nozzle eductors at the 15 in back pressure. For the Rivesville fabricated eductor, a higher supply pressure was necessary to attain the desired coal inlet conditions for a slightly lower mass flow compared with the four nozzle .25 in diameter design.

Table 40 is a summary of test results for eductors operating with a low air supply pressure near the 3 psig currently available at the feeder tables at Rivesville. As expected, the pressure reduction at the solids inlet is minimal and the air flow is less than required in the smaller nozzle sizes. Again the single nozzle eductor performance is better than the multiple nozzle designs.

Mathematical Model

In order to parameterize the nozzle selection for the eductor, a semi-empirical mathematical model was developed to describe the results of the test data. The following simplifying assumptions were made in constructing the model.

1) The flow is compressible.
2) The fluid in the eductor is an ideal gas.
3) The isentropic relation $p = (\text{constant})\ \rho^{\gamma}$ applies.
4) The mass flow through the eductor is conserved. The solids flow that would be present in a feed system is neglected.

Based on these four assumptions the following equations govern the system.

Continuity

$$\rho_A U_A a_A = \rho_B U_B a_B = \dot{M} \tag{1}$$

Bernoulli

$$g_c \frac{U_B^2}{2} - g_c \frac{U_A^2}{2} + \int_A^B \frac{dp}{\rho} + \frac{1}{2} g_c U_B^2 e_v = 0 \tag{2}$$

State

$$p = \rho RT \tag{3}$$

Isentropic Relation

$$p = (\text{constant})\ \rho^{\gamma} \tag{4}$$

where:

p is the absolute fluid pressure (psia)
ρ is the density of the fluid (lb_m/ft^3)
U is the fluid velocity (ft/sec)
a is the area (in^2)
T is the temperature (°R)
R is the ideal gas constant
γ is the ratio of specific heats $\frac{C_p}{C_v}$

A and B refer to stations along the flow path
e_v is a friction loss factor associated with viscous dissipation
M is the mass flow through the eductor (lb_m/hr)

If the friction loss factor, e_v, is assumed to be zero, the losses can be taken into account by the factor C_{dAB} in the equation which results from solution of equations (1) through (4):

$$\dot{M} = C_{dAB}\,\rho_B\,a_B \cdot \left\{ \frac{2g_c \left(\frac{P_A}{\rho_A}\right)\left(\frac{\gamma}{\gamma-1}\right)\left[1-\left(\frac{P_B}{P_A}\right)^{\frac{\gamma-1}{\gamma}}\right]}{1-\left(\frac{a_B}{a_A}\right)^2 \left(\frac{P_B}{P_A}\right)^{\frac{2}{\gamma}}} \right\}^{\frac{1}{2}} \tag{5}$$

This equation can be recognized as the general orifice equation for compressible flow where C_{dAB} is the discharge coefficient.

The eductor is divided into two separate regions. These regions are the flow through the nozzle portion and the expansion and pressure recovery sections between nozzle outlet and eductor outlet. By applying equation (5) to each of these regions, the following system of equations is generated.

$$\dot{M} = C_{d01}\,a_1 \left(\frac{P_1}{P_0}\right)^{\frac{1}{\gamma}} \left\{ \frac{2g_c \left(\frac{P_0^{\,2}}{RT_0}\right)\left(\frac{\gamma}{\gamma-1}\right)\left[1-\left(\frac{P_1}{P_0}\right)^{\frac{\gamma-1}{\gamma}}\right]}{1-\left(\frac{a_1}{a_0}\right)^2 \left(\frac{P_1}{P_0}\right)^{\frac{2}{\gamma}}} \right\}^{\frac{1}{2}} \tag{6}$$

and

$$\dot{M} = C_{d12}\,a_2 \left(\frac{P_2}{P_0}\right)^{\frac{1}{\gamma}} \left(\frac{P_1}{P_0}\right)^{\frac{\gamma-1}{2\gamma}} \cdot \left\{ \frac{2g_c \left(\frac{P_0^{\,2}}{RT_0}\right)\left(\frac{\gamma}{\gamma-1}\right)\left[1-\left(\frac{P_2}{P_1}\right)^{\frac{\gamma-1}{\gamma}}\right]}{1-\left(\frac{a_2}{a_1}\right)^2 \left(\frac{P_2}{P_1}\right)^{\frac{2}{\gamma}}} \right\}^{\frac{1}{2}} \tag{7}$$

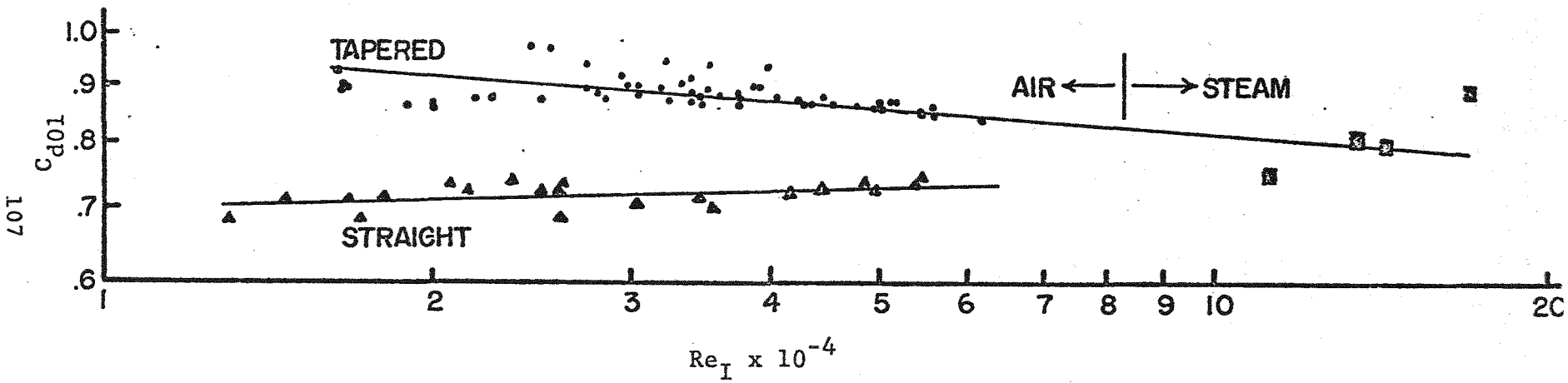
where the subscripts 0, 1, and 2 refer to eductor inlet, coal inlet and eductor outlet respectively. Using equations (6) and (7) and the test data generated when air and steam were the operating fluids, C_{d01} and D_{d12} were then correlated with different physical parameters of the system. A least squares fit was made for each correlation and optimum fits were determined.

The results of this work are shown in Figures 27 and 28. It is apparent that for both tapered and straight through nozzles (1), the logarithm of C_{d01} correlated with the logarithm of the inlet Reynolds number (2), the logarithm of C_{d12} correlated with the ratio of the nozzle throat area to the eductor outlet cross sectional area a_r and the logarithm of the Prandtl number, and (3) both correlations are independent of the back pressure. Therefore, for a given amount of taper C_{d01} depends upon the fluid dynamics of the eductor only, whereas, C_{d12} depends both upon the eductor fluid dynamics and the eductor geometry. Subsequent tests have shown in fact that for different eductor bodies, the logarithm of C_{d01} still correlated with log Re_I is very nearly the same manner as shown in Figure 28. However, although C_{d12} correlates with A_2/A_1 and Pr, it appears that the correlation changes with eductor body geometry. The mathematical expressions for the correlations for C_{d01} and C_{d12} are found in Table 41.

The correlations for C_{d01} and C_{d12} were used to predict coal inlet pressures and throat area (or diameters) for given mass flows in tapered nozzles with air and steam as the operating fluid. These calculations were then compared to the actual test data. Figures 29, 30 and 31 show the agreement of the test data and the calculated results.

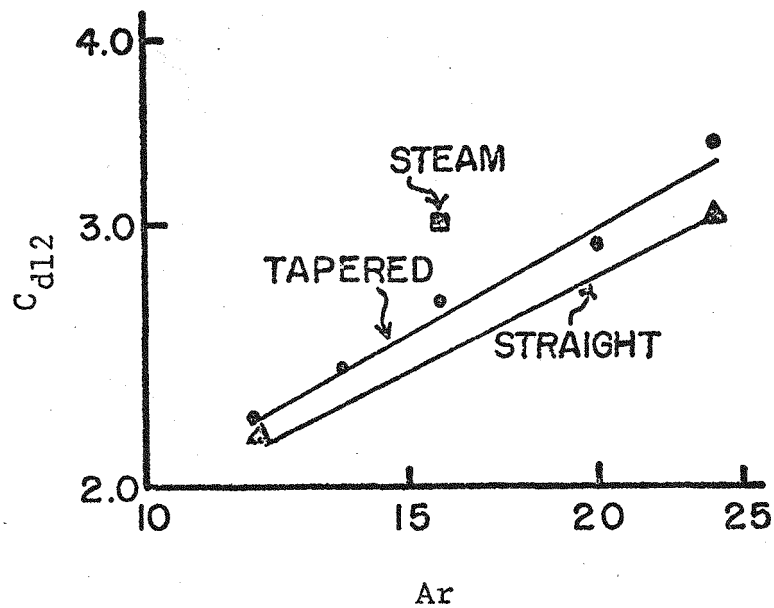
Use of the Model in Eductor Design

Design of an eductor involves the use of five variables, three of which are required to calculate the other two. The five design variables are the mass flow rate, the nozzle diameter and the pressures at the air inlet, the coal inlet and the air outlet. Typically, the pressure at the air outlet is set by the back pressure from the fluidized bed and the frictional losses downstream of the eductor. The coal inlet pressure should be set at atmospheric since a positive pressure would be less effective and too negative a pressure would require excessive blower horsepower. The mass flow of air through the eductor is also predetermined by the cooling requirements and proper fuel-air requirements in the needle. Once these three variables have been fixed, there is only one combination of nozzle diameter and air inlet pressure that will satisfy the preset conditions.



Plot of log C_{d01} versus log Re_I for tapered and straight nozzles

Figure 27



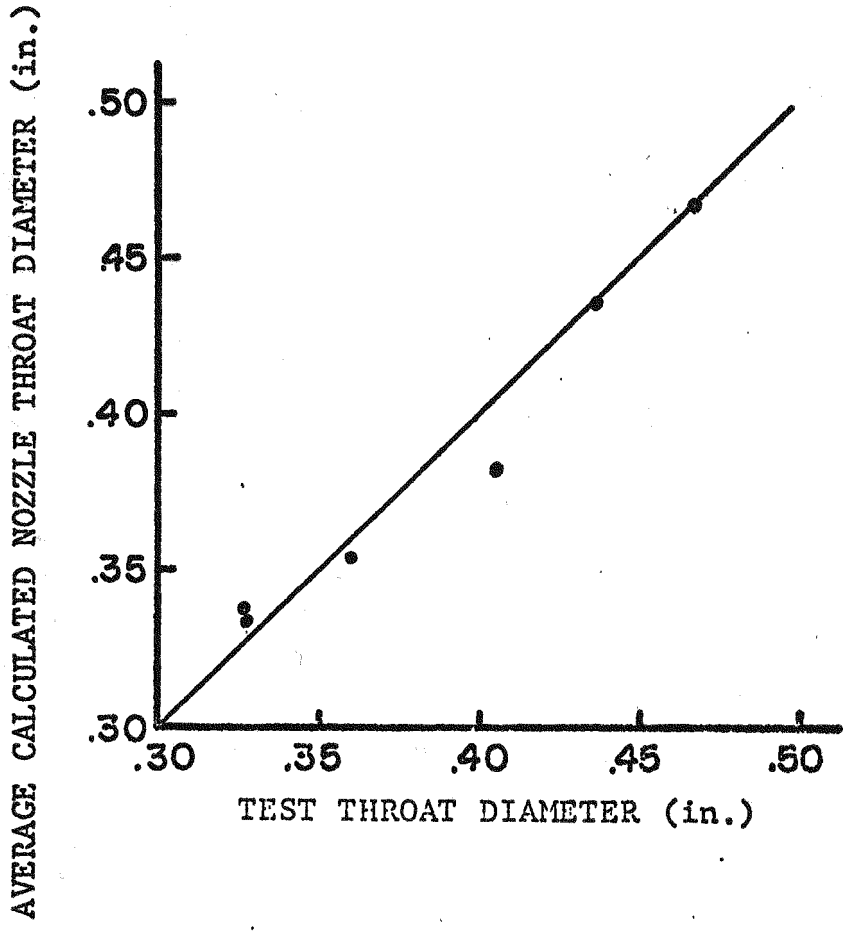
Log of average C_{d12} versus log of Ar for tapered and straight nozzles

Figure 28

Table 41

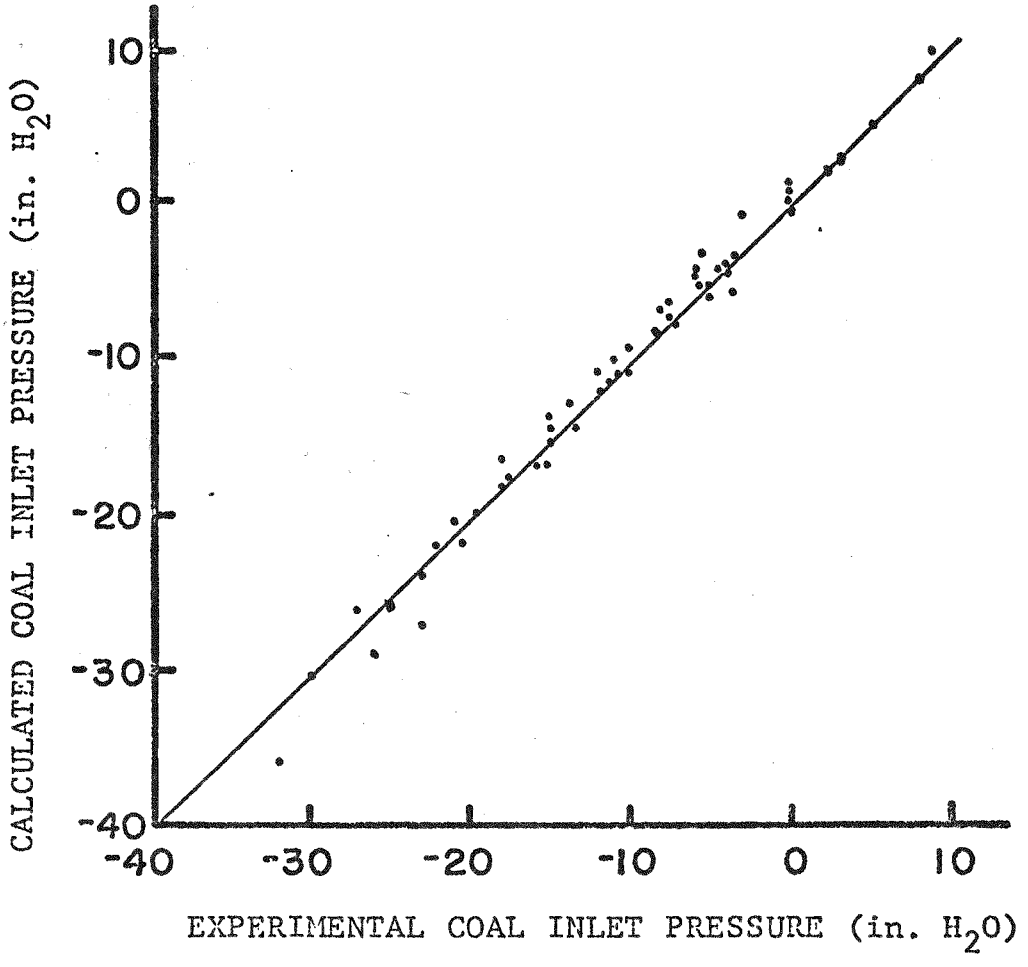
Mathematical Expressions for the Best Correlations Of C_{d01} and C_{d12} for Tapered and Straight Through Nozzles.

Tapered Nozzle
$\text{Log } C_{d01} = -.07287 \log Re_I + .275194$
$\text{Log } C_{d12} = .55223 \log a_r - .061 \log Pr - .249745$
Straight Through Nozzle
$\text{Log } C_{d01} = .017628 \log Re_I - .22395$
$\text{Log } C_{d12} = .48831 \log a_r - .061 \log Pr - .200278$



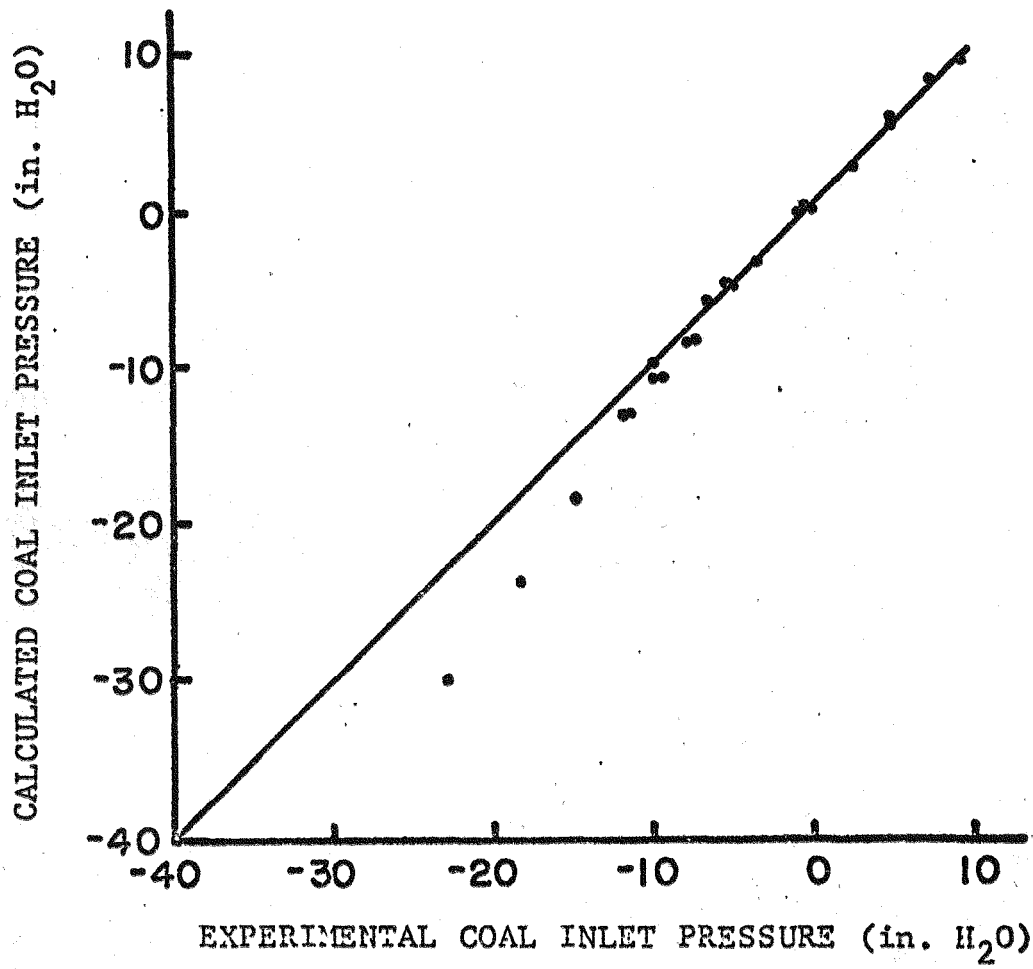
Average calculated nozzle throat diameter versus test throat diameter for tapered and straight nozzles

Figure 29



Experimental versus calculated solids inlet pressure for the tapered nozzle for air and steam

Figure 30



Experimental versus calculated solids inlet pressure for the straight nozzle with air

Figure 31

Solutions of equations 6 and 7 involves a trial and error solution for the nozzle diameter and the air inlet pressure. The solution can be simplified by recognizing the fact that equations 6 and 7 are solved for the mass flow of air as a function of the other variables. Thus, by specifying three of the variables, a plot of the mass flow obtained from each equation versus the fourth variable will yield two curves that intercept at the solution for the three specified variables.

An example of this procedure is the problem where a back pressure of 17 in w.c. and a coal inlet pressure of 0 in w.c. are specified by the design at a mass flow of 300 lb/hr. The above procedure was followed to calculate the nozzle area and the mass flow rate at various blower pressures. These results are plotted in Figure 32. From the figure it can be seen that for a 300 lb/hr air flow at these conditions a 1.8 psi air inlet pressure and a .74 in diameter (.43 sq in area) nozzle are required.

2.3.5 Coal Feed Needle Tests

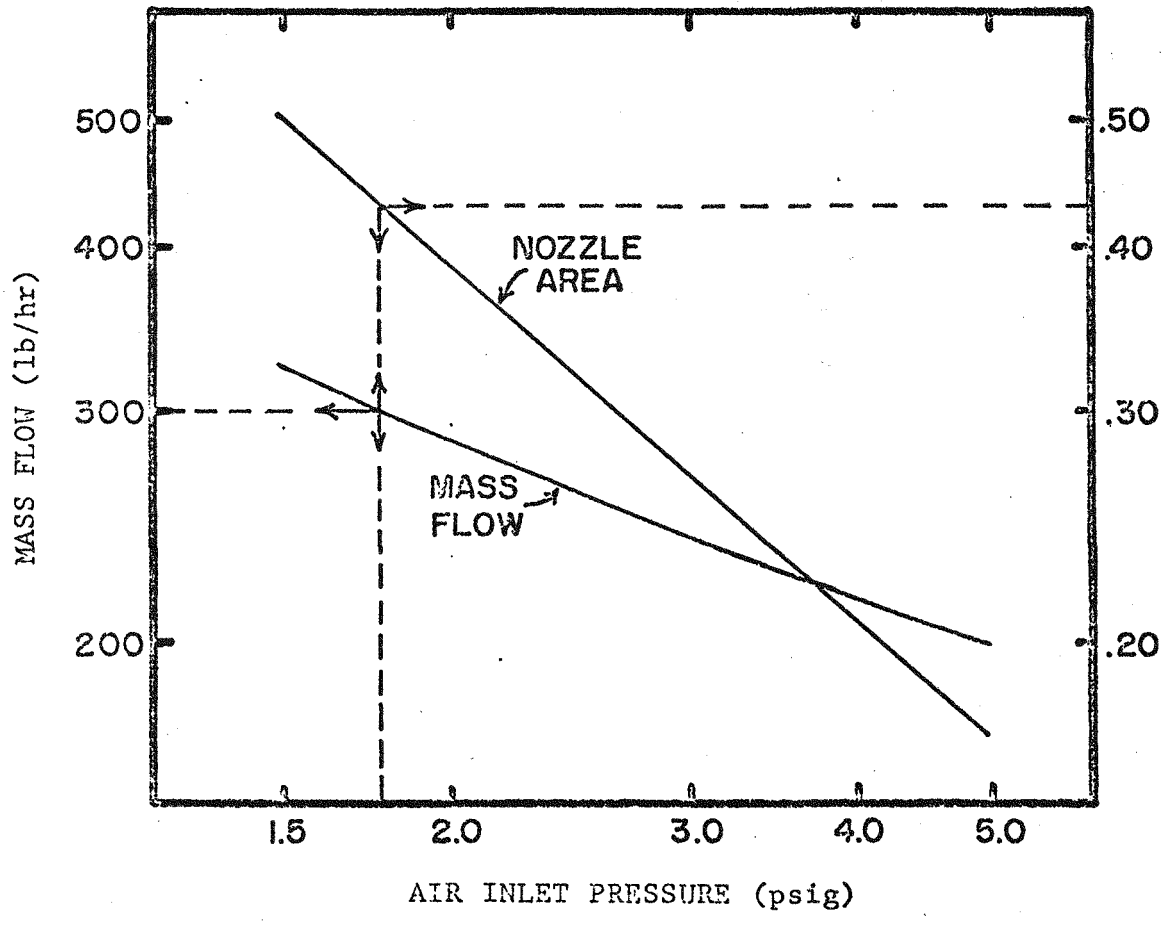
Tests have been conducted to optimize the design and location of the coal feed needle for the PDU. Four coal feed needle configurations have been tested; two straight needles shown in Figure 33, a "Y" shaped needle shown in Figure 34, and a short, straight needle shown in Figure 35.

The first straight needle shown in Figure 33 was used in Test No: PDU-4. The second straight needle shown in Figure 33 was used during Test Nos. 700-1, 2 and 3 with transport air rates of approximately 240 lb/hr and in Test Nos. 700-5 and 6 with transport air rates of approximately 340 lb/hr. During Test No. 700-5, the transport air rate was decreased from 340 to 240 lb/hr to observe the effect. The "Y" shaped needle was used in Test Nos. 700-4, 7, 8 and 9 with a transport air rate of 340 lb/hr.

The short, straight needle used in Test Nos. 700-10 and 700-11, exits 12 in to the right of the center of the boiler and 4 5/8 in above the air distributor. The purpose of this feeder was to prevent an excess amount of coal from being sprayed to the left side of the boiler which occurred with the "Y" feeder.

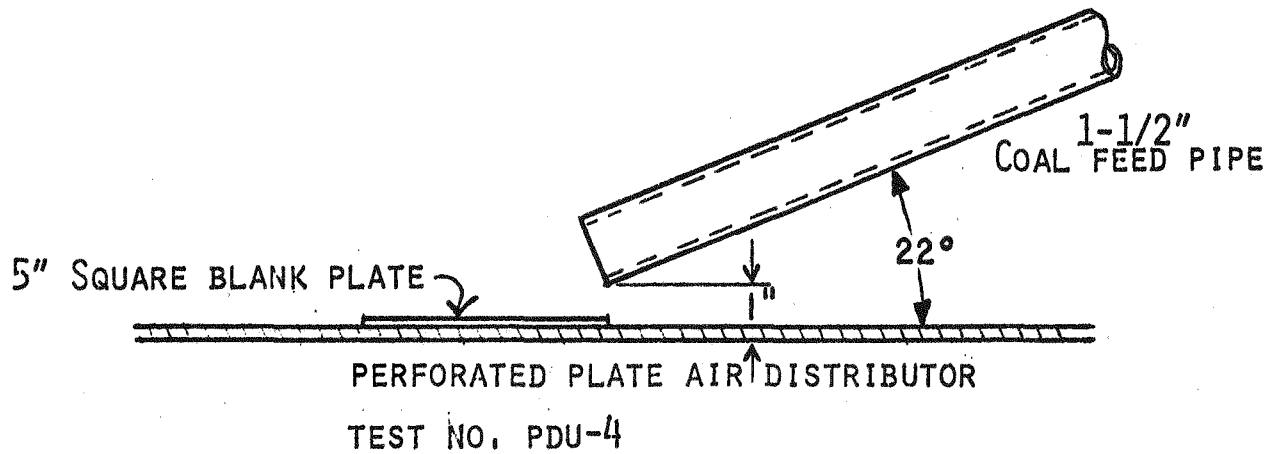
Operating conditions of Test Nos. PDU-4 and 700-1, 2 and 3 are shown in Table I*. Due to a malfunction in the fly ash collection equipment in Test Nos. 700-2 and 3, no combustion efficiencies were obtained for those tests. Tables 42, 43, 44, 45 and 46 show the chemical analyses of the solid samples taken during Test Nos. PDU-4, 700-1, 2, 4, 5 and 6 respectively.

*Shown in Appendix A

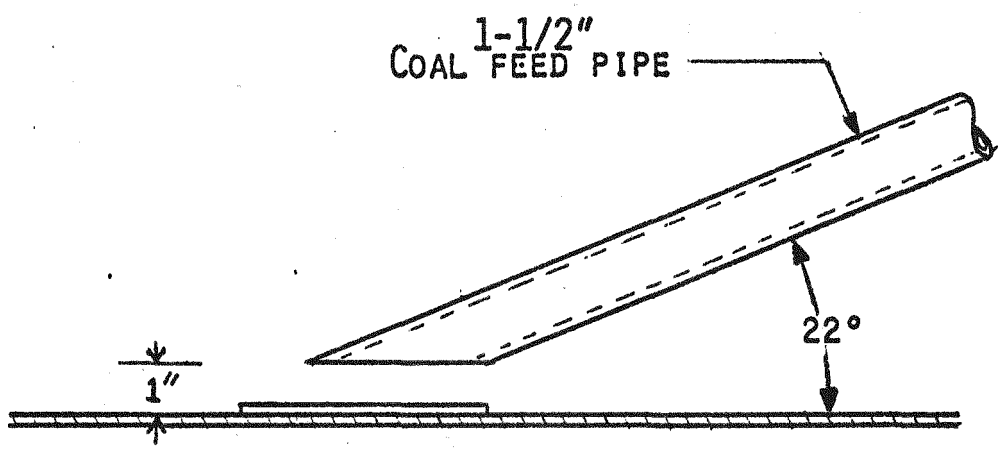


Design procedure for Rivesville eductor. Conditions: 17" w.c. back pressure, 0" w.c. coal inlet pressure

Figure 32



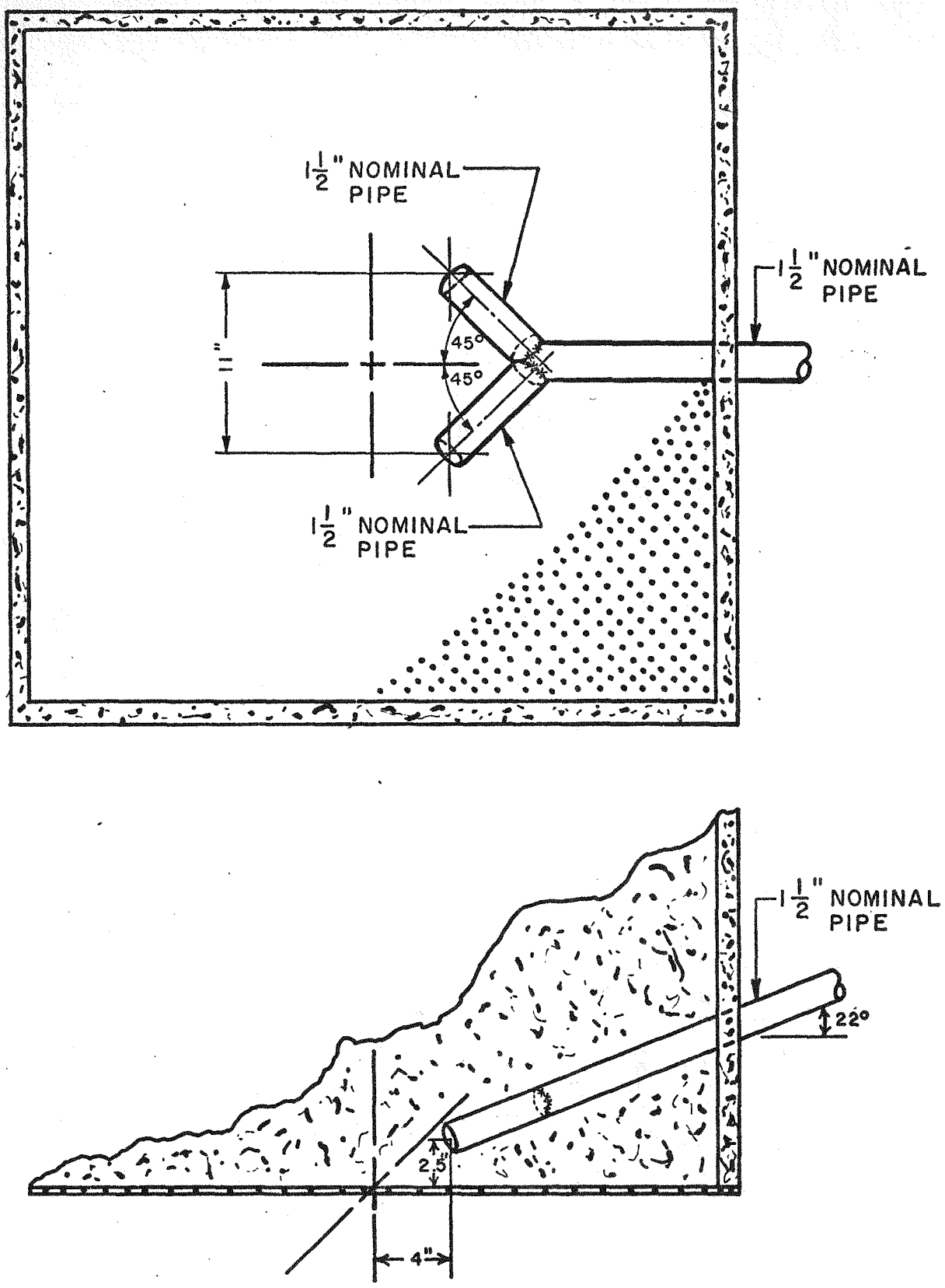
TEST NO. PDU-4



TEST NO'S. 700-1, 2 & 3

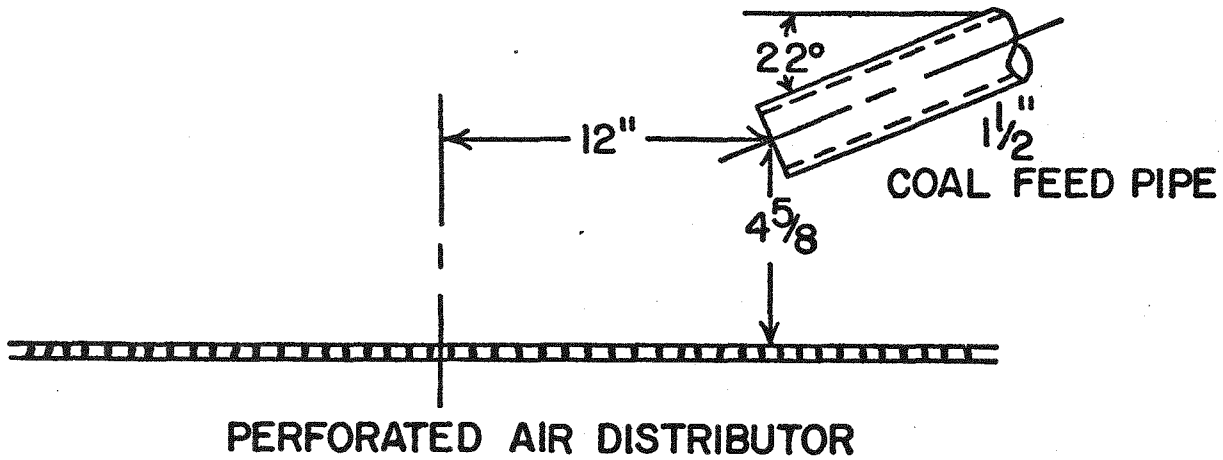
COAL FEEDER CONFIGURATIONS FOR TESTS PDU-4 AND 700-1, 2 AND 3.

Figure 33



"Y" SHAPED COAL FEED NEEDLE USED IN TESTS 700-4, 7 AND 8.

Figure 34



SHORT COAL FEED NEEDLE LOCATION

Figure 35

Table 42

Chemical Analysis of Solid Samples Taken During Test PDU-4
1-31-78

Material	Time (hr)	C (%)	H (%)	Ash (%)	Ca (%)	S (%)
Coal	(composite)	72.91	5.34	11.33	0.59	3.71
Limestone	(composite)	10.90	0.66	61.73	34.27	0.0
Bed	14:30	0.81	0.18	98.98	36.63	9.82
	15:00	0.52	0.18	98.78	34.42	9.31
Fly Ash	14:30	37.30	0.79	60.91	9.50	3.65
	15:00	43.89	0.92	53.42	8.69	3.61
Dust	14:30	25.73	0.34	68.26	9.13	2.25
	15:00	38.75	1.65	55.24	6.77	4.27

Table 43

Chemical Analysis of Solid Samples Taken During Test 700-1
2-22-78

Material	Time (hr)	C (%)	H (%)	Ash (%)	Ca (%)	S (%)
Coal	(composite)	70.56	5.16	13.67	0.0	4.06
Limestone	(composite)	9.54	0.13	62.41	31.31	0.0
Bed	14:00	1.01	0.50	99.22	33.26	9.92
Fly Ash	14:00	43.17	0.78	56.58	7.91	3.89
Dust	14:00	26.43	0.85	70.49	5.27	2.36

Table 44

Chemical Analysis of Solid Samples Taken During Test 700-4
3-15-78

Material	Time (hr)	C (%)	H (%)	Ash (%)	Ca (%)	S (%)
Coal	(composite)	69.34	4.73	13.72	1.43	3.98
Limestone	(composite)	10.51	0.51	62.80	32.07	0
Bed	12:00	0.72	0.38	99.50	37.22	11.62
	13:00	0.95	0.29	99.24	33.11	11.19
	14:00	0.34	0.28	99.27	42.95	10.04
Fly Ash	12:00	15.62	0.57	82.91	19.54	4.52
	13:00	16.10	0.20	83.57	21.25	4.63
	14:00	16.25	0.42	82.65	24.00	5.23
Dust	12:00	17.50	0.28	81.22	8.31	2.45
	13:00	18.68	0.27	81.97	8.27	2.62
	14:00	17.78	0.21	82.01	13.43	3.47

Table 45

Chemical Analysis of Solid Samples Taken During Test 700-5
3-21-78

Material	Time (hr)	C (%)	H (%)	Ash (%)	S (%)
Coal	(composite)	72.30	5.53	13.04	4.16
Limestone	(composite)	11.19	0.93	67.42	0
Bed	12:00	2.17	0.15	98.31	9.12
	13:00	2.67	0.20	99.60	9.27
	14:00	2.26	0.94	98.79	9.10
	15:00	1.57	0.50	99.28	8.73
Fly Ash	12:00	18.67	1.86	82.13	5.90
	13:00	16.27	0.03	83.65	5.07
	14:00	15.40	0.34	83.64	4.93
	15:00	34.49	0.36	64.90	3.11
Dust	12:00	16.42	1.70	80.30	2.96
	13:00	24.19	1.44	81.69	3.13
	14:00	23.43	1.33	83.53	2.74
	15:00	18.76	0.14	64.03	2.12

Table 46

Chemical Analysis of Solid Samples Taken During Test 700-6

3-22-78

Material	Time (hr)	C (%)	H (%)	Ash (%)	S (%)
Coal	(composite)	72.30	5.53	13.32	4.86
Limestone	(composite)	11.19	0.93	62.26	0
Bed	11:00	0.64	1.80	97.25	7.85
	12:00	0.13	0.05	97.94	5.81
	13:00	0.86	0.46	97.55	7.85
	14:00	0.50	0.38	97.70	8.40
	15:00	0.22	0.36	98.44	8.36
Fly Ash	11:00	16.99	0.67	83.37	5.43
	12:00	14.16	0.81	85.57	4.85
	13.00	21.23	0.11	74.12	4.56
	14:00	15.87	1.79	85.70	5.46
	15:00	13:30	0	84.37	5.91
Dust	11:00	20:01	0.98	81:77	3.10
	12:00	21.83	2.08	82.26	2.70
	13:00	17.94	1.66	79.78	2.45
	14:00	14.43	0.64	82.30	2.27
	15:00	24.23	1.95	76.01	2.26

Based on the data obtained so far, it appears that an important parameter in designing a coal feed needle is the fuel to air ratio. A deficiency of air in the needle can result in reduced combustion efficiency. This is more evident from Test No. 700-5, shown in Table II* where the data taken at 1400 hrs had a fuel to air ratio of 2.2 lb coal/lb air and the data taken at 1500 hrs was with a fuel to air ratio of 3.2 lb coal/lb air. Table 45 shows that the carbon content of the fly ash increased from 15.4 percent at 1400 hrs to 34.5 percent at 1500 hrs. The combustion efficiency dropped from 91 percent to 85 percent during that time. The data shows that with proper fuel to air ratio, the straight needle performed about as well as the "Y" feeder.

Gas Sample Traverse

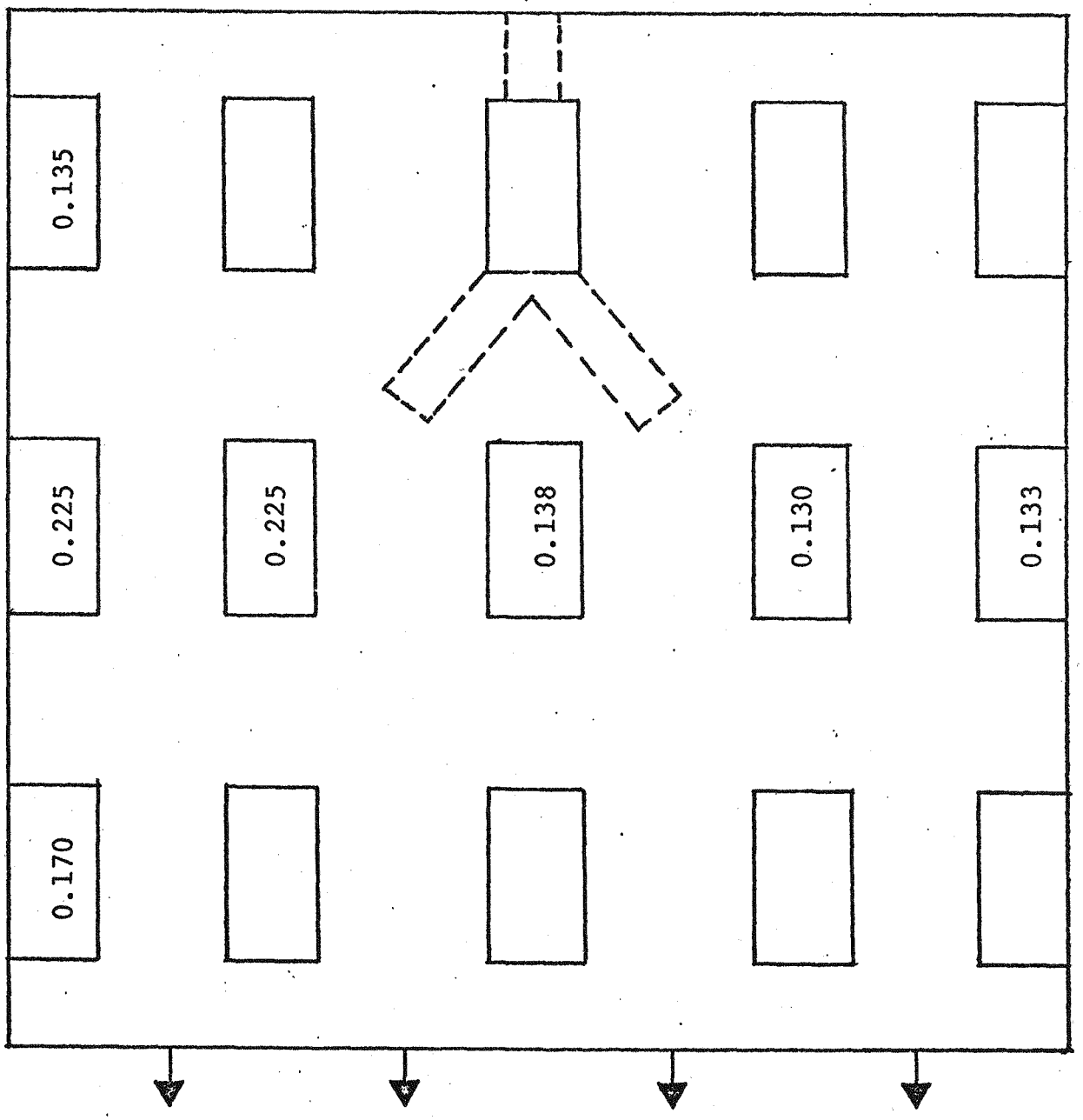
To further study the effects of coal feeder type and location, a traverse arrangement was added to the flue gas sample probe. Figure 2 shows the external piping arrangement used to position the probe. By rotational and linear movement of the probe, the freeboard can be sampled in a two dimensional pattern parallel to the air distributor, between the upper tube bundle and the steam drum. The traversing assembly was used in Test Nos. 700-7 and 700-8 with the "Y" feeder to obtain gas sample profiles. Due to the time required to do a traverse (usually 15 minutes is required at each position) each test represents a single traverse. Care was taken during the tests to maintain steady conditions during the time of traverse.

In Test No. 700-7, a traverse across the center of the freeboard from front to back was made. Figure 36 shows the CO concentration in the sample position for this test. The traverse indicated higher CO concentrations in the back half of the boiler, which corresponds to the location of the in-bed economizer bundle. The traverse for oxygen showed the opposite trend, as expected. The traverse showed no apparent trend for NO_x. During the sampling period, the fixed position Westinghouse O_2 probe read between 2.5 and 2.7 percent O_2.

In Test No. 700-8, thirteen locations were sampled in the freeboard. Figures 37, 38 and 39 show the concentration profiles for CO, hydrocarbons, and NO_x respectively. Due to the length of time required to complete the traverse, the fixed position O_2 readings fluctuated during the traverse. Figure 40 shows the fixed position O_2 readings obtained at the time the sample probe was at the location indicated. Most of the readings were taken when the fixed O_2 probe was between 3.0 and 3.4 percent. However, three samples were obtained when the oxygen was between

*Shown in Appendix A

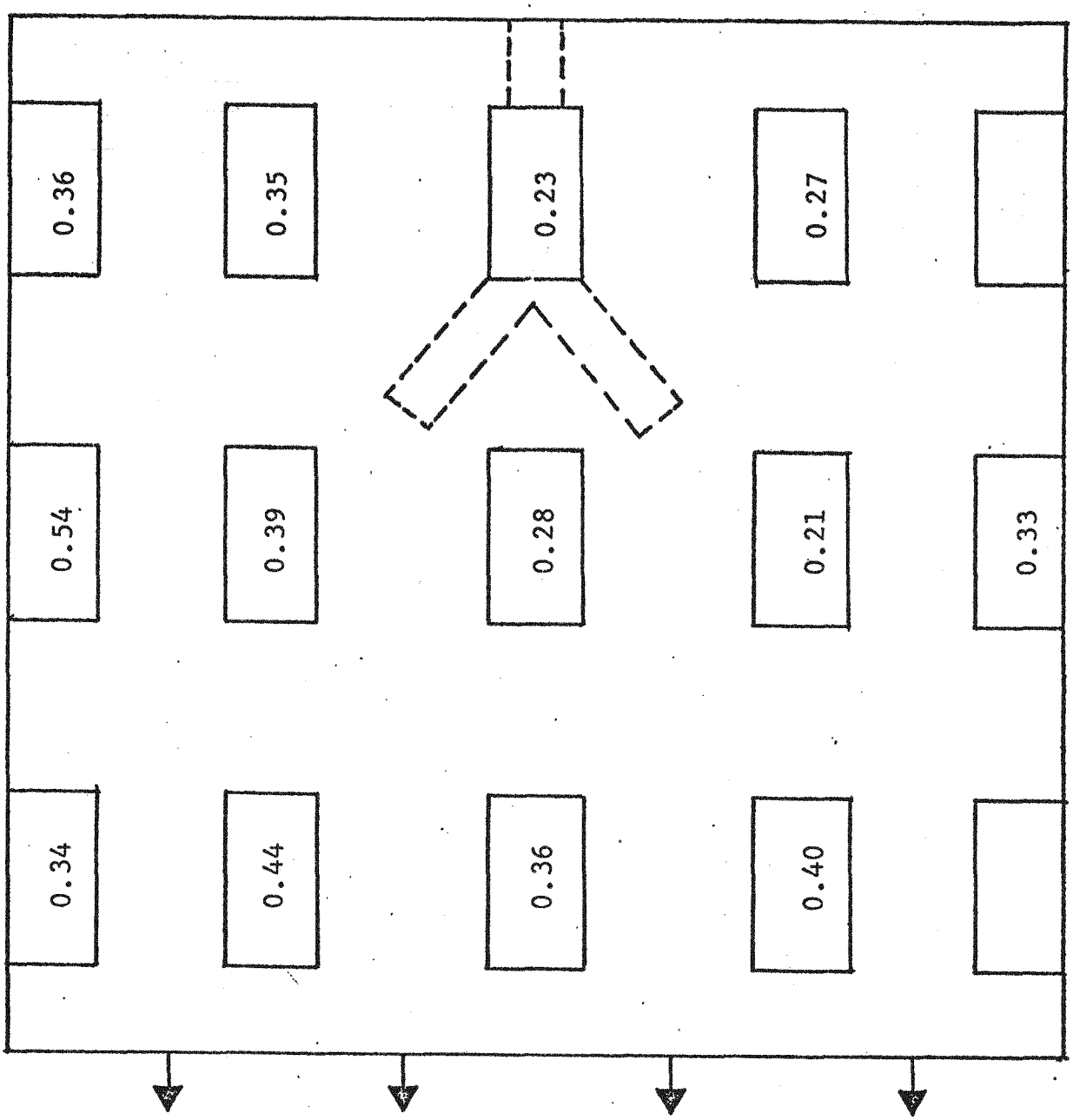
RELATIVE GAS SAMPLING POSITIONS PLAN VIEW



Test 700-7 Gas Sample Traverse Pattern for CO Volume %.

Figure 36

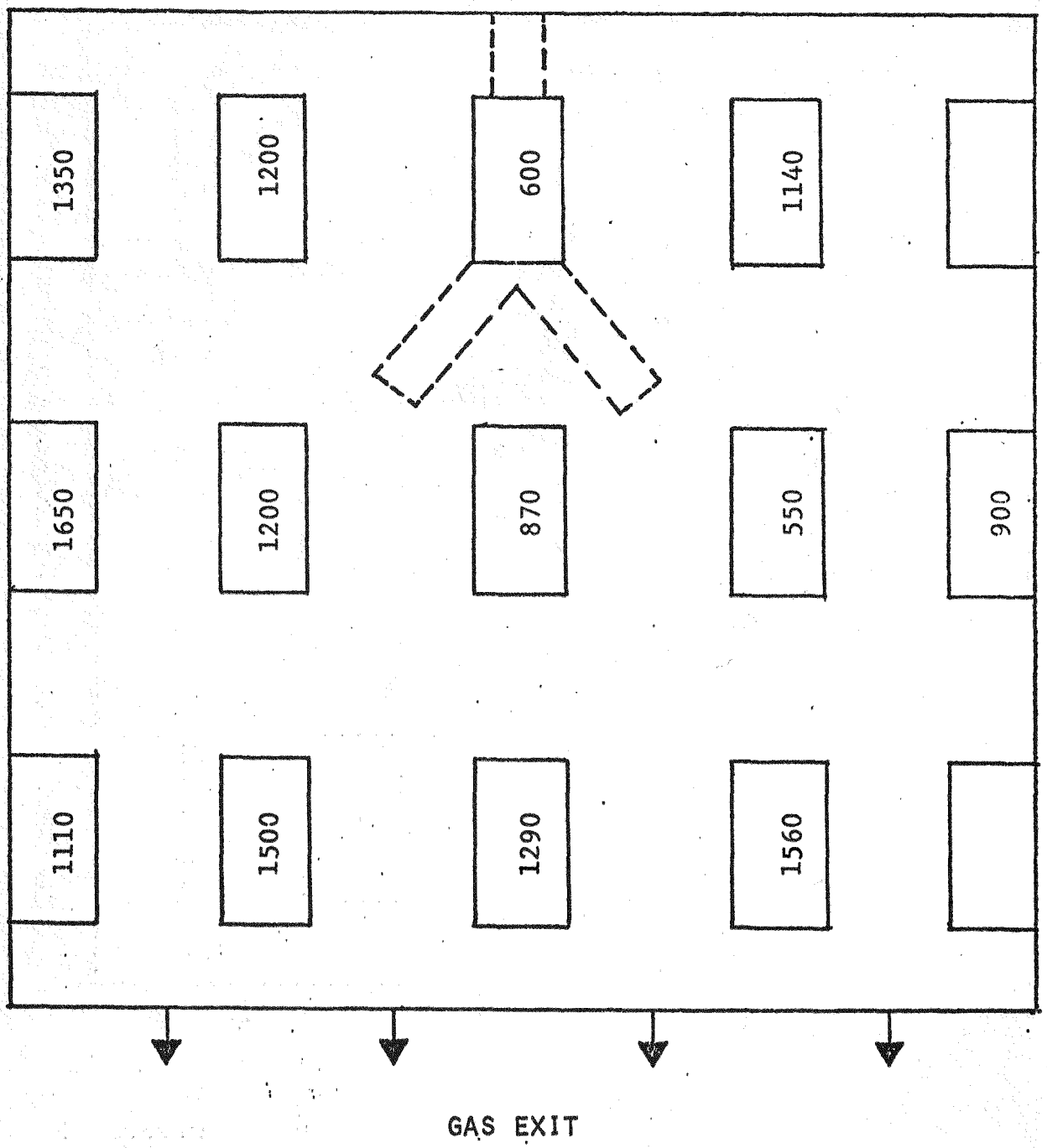
RELATIVE GAS SAMPLING POSITIONS PLAN VIEW



Test 700-8 Gas Sample Traverse for CO Volume %.

Figure 37

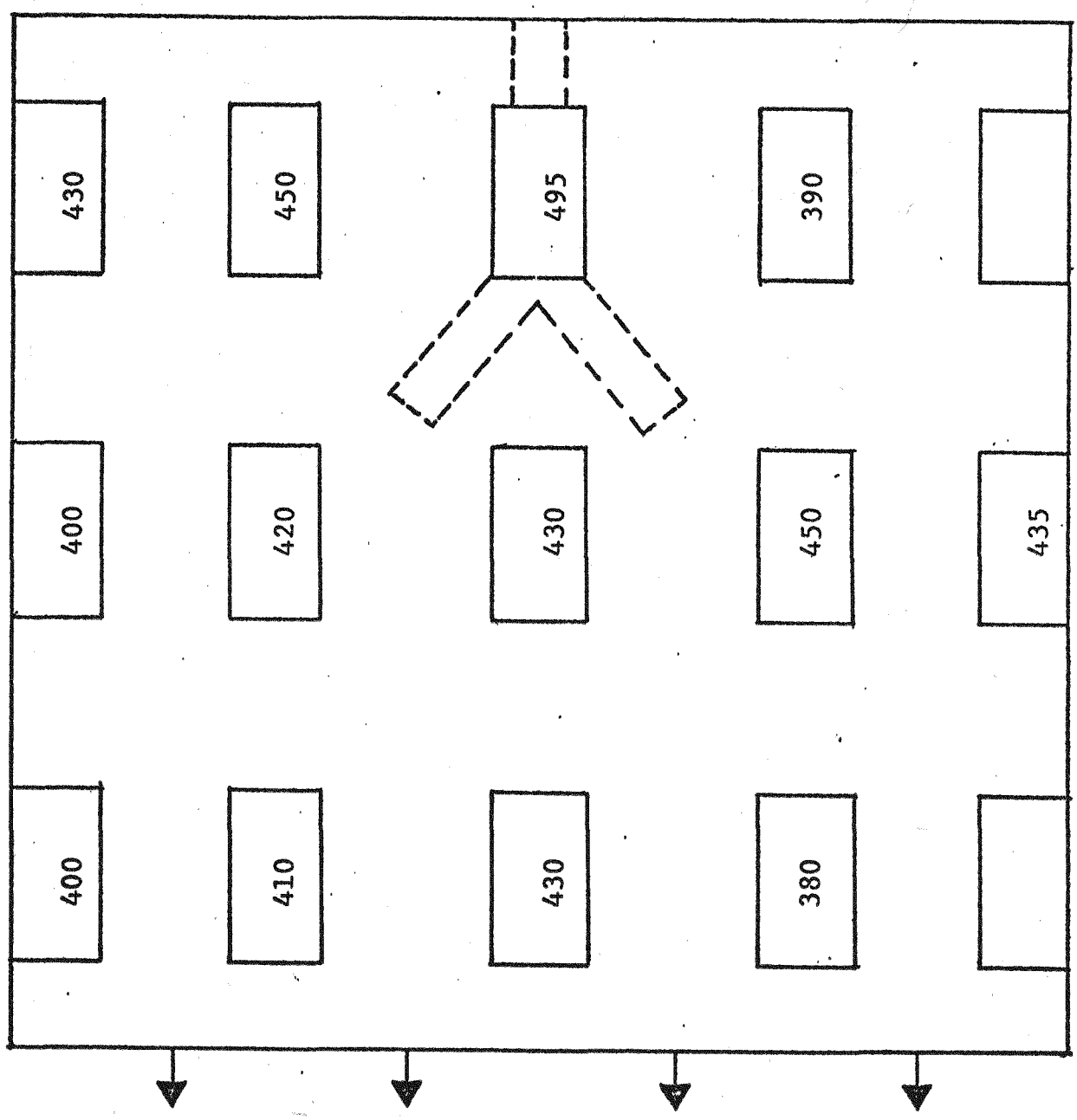
RELATIVE GAS SAMPLING POSITIONS PLAN VIEW



Test 700-8 Gas Sample Traverse for Hydrocarbons ppm.

Figure 38

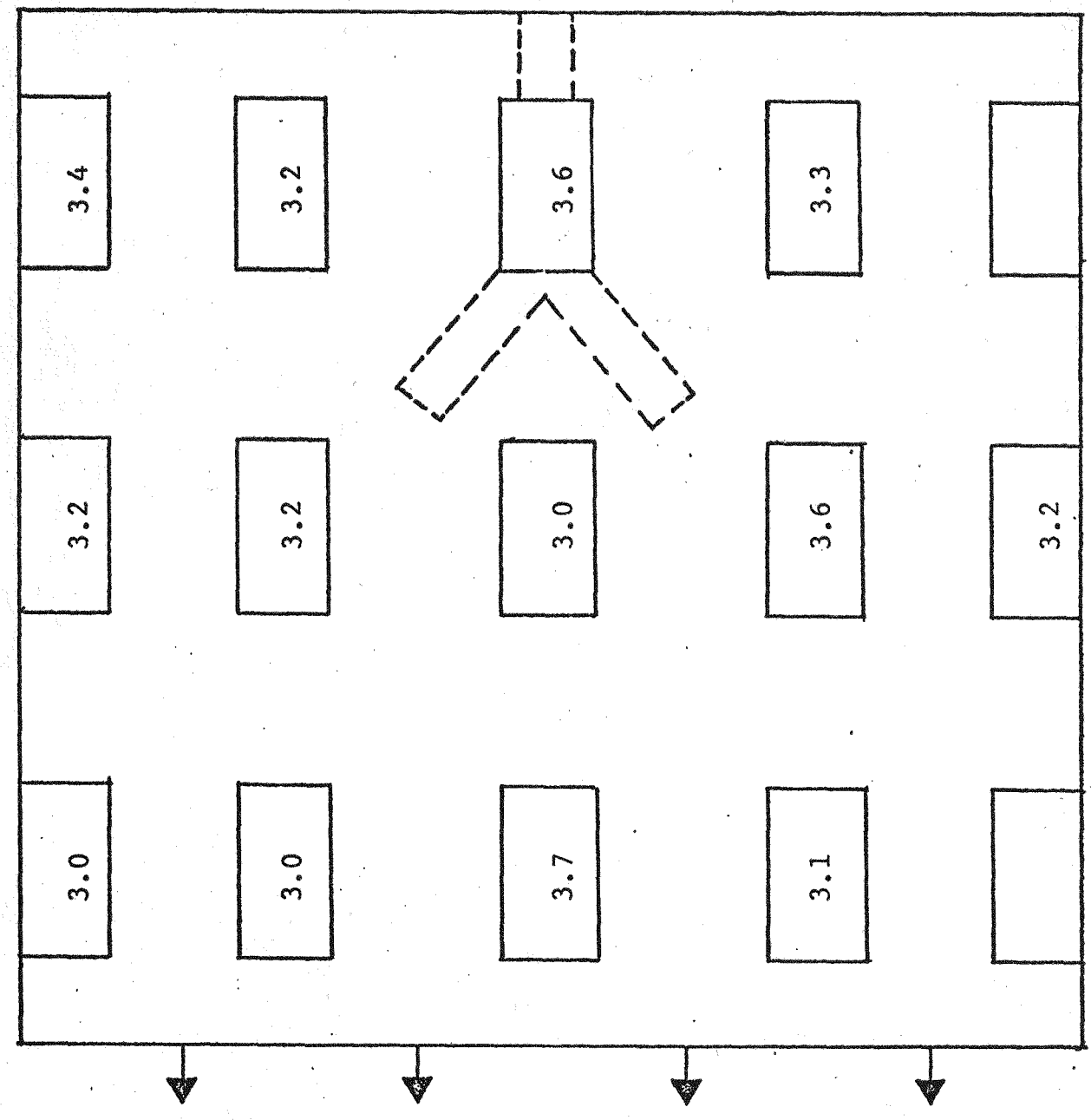
RELATIVE GAS SAMPLING POSITIONS PLAN VIEW



Test 700-8 Gas Sample Traverse for NO_x ppm.

Figure 39

RELATIVE GAS SAMPLING POSITIONS PLAN VIEW



Test 700-8 Fixed O_2 Probe Value During Sampling at the Various Locations.

Figure 40

3.6 and 3.7 percent. The same trends observed in Test No. 700-7 were again observed for Test No. 700-8. In addition, CO, hydrocarbons, CO_2 and SO_2 were found to be higher and the O_2 was found to be lower along the left side of the boiler. This indicates excessive fuel burning along this side of the bed which is expected since the coal feeder sprays the coal to the left side of the boiler. Figure 39 shows no observable trend in NO_x values.

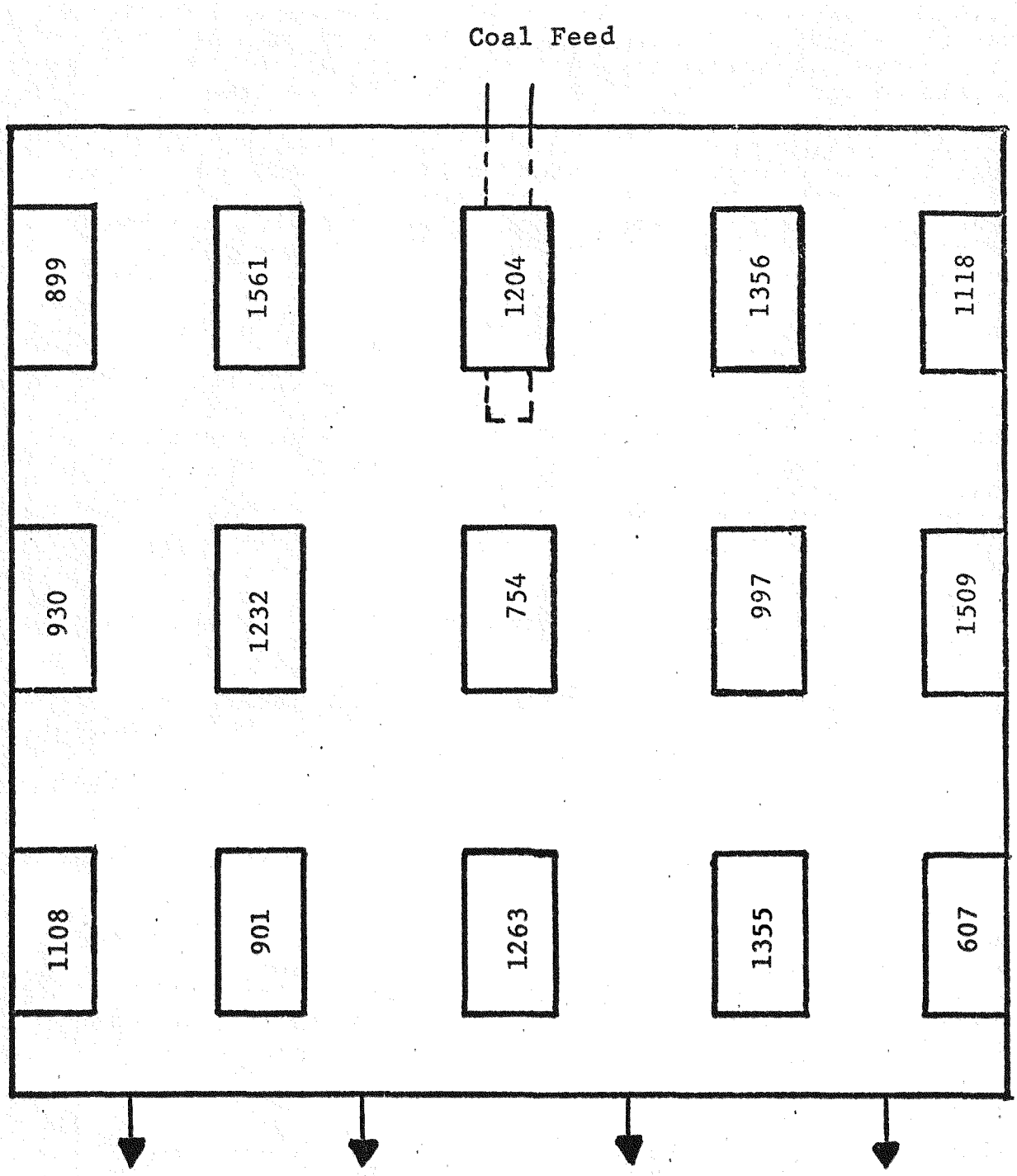
Test No. 700-9 used the "Y" feeder at 6 ft/sec. superficial velocity. Test conditions for the coal feed tests are given in Table III*. As noted later in this section, combustion efficiency for Test No. 700-9 is uncertain due to fly ash sampling discrepancies.

During Test No. 700-9, a gas sample traverse was performed. The same trends were observed for this test as was observed in Test Nos. 700-7 and 8. The CO, hydrocarbons, CO_2 and SO_2 were again found to be higher and the O_2 lower along the left side of the boiler where the coal is sprayed.

Figure 41 shows the gas sample traverse for hydrocarbons taken during Test No. 700-10 using the short, straight needle. From the data, there is no trend from the right side to the left side of the boiler. Most of the differences in the point readings are due to differences in the overall operating conditions during the test. However, the hydrocarbons are higher than was previously obtained and it appears that this needle configuration gives more uniform distribution of coal to the entire bed area compared to needles tested thus far. The combustion efficiencies shown in Table III* indicate no apparent loss of efficiency as a result of the short needle. Test No. 700-11 showed a similar gas distribution as Test No. 700-10. Further tests will be conducted with the short needle to reduce the amount of hydrocarbons produced.

The effect of flue gas composition on combustion efficiency can be estimated using the operating conditions obtained from Test No. 700-10 at 1500 hrs. Using the highest observed values of CO and hydrocarbons during the traverse, gives a combustion efficiency of 90.9 percent. Using the lowest observed values for the traverse at the same operating conditions, yields a combustion efficiency of 92.5 percent. The average values of the CO and hydrocarbons during the traverse gives a combustion efficiency of 91.7 percent. The position of the probe may effect the measured combustion efficiency by a maximum of about 2 percent. It should be noted, however, that some of the variations in gas readings were due to fluctuations in operating

*Shown in Appendix A



Test 700-10 Gas Sample Traverse for Hydrocarbons ppm.

RELATIVE GAS SAMPLING POSITIONS PLAN VIEW

Figure 41

conditions rather than to positional variations. The actual difference in combustion efficiency due to sample location would be less than 2 percent. In either case, this is within the accuracy of the combustion efficiency measurement. The gas sample probe can be located in the center of the bed for most tests without adversely affecting results.

An error was discovered in the fly ash sampling technique used in some tests. Prior to Test No. 700-7, the fly ash was sampled outside of the combustion laboratory using diverter valves. In this way, the entire fly ash stream was diverted to the sample bag and was representative of fly ash being produced. However, due to a malfunctioning diverter valve, fly ash samples were taken using a small pipe that was welded tangentially to an elbow in the fly ash transport line during Test Nos. 700-7, 8 and 9.

When fly ash from these tests was analyzed, unusually high carbon contents were noted. For this reason during Test No. 700-10, fly ash was sampled at both locations during one of the sample periods. Chemical analysis showed the sample taken at the tangent of the elbow had a carbon content of 34.1 percent while the fly ash sampled by diverting the entire stream had a carbon content of only 16.7 percent. It appears that the tangential elbow sampler was sampling only a certain size fraction of the fly ash which was not representative of the total stream. Because of this discrepancy, no combustion efficiencies are reported for Test Nos. 700-7, 8 and 9.

2.3.6 PDU Performance Characterization Tests

Testing of the PDU has been directed toward comparing its characteristics with the FBM. The purpose of this comparison is to identify the process related effects of boiler configuration and to verify process characteristics which are believed to be independent of bed geometry. For most of the comparisons, the data from Test No. 621 on the FBM was used. This test was a well documented test with the horizontal tube bundle. Table 47 shows the summary of operating conditions for Test No. 621. Table 48 shows the chemical compositions of the solid samples from Test No. 621.

Combustion Efficiency

Combustion efficiencies for the PDU range between 5 and 10 percent higher than those for the FBM. The combustion efficiency for Test No. 621 was 81.9 percent. The higher superficial velocity (14.6 ft/sec) could be responsible for the low efficiency. Typical combustion efficiencies

Table 47

Summary of Operating Conditions for Test 621

Test No.	621
Date of Test	4/24/75
Time	15:00
Bed temp., °F	1491
O_2 in flue gas, %	3.4
Coal feed rate, lb/hr	723
Fly ash collection rate, lb/hr	161
Limestone feed rate, lb/hr	291
Bed drain rate, lb/hr	208
Stack dust rate, lb/hr	50.6
Main air flow rate, lb/hr	7017
Total air flow rate, lb/hr	7337
Superficial velocity, ft/sec	14.6
Static bed depth, inches	18.0
SO_2, ppm	1200
Ca/S	2.5
Combustion efficiency, %	81.9
Sulfur Capture	67.9
lb $SO_2/10^6$ Btu	2.49

Table 48

Chemical Analysis of Solid Samples Taken During Test 621
4-24-75

Material	Time (hr)	C (%)	H (%)	Ash (%)	Ca (%)	S (%)
Coal	(composite)	70.1	4.68	13.59		4.04
Limestone	(composite)					
Bed	15:00	0.6	0.06		36.81	7.57
Fly Ash	15:00	50.39	0.99	49.61	7.02	2.97
Dust	15:00	37.55	0.90	62.45	6.32	2.61

for the FBM range between 80 and 87 percent. From Tables I-VI*, it can be seen that combustion efficiencies range from 86 to 93 percent when a proper fuel/air ratio is maintained in the feed needle.

The most dramatic difference between the two units is seen in the carbon content of the fly ash. Table 48 shows the carbon content of the FBM fly ash to be 50.4 percent. Typical fly ash values from the FBM ranged from 35 to 50 percent carbon. In the PDU, the fly ash carbon content ranged from 12 to 20 percent with proper fuel/air ratios in the needle and between 35 and 43 percent when the fuel/air ratio was high. This decrease in fly ash carbon content is probably due to the higher freeboard in the PDU which allows the fly ash more time in the combustion zone to burn out.

Sulfur Capture

Test Nos. 700-6 had a steady Ca/S feed ratio for a sufficiently long period to obtain reliable data on sulfur capture. From Table II* it can be seen that for Test No. 700-6, sulfur capture percentages greater than 86 percent were achieved with Ca/S molar ratios of 2.5. This is slightly higher than the results obtained from the FBM. There is insufficient data at this time, however, to determine if the boiler configuration is responsible.

Test No. 700-10 was the first of a series of long duration tests that are planned to study sulfur capture. These tests will be run at steady conditions for a minimum of 16 hours to insure steady state bed composition. Test No. 700-10 lasted 17 hours at an average Ca/S molar feed ratio of 3.1 mole Ca/mole S. At the end of the test, SO_2 in the stack was 850 ppm. This corresponds to a percent sulfur capture of 80.7 percent. The percent sulfur capture is defined as the pounds of sulfur in the fuel less the pounds of sulfur emission in the flue gas as SO_2.

Heat Transfer

The outside coefficients in the characterization tests ranged from 39.4 to 49.6 Btu/hr/ft^2/°F. When comparing heat transfer of the FBM and the PDU, several factors should be considered. The most important variables are bed mean particle size, superficial velocity, and tube diameter. In Test No. 700-4, the bed mean particle size was 1350 microns, the superficial velocity was 12 ft/sec at 1400 hr , and the outside coefficient for the economizer was measured at 45.8 Btu/hr/ft^2/°F. The only data from the FBM at bed particle sizes of 1350 microns

*Shown in Appendix A

was obtained at 14.6 ft/sec to be 36.5 Btu/hr/ft^2/°F. Data taken at 700 microns shows that the outside coefficient will increase by 4.2 percent when the velocity is reduced to 12 ft/sec. Thus the outside coefficient in the FBM should be 38.0 Btu/hr/ft^2/°F. The Vreedenberg[1] equation for correlating the heat transfer coefficient between a fluidized bed and an immersed horizontal tube shows the outside coefficient to be inversely proportional to $d_t^{0.56}$ where d_t is the outside diameter. The tubes in the FBM were 2 in O.D. while the tubes in the PDU are 1.5 in O.D. The heat transfer coefficients for the PDU should be larger by a factor of $\left(\frac{2.0}{1.5}\right)^{56} = 1.175$. Based on the FBM results, the outside coefficient for the PDU for this test should be 1.175 x 38.0 Btu/hr/ft^2/°F = 44.6 Btu/hr/ft^2/°F. This is in good agreement with the observed value of 45.8 Btu/hr/ft^2/°F and indicates that a Vreedenberg type of equation may be useful in correlating outside heat transfer coefficients. More complete heat transfer results are given in section 2.4.1.

Complete Heat and Material Balance

Test No. 700-19 was conducted to provide data on sulfur capture at steady state conditions, and to collect the data necessary for a complete heat and material balance on the PDU. The operating conditions for this test, Table VI*, were set up so that the results can be compared to previous FBM test results. This will be done in an attempt to determine what significant influence the combustor geometry may have on the process performance.

Over a sixteen hour test period, samples of coal and limestone were taken every four hours. These samples were screened and and analyses for carbon, hydrogen, sulfur, ash, calcium, iron and water are being performed on each of the samples. A sample of coal from this test was coked; this sample is being analyzed to determine the sulfur, carbon, hydrogen, ash, calcium and iron. At the end of the test period after steady state conditions had been maintained for eleven hours, samples of fly ash, bed material, and dust were taken. These samples have been screened and are being analyzed by size distribution. The size distribution for each material are given in Tables 49 through 53.

[1] H. A. Vreedenberg, J. Appl. Chem. (London), 2, Suppl. 1, S26 (1952); Chem. Eng. Sci., 9, 52 (1958); 11,274 (1960).

*Shown in Appendix A

Table 49

Size Distribution of Sewickley Coal

MEAN PARTICLE SIZE CALCULATIONS

TEST NO. 700-19
TIME 0
COAL
SAMPLE WEIGHT BEFORE SCREENING 288.90(GRAMS)
SAMPLE WEIGHT AFTER SCREENING 287.15(GRAMS)
PERCENT DIFFERENCE .606%

U.S.S.	WT(GRAMS)	FRACTION (XI)	FRACTION SMALLER THAN SMALLEST SCREEN	MEAN PARTICLE DIAMETER
-.265"+.25"	1.65	.0057	.99	6525.
-.25"+ 6	83.25	.2899	.70	4855.
- 6+ 8	49.20	.1713	.53	2870.
- 8+ 12	40.65	.1416	.39	2030.
- 12+ 16	26.50	.0923	.30	1435.
- 16+ 20	23.15	.0806	.22	1015.
- 20+ 30	16.65	.0580	.16	715.
- 30+ 40	12.35	.0430	.12	505.
- 40	33.75	.1175	0.	210.

SAMPLE MEAN PARTICLE SIZE (MICRONS) 943.64
SAMPLE MEAN PARTICLE SIZE (INCHES) .03715

Table 50

Size Distribution of Greer Limestone

MEAN PARTICLE SIZE CALCULATIONS

TEST NO. 700-19
TIME 0
LIMESTONE
SAMPLE WEIGHT BEFORE SCREENING 350.50(GRAMS)
SAMPLE WEIGHT AFTER SCREENING 350.33(GRAMS)
PERCENT DIFFERENCE .049%

U.S.S.	WT(GRAMS)	FRACTION (XI)	FRACTION SMALLER THAN SMALLEST SCREEN	MEAN PARTICLE DIAMETER
-.265"+.25"	.30	.0009	1.00	6525.
-.25"+ 6	71.35	.2037	.80	4855.
- 6+ 8	170.40	.4864	.31	2870.
- 8+ 12	82.50	.2355	.07	2030.
- 12+ 16	9.75	.0278	.05	1435.
- 16+ 20	3.50	.0100	.04	1015.
- 20+ 30	.90	.0026	.03	715.
- 30+ 40	.53	.0015	.03	505.
- 40	11.10	.0317	0.	210.

SAMPLE MEAN PARTICLE SIZE (MICRONS) 1944.51
SAMPLE MEAN PARTICLE SIZE (INCHES) .07656

Table 51

Size Distribution of Spent Bed Material

MEAN PARTICLE SIZE CALCULATIONS

TEST NO. 700-19
TIME 1530
BED
SAMPLE WEIGHT BEFORE SCREENING 321.42(GRAMS)
SAMPLE WEIGHT AFTER SCREENING 321.93(GRAMS)
PERCENT DIFFERENCE -.159%

U.S.S.	WT(GRAMS)	FRACTION (XI)	FRACTION SMALLER THAN SMALLEST SCREEN	MEAN PARTICLE DIAMETER
-.265"+.25"	3.60	.0112	.99	6525.
-.25"+ 6	45.00	.1398	.85	4855.
- 6+ 8	113.40	.3523	.50	2870.
- 8+ 12	91.63	.2846	.21	2030.
- 12+ 16	38.50	.1196	.09	1435.
- 16+ 20	23.10	.0718	.02	1015.
- 20+ 30	5.65	.0176	.00	715.
- 30+ 40	.65	.0020	.00	505.
- 40	.40	.0012	0.	210.

SAMPLE MEAN PARTICLE SIZE (MICRONS) 2074.93
SAMPLE MEAN PARTICLE SIZE (INCHES) .08169

Table 52

Size Distribution of Fly Ash

MEAN PARTICLE SIZE CALCULATIONS

TEST NO. 700-19
TIME 1530
FLYASH
SAMPLE WEIGHT BEFORE SCREENING 24.50(GRAMS)
SAMPLE WEIGHT AFTER SCREENING 24.61(GRAMS)
PERCENT DIFFERENCE -.449%

U.S.S.	WT(GRAMS)	FRACTION (XI)	FRACTION SMALLER THAN SMALLEST SCREEN	MEAN PARTICLE DIAMETER
- 6+ 16	.60	.0244	.98	2275.
- 16+ 30	3.81	.1548	.82	890.
- 30+ 40	3.10	.1260	.69	505.
- 40+ 80	5.40	.2194	.48	299.
- 80+100	1.12	.0455	.43	163.
-100+140	2.70	.1097	.32	127.
-140+200	4.50	.1829	.14	89.
-200+400	2.90	.1178	.02	56.
-400	.48	.0195	0.	19.

SAMPLE MEAN PARTICLE SIZE (MICRONS) 133.26
SAMPLE MEAN PARTICLE SIZE (INCHES) .00525

Table 53

Size Distribution of Dust

MEAN PARTICLE SIZE CALCULATIONS

TEST NO. 700-19
TIME 1530
DUST
SAMPLE WEIGHT BEFORE SCREENING 1.70(GRAMS)
SAMPLE WEIGHT AFTER SCREENING 1.60(GRAMS)
PERCENT DIFFERENCE 5.882%

U.S.S.	WT(GRAMS)	FRACTION (XI)	FRACTION SMALLER THAN SMALLEST SCREEN	MEAN PARTICLE DIAMETER
-140+200	.50	.3125	.69	89.
-200+400	1.05	.6563	.03	56.
-400	.05	.0313	0.	19.

SAMPLE MEAN PARTICLE SIZE (MICRONS) 59.19
SAMPLE MEAN PARTICLE SIZE (INCHES) .00233

The analysis of coal, limestone, coke, coal ash, fly ash, bed material and dust size cuts for carbon , hydrogen, sulfur, ash, calcium, iron, and water will permit the results to be evaluated by the simultaneous component balance technique presented previously in Interim Report No. 3. Using this technique, the distribution of the elements can be better defined; therefore, the final percentages of carbon, hydrogen, sulfur, ash, calcium, iron and water can be associated with initial macroscopic components, i.e., carbon in the form of carbon dioxide from uncalcined limestone versus carbon in the form of coal. These balances will give the distribution of coal, coal ash, coke, limestone, calcined (but unsulfated) limestone and sulfated limestone in the various size cuts of the bed and the fly ash.

2.3.7 Air Distributor Tests

A test program of evaluating protective air distributor materials suitable for installation over the perforated metal plate air distributors at Rivesville was conducted at the Alexandria laboratory.

It has been reported that the grid plates in the Rivesville MFB deformed and cracked in several places. A study of samples of the grid plates conducted by Foster Wheeler Corporation, concluded that buckling resulted from grid plate restriction during thermal expansion and exposure to excessive operating temperatures.

Protective materials tested include trap rock, porous ceramic plates, 1 1/2 in alumina stones and 3/4 in alumina spheres. A 3 in layer of 3/4 in alumina spheres with an expanded metal hold down screen proved to be better than other materials.

Trap Rock

A suggested method to reduce grid plate temperature was to place a layer of rocks on the grid plate which would not fluidize at operating conditions. The static rock layer would prevent direct contact of the hot bed on the grid plate, thereby insulating the grid plate from the fluidized bed environment.

In order to determine the feasibility of this approach, a test program was initiated at the Alexandria laboratory. For the layer of rocks to perform successfully they must satisfy several requirements:

- remain static and not shift on the air distributor,
- withstand bed temperature,
- be non-reactive,
- not adversely affect fluidization.

To estimate the size of rocks which would not fluidize, a minimum fluidizing velocity calculation was performed using the modified Ergun equation. For rocks having a specific gravity of 2.96 and mean particle size of 1 in the minimum fluidizing velocity is 31.2 ft/sec. Due to these high fluidizing velocities, a 1 in to 2 in rock size appeared to be appropriate.

A survey of the quarries, crushed stone companies and construction materials companies in the northern Virginia, Washington, D.C. and Baltimore areas, resulted in a selection of eight samples of rocks to be tested. The rocks were furnished by Luck Quarries Incorporated of Leesburg, Virginia.

Several of the largest stones of each sample were put in crucibles, weighed and left in the chemistry laboratory muffle furnace for sixteen hours at 1500°F. The trap rock sample was wet, having been removed from a snow covered pile. To remove surface moisture, it was pre-heated for half an hour at 212°F before weighing. The samples were weighed after heat treating to determine weight loss. Based on the weight loss, the trap rock and river gravels appeared suitable for fluidized bed application.

The first three hot tests were conducted under normal operating conditions. The purpose of Test Nos. 700-1, 700-2 and 700-3 was to provide baseline grid plate temperature data to evaluate the effectiveness of protecting the grid plate with a layer of stone. The results of these tests are shown in Table I*.

To measure grid plate temperature, five thermocouples were installed in contact with the bottom side of the grid plate and cemented in place with a heat transfer cement. In all three tests, three of the thermocouples broke loose from the grid plate and read considerably lower than the two remaining in contact. For this reason the grid plate temperature was taken to be the average of the two highest temperature readings. In order to prevent this problem in future tests, thermo-couples were welded to a small metal plate. This plate was then bolted to the grid plate with a layer of heat transfer cement between the grid plate and the thermo-couple plate.

During all three tests, the steam air preheater was in operation to provide a plenum temperature of 280°F. The observed grid plate temperature ranged from 1048°F to 1168°F. These grid plate temperatures are in general agreement with the more extensive test air distributor operating temperatures reported in Interim Report No. 3.

*Shown in Appendix A

Test No. 700-4 was conducted to determine the effectiveness of protecting the grid plate with a 2 in deep layer of trap rock in the size range of 2 in x 1 in During this test, gas superficial velocities ranged from 11 to 12 ft/sec. Air preheat for this test was 280°F. Grid plate temperatures averaged 663°F for this test with thermocouple readings ranging from 612°F to 719°F. This temperature average is at least 300°F lower than the grid plate temperatures measured in Test Nos. 700-5, 700-6, 700-7 and 700-8, using the same thermocouples with no rocks on the grid plate. The measured average grid plate temperature for these tests are shown in Table II*.

At the end of Test No. 700-4, the bed material was vacuumed out of the PDU to expose the layer of rocks. It was found that some rocks had piled up in the corners of the unit and under the coal feed needle. More importantly, a large area of grid plate (approximately 15 percent of the total area) was completely uncovered. Since this area was not above any of the grid plate thermocouples, the recorded temperatures did not reflect the exposed conditions. This indicates that in order for rocks to effectively protect the entire grid plate, some method of restricting their motion is required. Additionally, about 5 percent of the rocks had cracked apart.

Porous Ceramic Plates

The next tests used permeable ceramic plates on top of the grid plate. The expected advantages of this composite material air distributor are; reduced grid plate temperature and dimensional stability of the insulating layer material. A disadvantage of this type of composite, is the potential problem of plugging the pores of the permeable ceramic with airborne dirt or carryover from an air preheater.

A piece of the porous ceramic plate selected for the air distributor tests was used to filter a stream of flue gas from the PDU for about one hour. Particulates in the flue gas did not plug the ceramic plate.

Prior to the test using the ceramic plates, it was found that leaks had developed around the grid plate. This discovery necessitated a revision to the grid plate peripheral seal design. Initially, 1 in x 1/2 in ceramic fiber felt strips were pressed between the grid plate and scallop bar seal edge which is welded to the boiler waterwall. Erosion of the compressed felt and movement of the grid plate due to thermal expansion, resulted in leaks developing in this seal after a few tests. To eliminate this problem, a new seal was designed and installed. To date, the new seal design shows no signs of air leakage.

*Shown in Appendix A

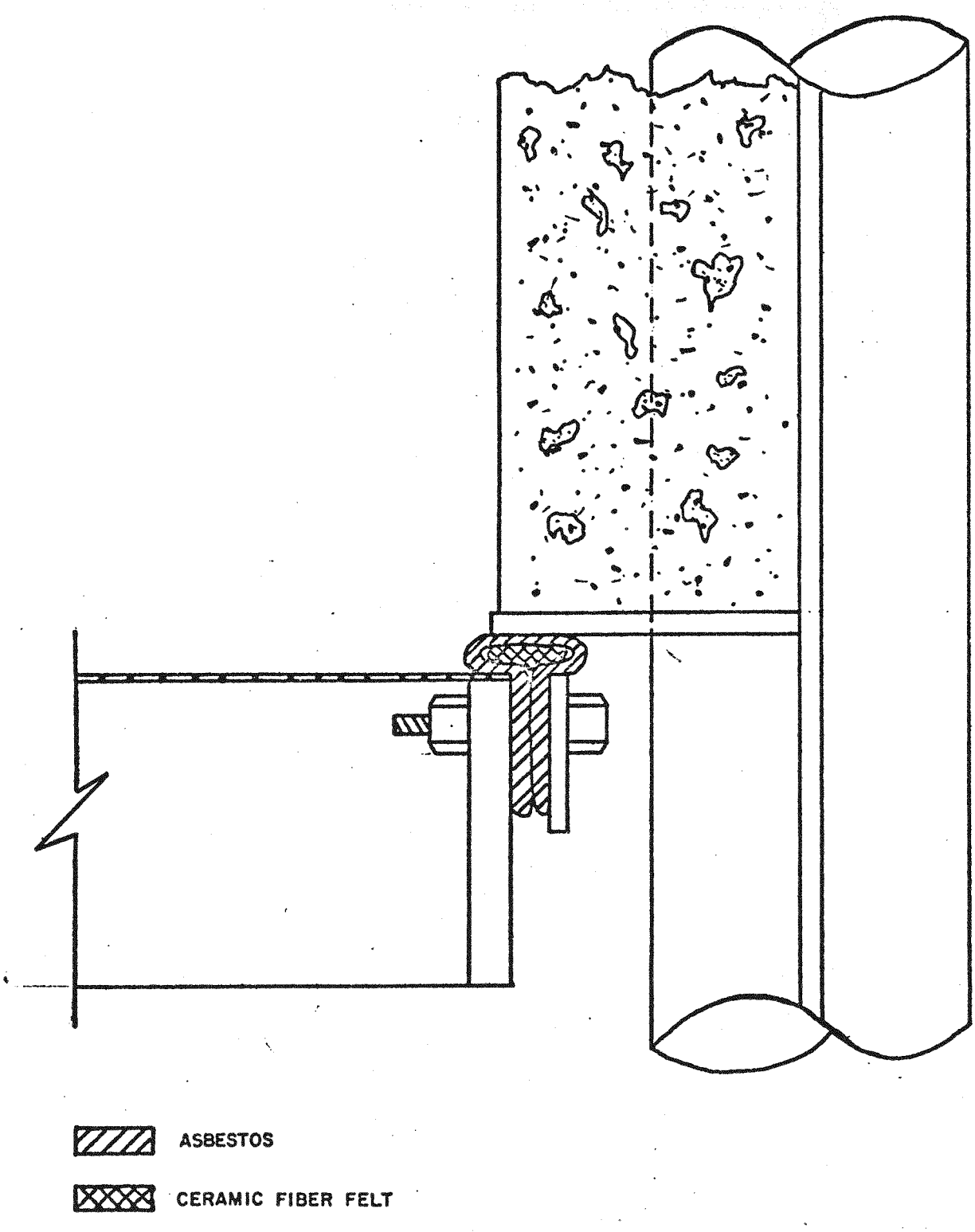
Figure 42 shows a cross section of the new seal. The object of the design was to develop a seal which would not be damaged during installation and which would increase the contact area between the grid plate and the boiler frame. To make this seal, a 6 in wide asbestos strip was folded over a 1 in x 1/2 in piece of ceramic fiber felt. The ends of the asbestos strip were then bolted between the frame of the grid plate and a 1 in x 1/8 in flat rolled steel bar to form a gasket. The gasket was made from a single piece of asbestos strip and felt which was wrapped completely around the entire grid plate. The two ends of the gasket were joined in the center of one side of the grid plate. In this way, all four corners, the most difficult areas to seal, were sealed with a continuouse piece of gasketing.

Nine 11 7/8 in square plates were installed on top of the grid plate. The corners of the plates were rounded and a bolt used to secure the corners of four plates to the grid plate. The space between the plates was filled with a refractory mortar. Cold testing of the composite air distributor showed no signs of air by-passing the plates.

Figure 43 shows the air distributor pressure drop as a function of mass flow under various conditions. The upper curve in Figure 43 was obtained after the plates had been installed but prior to testing. The middle curve was obtained after the first hot test. The lower curve in Figure 43 shows the pressure drop of the unprotected grid plate measured during hot Test Nos. 700-5 through 700-11. The scatter of this data is probably the result of different operating conditions during these tests. From this data, it appears that the ceramic plates would add about 6 in W.C. of pressure drop to the grid plate at 12 ft/sec with 250°F air preheat.

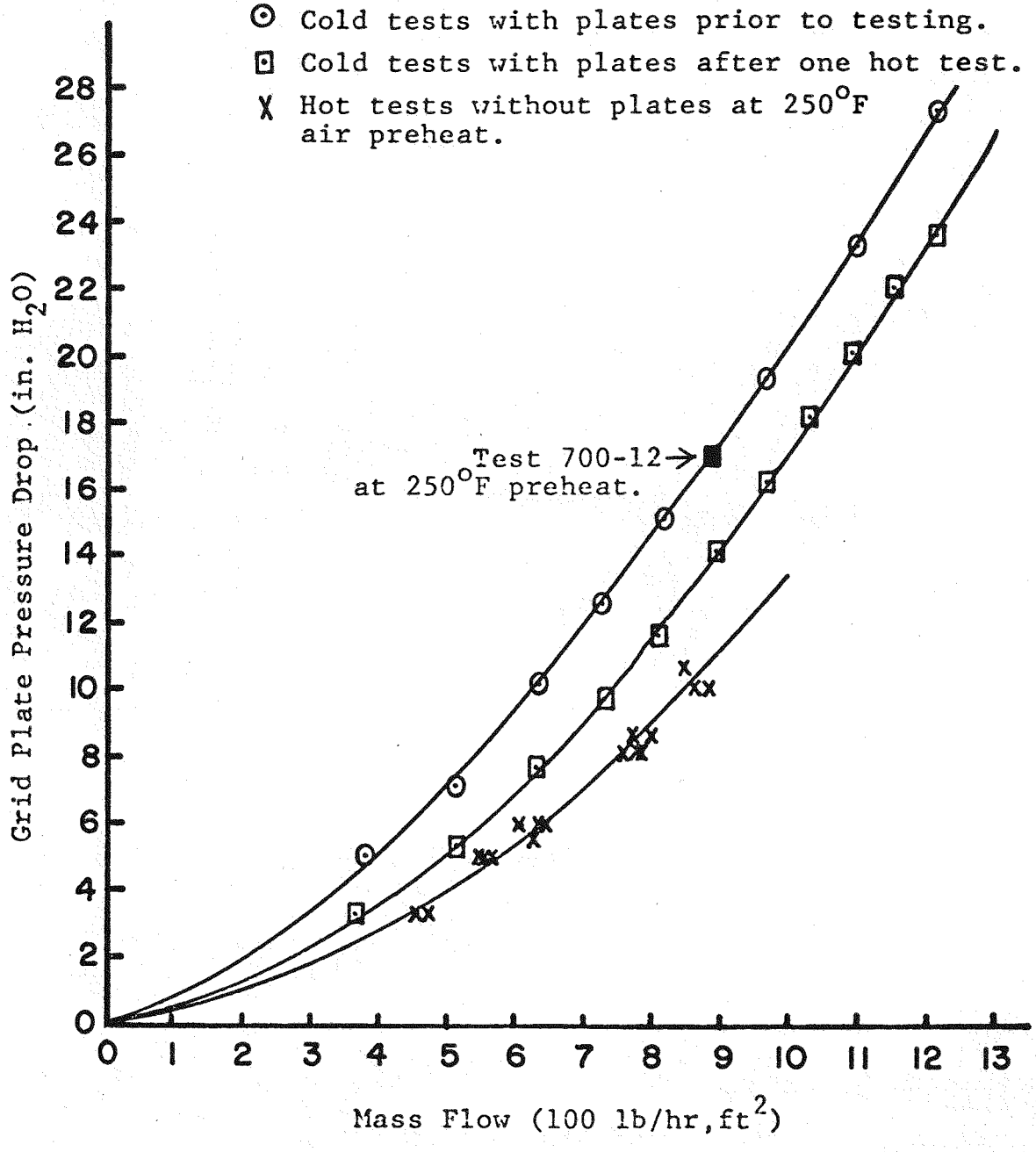
Test No. 700-12 was the first hot test using the composite air distributor. Operating conditions achieved during Test No. 700-12 are shown in Table III*. The insulating properties of the porous ceramic are clearly evident from the 275°F grid plate temperature. Air preheat for this test was set at 250°F and the resulting temperature rise was less than 30°F. The grid plate pressure drop shown in Figure 43, shows that after heating, the plates apparently lost some of their flow resistance. The pressure drop measured during Test No. 700-12 indicates that the resistance loss apparently occurred shortly after heating, because the pressure drop during the hot condition should be higher than the pressure drop when cold.

*Shown in Appendix A



AIR DISTRIBUTOR SEAL

Figure 42



EFFECT OF POROUS CERAMIC PLATES
ON GRID PLATE PRESSURE DROP

Figure 43

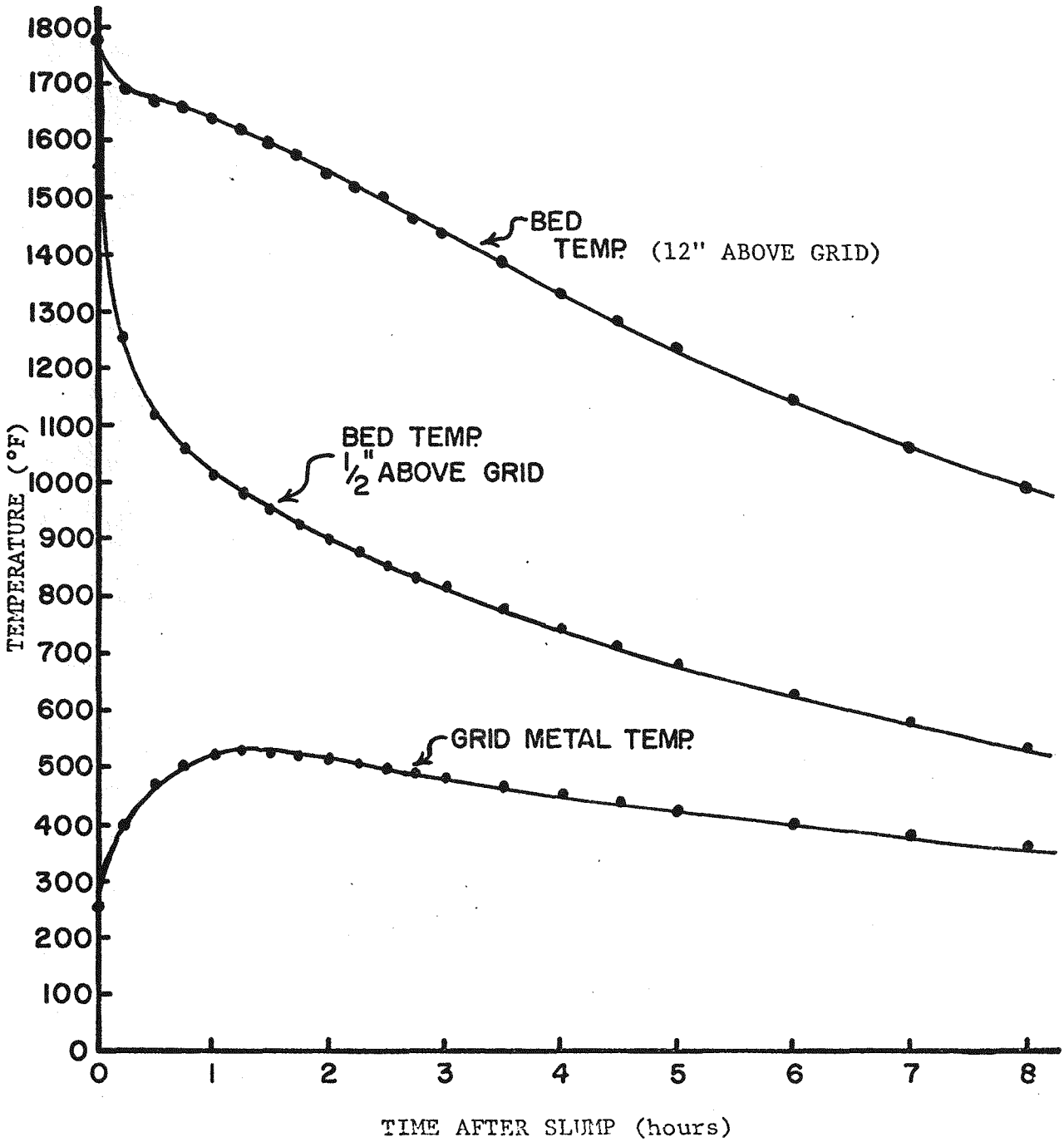
In Test No. 700-13, the bed temperature was increased to 1780°F at which time the bed was slumped. Figure 44 shows the temperature-time curves for the bed temperature 12 in. off the air distributor, the bed temperature 1/2 in off the air distributor, and the air distributor metal temperature after the hot slump. As noted, the metal temperature never rose above 550°F. After the bed had cooled, the bed was drained and the air distributor inspected. Some small cracks were found in two of the plates. The cracks were patched with a refractory cement.

Figure 45 shows the pressure drop across the air distributor measured under a variety of test conditions. For comparison, the data of Test No. 700-17 (no grid plate protection) is included in the graph. There is no apparent change in the air distributor pressure drop under cold conditions after three hot tests.

Another interesting aspect of Figure 45 is that the pressure drop under hot conditions (Test No. 700-13 and 16 were at bed temperatures between 1500°F and 1600°F with air preheat of 220°F - 270°F) was equal to the cold test data. A possible explanation is that the cold tests were conducted without bed material on the air distributor. It was reported in Interim Report No. 3 that the presence of bed material on the grid plate reduces the grid plate pressure drop. Thus, the increased pressure drop resulting from hot operation was apparently offset by the decrease in the pressure drop caused by the presence of bed material.

In Test No. 700-14 the effect of air preheat temperature was examined. Figure 46 shows the results of this test at three different air preheat temperatures, 100°F, 215°F and 310°F. The data at 215°F agrees very nearly with the data in Figure 46. It can be seen by comparing Figures 45 and 46 that the pressure drop at 100°F is lower than the cold test data, confirming the effect of bed material on the air distributor pressure drop.

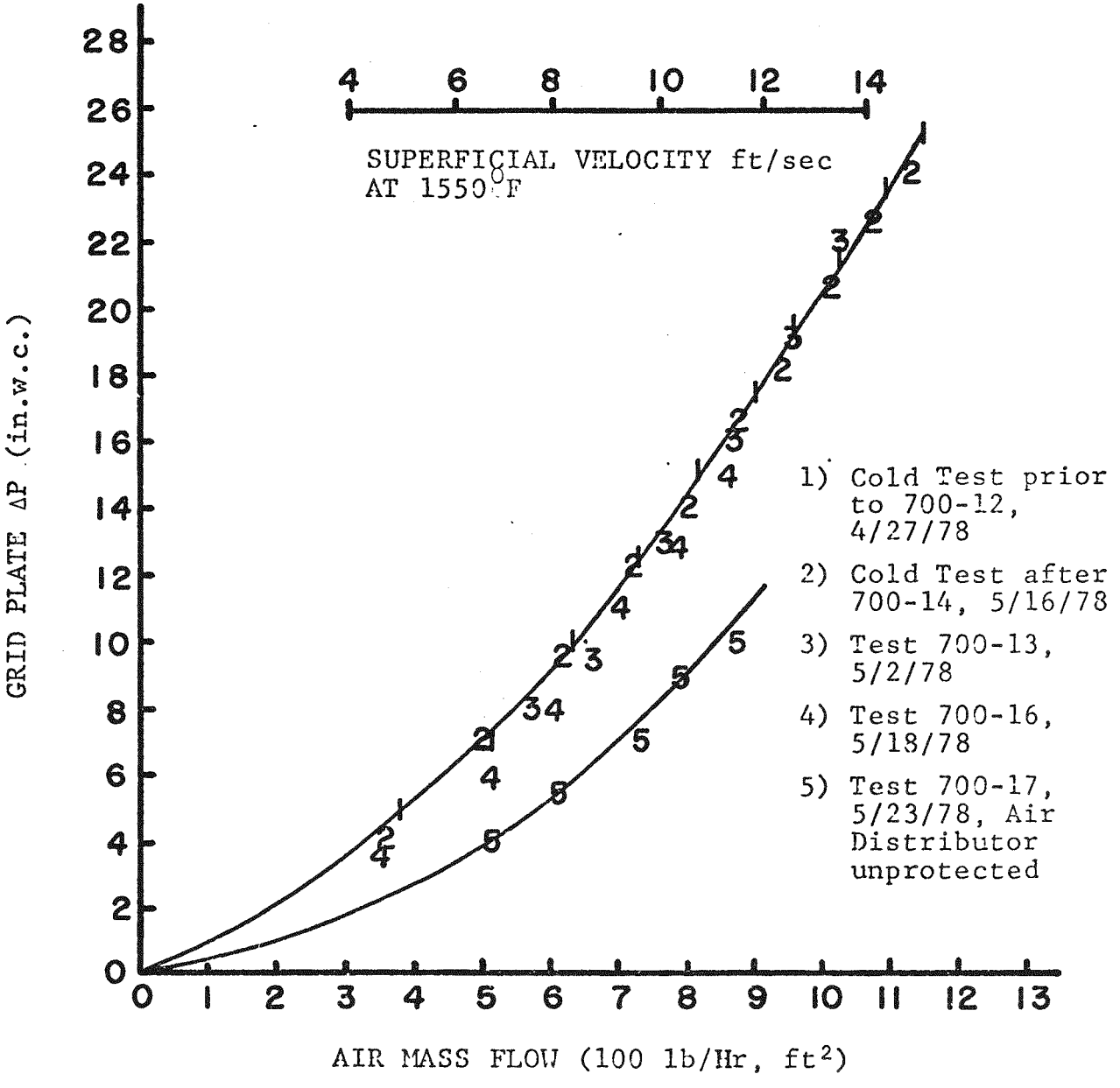
As a final test of the porous ceramic plates, the bed was clinkered at the end of Test No. 700-16. During the clinkering operation, bed temperatures as high as 2270°F were recorded. However, the grid plate metal temperature never rose above 300°F. Inspection of the air distributor after the test showed one of the plates had broken into several large pieces and separated from the grid plate. This failure under these extreme conditions was the only limitation of the porous plates noted during the five hot tests. Based on this failure, placement of the porous plates over previously used perforated plate air distributors at Rivesville would not



EFFECT OF USING POROUS CERAMIC PLATES ON P.D.U.

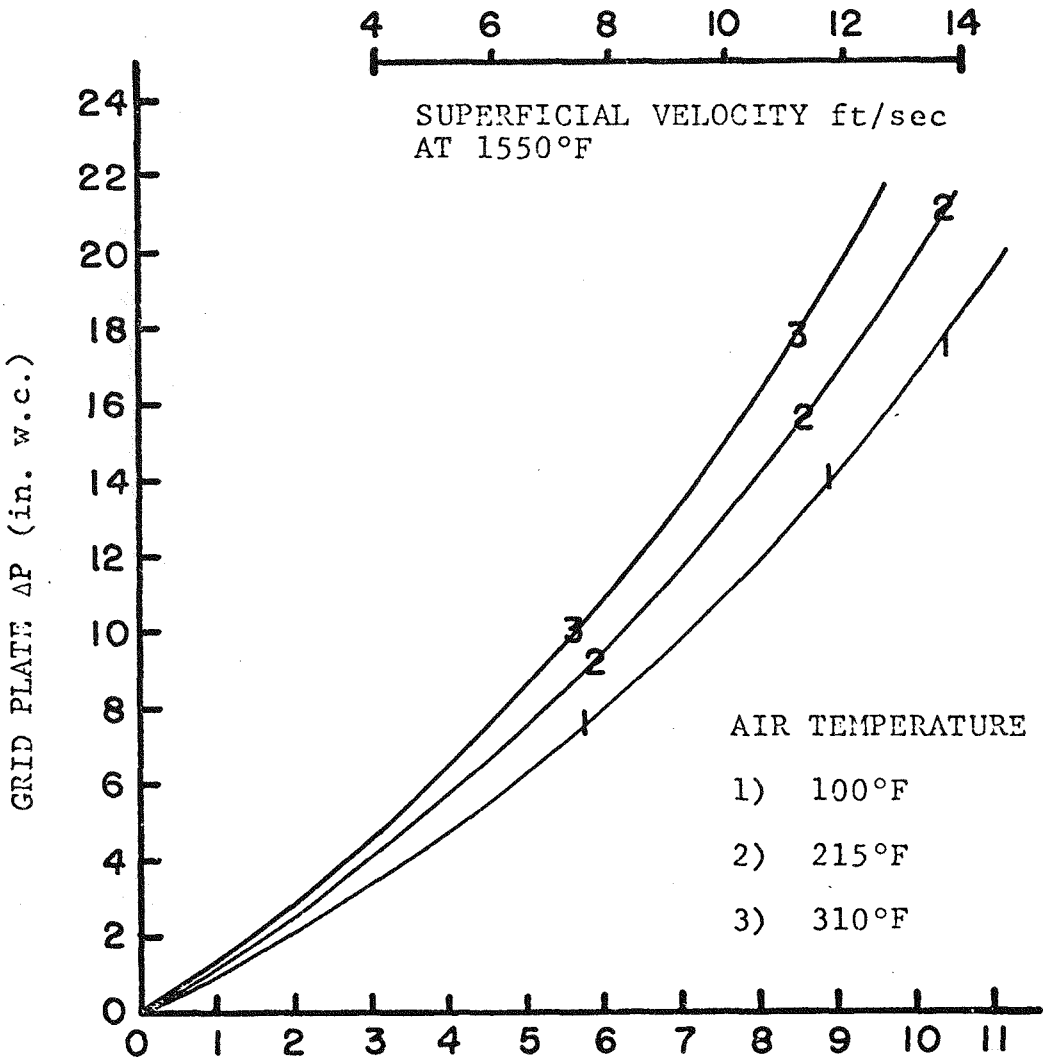
TEMPERATURES AFTER A HOT SLUMP TEST 700-13, 5/2/78

Figure 44



AIR DISTRIBUTOR PRESSURE DROP COMPARISON

Figure 45



EFFECT OF AIR TEMPERATURE ON GRID PLATE ΔP
TEST 700-14, 5/16/78

Figure 46

be recommended. To best apply the porous plates as an air distributor, a support frame designed specifically for this purpose would be required.

Alumina Stones

An alternative air distributor protection cover tested consisted of a 2 to 3 in deep layer of rounded alumina stones held in place by a metal screen. The major difficulty observed in the use of the previously tested 1 in trap rock was the tendency of the rocks to shift and leave portions of the air distributor uncovered. To eliminate this problem, rounded alumina stones of nominal 1 1/2 in size were used and a sheet of expanded metal screen was placed on top of the stones to hold them in place. The metal screen was made of a special stainless steel (Rolled Alloys Co. R.A. 330) to withstand the bed temperatures. Test No. 700-18 was conducted using this composite cover. The results of Test No. 700-18 are shown in Table IV*. As can be seen from Table IV, the air Distributor metal temperatures were significantly reduced compared to the unprotected air distributor temperatures. For this test the metal temperatures ranged between 50 and 70° hotter than the plenum air temperatures. This compares favorably with the porous ceramic plates which recorded about a 30°F rise.

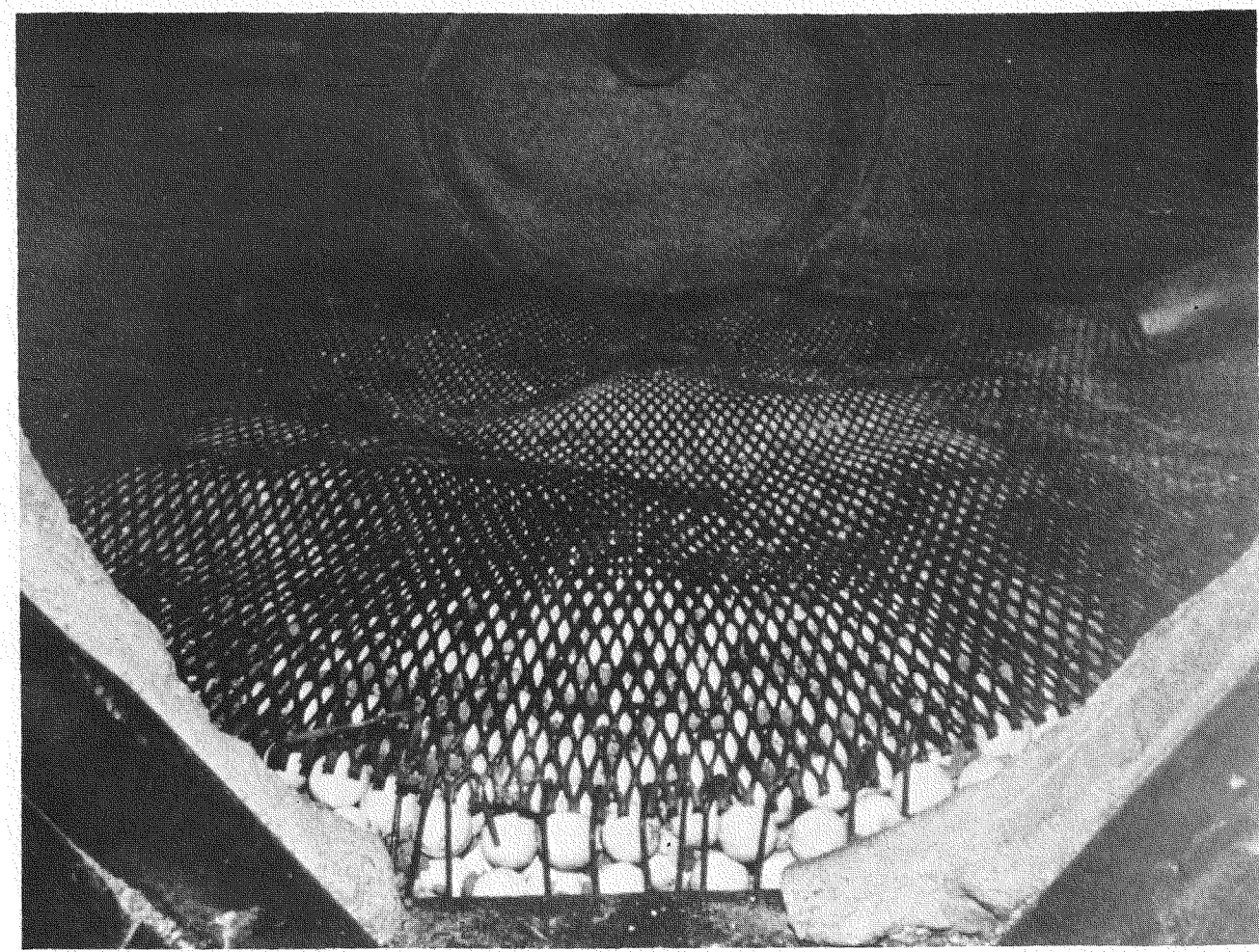
An expected disadvantage of this type of protection is the exposure of the expanded metal screen to the hot bed material. Examination of the screen after the test showed no visible signs of deterioration due to the short exposure. It is possible that this metal was cooled somewhat by the incoming air.

After Test No. 700-18, cold testing of the composite grid revealed several dead zones at low superficial velocities. The dead zones were due to the uneveness of the stone layer. It is not known if these dead zones will be detrimental at higher velocities or not. There was no evidence of clinker material in the bed. If the dead zones were present, it is likely that some clinkers would have formed.

Alumina Spheres

In order to eliminate the problem of the dead zones, a 3 in deep layer of 3/4 in alumina spheres was tested with an expanded metal hold down screen. The 3/4 in diameter spheres were high purity alumina manufactured by Coors Poreclain Company and the metal screen was the same screen used in Test No. 700-18. Figure 47 shows the spheres and screen installed in the PDU from the open door.

*Shown in Appendix A



3/4" Alumina Spheres with Metal Screen
Viewed from PDU Door

Figure 47

The metal screen was used in a similar arrangement with larger alumina stones in Test No. 700-18.

Test Nos. 700-21 and 22 were conducted using this air distributor combination. Temperatures were measured at the bottom side and top side of the composite protective cover during Test Nos. 700-21 and 700-22 respectively. Table VI* gives these temperatures and the conditions for two tests.

In order to compare the temperature insulating property of the protective materials, the results of a base line test with no grid protection is included in Table 54 along with selected previous test results. Table 54 shows the temperature measured 1 in above the perforated grid plate and the temperature difference between the grid plate and plenum for Test Nos. 700-16, 17, 18 and 21. The temperature rise in all these tests ranged from 27°F to 72°F with the 3/4 in alumina spheres performing slightly better than the other materials with the lowest average rise being 29°F.

Figure 48 shows the pressure drop across the grid plate measured for different air flows in Test Nos. 700-17, 18, 21 and 22 for comparison purposes. As expected, pressure drops are higher for Test Nos. 700-21 and 22 than for Test No. 700-17, the baseline test, and for Test No. 700-18, which had larger size particles and hence more voids in the protective layer. All the protective materials tested would require adjustment of the perforated plate pressure drop prior to installation due to the relatively high additional drop.

In review of the protective materials air distributor tests, the findings are as follows: trap rock would be considered unacceptable since it was found that about 5 percent of the rock was cracked or broken. The porous ceramic plates were unacceptable for use on top of an existing grid due to breakage of these plates under severe conditions. The rounded 1 1/2 in alumina stones were observed to cause dead zones during cold tests at low superficial velocities. None of these problems were observed during nor after Test Nos. 700-21 and 22 with the 3/4 in alumina spheres.

Use of the 3/4 in alumina spheres would add an additional pressure drop to the air distributor. This could be compensated for in the design of the perforated plate air distributor, thereby resulting in a composite air distributor with the same pressure drop characteristics of the present perforated plate air distributor.

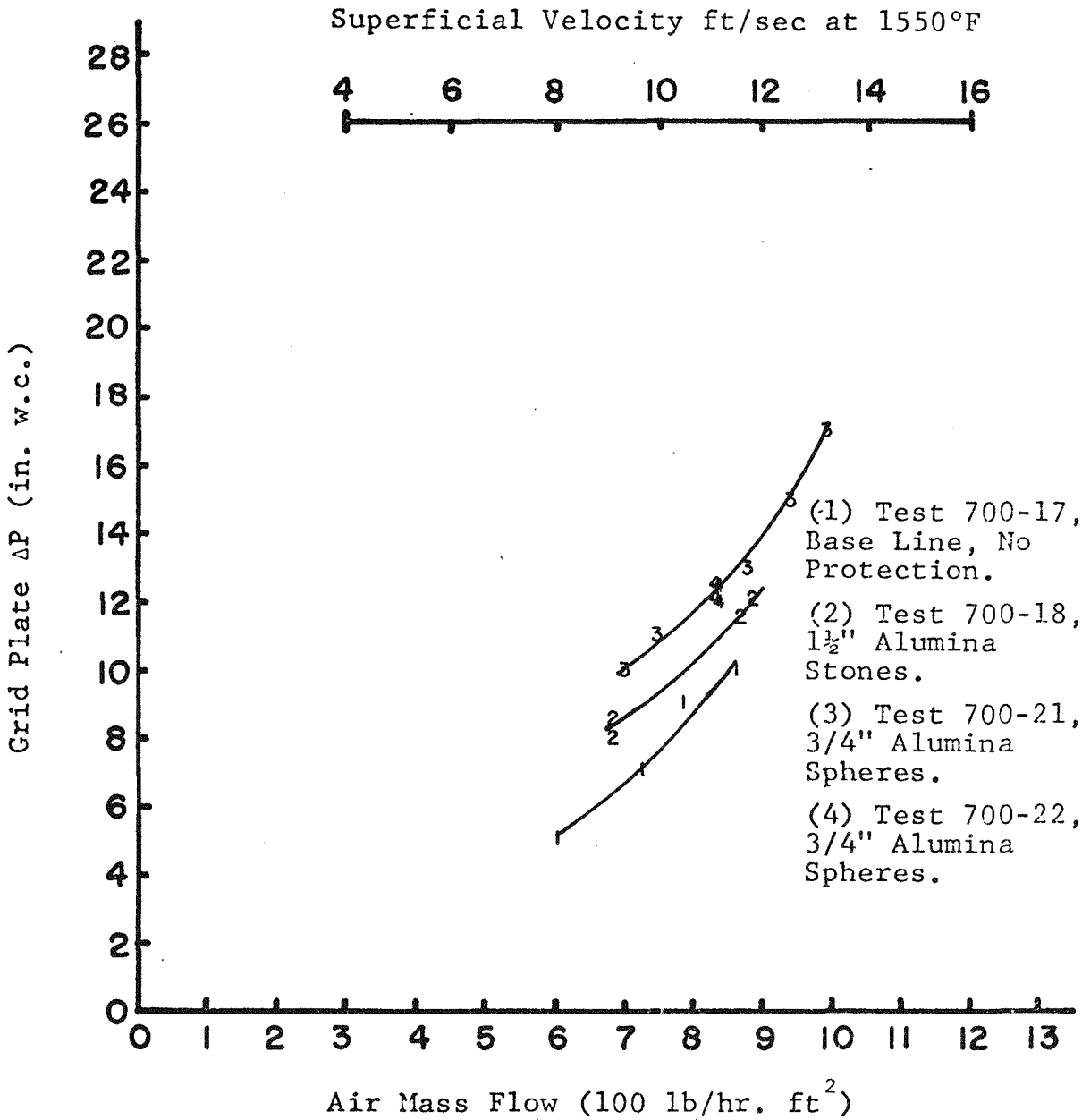
*Shown in Appendix A

Table 54

Comparison of Air Distributor Temperatures for Tests 16, 17, 18, and 21

	Test 700-17 Base Line, No Protection					Test 700-18 1½" Rounded Alumina Stones				
Time /Hr	13:00	13:30	14:00	14:30	Ave.	10.25	11:40	13:30	15:00	Ave.
1" above grid plate, °F	1530	1530	1520	1450	1508	645	630	698	720	673
Grid plate temp., °F	947	930	944	876	924	313	311	291	293	302
Plenum temp., °F	345	351	351	333	345	254	239	247	249	247
Grid plate to plenum, ΔT, °F	602	579	593	543	579	59	72	44	44	55
Bed temp., °F	1535	1535	1521	1440	1508	1541	1565	1765	1680	1638

	Test 700-21 3/4" Alumina Spheres						Test 700-16 Porous Plates					
Time /Hr	12:30	13:10	13:45	14:28	14:52	Ave.	11:00	12:30	13:00	13:15	14:00	Ave.
1" above grid plate, °F	275	295	325	330	365	318	Same as bed temperature					
Grid plate temp., °F	291	278	273	274	276	278	338	283	314	287	281	301
Plenum temp., °F	257	251	245	245	247	249	275	254	258	255	250	258
Grid plate to plenum, ΔT, °F	34	27	28	29	29	29	63	29	56	32	31	42
Bed temp., °F	1648	1670	1610	1668	1630	1645	1558	1569	1548	1555	1706	1587



AIR DISTRIBUTOR PRESSURE DROP COMPARISON

Figure 48

2.3.8 Corrosion/Erosion Studies

Test Racks

Two uncooled corrosion/erosion test racks, identical to the test rack previously installed in the FBM and reported in Interim Report No. 3 were furnished by International Nickel Company, Inc. (INCO) and installed in FBM.

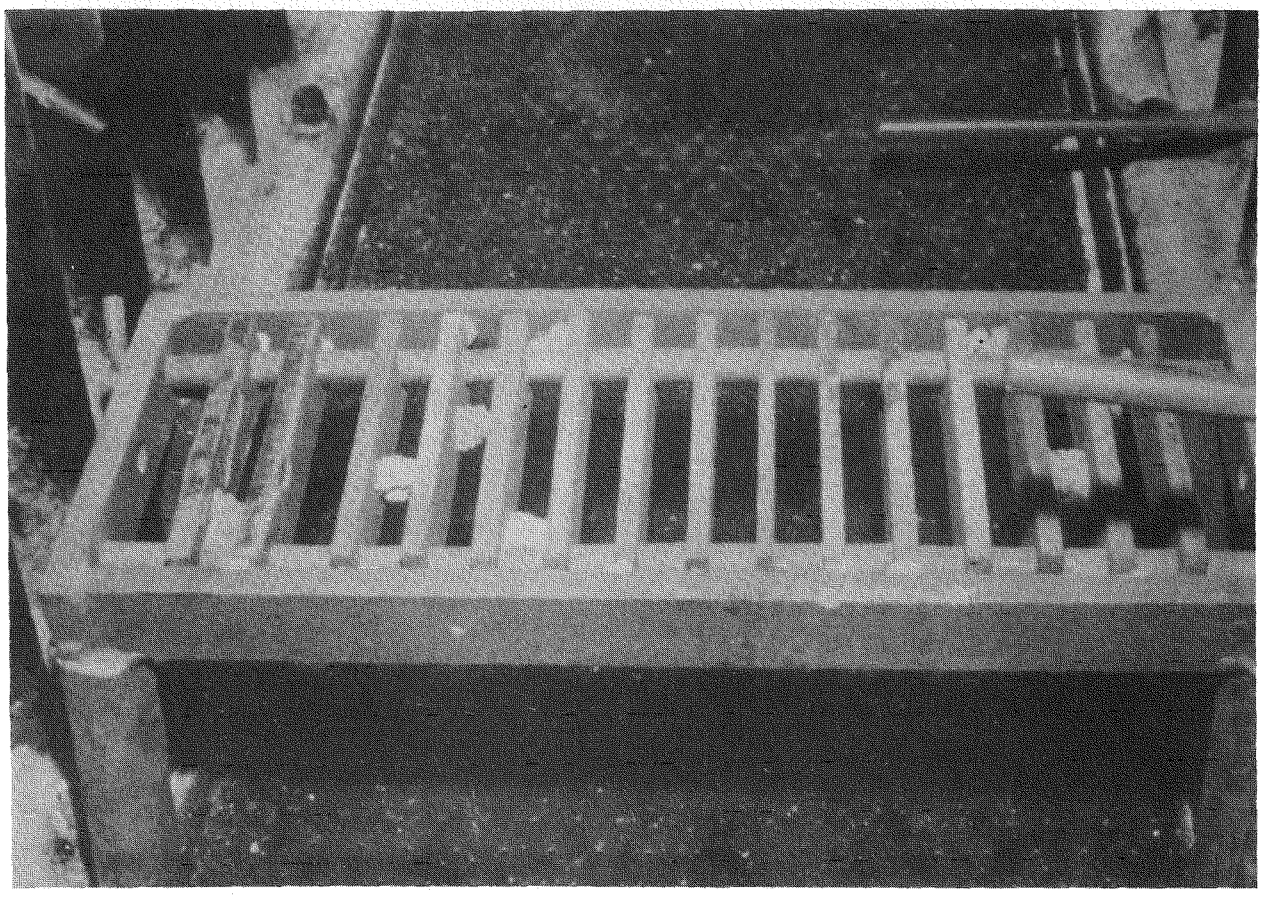
The test racks measured 14 in x 6 in x 1 in and contained metal coupons measuring 4 in x 1 in x 1/4 in An in-bed rack, No. 434, shown in Figure 49 was located 6 1/2 in above the grid plate in the center of the furnace in a normally submerged position. An upper rack, No. 452, shown in Figure 50, was located 6 in. below the steam drum in the middle of the furnace freeboard area above the bed. Both photographs were taken after 190 hours of operation.

The test rack previously reported in Interim Report No. 3, No. 433, was mounted below the horizontal tube bundle near the front of the bed and above the front coal feeder. This uncooled rack was subjected to hot operation under a variety of conditions, including bed agglomeration tests requiring many light-offs. Sodium Sulfate (Na_2SO_4) was fed during two runs, each time for approximately 30 minutes at a rate of 18 lbs/hr. The first time was after 33 hr of hot operation and the second time was after 55 hr of hot operation.

The most recently tested racks, Nos. 434 and 452 were not subjected to sodium sulfate. These racks were exposed without a submerged tube bundle in the bed.

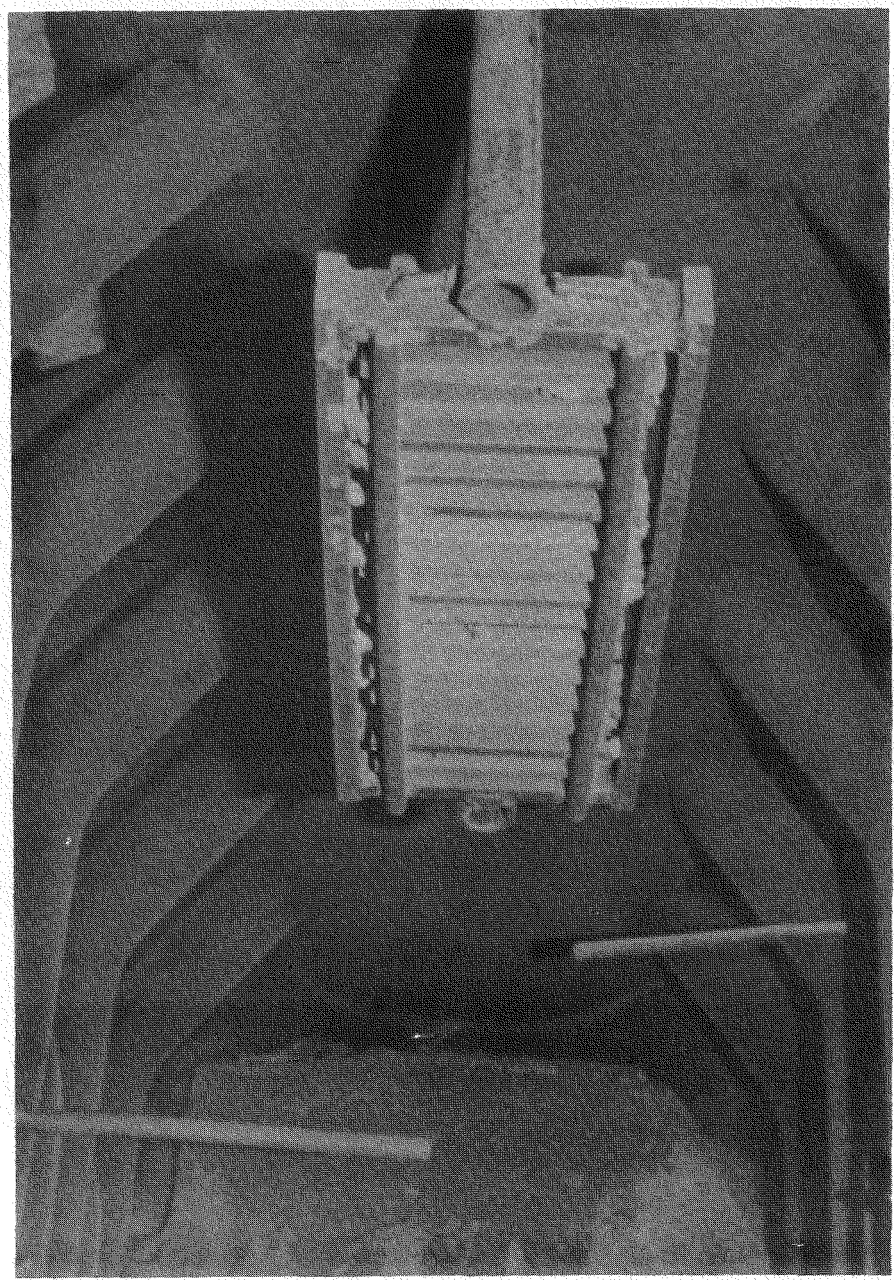
Table 55 and 56 show the corrosion data as supplied by INCO for test rack nos. 434 and 452. Generally, all alloys except T-22 and 1010 carbon steel lost less than 10 mg/cm^2 in both locations. The data also shows that corrosion to the above bed test rack No. 452 was less severe than to the in-bed rack No. 434. INCO also reported an accelerated rate of attack at the corner of the coupons of Alloy 600 and 309 stainless located in the bed and 304 stainless located above the bed. They caution that this could cause problems for longer exposure periods.

The comparisons of coupon weight loss and test conditions is shown in Table 57. Also shown in Table 57, is the ranking of each coupon's weight loss relative to other coupons in the racks. Table 57 shows that regardless of longer exposure time and increased number of



Corrosion Test Rack No. 434, above bed

Figure 49



Corrosion Test Rack No. 452, in bed

Figure 50

Table 55

Test Rack No. 434 Data

MATERIAL	WEIGHT LOSS (Mg/Cm_2)	METAL LOSS (ONE SIDE) WIDTH (1")	METAL LOSS (ONE SIDE) THICKNESS (1/4")	PENETRATION (ONE SIDE) WIDTH (1")	PENETRATION (ONE SIDE) THICKNESS (1/4")	TYPE OF ATTACK*	COMMENTS**
IN-811	1.23	< .0001	< .0001	.002	.003	O,SS	
IN-809	1.64	.0005	.0007	< .001	< .001	O	
HASTX	2.49	< .0001	.0005	< .001	.015	O	
671	2.57	.0012	.0012	< .001	.003	O	
T-310SS	2.92	.0004	< .0001	< .001	< .001	O	
ALLOY800	2.93	.0015	.0004	.006	.002	O	
ALLOY601	3.11	.0017	.0004	.011	.004	O,IS	
ALLOY804	3.16	.0006	.0010	.001	.002	O,SS	
ALLOY600	4.30	.0028	.0010	.005	.003	O,IS	7
T-309SS	4.49	.0021	.0006	.003	< .001	O,SS	7
T-314SS	6.73	.0018	.0007	.004	< .001	O,SS	
IN-890	7.92	.0005	.0030	< .001	.002	O	8
T-304SS	9.13	.0027	.0007	.003	.002	O,SS,IS	
T-22	687.42	.0435	.0499	< .001	.001	O,SS	
1010	930.33	.0767	.0637	< .001	< .001	O	

*O Oxidation
S Sulfidation
SS Some Sulfidation
IS Intergranular Sulfidation
I Intergranular

**3. Non-uniform attack, coupon measured at the area of greatest attack.

7. Non-measurable accelerated attack at coupon edge.

8. Some possible sulfidation attack was observed.

Table 56

Test Rack No. 452 Data

MATERIAL	METAL LOSS (ONE SIDE) WEIGHT LOSS (Mg/Cm^2)	METAL LOSS (ONE SIDE) WIDTH (1")	METAL LOSS (ONE SIDE) THICK-NESS (1/4")	PENETRATION (ONE SIDE) WIDTH (1")	PENETRATION (ONE SIDE) THICK-NESS (1/4")	TYPE OF ATTACK*	COMMENTS**
IN-811	.81	.0016	.0004	<.001	<.001	O	
ALLOY601	.85	.0005	.0009	.075	.001	O,S,I	3
HASTX	.87	.0004	.0001	<.001	<.001	O	
T-310SS	.91	.0010	.0003	<.001	<.001	O	
ALLOY800	1.05	.0009	.0131	<.001	<.001	O	
ALLOY804	1.16	.0023	.0009	.004	<.001	O,S	
ALLOY600	1.17	.0010	.0005	<.001	<.001	O	
T-309SS	1.34	.0005	.0003	<.001	<.001	O	
T-314SS	1.80	.0012	.0004	<.001	<.001	O	
IN-809	2.28	.0008	.0005	<.001	.002	O,S	3
T-304SS	2.30	.0007	.0010	<.001	<.001	O	7
IN-890	4.20	.0013	.0004	.005	.004	O,S	3
671	6.54	.0017	.0010	<.001	.015	O	3
1010	601.66	.0324	.0342	.002	.004	O,S,IS	
T-22	953.50	.0233	.0655	<.001	<.001	O,SS	

*O Oxidation
S Sulfidation
SS Some Sulfidation
IS Intergranular Sulfidation
I Intergranular

**3. Non-uniform attack, coupon measured at the area of greatest attack.

7. Non-measurable accelerated attack at coupon edge.

8. Some possible sulfidation attack was observed.

Table 57

Comparison of Coupon Weight Loss and Test Conditions for Racks

Material	Rack No. and Location 433 In Bed Rank	433 In Bed mg/cm²	434 In Bed Rank	434 In Bed mg/cm²	452 Above Bed Rank	452 Above Bed mg/cm²
DATE:	1/76 - 4/76		1/17 - 3/77		1/77 - 3/77	
Incoloy 800	1	2.32	6	2.93	5	1.05
Type 314 s.s.	2	4.24	11	6.73	9	1.80
Type 310 s.s.	3	4.29	5	2.92	4	0.91
Type 309 s.s.	4	8.21	10	4.49	8	1.34
Hastelloy X	5	29.93	3	2.49	3	0.87
Inconel 811	6	31.18	1	1.23	1	0.81
Type 304 s.s.	7	36.05	13	9.13	11	2.30
Inconel 671	8	92.86	4	2.57	13	6.54
Incoloy 804	9	147.85	8	3.16	6	1.16
Inconel 601	10	430.43	7	3.11	2	0.85
Inconel 809	11	686.51	2	1.64	10	2.28
Alloy T-22	12	939.62	14	687.42	15	953.50
Inconel 600	13	1003.54	9	4.30	7	1.17
1010 c.s.	14	1135.14	15	930.33	14	601.66
Inconel 690	15	1209.44	12	7.92	12	4.20

Test Conditions			
Hours of operation (hrs)	140	210	210
Period:	Jan 76-Apr 76	Jan 77-Mar 77	Jan 77-Mar 77
Temperature (°F)	1450-1750	1500-1600	1500-1600
No. of light-offs	13	26	26
NA_2SO_4 Addition	Yes	No	No

light-offs, all but two of the samples had lower weight losses in the present study. It appears that sodium sulfate additions which occurred during exposure of speciments in rack No. 433, had a significant detrimental effect on all samples except Incoloy 800 and 314 stainless.

Examination of Waterwall Tubes from the FBM

Sections of the waterwall tubes (2 1/2 in I.D.), the water door and the horizontal tube bundles (2 in O.D.) of the dismantled FBM were cut out, marked, labeled and sent to Hercules, Inc. Table 58 lists the samples and their respective locations within the FBM. Figures 51 and 52 show the overall view and closeup view of the partially dismantled FBM from the west side; chalk lines indicate the sections cut out.

A metallographic examination of portions of the old FBM was conducted by Hercules, Inc. The report was prepared by R.D. Crooks of Hercules, Inc. and the following observations were made in the summary of that report. The full report is included as Appendix B.

Examination and analysis of the tubes has shown remarkedly little change in the surface or subsurface characteristics of the tube material as a result of exposure to the high temperature fluidized coal bed or its combustion products. The most probable reasons for the absence of change are the relatively low operating temperature of the bed coupled with low tube wall temperatures via adequate water flow.

Specifically it was found that:

- Surface decarburization appeared to be a pre-existing condition in some tubes and not present in others, leading to the conclusion that decarburization did not occur in service.

- Despite the presence of a carbonaceous layer deposit on the fireside of the tubes, no carburization of the tube surfaces was observed.

- Despite the presence of high sulfur levels in the tube fireside deposits, no evidence of tube surface sulfidation was observed.

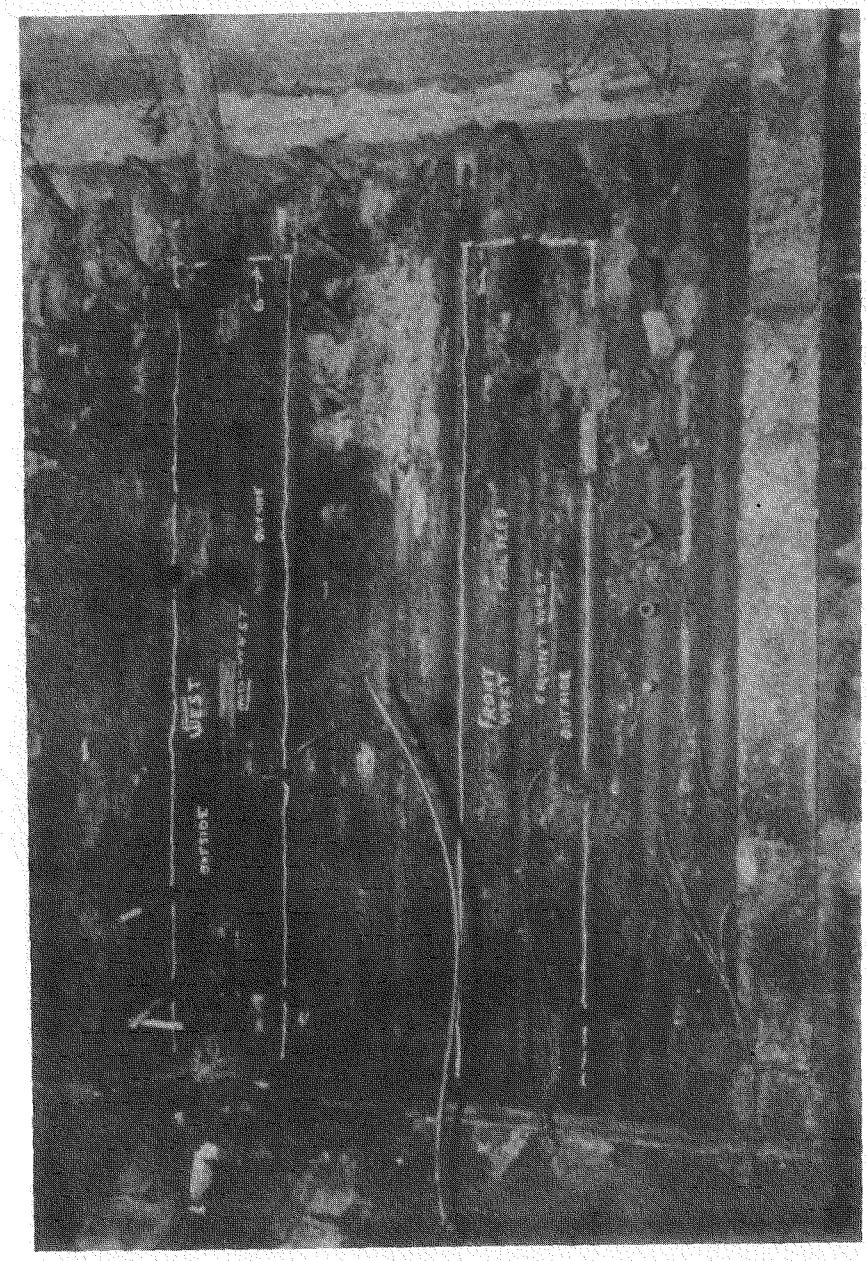
- The microstructure of the tubes showed no evidence of overheating nor was there any significant surface attack on the fireside.

Table 58

Corrosion/Erosion Samples Taken from the FBM

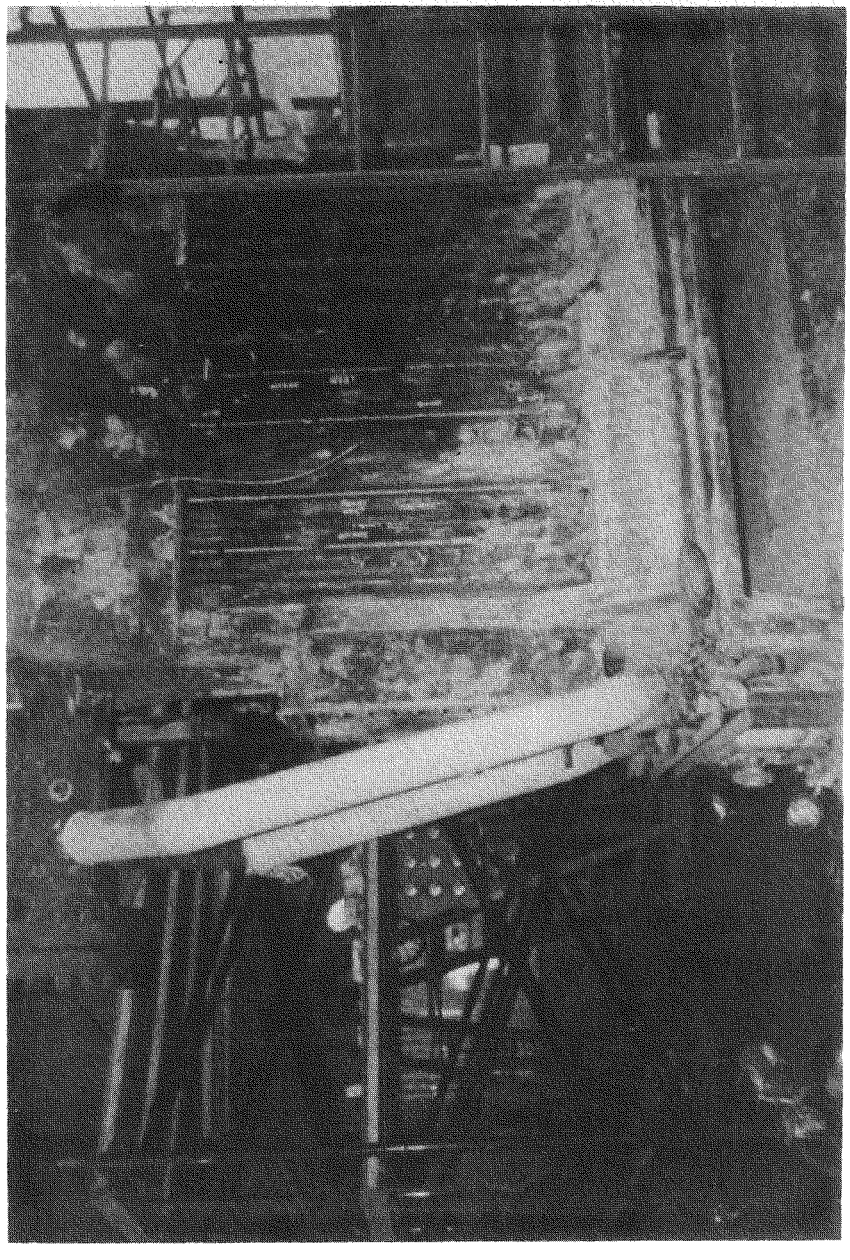
Part No.	Location	Tube No.*	Description
Waterwall Tubes			
1	Front-West	5 & 6	6" above air distributor, coal feeder at bottom marked "outside"
2	Mid-West	10 & 11	6" above air distributor
3	Front-East	5 & 6	6" above air distributor
4	Mid-East	13 & 14	6" above air distributor
5	Back-East	---	Connection between waterwall tube and steam drum, covered with refractory, "top" means to the steam drum
Water Door			
6	Water door, in bed		with duct burner pipe
7	Water door ledge		water cooled shelf for supporting the water door
Horizontal Tube Bundle			
8	Horizontal tube bundle		marked fire side, outside, top side and bottom side

* Tube numbers are counted starting from the front of the FBM.



CLOSEUP VIEW FROM THE WEST SIDE; CHALK LINES INDICATE THE SECTIONS TO BE CUT OUT.

Figure 51



PARTIALLY DISMANTLED FBM, VIEWED FROM THE WEST SIDE.

Figure 52

- Small losses of tube wall thickness were observed on the tube fireside concentrated at the mid tube location as a result of abrasive erosion.

These observations lead to the conclusion that for service conditions in this fluidized bed boiler, no serious deterioration of the boiler tubes occurred and it appears that plain carbon steel, ASTM A192, is a satisfactorily serviceable material.

2.4 Process Analysis Results

2.4.1 Horizontal Tube Bundle Heat Transfer Studies

Calculation Procedure

The new PDU is equipped with an in-bed superheater bundle and an in-bed economizer bundle to allow measurement of overall and outside heat transfer coefficient under different operating conditions. With these two bundles, it is possible to generate a significant amount of heat transfer data under various test conditions. The following calculation procedure outlines how the data from the test runs will be reduced to obtain heat transfer coefficients.

Overall heat Transfer Coefficient

The overall heat transfer coefficient based on outside area, is calculated as follows

$$U_o = \frac{W(H_o - H_i)}{A_o \, \Delta T_{LM}} \quad \text{Btu/hr/ft}^2/^\circ\text{F} \qquad (1)$$

where:

W = the fluid flow rate through the bundle, lb/hr.
H_o = enthalpy of the fluid at outlet conditions, Btu/lb.
H_i = enthalpy of the fluid at inlet conditions, Btu/lb.
A_o = outside surface area of the tube bundle, ft^2.
ΔT_{LM} = log mean temperature between the bed and fluid, °F

The overall heat transfer resistance $1/U_o$ is the sum of several heat transfer resistances. It is generally assumed that there are five resistances in series; inside fluid film, inside scale, tube wall, outside scale and outside film coefficient. To calculate the outside film

coefficient from the overall coefficient, the other four resistances must be either known or estimated.

Inside Fluid Film Coefficient

Several correlations exist in the literature which predict the inside film coefficient. The correlations chosen were those recommended by Foster Wheeler Corporation and used in the design of the Rivesville MFB. The following correlation was used to calculate the inside film coefficient in the economizer:

$$h_i = C_W \times 0.0378 \frac{(G)^{0.8}}{(D_i)^{0.2}} \text{ Btu/hr/ft}^2/^\circ\text{F} \quad (2)$$

where:
- G = the fluid mass flow, lb/hr/ft^2
- D_i = tube inside diameter in inches
- C_W = the physical property constant for water evaluated at the average film temperature.

Figure 53 shows the value of C_W for the ranges of temperature and pressure encountered in these tests.

For the superheater, the correlation used is:

$$h_i = C_S \frac{.0198(G)^{0.84}}{(D_i)^{0.16}} \text{ Btu/h/ft}^2/^\circ\text{F} \quad (3)$$

where:
- C_S = the physical property constant for steam elevated at the average film temperature.

The values of C_S are also shown in Figure 53.

Metal and Scale Resistances

The metal heat transfer resistance is given by:

$$h_m = \frac{k\, r_{LM}}{\Delta t\, r_o} \text{ Btu/hr/ft}^2/^\circ\text{F} \quad (4)$$

where:
- r_o = the outside tube radius, inches
- Δt = tube wall thickness, inches
- r_{LM} = log mean tube radius, inches
- k = thermal conductivity of the tube metal, Btu/in/hr/ft

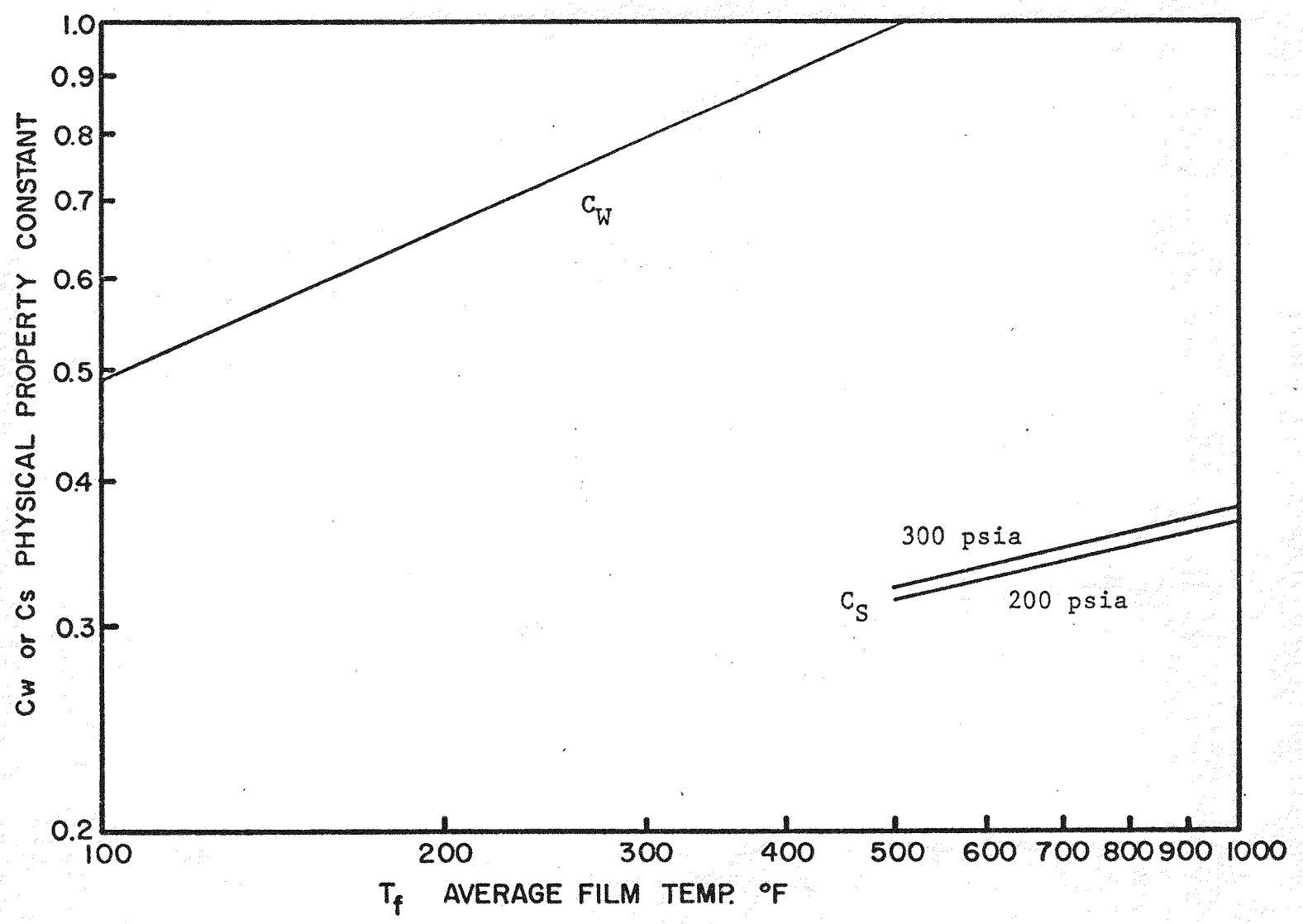


Figure 53

PHYSICAL PROPERTY CONSTANTS FOR COMPRESSED WATER (C_W) AND SUPERHEATED STEAM (C_S)

The thermal conductivity for the type T11 steel tubes is 210 Btu/in/hr/ft^2/°F. This yields a metal heat transfer coefficient of 1300 Btu/hr/ft^2/°F.

From previous heat transfer work reported in the Appendix of PER Monthly Progress Report No. 41, (February 1976) the inside scale coefficient was taken to be 3250 Btu/hr/ft^2/°F and the outside scale coefficient was taken to be 5000 Btu/hr/ft^2/°F. The three resistances result in a total wall coefficient of 784 Btu/hr/ft^2/°F.

Outside Coefficient

The outside resistance is then given by:

$$\frac{1}{h_o} = \frac{1}{U_o} - \frac{1}{h_w} - \frac{r_o}{h_i\ r_i} \qquad (5)$$

where:

h_w = wall coefficient, Btu/hr/ft^2/°F
h_i = inside coefficient, Btu/hr/ft^2/°F

The solution of equations 2 through 5 involves a trial and error calculation. An average wall temperature must be assumed to calculate the average film temperature. Next h_i and h_o are calculated from equations 2 or 3 or 5. The wall temperature is then recalculated using the following relationship:

$$\bar{T}_w = U_o \left(\frac{\bar{T}_s}{h_o} + \frac{T_b}{h_i} \right) \qquad (6)$$

where:

$\bar{T}_s$ = the average steam (or water) temperature, °F

$\bar{T}_w$ = the average wall temperature, °F

T_b = the temperature of bed, °F

The calculation is repeated until the average wall temperature coverges. Usually only 3 or 4 iterations are required for convergance.

Test Results

Table 59 shows the result of the above calculations from Test Nos. 700-4 thru 700-7, in which bed temperatures ranged from 1511°F to 1565°F and superficial velocities ranged from 10.7 to 12.2 ft/sec.

In performing these calculations the following tube bundle data was used:

Table 59

Calculated Heat Transfer Coefficient for Tests 700-4, 5, 6 and 7

Test No.	700-4 3/15/78			700-5 3/21/78				700-6 3/22/78					700-7 3/28/78		
Time/Hr.	12:00	13:00	14:00	12:00	13:00	14:00	15:00	11:00	12:00	13:00	14:00	15:00	13:00	14:00	15:00
Bed temperature °F	1538	1535	1538	1550	1516	1532	1565	1553	1545	1543	1541	1511	1555	1552	1548
Bed depth (static)	23.2	24.9	24.5	23.6	24.9	24.6	26.4	26.4	26.5	25.5	26.5	25.1	26.1	26.2	26.3
E.O.B.H.	33.5	37.5	39.0	38.3	40.1	40.5	41.2	43.1	38.3	38.0	39.4	36.2	40.9	39.3	43.7
Superficial velocity ft/sec	10.8	11.0	12.0	11.6	11.6	11.7	11.7	11.4	10.7	10.9	10.7	10.7	12.0	12.2	12.0
Economizer															
Overall coefficient Btu/hr ft^2 °F	35.1	35.2	36.8	32.6	32.9	32.5	31.8	33.5	34.0	33.2	32.3	32.3	39.8	32.7	30.4
Inside coefficient Btu/hr ft^2 °F	292.8	291.6	302.3	305.0	277.2	267.8	275.1	268.2	271.4	265.7	255.4	259.2	285.8	271.2	254.3
Outside coefficient Btu/hr ft^2 °F	43.6	43.8	45.8	39.5	40.5	40.2	39.0	41.8	42.5	41.4	40.2	40.1	51.2	40.4	37.3
Calculated avg. wall temp. °F	402.7	404.8	405.5	397.7	415.5	423.3	417.8	419.4	420.7	418.7	415.1	407.6	423.9	406.8	413.5
Superheater															
Overall coefficient Btu/hr ft^2 °F	35.5	35.8	34.6	30.8	34.2	35.1	35.9	34.6	34.5	33.1	34.0	33.3	37.7	34.6	37.2
Inside coefficient Btu/hr ft^2 °F	242.3	237.4	233.2	211.6	224.7	231.9	243.5	241.6	239.5	233.1	239.5	225.9	242.9	243.8	244.4
Outside coefficient Btu/hr ft^2 °F	46.0	46.6	44.8	39.4	44.5	45.7	46.5	44.5	44.3	42.3	43.9	43.0	49.6	44.3	48.6
Calculated avg. wall temp. °F	588.2	594.1	593.1	589.4	588.4	591.4	594.1	588.2	588.5	587.9	587.9	587.3	604.8	589.8	598.3

	Superheater	Economizer
Surface Area	11.316 ft^2	11.316 ft^2
Tube O.D.	1.506 in	1.506 in
Tube I.D.	1,216 in	1,216 in
No. of Tubes Total	4	4
No. of Tubes/Pass	2	2

As more tests are run, additional test data will be developed which will allow correlation of the various factors to estimate h_o.

Test No. 700-17 was conducted to determine the effect of superficial velocity on the in-bed heat transfer rate. During this test the superficial velocity was varied from 11.9 to 6.5 ft/sec with sufficient bed material to maintain the Effective Operating Bed Height (E.O.B.H.) above the top of the Tube Bundle. Table 60 shows the operating conditions maintained during the test. The calculated heat transfer coefficients are shown in Table 61.

The data in Tables 60 and 61 show that the heat transfer rate increases with increasing superficial velocity up to about 11 ft/sec. Above 11 ft/sec, the heat transfer rate declines slowly. This is consistent with previous studies performed in the FBM (see Progress Report No. 37, October 1975). The increase in the outside coefficient is small in the range of velocities tested, amounting to a maximum increase of only 18 percent for the economizer and 15 percent for the superheater.

These heat transfer coefficients can be used to explain the decrease in bed temperatures observed at lower superficial velocities. In going from 12 to 6 ft/sec, the amount of heat released is cut in half. However, the in-bed heat transfer coefficient only decreases 15 to 18 percent. If the bed temperature remained constant, more heat would be removed by the in-bed tube bundle than was available. Thus to maintain thermal equilibrium, the bed temperature drops.

2.4.2 Hot Bed Transfer Light-Off Studies

A study was made of the light-off procedure used at Rivesville for lighting Cell C by hot bed transfer from Cell D. This study resulted in a mathematical model of the hot bed transfer technique. The results of this model should be useful for future designs.

The initial condition for light-off of Cell C by transfer of hot material from Cell D consists of a high hot bed in Cell D and a shallow cold bed in Cell C. The transfer occurs rapidly with respect to any inflow of feed or

Table 60

Operating Conditions for Test 700-17 5/23/78

Time / hr	13:00	13:30	14:00	14:30	15:00			
Bed Temp., ^{o}F	1535	1535	1521	1440	1383			
O_2 in flue gas, %	3.3	2.0	1.4	2.1	2.5			
Coal feed rate, lb/hr	697	715	653	555	*			
Fly ash collection rate, lb/hr	219	213	177	115	115			
Limestone feed rate, lb/hr	431	390	390	348	358			
Bed drain rate, lb/hr	310	310	310	208	208			
Stack dust rate, lb/hr	15.6	*	18.4	*	22.8			
Main air flow rate, lb/hr	7779	7082	6543	5460	4572			
Total air flow rate, lb/hr	8080	7383	6844	5761	4873			
Superficial velocity, ft/sec	11.9	10.9	10.1	8.1	6.5			
Static bed depth, inches	26.8	26.8	24.8	24.8	25.3			
EOBH (effective operating bed ht.), inches	40.0	42.5	35.0	35.0	35.0			
Steam Flow lb/hr	4191	4212	3930	3592	3127			

* Data not taken.

Table 61

Calculated Heat Transfer Coefficients
for Test 700-17 5/23/78

Bed mean particle size = 1926 microns

Time /hr	13:00	13:30	14:00	14:30	15:00
Bed temperature °F	1535	1535	1521	1440	1383
Bed depth (static) inches	26.8	26.8	24.8	24.8	25.3
E.O.B.H. inches	40.0	42.5	35.0	35.0	35.0
Superficial velocity ft/sec	11.9	10.9	10.1	8.1	6.5
Economizer					
Overall Coefficient *	31.3	31.6	30.0	28.9	27.1
Inside Coefficient *	271.1	276.0	263.1	257.4	244.3
Outside Coefficient *	38.2	38.8	36.6	35.0	32.8
Calculated Avg. wall Temp °F	404	402	399.	384	383
Superheater					
Overall Coefficient *	33.8	34.2	32.9	31.3	29.2
Inside Coefficient *	222.6	223.7	211.3	195.2	173.7
Outside Coefficient *	43.9	44.5	43.1	41.1	38.6
Calculated Avg. wall Temp °F	609	609	614	602	601

*All Heat Transfer Coefficients have units of: BTU/HR/ft^2/°F.
All coefficients are uncorrected for Bed Temperature.

outflow of solids. Therefore, the sum of the mass of solids contained in both cells is assumed to be constant during transfer. The mass of bed solids contained in a cell is a linear function of bed pressure drop above 2 in bed height as was found from calibration data.

The mass balance above 2 in bed height can, therefore, be written in terms of the static pressures observed as follows:

Final bed mass in both cells = Initial bed mass in both cells.

$$A_C(P_C - P_{FC}) + A_D(P_D - P_{FD}) = A_C(P_{CI} - P_{FCI}) + A_D(P_{DI} - P_{FDI})$$

where:

A_C and A_D are the respective grid plate areas of Cell C and D; P is a static pressure in inches of water and; the subscript F refers to the freeboard static pressure in each cell.

Upon opening the slide gate, it is assumed that a dynamic energy balance can be written such that the kinetic energy of the material passing through the slide gate is equal to the change in potential energy of the fluid. Losses are related by an orifice discharge coefficient, C, as follows:

$$V = C\sqrt{2G\Delta H}$$

where:

V is the velocity of discharge through an orifice of area, A_o, G is the gravitational constant, ΔH is the difference in potential energy of the top of the two beds. Relating P_C to P_D by this equation and solving for P_D, a quadratic equation in $C(T-T_o)$ is obtained where T_o is the lag time (seconds to open the slide gate. Solving this by the quadratic formula the following equation is obtained.

$$C(T - T_o) = \frac{\sqrt{2}}{\sqrt{G}\left(1 + \frac{A_D}{A_C}\right)\left(\frac{A_o}{A_C}\right)} \left[(P_{DI} - P_{CI})^{\frac{1}{2}} - \left(\left(1 + \frac{A_D}{A_C}\right) P_D - \frac{A_D}{A_C} P_{DI} - P_{CI}\right)^{\frac{1}{2}}\right]$$

For light-off of Cell C this becomes:

$$C(T - T_o) = 9.77 \left[(P_{DI} - P_{CI})^{\frac{1}{2}} - (1.62\ P_D - 0.62\ P_{DI} - P_{CI})^{\frac{1}{2}} \right]$$

The right-hand side of this equation can be calculated as a function of time, given experimental values of P_D at various times after initiation of hot bed transfer. This can be plotted as a "calculated" $C(T - T_o)$ versus observed time. The discharge coefficient can be obtained directly as the slope of the straight line and the abscissa intercept yields the observation lag time.

It was found that the straight line fit of the data (tabulated in Table 62) to the correlation is very good. The observed lag times of 2.0 and 7.0 sec agree well with the specified actuator lag time. The orifice coefficients of 0.49 and 0.22 demonstrate the range of values to be expected. The lower value may indicate incomplete slide gate opening or partial blockage of the opening.

The conclusions obtained from the Rivesville operating experience are that hot bed transfer light-off has been demonstrated to be highly successful and reliable. Light-off of a single cell or multiple cells simultaneously is rapid and practical. The use of a dynamic energy balance allows the successful correlation of the flow of hot solids between compartments. Orifice discharge coefficients between 0.2 and 0.5 are obtained. Lag times observed are consistent with the actuator stroke time. It can be further speculated that light-off is enhanced by high initial bed height differences and high initial bed temperatures. The high rate of bed transfer with respect to the expected mixing rates implies that a uniform bed temperature is not likely in the receiving bed. This nonuniformity may cause actual lightoff to be faster than that calculated by the present method. The total pressure difference between cells appears to govern the mass flow rate and the final heights. The method of pressurizing the freeboard of the delivery cell with respect to the receiving cell, would lead to greater rates of bed transfer and faster light-off without the necessity of actually having a high initial bed height difference.

2.4.3 Sulfur Capture in the CBC

A great deal of concern arises when considering the overall emission of sulfur oxides from a fluidized-bed boiler if substantial sulfur is combusted in the CBC and if no sulfur capture occurs under those conditions. For example, if combustion of a 4.5 percent sulfur coal were anticipated and if 10 percent of that sulfur were emitted uncaptured

Table 62

Test Data for Hot Bed Transfer from Cell D

Test No. 1

Time (sec)	ΔP (inches of water)
0	33
10	28
20	22
30	18
40	17
50	17
60	16

h_{ci} = 7.5 inches of water

Test No. 2

Time (sec)	ΔP (inches of water)
0	32.5
10	28.5
20	26.5
30	21.5
40	17.5
50	16.5

h_{ci} = 5 inches of water

Gate opening = 169 inches^2

A_d = 72 ft^2

A_c = 120 ft^2

from the CBC, a 93 percent sulfur capture would be required in the main cells to obtain the overall sulfur capture of 83 percent required by present regulations. If, on the other hand, 80 percent sulfur capture occurred in the CBC, only 84 percent sulfur capture would be required in the main cells. This difference could cause a large error in limestone consumption estimates for the main cells. An analysis of the sulfur capture reaction was carried out using thermodynamic calculations, along with data from the CBC test series (Test Nos. 644-1 to 644-12). The analysis indicates that sulfur capture at high efficiencies can be obtained in the CBC if it is designed and operated properly.

The free energy of the following reaction was calculated as a function of temperature.

$$SO_2 + CaO + 1/2O_2 \rightleftarrows CaSO_4$$

From this, the equilibrium constant was determined and equilibrium concentrations of SO_2 and O_2 were calculated. The results are shown in Figure 54. It can be seen that at 1800°F the conditions are favorable for sulfur capture down to equilibrium levels of 550 ppm at all oxygen concentrations above 10^{-7} atm. Furthermore, as long as oxidizing conditions are maintained in the bed at all places, sulfur capture can be expected to occur up to above 2000°F.

Considerable data has been presented in the literature describing the optimum temperature for sulfur capture at between 1500°F and 1600°F whereas very little information is available at more extreme temperatures. Early results from the Alexandria FBM indicated that sulfur capture at 1800°F was between 30 and 60 percent of that at 1550°F. This is within the range that can be obtained by extrapolating most literature sulfur capture data at 1800°F. During the recent Alexandria FBM CBC test series, Test Nos. 644-1 to 644-12, records were kept of sulfur capture as a function of operating conditions. These data are presented in part in Table 63 and Figure 55. The main results of these data are summarized in Table 64. It was found that when fly ash (approximately 40 percent carbon) was fed at average rates of 323 lb/hr/feed point, 90-98 percent sulfur was retained in the bed. At higher feed rates, less sulfur was captured and under conditions of very high feed rates per feed point, no sulfur was captured. These results indicated that it was extremely important to have many feed points and a low feed rate per feed point in order to obtain good sulfur capture under CBC conditions. Also, this distributes the recycled limestone most effectively in the CBC freeboard when significant sulfur capture occurs under CBC conditions. Another important parameter was excess O_2 which should remain above 2 percent in the flue gas.

THERMODYNAMIC EQUILIBRIUM CONCENTRATIONS OF SO_2 AND O_2 IN AN FBM ENVIRONMENT

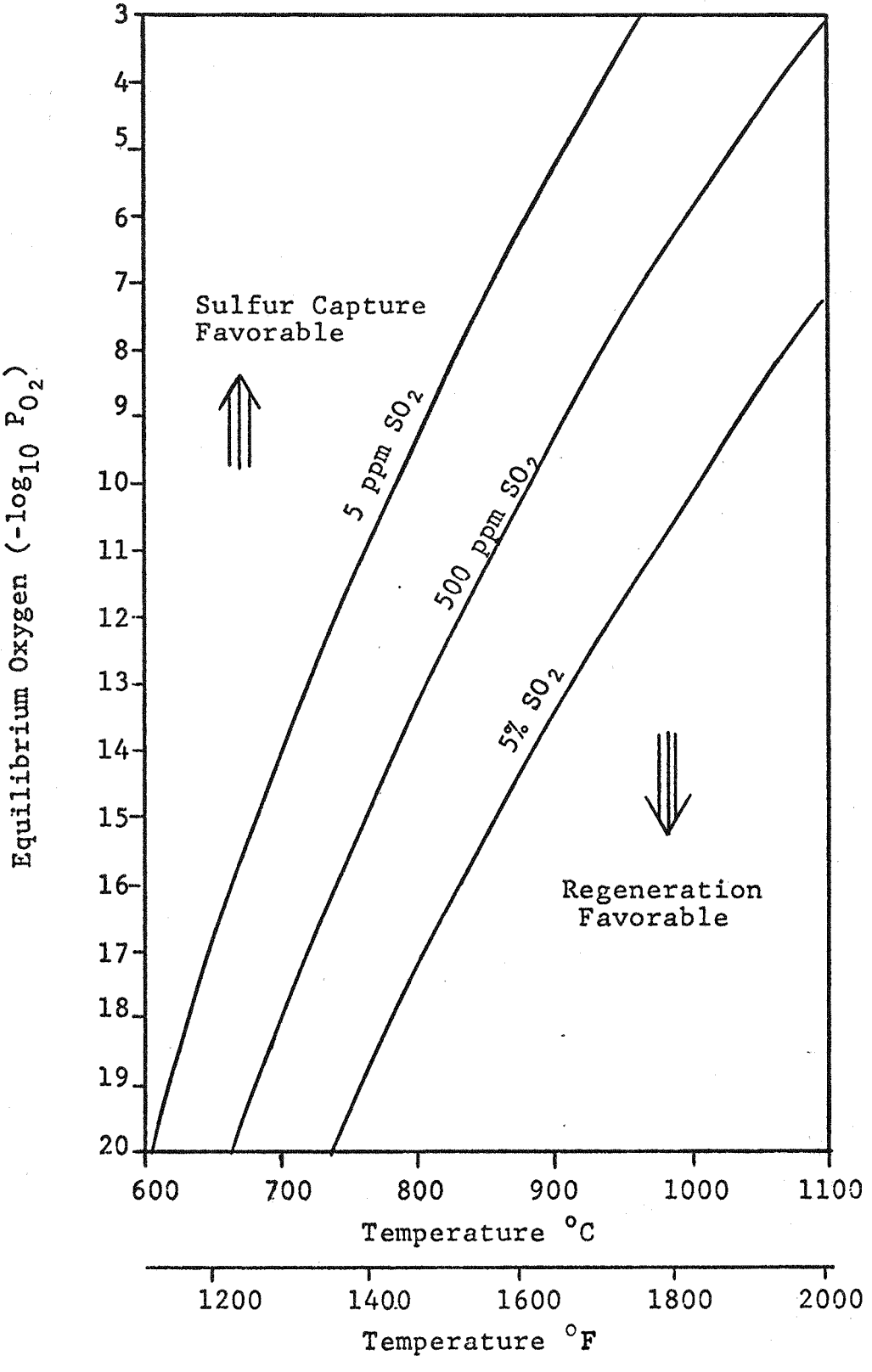


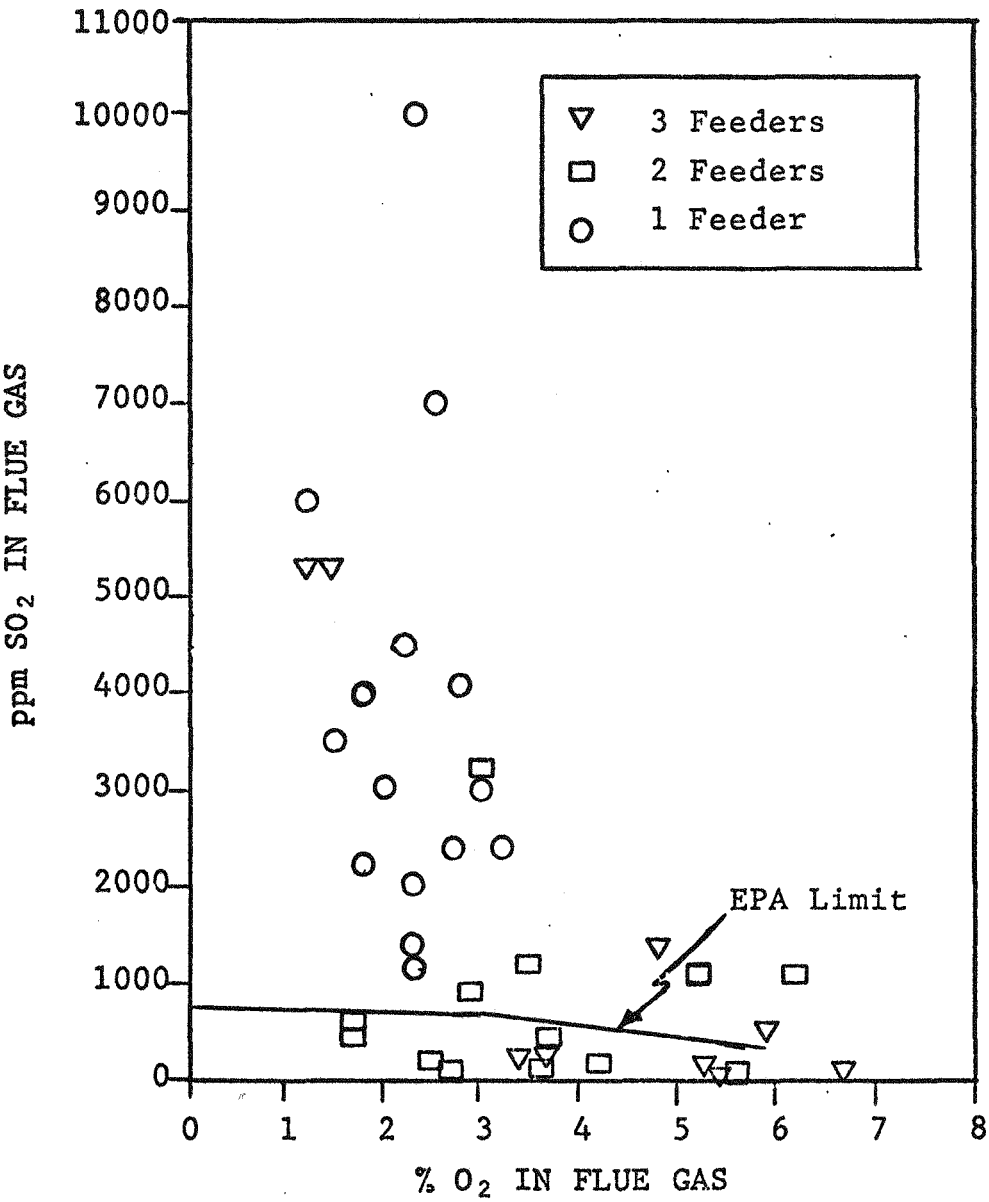
Figure 54

Table 63

Number 644 Test Series Results

Test No.	Time	Feed Points	Flue Gas O_2 (%)	Fly Ash Feed Rate Lb/hr/ feed point	Equivalent Feed SO_2 (ppm)*	SO_2 Emissions (ppm)
4	13:00	3	5.4	320	5155	70
6	13:30	3	3.6	300	4992	260
	15:00	3	1.4	310	4598	5000
	15:40	3	1.6	300	4384	5000
7	11:30	2	2.9	431	4541	900
	12:00	2	3.5	525	4531	1200
	13:00	2	1.7	420	4922	550
	13:30	2	2.5	430	5059	150
	14:00	2	2.6	525	4768	100
	14:30	2	1.7	420	4800	450
8	10:00	1	3.0	540	4640	3000
	11:00	1	2.7	620	4672	2400
	12:00	1	2.2	650	4544	4500
	12:30	1	3.2	590	4736	2400
	13:30	1	2.3	730	4576	10000
	14:00	1	2.5	660	4640	7000
	14:30	1	2.8	630	4704	4100
9	12:30	2	3.6	425	4544	190
	13:00	2	5.6	270	5088	60
	13:30	2	3.7	270	5120	330
	14:00	2	4.2	390	4992	130
10	11:00	1	2.0	660	4768	3000
	11:30	1	2.3	900	4960	2000
	12:00	1	1.7	803	4928	3500
	12:30	1	1.4	780	4704	6000
	13:00	1	1.8	940	4736	4000
	13:30	1	1.8	930	4544	2200
	14:00	1	2.3	940	5024	1400
	15:00	1	2.3	960	4960	1150
11	10:30	3	3.4	190	4672	170
	11:00	3	6.7	130	5088	170
	11:30	3	5.3	205	5056	180
	12:00	3	5.9	205	4864	570
	12:30	3	4.8	205	4608	1440
	13:00	2	5.2	308	4800	1100
	13:30	2	6.2	308	4928	1100
	14:00	2	3.0	308	4608	3200

*For explanation see text (Recent CBC Test Series Results)



SULFUR DIOXIDE EMISSIONS FROM CBC TESTS (644 SERIES)

Figure 55

Table 64

Summary of Test Results
Sulfur Capture in CBC
(Tests performed in PER Alexandria FBM, 1976-77)

Sulfur Retained In Bed* (%)	No. of Data Points	Feed Rate**
90 - 98	12	323
60 - 90	9	478
<60	16	647

* Percentage of feed sulfur minus uncombusted sulfur.
** Average fly ash feed rate: lb fly ash/hr/feed point.

III. RIVESVILLE, W. VA., MFB PLANT OPERATIONS

3.1 General

During the period from May 15 through June 27, 1978, the 30 MW_e MFB operated for 850 cell hours including 4 in-turbine commercial runs of 24-1/2 hrs, 50 hrs, 26 hrs and 3/4 hrs.

The period included seven (7) performance runs directed at trouble shooting the newly installed data acquisition system and obtaining preliminary performance results.

Through July and into August, the first phase performance testing was completed including stabilized testing of 2 and 3 cell operation to obtain:

- combustion efficiency
- SO_2 and NO_x reduction
- CA/S
- other performance parameters

Major progress has been made in identifying and solving both site specific and technology problem areas. (See Exhibit I and II).

Rivesville is, and will be for several years, the only large (400 sq ft) unit available and suitable to support the Atmospheric Fluidized Bed Demonstration Program. Rivesville will be utilized for maintaining and generating growing interest in both industrial and utility AFB coal burning plants. As a reliable multipurpose facility, Rivesville will be capable of testing key components of the AFBC processes and component systems. For the 1979 - 1983 time period, Rivesville can be the single most important test facility in which technology developers can test their process and component systems without the need to construct their own separate facilities.

3.2 Operational Highlights

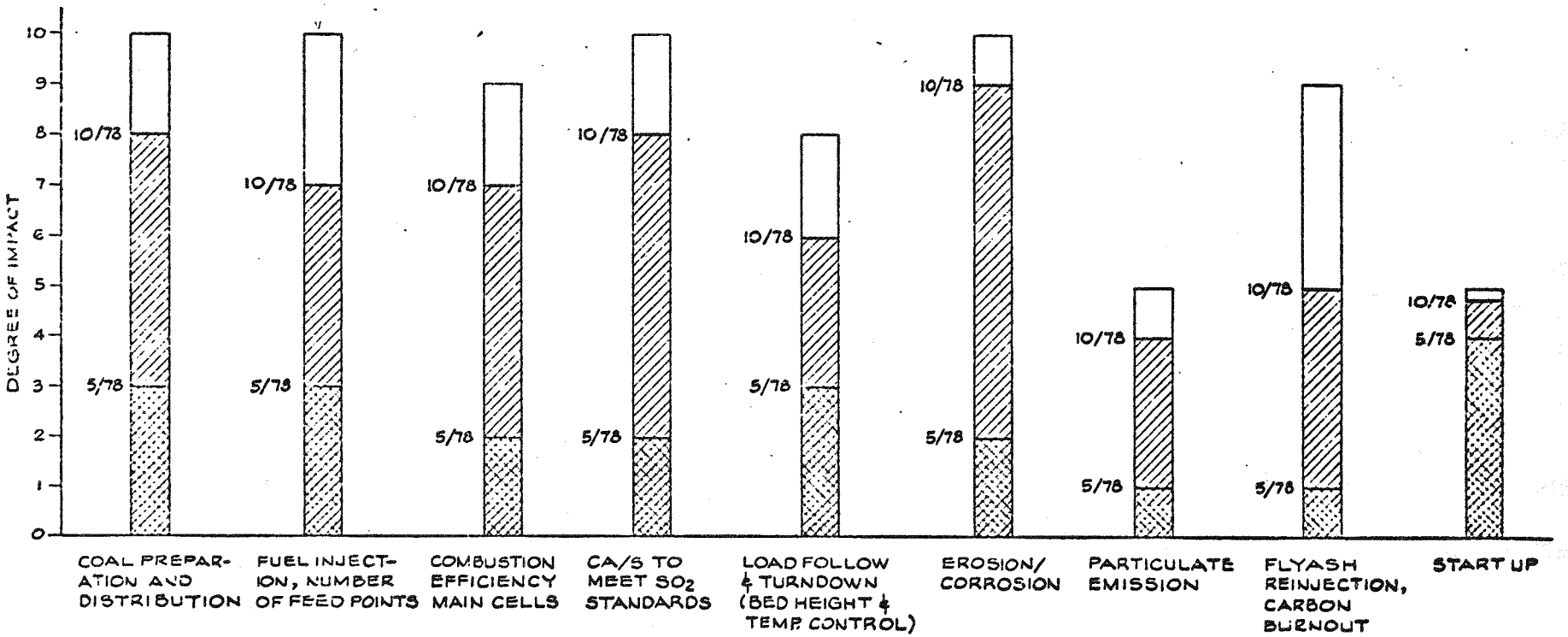
During the period of July 1977 through June 1978, several important events occurred. Highlights of these events in chronological order follow.

In July 1977, the steam safety valves were set. Safety valve SV-2 was lifted on July 13, 1977 and set at a relief pressure of 1600 psig, with a reseating pressure of 1550 psig. Valve SV-1 was lifted on July 14, 1977 and set at a relief pressure of 1545 psig with a reseating pressure of 1500 psig. The superheater safety valve, SV-3 was lifted on July 15, 1977, with saturated steam set at a relief pressure of 1420 psig with a reseating pressure of 1340 psig.

U.S. DEPARTMENT OF ENERGY

RIVESVILLE 30 MWe AFBC PILOT PLANT

COMMERCIALIZATION DEMONSTRATION
PROOF OF TECHNOLOGY APPRAISAL INDEX



CONFIDENCE LEVEL:

DATE: 5/78 LEVEL 21÷79 = 27 %

10/78 LEVEL 56÷79 = 71 %

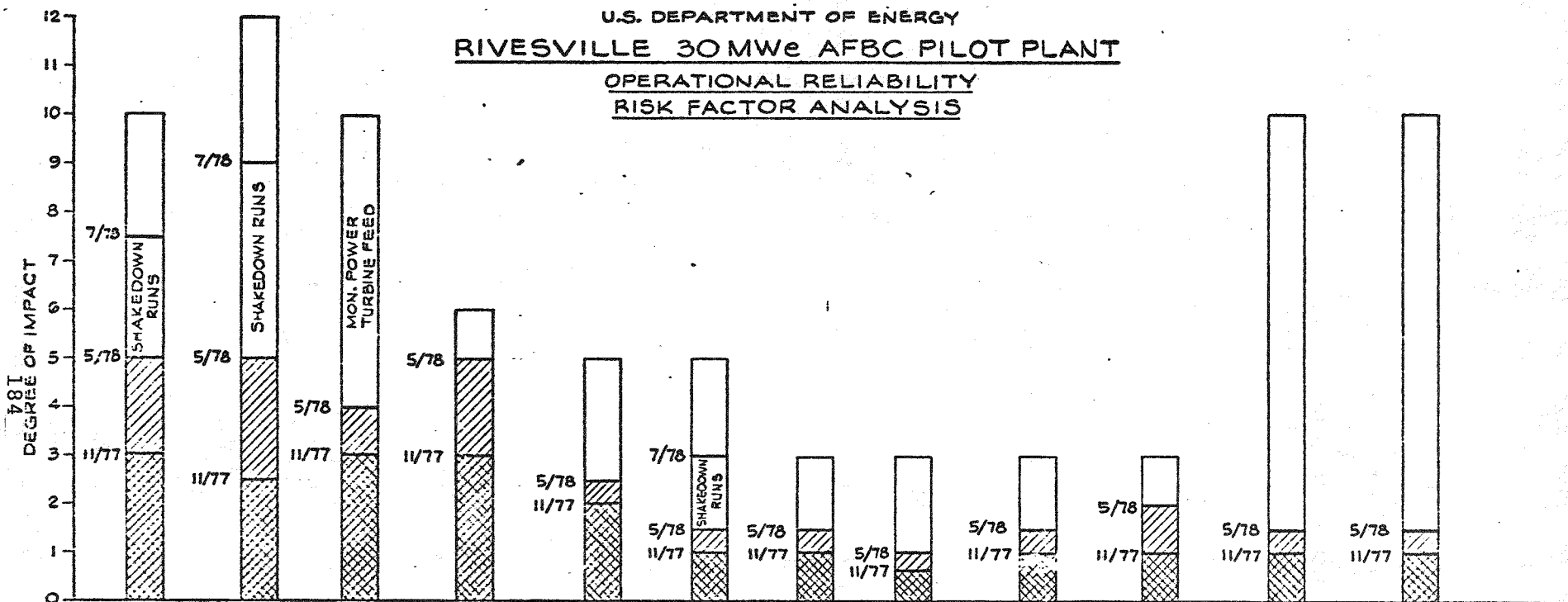
POPE, EVANS AND ROBBINS INC.

EXHIBIT I

U.S. DEPARTMENT OF ENERGY

RIVESVILLE 30 MWe AFBC PILOT PLANT

OPERATIONAL RELIABILITY
RISK FACTOR ANALYSIS



RIVESVILLE "PILOT PLANT"
OPERATIONAL AVAILABILITY OBJECTIVE 50 %

"COMMERCIAL PLANT"
OPERATIONAL AVAILABILITY 80%
(REQUIRES HIGH COST AND LONG SHUT DOWN PERIOD)

CONFIDENCE LEVEL:

DATE: 11/77 LEVEL 20 ÷ 80 = 25 %
5/78 LEVEL 32 ÷ 80 = 40 %
7/78 LEVEL 40 ÷ 80 = 50 %

POPE, EVANS AND ROBBINS INC.

EXHIBIT II

During the month of August 1977, the first operation of A cell and the plant dedication took place. A cell was ignited for the first time on August 20, 1977. The unit operated with four cells for 1 hr and 46 min. Dedication of the MFB unit took place on August 26, 1977. Senator Robert Byrd keynoted the event attended by an estimated 400 people.

During the month of September 1977, three goals were achieved. On September 16, 1977, the MFB entered into its first commercial operation, operating in parallel with a conventional boiler and condensing turbine to generate electricity. The unit remained in commercial operation for a period of 1 hr 43 min. The second goal achieved during September 1977 was the completion of five "confidence" runs for Monongahela Power Company, designed to demonstrate MFB load following capabilities. The final achievement made during September occurred on September 30, 1977, when full design load operation was attained and maintained for approximately 29 minutes.

The next significant event occurred in November 1977 with the start and completion of the "Interim Test Plan". This was designed to give preliminary results on the performance of the MFB. Performance factors which were studied were boiler efficiency, combustion efficiency, sulfur capture and gas emissions. Preliminary results were presented at the Fifth International Symposium for Fluidized Bed Combustion held in December 1977.

A major outage occurred from December 1977 to March 1978 due to the United Mine Workers of America's strike. Advantage of this shutdown was taken to install an extensive instrumentation package for unit testing.

On completion of the outage, the unit performed an extended run which yielded 202 hours of continuous operation of the CBC. This occurred between April 24, 1978 and May 3, 1978.

Beginning May 15, 1978, an extensive test program was initiated to characterize unit performance. Results analyzed included boiler efficiency, combustion efficiency, sulfur capture, gas emissions and particulate emissions. The test plan continued through June 1978.

Between May 11, 1978 and May 13, 1978 the unit maintained steady commercial operation for 50.3 hrs, the longest period to date.

3.3 Operations

During the period from July 1977 to June 1978, MFB operations accomplished several milestones. Exhibit III shows the total operating hours for the reporting period by month and brief narratives on specific MFB operations follow.

On August 20, 1978, the MFB entered into the first four cell operation. The initial lightoff of A cell was routine as the result of experience gained from previous C and B cell lightoffs. Routine lightoff consists of building bed height in operating cells to 30 - 38 in. A 5 percent coal charge, by weight, is added. The slide gate is then opened and ignition occurs approximately 1 minute later.

A cell operated for 1 hour 48 minutes. Bed temperature was 1400°F and feedwater flow 305,000 lbs/hr. A cell was shutdown due to overloading of Dust Collector No. 1.

On September 16, 1977 the MFB entered into the first commercial operation of a fluidized bed boiler. Steam entered the turbine at 1150 psig, 900°F and 132,000 lb/hr. Exhibit IV illustrates operating conditions during this run. Commercial operation halted after 1 hour and 43 minutes.

Between July 1977 and June 1978, there have been a total of seven commercial runs, the longest being 50 hours and 20 minutes on May 1978.

Following the first commercial operation, Monongahela Power Company requested demonstration of the load following capability of the MFB. Five "confidence" runs were performed in September and October 1977 for this purpose. Conditions for each run are shown in Exhibit V. The runs were completed in three (D, C and B) cell operation.

Because of a tube leak in MPC's No. 7 boiler after Run D, the decision was made to operate the MFB in various conditions and to observe performance. A cell was ignited and steam flow increased to 304,000 lbs/hr, exceeding the MFB's nameplate rating. D cell was then shut down due to high fly ash temperature leaving DC-1. This would determine whether steam temperature could be controlled with only A and C cells in operation. It was observed that there was little or no control of steam temperature in this mode. Consequently, A cell was shut down due to high tube metal temperatures. C cell remained in operation and an attempt was to be made to re-ignite D from C, but was aborted due to a low condensate tank level.

RIVESVILLE MFB BOILER PLANT
SUMMARY OF COAL FIRED OPERATING HOURS
PERIOD JULY 1977 THRU JUNE 1978

Month	Hours Of Operation				
	CBC Cell	C Cell	B Cell	A Cell	Steam To Turbo-Generator
July	102.5	26.2			
August	143.8	38.0	20.3	1.8	
September	152.3	67.1	30.7	.7	1.8
October	172.3	36.5	20.2		1.8
November	183.6	53.1	28.6		9.2
December	87.7	13.1	8.7		5.0
January, February, March	Coal Strike Outage				
April	145.5	58.2	.2		
May	239.4	102.2	78.5		50.3
June	198.5	55.9	26.7		28.8
Total	1425.6	450.3	213.9	2.5	96.9

EXHIBIT III

EXHIBIT IV

SUMMARY OF OPERATIONS DURING FIRST COMMERCIAL RUN

MFB CONFIDENCE RUNS
OPERATING CONDITIONS

Run	Run Time	Date	Steam Press. PSIG	Steam Temp. °F	Steam Flow lb/hr
A	2 hrs.	9-26-77	1225	935	150,000
B	4 hrs.	9-29-77	1190	935	160,000
C	2 hrs.	9-29-77	1280	900	140,000
D	4 hrs.	9-30-77	1240	930	145,000
E	4 hrs.	10-04-77	1265	930	185,000

EXHIBIT V

A new technique was developed for CBC lightoff. Instead of establishing "reverse flow" through CBC to heat the bed, a 7 percent coal charge, by weight, is placed in CBC after the burners reach high fire. The bed is then slumped and combustion air pressure reduced. Bed temperature is monitored and the bed is "fluffed" periodically. With this method, CBC can be ignited in 1.5 hours from the time of burner ignition. Exhibit VI shows D cell temperature during typical lightoff.

With the addition of new seven million Btu/hr burners, CBC lightoff was again modified. The burners are brought to high fire. A coal charge is injected into the bed and the bed is allowed to remain fluidized. When bed temperature reaches 800°F, coal is fed and ignited. This method has proved extremely fast, approximately 1 hr from burner lightoff.

The origin of many MFB operational and mechanical problems has been coal fines. When coal containing a high percentage of fines enters the vibrating feeder, the fines are entrained in the system. This may eventually cause pluggage which has a "slugging" effect on the boiler. Pictured in Exhibits VII and VIII are examples of B, C and D cell excess O_2 and bed temperatures. Particularly noteworthy is the extreme fluctuations of the excess O_2 in C cell. Such fluctuations are indicative of normal boiler operation hampered by the "slugging" effect of the fines.

An important aspect of MFB operation is unit response to "trip" or emergency situtations. It has been discovered that the unit can be restarted after a trip-out. This is illustrated in Exhibit IX. From the time the unit tripped until both cells were back in operation was approximately 40 minutes. Exhibit X shows a unit trip on November 30, 1977 caused by electrical failure in the boiler safety circuit. The unit was down for 2.5 hours in order to correct the failure. Another trip situation is shown in Exhibit XI when, on November 27, 1977, the Auxiliary F.D. fan tripped due to faulty electrical overload protectors. In this case, the unit could not be restarted due to Auxiliary F.D. fan problems and coal became coked in the feed needles.

As the experience of the operators improved, the operating reliability of the unit will improve. Expected goals are 3 - 5 hours startup while "on line", increased efficiency, both boiler and combustion; fewer mechanical problems and improved overall unit performance.

D-CELL BED TEMPERATURE

DATE: NOV. 9, 1977

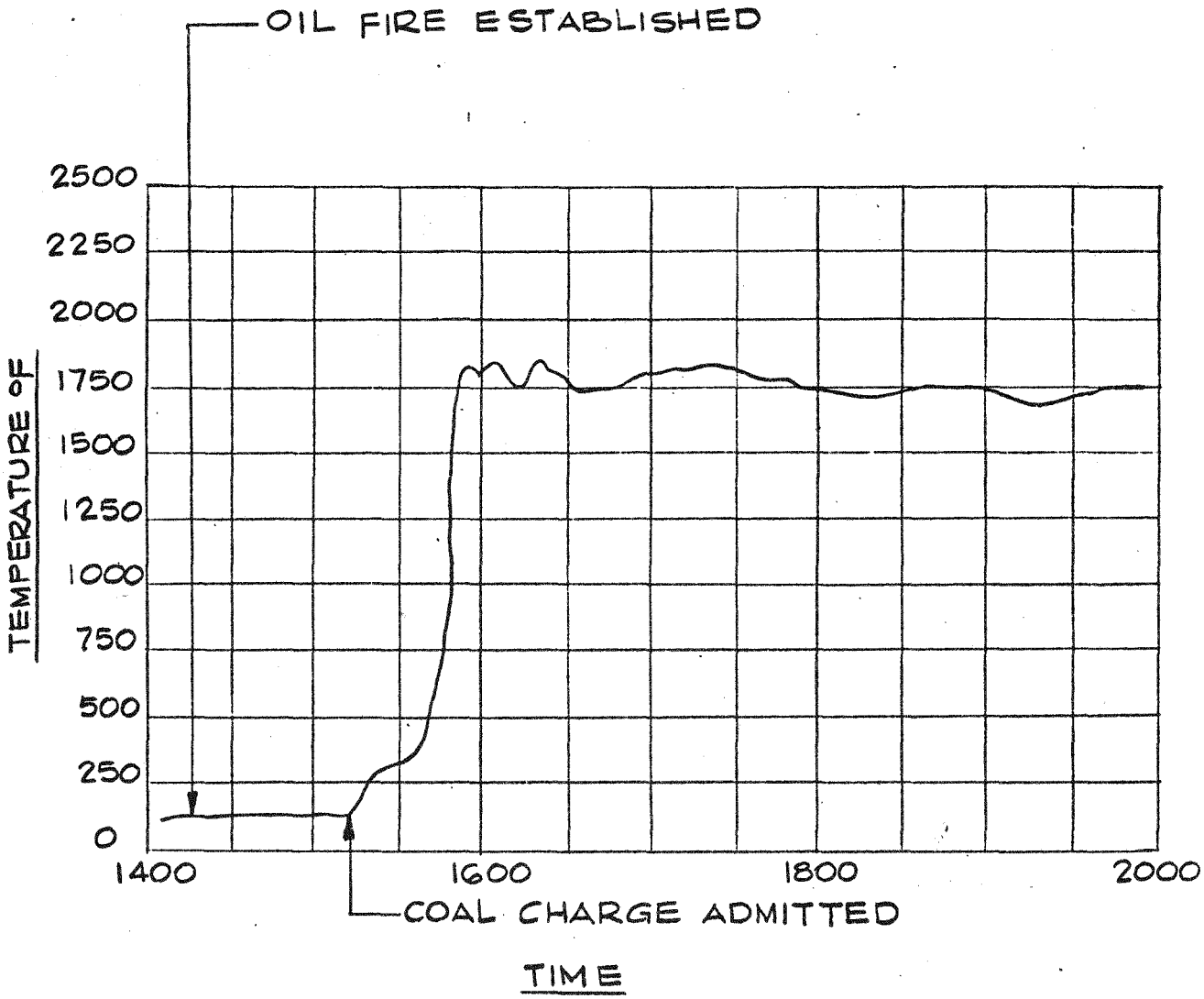


EXHIBIT VI

EXCESS O_2 READINGS FOR "CBC", "C" AND "B" CELLS DURING CONFIDENCE RUN "E"

CONDUCTED: OCT. 4, 1977

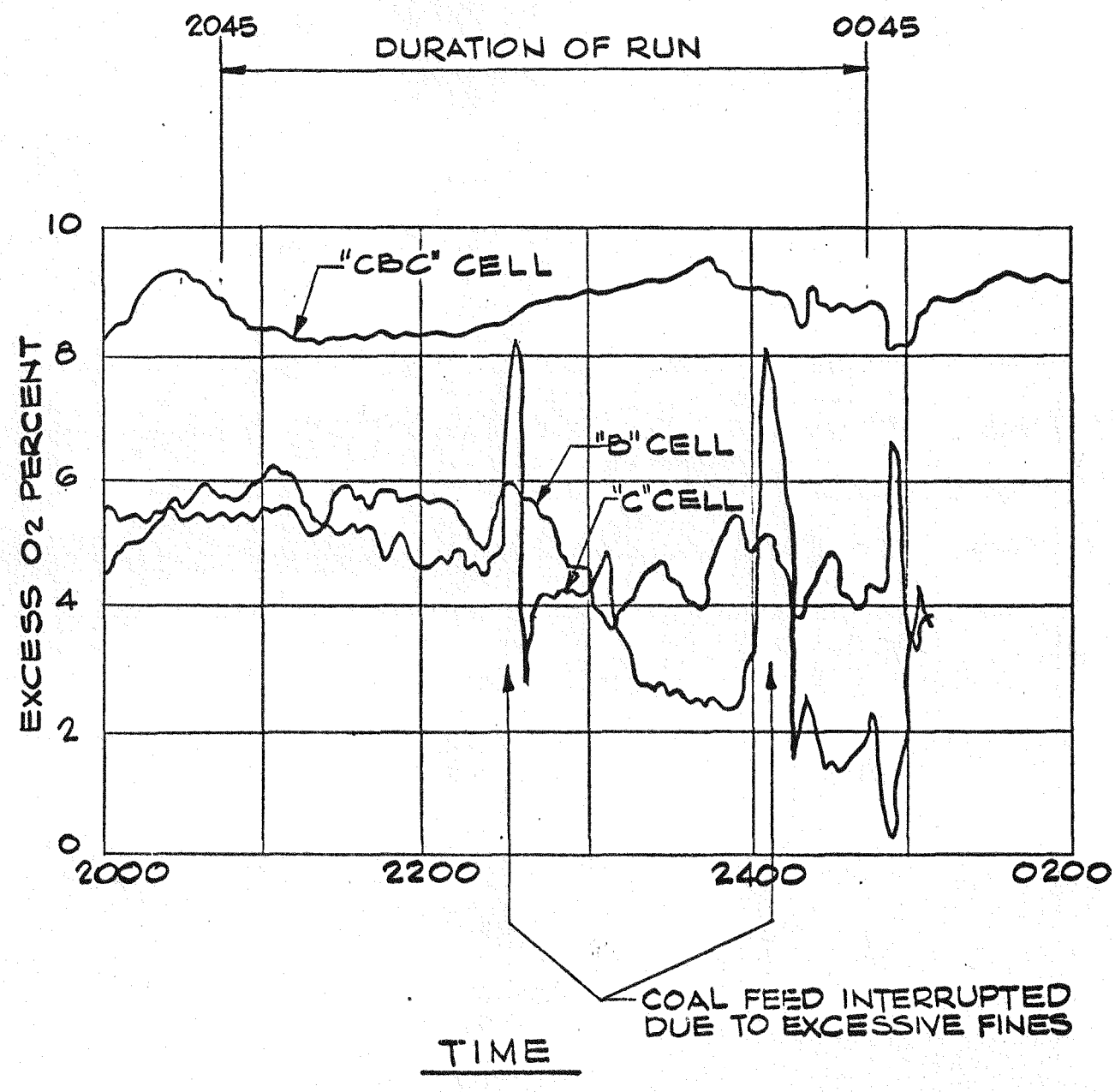


EXHIBIT VII

CONFIDENCE RUN "E"

DATE: OCT. 4, 1977

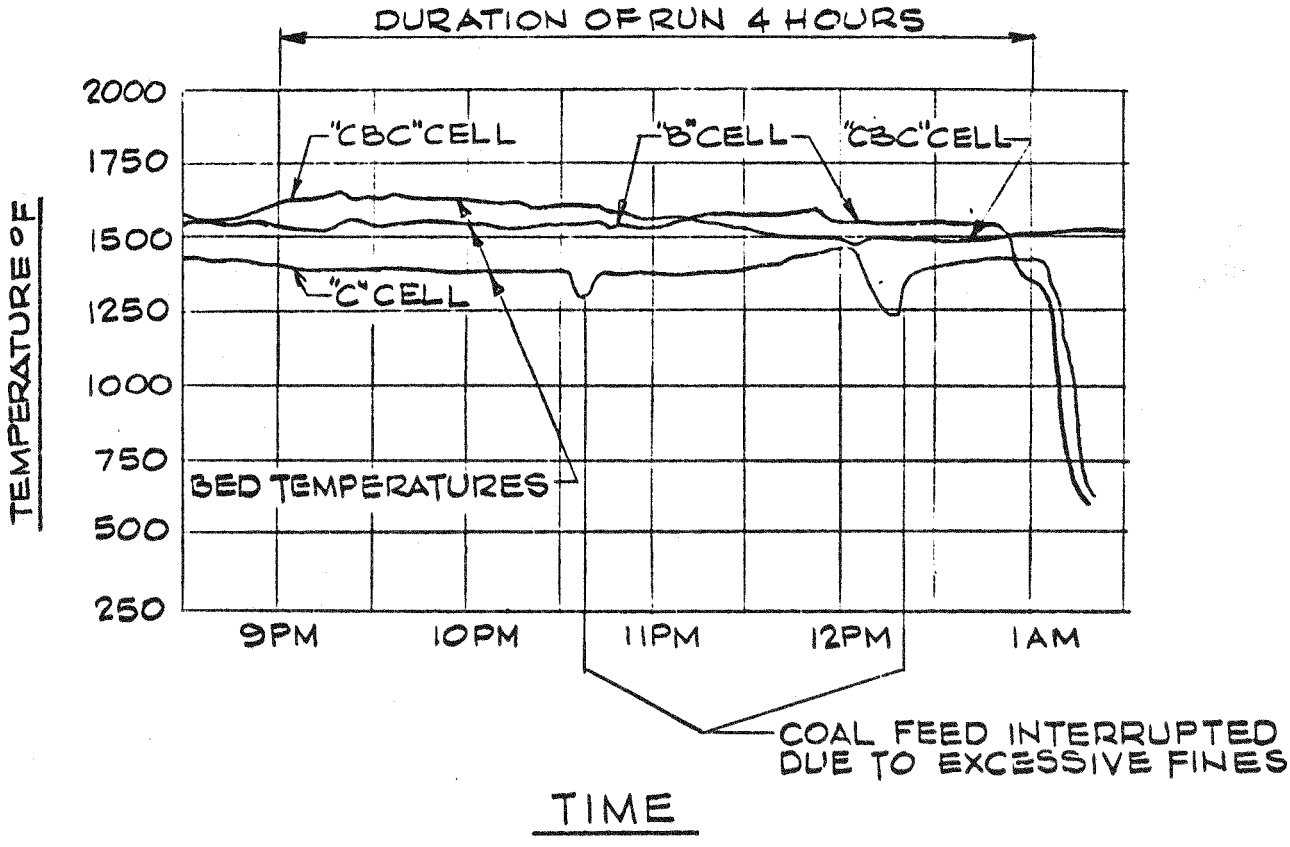


EXHIBIT VIII

CELLS "C" AND "D" BED TEMPERATURE

DATE: JULY 13, 1977

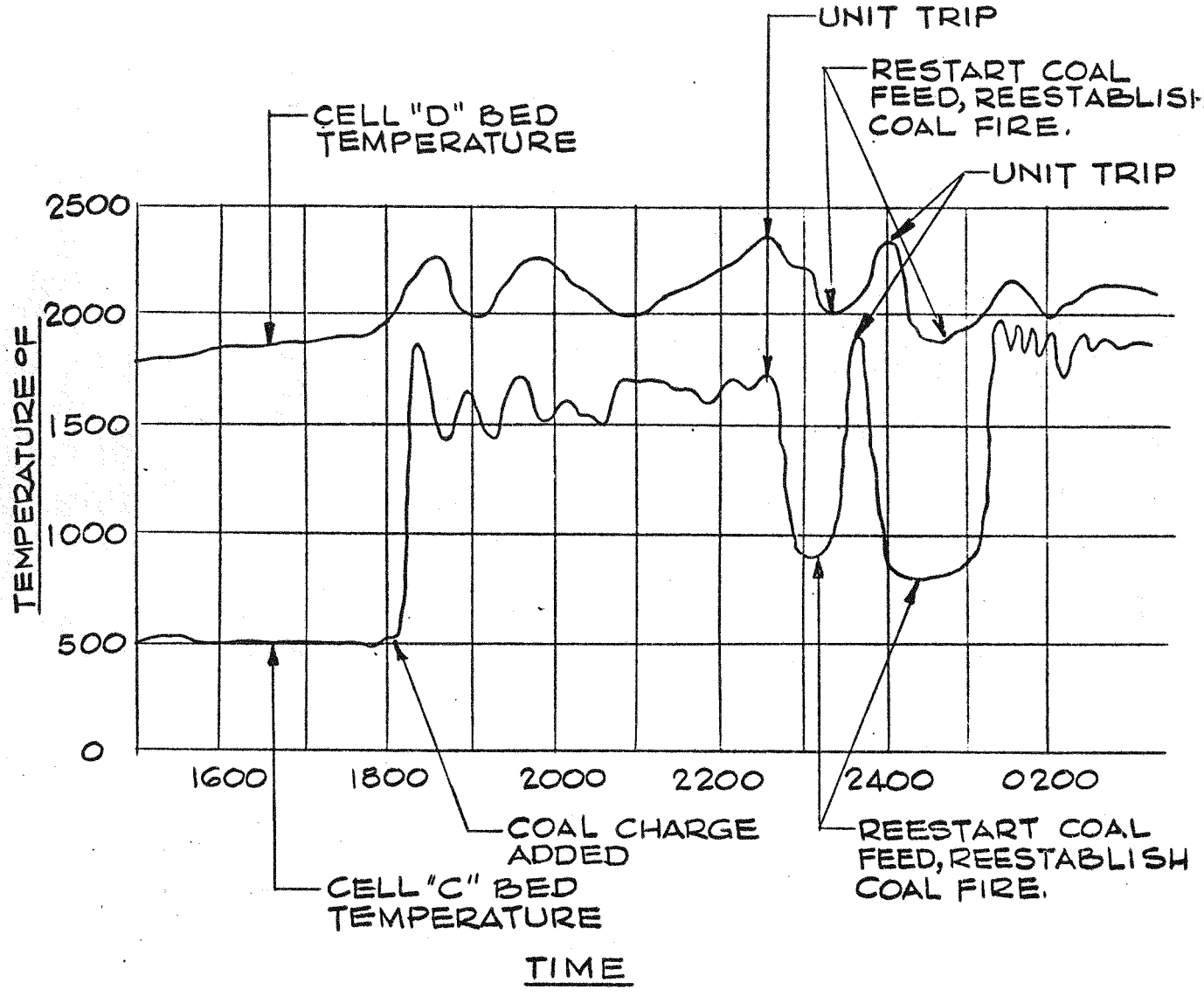


EXHIBIT IX

"C" AND "CBC" CELL BED TEMPERATURES AFTER A UNIT TRIP

TEST CONDUCTED NOV. 30, 1977.

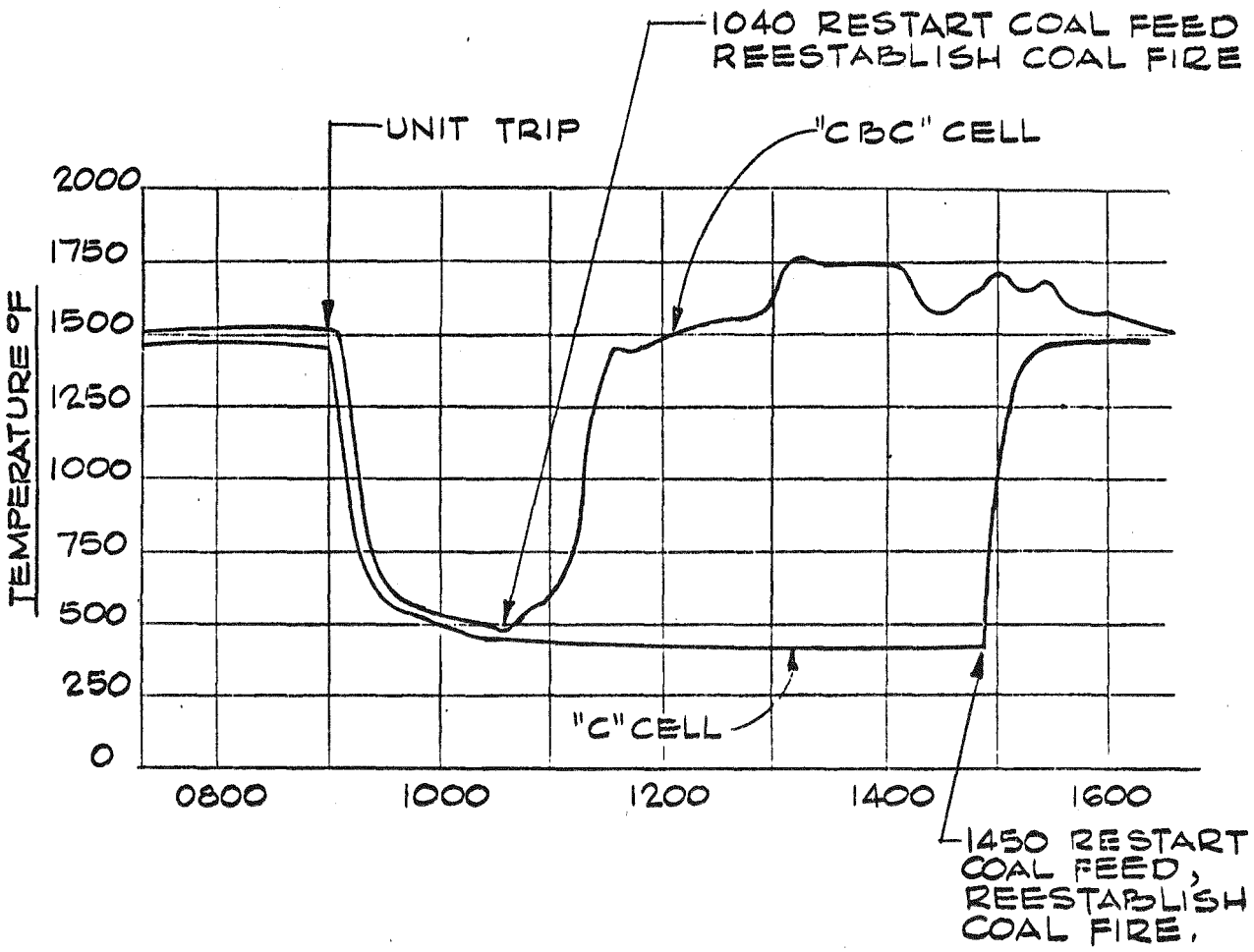


EXHIBIT X

"CBC" CELL BED TEMPERATURE AFTER A UNIT TRIP

DATE: NOV. 27, 1977

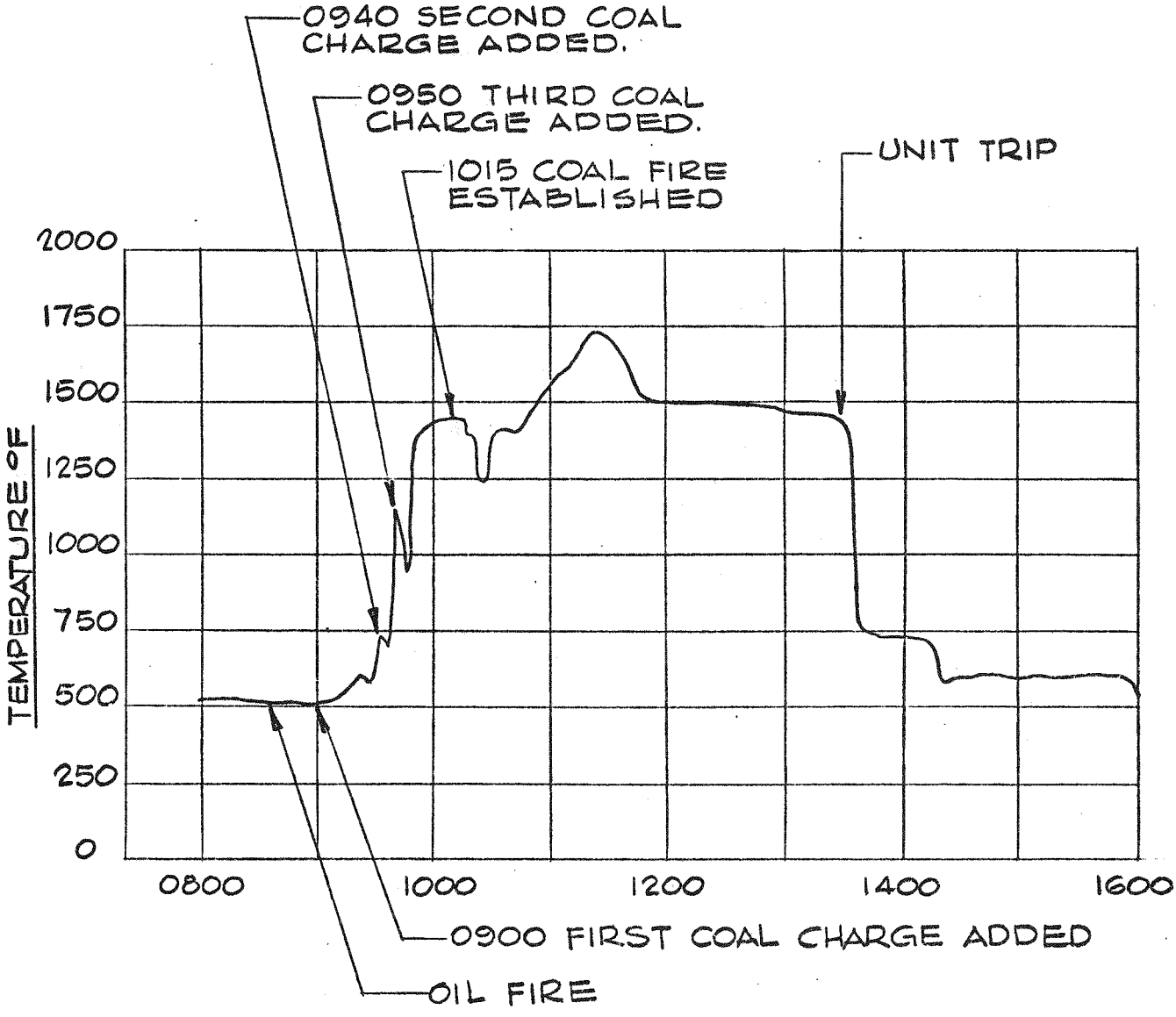


EXHIBIT XI

3.4 Plant Modifications

Operation during the past year has produced significant advances in fluidized bed combustion technology. Operation in excess of 2,000 hours has also indicated the necessity for modifications because of limitations imposed by the interface with Monongahela Power Company. Modifications to reduce equipment failure, have now been time-proven successful. Other major modifications are in the form of additional equipment used to better monitor the process of fluidized bed combustion and provision for redundant auxiliaries.

3.4.1 Fuel Feed System

As coal fines continued to create pluggage in the feed pipe above the "three" series rotary feeders, the existing Allen-Sherman-Hoff units were replaced by a larger capacity vented Newman rotary feeder. The motor gear reduction ratio was also increased to aid in prevention of coal back-up in the feed pipe. The venting of coal fines reduced the pluggage problem but still larger feed capacity was deemed necessary. During the UMW strike, the entire coal feed system was redesigned excluding the three series rotary feeders.

An 18 inch standpipe provided a straight drop from the coal bins to gravimetric belt feeders. Each belt feeder is provided with a digital integrator that indicates both mass rate of coal feed and an accumulated weight fed across the belt. The discharge of the belts is to the existing vibrating feeders. Subsequent to the installation of the belt feeders, each belt and its integrator was calibrated by distribution tests similar to the distribution tests performed on the rotary feeder system. Coal fines accumulation was noticed in the housings of these belt feeders after continuous operation. In April 1978, pressure equalization lines were installed between the belt and vibrating feeders. These lines keep the belt feeder pressure slightly higher than the vibrating feeder pressure and the fines from backing up into the belt feeder housings. Exhibits XII and XIII show the Feed System at the beginning and end of the Reporting Period.

3.4.2 Feedwater Supply

The original feedwater control valve (BFWV-4) was designed for a flow coefficient (C_v) of 195 which proved too much for control at low flow operation. This was confirmed by tests during which the feedwater isolation valve (BFWV-2) was throttled down to reduce BFWV-4 inlet pressure.

RIVESVILLE MFB
FUEL FEED SYSTEM CONFIGURATION
JUNE 1977

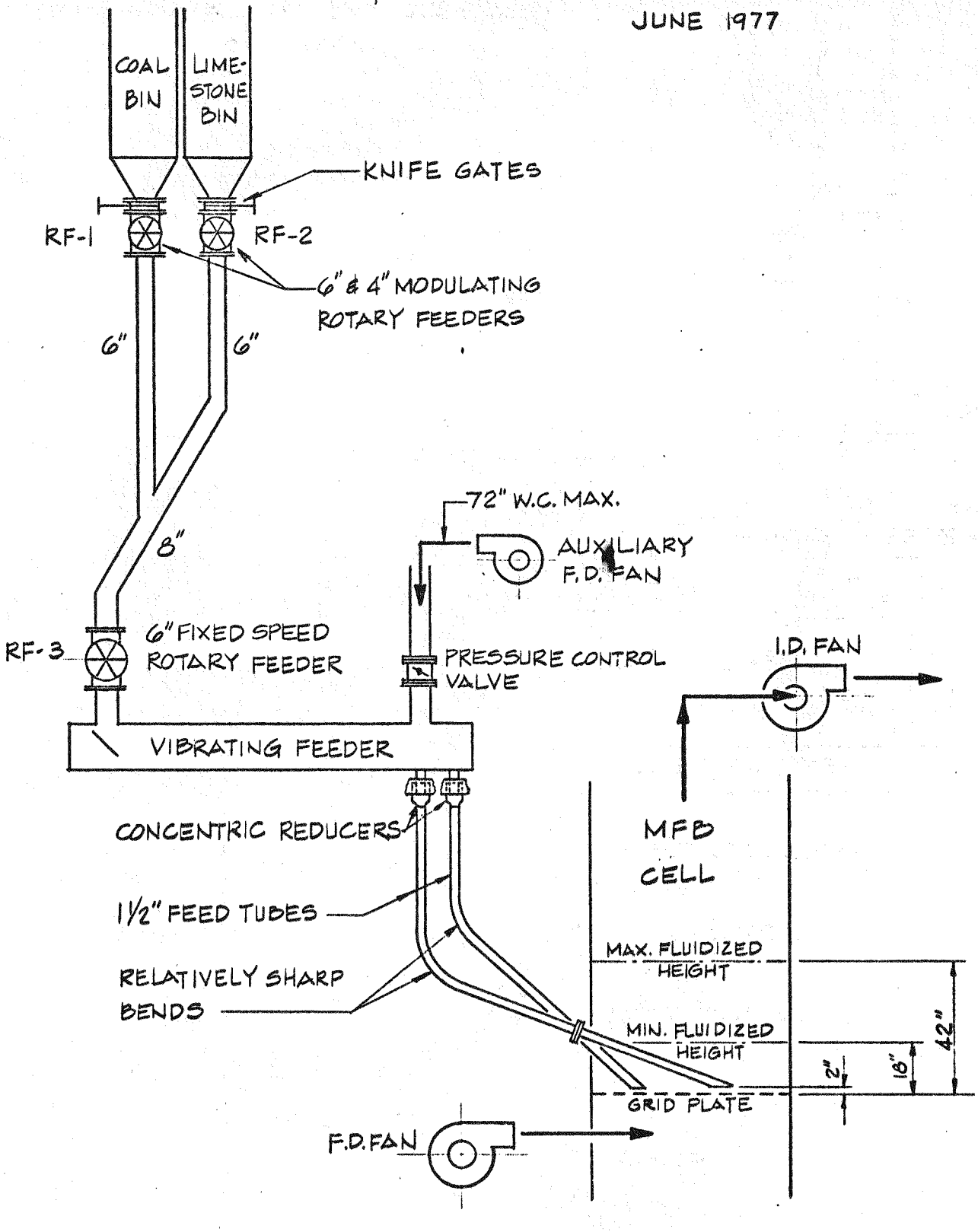
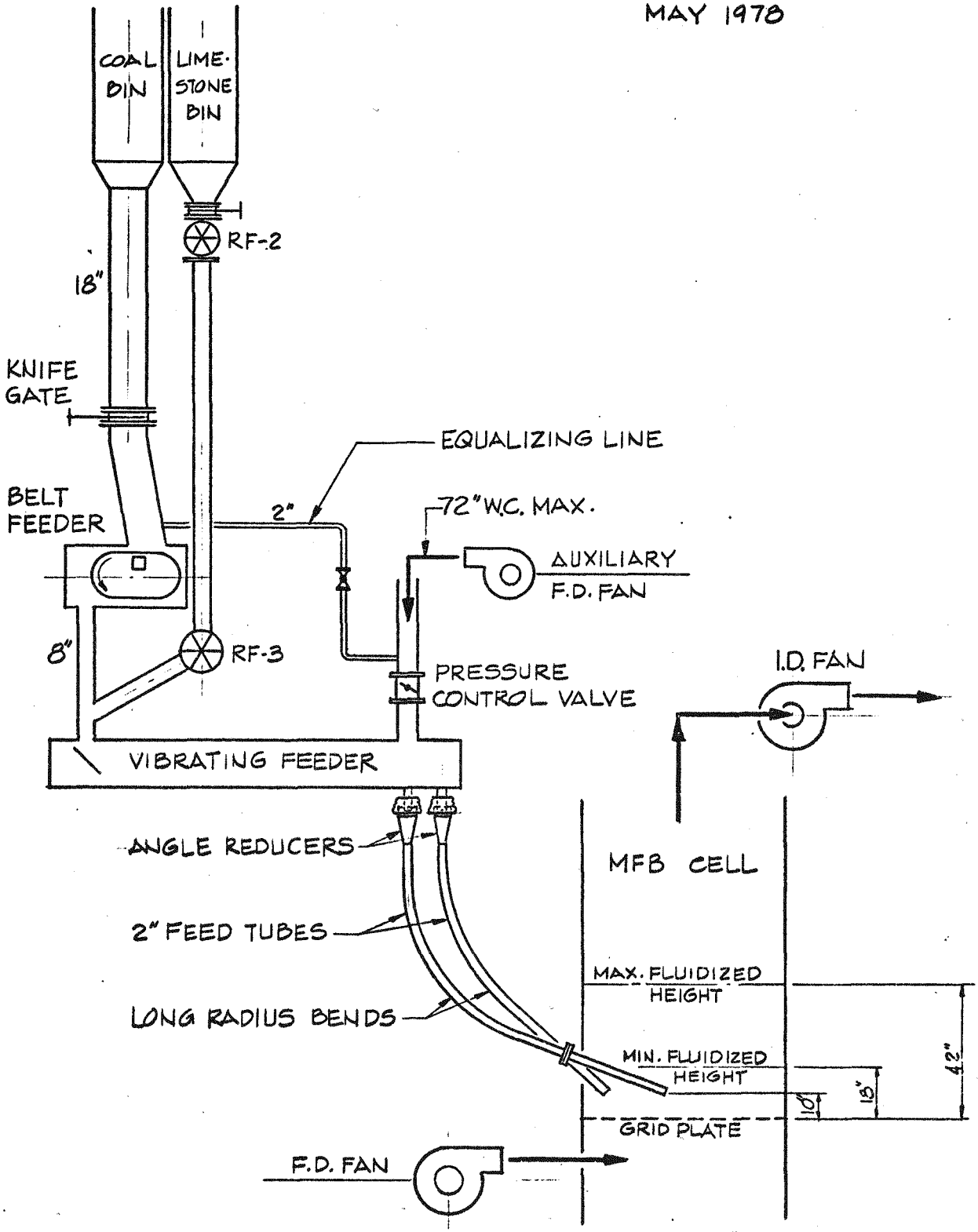


EXHIBIT XII

RIVESVILLE MFB
FUEL FEED SYSTEM CONFIGURATION
MAY 1978



To compound the problem, closed valve leakage was estimated to approach 50,000 lb/hr. The valve was disassembled for inspection and several factors noted which contributed to the excessive leakage. The seats had several nicks and scratches apparently caused by feedwater line debris. Additionally, the flexitallic seat ring gasket was badly damaged and provided virtually no seal. Further cavitation on the body's seat ring gasket surface prevented proper gasket compression of the seat ring and pressure balance sleeve gaskets. Similar cavitation contributed to closed valve leakage on the attemperator spray control, surface and bottom blowdown and the economizer recirculation valves.

The valve plug of BFWV-4 has been changed from a linear C_v of 195 to an equal percentage C_v of 110. The flow coefficient of BFWV-6 was also decreased from 28 to 12.

A secondary, or valve bypass system will be installed for operation at low flow. BFWV-4 will be bypassed by a valve with a C_v of 2 and used during one cell operation. BFWV-6 bypass will control steam temperature in three cell operation. The original BFWV-6 will be used for four cell operation.

3.4.3.1 By-pass Chutes

Coal conveyed through the existing hammermill crusher was degraded to fines that created plugging in Redlers downstream of the crusher. This necessitated construction of a by-pass chute. Manually operated at the crusher, this chute bypasses only the crusher. A similar bypass has been constructed around the MPC crusher.

3.4.3.2 Limestone Blower

To assit unloading limestone into the storage bunker, a diesel powered blower has been installed. This blower acts as a booster to the blower on the limestone delivery trucks.

3.4.3.3 Limestone Baghouse

To collect dust created by unloading of limestone, a baghouse has been added which captures this dust and conveys it to the new ash silo. As dust-laden air enters the fabric filter, the air velocity decreases allowing large particles to fall into the hopper. Fine particles are borne into the bag area. The air passes through the felt media, depositing the dust on the outside of the bag. A timer operates solenoid valves for momentary pulses of compressed air for sequential filter bag cleaning.

3.4.3.4 Coal Dryer

Due to the possibility of F.D. air transporting coal dust into the precipitator and creating a potential explosion hazard, the coal dryer has been sealed at its existing inlet and outlet dampers.

3.4.4 Silencers

Initial multicell operation of the MFB required a continuous steam blow-off to atmosphere. This condition resulted in excessive noise to the nearby community. A consultant noise control group recorded sound levels at various locations and elevations within the town limits of Rivesville and out-lying areas. During an outage period in August 1977, an ambient sound level was established. Measured and projected sound levels from steam blow-off were compared to the ambient level and it became apparent that the existing silencers did not have the required attenuating capability.

As a result of these sound level studies, the existing silencers have been replaced with upright AAF-Pulsco blowdown mutes. These silencers are a multi-tube design with a floating module of highly absorbent fiberglass. Blow-off noise levels with these mutes are still unacceptable and further studies are being conducted.

3.4.5 Air Compressors

Due to the increased demand for control air during multicell operation, the existing Dresser air compressors were unable to maintain adequate pressure. To supplement damper control air supply, two Ingersoll Rand Model ESV 9 1/2 x 7 NL oil free vertical air compressors have been installed. Each unit has a capacity of 199 ACFM at 100 psig, 514 rpm and is driven by a 50 H.P., three phase, 460V motor. The operation of this air system is such that one unit functions at a base load, and the second starts on demand. This sytem is designed with inherent load control and high temperature protection circuitry.

3.4.6 Observation Doors

Observation Doors

Six existing observation doors located on the front and rear walls of the three main cells, have been modified to provide more complete viewing of the bed surface. The wallbox sleeves were reworked to increase the field of vision from outside the boiler. Exhibit XIV illustrates the observation door installation.

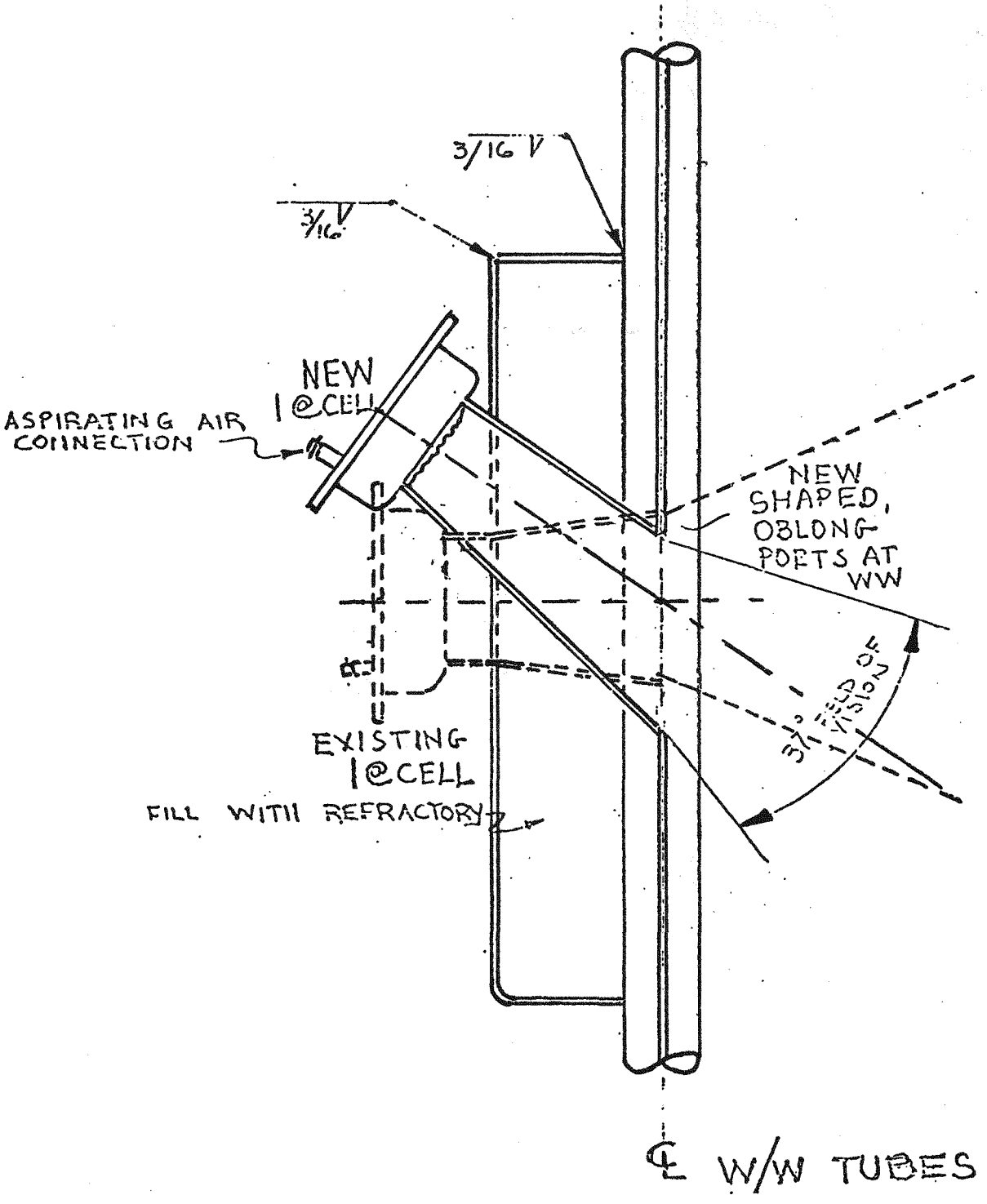

3/16 V
3/16 V
NEW
I@CELL
ASPIRATING AIR
CONNECTION
NEW
SHAPED,
OBLONG
PORTS AT
WW
31½° FIELD OF
VISION
EXISTING
I@CELL
FILL WITH REFRACTORY
℄ W/W TUBES

EXHIBIT XIV

3.4.7 Instrumentation Package

During the UMW strike, an extensive instrumentation package was installed. This package was designed to provide data on all phases of MFB operation. The instrumentation package is broken up into three sections; centrally located gauge boards, temperature monitoring and flue gas analysis.

- Centrally Located Gauge Boards

  Pitot venturi airflow measuring devices are installed in the plenum of each cell. Magnehelic and manometer gauges have been constructed in two central locations. These gauge boards are detailed in Exhibit XV.

- Temperature Monitoring

  Various temperatures throughout the entire process are recorded by a continuous scan of thermocouples by the Doric Data Logger. The Doric provides both a printout and a data cassette from the input of 392 thermocouples. Operating temperatures are monitored by thermocouples in the bed and at three elevations within the boiler. Water temperatures are monitored by chordal thermocouples implanted into the vertical waterwalls.

- Flue Gas Analysis

  Flue gas analysis is obtained from the Beckman Analyzer. Seven streams from the ductwork are brought to the analyzer by heat traced lines. The seven streams are:

  - gas leaving A cell
  - gas leaving B cell
  - gas leaving C cell
  - gas leaving CBC cell
  - gas in the combined main uptake
  - gas entering the air preheater
  - gas leaving the air preheater

Gases are analyzed for percentage of: SO_2, CO_2, CO, O_2 NO_x and HC.

3.4.8 Fuel Oil Burners

The original four million Btu/hr lightoff burners have been replaced with seven million Btu/hr burners. The newly installed burners utilize propane pilots for ignition of

TEST PLAN GAUGE BOARD 'A'

-Point#	Description	-Gauge-
2028	E.S.P. outlet draft	U-tube manometer
2029	A.P.H. gas out draft	U-tube manometer
2009	F.D. fan discharge pressure	Magnehelic 0-150" w.c.
2010	A.P.H. air out pressure	Magnehelic 0-150" w.c.
2030	I.D. fan out draft	Inclined manometer

TEST PLAN GAUGE BOARD 'B'

2012	Cell 'A' plenum pressure	Magnehelic 0-100" w.c.
2013	Cell 'B' plenum pressure	"
2014	Cell 'C' plenum pressure	"
2015	Cell 'D' plenum pressure	"
2031	Aux. F.D. fan out pressure	Magnehelic 0-150" w.c.
2020	Cell 'A' bed P	U-tube manometer ±24" w.c.
2021	Cell 'B' bed P	"
2022	Cell 'C' bed P	"
2023	Cell 'D' bed P	"
3065	Aux. F.D. fan airflow	U-tube manometer ±18" w.c.
3057	Cell 'A' airflow	Well manometer 0-8" w.c.
3058	Cell 'A' airflow	"
3059	Cell 'B' airflow	"
3060	Cell 'B' airflow	"
3061	Cell 'C' airflow	"
3062	Cell 'C' airflow	"
3063	Cell 'D' airflow	"
3064	Cell 'D' airflow	"
2024	Main uptake pressure	Inclined manometer
2027	Cell 'D' uptake pressure	"
2037	Cell 'A' uptake pressure	"
2038	Cell 'B' uptake pressure	"
2039	Cell 'C' uptake pressure	"
3054	Main airflow - Air foil	Inclined manometer

EXHIBIT XV

fuel oil instead of the high voltage spark that ignited the original burners. Boiler waterwalls were re-routed to accommodate the larger burners. Construction was accomplished during the strike outage early in 1978.

These greater heat generating burners have reduced cold startup ignition time.

3.4.9 Flyash Reinjection

The original fly ash header has been removed and a redesigned header with four reinjection ports has been installed.

Vibrating feeder VF-DE has been internally modified with diverter plates to feed only four needles of the CBC. Reinjection from the electrostatic precipitator has been blanked off for initial reinjection operation.

Redesign of the fly ash header was necessary to provide sufficient transport velocity for the four reinjection needles.

3.4.10 CO_2 Fire Inerting System

The carbon dioxide fire inerting system for coal bunkers #2 and #6 provides flowing gaseous CO_2 into the bunker through fixed nozzles on indication of a temperature rise in the bunkers. Carbon dioxide is supplied from a refrigerated low pressure storage unit located outside the building and passes through an electric vaporizer for conversion from a liquid to a gas for use as required.

Three control panels monitor the fixed thermocouples and provide visual and audible alarms as well as manual control of CO_2 flow to the bunkers. An alarm panel indicating the positions of 80 thermocouples throughout the three bunkers provide digital readout of temperature. Graphic lights represent each thermocouple location and alarm at temperatures exceeding 160°F.

3.4.11 Ash Silo

The original installation of the MFB provided for bed material transport to Monongahela Power Company's ash silo. Due to the size and nature of MFB waste, it became necessary to erect a separate waste silo. This silo has an 80 ton capacity and an ash conditioner at the discharge for dust control.

3.4.12 Environmental Protection Agency (EPA)

An emissions test program sponsored by EPA is scheduled to commence in late 1978. The program will be divided into five separate elements;

1) Comprehensive analysis of emissions,
2) Identification of atmospheric trace elements,
3) Performance analysis of Electrostatic Precipitator,
4) Emission control alternatives,
5) Analysis of bed material leachate.

In preparation for the program, various sampling ports and access platforms have been provided.

3.5 Maintenance

Repair or replacement of the MFB related systems was performed by the Maintenance Department on an as required basis throughout the reporting period. This department is divided into three specialized groups; Electrical, Instrumentation and Mechanical and work in their respective disciplines.

3.6 Performance Testing

During the latter portion of the reporting period, the MFB underwent two extensive testing periods encompassing thirteen tests. A list of each test and test conditions is shown in Exhibit XVI.

The first test period, the Interim Test Plan, was held during November 1977. This consisted of six tests, Nos. 1 through 6. The tests were designed to give preliminary indications of boiler efficiency, combustion efficiency, sulfur capture and gas emissions. Sulfur capture could not be effectively studied due to operational problems with the bed material handling system. After intensive study of the test results, it was decided that only D cell results would be reported. These results are summarized in Exhibit XVII.

Carbon burnup results appear favorable at this time. The highest carbon content of any of the flyash samples leaving D cell was found to be 22 percent carbon in Test #1. The lowest was 2.5 percent in Test #4. The variation in flyash carbon content can possibly be explained by the variation of bed temperature and excess oxygen. For example, the highest excess oxygen was in Test #1 and the lowest carbon content was conincident with the combination of high bed temperature and high excess oxygen.

While the range of carbon contents was found to vary widely from test to test, all of the D cell flyash carbons were found to be consistently lower than the main cell flyash carbon content and also lower than the fly ash carbon content for the Alexandria FBM tests. This indicated that

Test Number	Date	Duration(hrs)	Cells in Service
1	11/09/77	2.5	D
2	11/14/77	5	D & C
3	11/16/77	3.5	D, C & B
4	11/23/77	5.5	D, C & B
5	11/30/77	1.5	D & C
6	11/30/77	1.5	D, C & B
7	5/16/78	4	D
8	5/22/78	4	D & C
9	5/23/78	2.5	D, C, B & A
10	5/25/78	4	D, C & B
11	6/13/78	4	D
12	6/14/78	4	D
13	6/15/78	4	D

Exhibit XVI

D-Cell Operating Conditions and Test Results

Test	1	2	4	5
Bed Temp., °F	1750	1625	1790	1550
O_2 in flue gas, %	3.9	10	7.5	10*
Coal feed rate, lb/hr	2580	1725	2130	2220
Fly ash collection rate, lb/hr	1500	2160	1360	1030
Limestone feed rate, lb/hr	3900	1560	1250	1510
Bed drain rate, lb/hr	0	0	0	0
Main air flow rate, lb/hr	28,000	24,000	23,500	33,000
Total air flow rate, lb/hr	29,580	25,700	25,460	34,750
Superficial velocity, ft/sec	6	4.9	5.3	6.4
Static bed depth, inches	29	27	26	27
NO_x, ppm	350	682	284	120
SO_2, ppm	50	48	51	60
Ca/S	13.5	9.2	5.6	9.2
Combustion efficiency, %	80	74	95	90
Fly ash carbon content, %	22.0	6.5	2.5	12.0

* Based on oxygen balance.

Exhibit XVII

the high freeboard and high excess air design of the CBC should permit carbon burnup. It should be noted that the lowest carbon content in the unburned flyash from the CBC tests at Alexandria was approximately 10 percent carbon. In Tests #2 and #4, D cell results indicated flyash carbon contents below 10 percent.

During the Interim Test Plan, several test limitations were imposed by operating conditions. Due to limited feedwater availability, data acquisition was initiated 1/2 hour after the last operating cell was ignited. Limestone could only be fed in limited amounts due to the inoperative bed material handling system. Another limitation was the isokinetic sampling. Because of the dependency of combustion efficiency on flyash flow rates, the isokinetic must be reliable. During this testing period, reliability of the isokinetics was uncertain.

The second test period Test Plan was begun in May 1978 and encompassed several tests (7-13). Again, the tests were designed to study boiler efficiency, combustion, efficiency, sulfur capture, gas emissions and particulate emissions. Prior to this period, an extensive instrumentation package was installed (see Section 3.4.7). This package allows more accurate measurement of gas analysis, air flows, pressures and temperatures. Isokinetic sampling was subcontracted to GCA, who used a modified ASME method for sampling.

Tests were to be four hours in duration with data taken every 15 minutes. A detailed account of test procedures, methodology, and calculations has been prepared.[1]

Test Nos. 7-10 are preliminary or dry run tests to check out instrumentation, computer programs and procedures. Test Nos. 11-13 represent D cell combustion efficiency with 4 feeders. Exhibit XVIII compares the results of 8 and 4 feeders. Other results include: confirmation of fly ash flows, confirmation of NO_x emissions and the effective increase of combustion efficiency when needle airflow is increased. A detailed report on Test Nos. 7-13 has been prepared.[2]

Exhibit XIX summarizes the results of Test No. 10 which are typical of this series of test runs.

[1] "Performance and Emission Testing Procedure for the U.S. DOE Rivesville MFB", Stone & Webster Engineering Corporation

[2] Rivesville Report on Test Nos. 7 thru 13, July 19, 1978.

COMPARISON OF COAL FEED POINT SPACING IN D-CELL

Test No.	7	8	9	10	11	12	13
Number of feeders	8	8	8	8	4	4	4
Bed area ft^2/feeder	9	9	9	9	18	18	18
Coal feed lb/hr/feeder	230	159	192	174	299	311	300
Needle fuel/air ratio	.42	.37	.61	.47	.51	.50	.48
Bed temp. °F	1390	1390	1650	1470	1275	1365	1438
Superficial Velocity ft/sec	4.7	4.8	5.5	4.2	3.8	4.6	5.1
Bed depth inches (static)	25	24	26	33	27	26	27
Percent O_2	11	12	12	10	12	12	12
lb Fly ash out/lb coal in	.365	.352	.357	.337	.472	.439	.417
Carbon content in Fly ash %	24.8	20.3	12.9	24.0	32.6	40.0	37.1
Combustion efficiency %	88	90	93	89	81	77	79

EXHIBIT XVIII

POPE, EVANS, AND ROBBINS, INC.
RIVESVILLE, WEST VIRGINIA
THREE-CELL TEST SUMMARY

TEST REPORT October 30, 1978
TEST NUMBER 10
TEST DATE May 25, 1978

OPERATING CONDITIONS	CELL			NOTES*
	D	C	B	
Coal Feed, lb/hr	1,340	9,565	8,871	1
Limestone Feed, lb/hr	754	1,264	1,462	2
Coal % S	1.63	1.63	1.63	3
Ca/S (mol ratio)	10.8	2.3	2.9	4
Main Air Flow, lb/hr	23,000	111,000	98,200	5
Steam Flow, lb/hr	164,400			6
Steam Conditions, psig/°F	1,324/921			
Bed Temperature, °F	1,470	1,490	1,390	7
Vs, fps	4.0	11.8	8.3	8
Static Bed Height, inches	33.4	33.1	34.0	9
Oxygen in Flue Gas, %	9.7	1.1	1.7	10
Bed Drain, lb/hr	250	97	0	11
Flyash, lb/hr	468	3489		12
TEST RESULTS				13
Sulfur Capture, %	58	27	38	
SO_2, lb/10^6 Btu	1.13	1.95	0.97	
NO_x, lb/10^6 Btu	0.55	0.29	0.16	
Combustion Efficiency, %	88	83		
Boiler Efficiency, %	76			

*See Next Page

EXHIBIT XIX

NOTES

1. Coal feed was calculated due to malfunctioning belt feeders. Based on Boiler Duty, Unit Heat losses and additional heat inputs in accordance with ASME standards (see sample calculation).

2. Limestone feed rate was measured by revolutions of a rotary feeder.

3. Percent S in the coal is an analytical measurement.

4. Ca/S ratio is a calculated number based on as-fed molar conditions.

5. Main air flow is a measured number, based on pitots in each cell's inlet air.

6. Steam flow is calculated based on feedwater, attemporator, sealwater inlet and outlet flows. Steam conditions are as measured.

7. Bed temperature is an average value of 18-27 in-bed thermocouples.

8. Superficial velocity is based on airflow at bed conditions.

9. Static bed height is a measured value.

10. Oxygen in the flue gas is as measured from the Hagan analyzer due to system leakage in the Beckman gas analysis system. Other gas constitutents were calculated based on leakage derived from O_2.

11. Bed drain was estimated based on rotary feeder speeds.

12. Fly ash was measured by isokinetic sampling.

13. All test results are calculated numbers.

3.7 Organization, Training

Several changes were made in the Rivesville organization. In the Operations Department, one equipment operator, one lead shift supervisor and two relief (Senior) equipment operators were added. In the Maintenance Department, four mechanics and one electrician were added. Two groups of technicians were added to the Results Department. In November, nine part-time technicians were added to take data during the Interim Test Plan. Twenty-five part-time technicians were hired in May as data takers for the Test Plan. These technicians received indoctrination into the MFB systems and presentations on firefighting and safety.

Exhibit XX illustrates the current Rivesville manning schedule. To date, 40 full time employees from operations and maintenance have received full training on the MFB and its support systems. Each class lasted for approximately two weeks. Classes consisted of the following presentations:

1. Introduction to power plants
2. General construction of the MFB
3. Feedwater, condensate and main steam
4. Coal and limestone systems
5. Fuel oil system
6. Bed material handling system
7. Service water system
8. Instrument air system
9. Air supply and flue gas system
10. Chemical feed and blowdown
11. Fly ash collection and reinjection
12. Turbine generators
13. Electrical system
14. Fire and safety
15. Start-up

These classes were designed to acquaint personnel with plant operations as an aid to operator training.

3.7.1 Management Advisory Group

In order to shorten communication paths, a Management Advisory Group was formed which met monthly to review operations, counsel plant management and keep all involved organizations informed. This group consisted of representatives of:

- U.S. Department of Energy
- Stone & Webster Engineering Corporation
- Foster Wheeler Energy Corporation
- MITRE/METREK
- Pope, Evans and Robbins Incorporated

POPE, EVANS AND ROBBINS, INCORPORATED
MULTICELL FLUIDIZED BED BOILER
Rivesville, West Virginia

Revised - 9/20/78 (DH)

Organization Chart

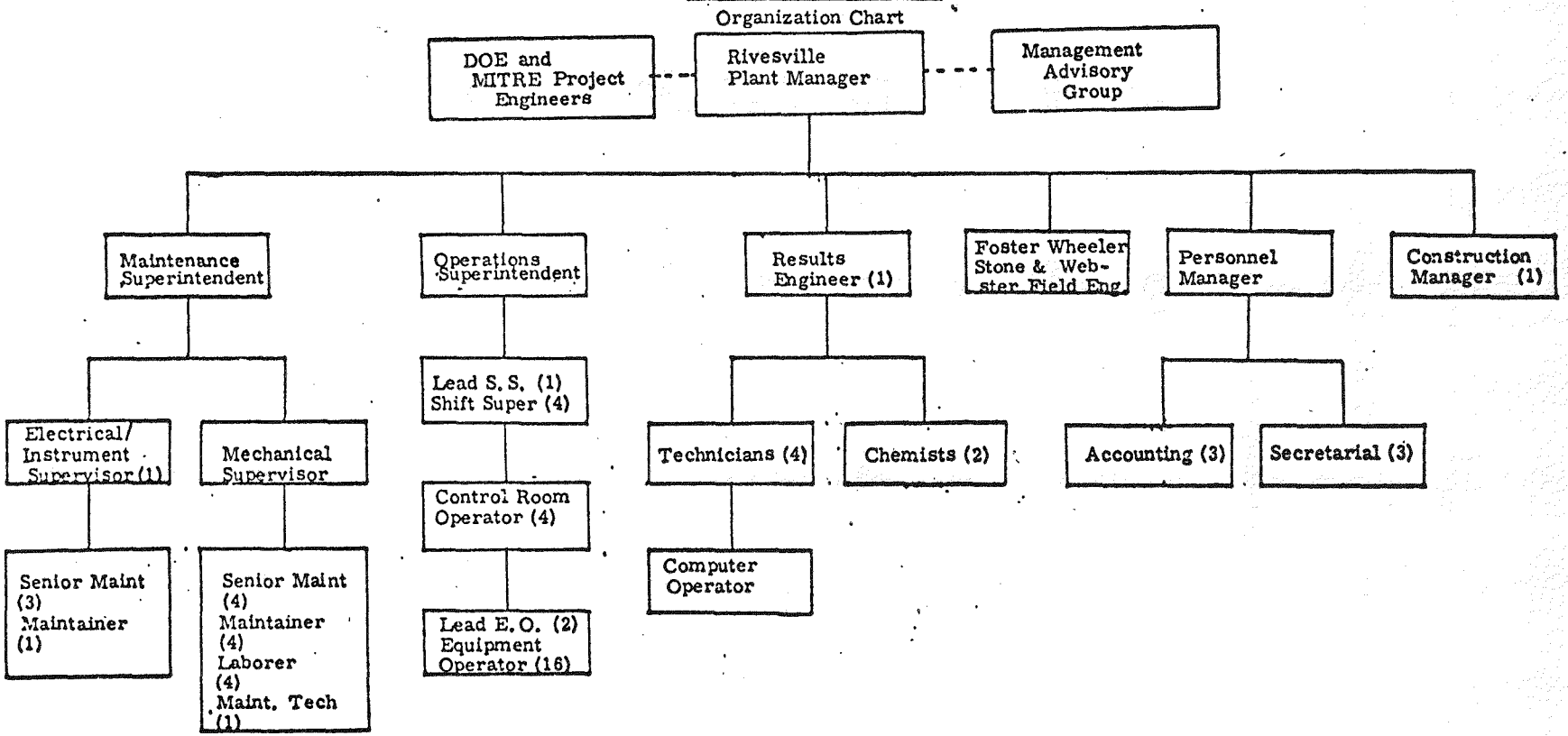


EXHIBIT XX

TOTAL 65

3.7.2 Mitre Corporation/Metrek Division

In accordance with METREK's contract with the Department of Energy, MITRE has maintained a field office on the job site staffed by one member of the METREK technical staff and a part-time secretary.

The objective of METREK's role on site is to independently assess operations of the MFB. The responsibility of the site representative is two-fold:

1. To monitor project activities and consult with the operating contractor on behalf of the DOE program manager.

2. To gather technical information so that METREK staff members can perform technical and program assessment for the DOE program manager.

In carrying out these duties the METREK representative has worked with the operating contractors forces and with their subcontractors, rendering assistance based on operating experience wherever possible.

3.8 Visitors, Dedication and Symposium

During the period of July 1977 and June 1978, the Rivesville site had a total of 852 visitors. These visitors encompassed a wide-range of interest including vendors' representatives, House and Senate members, foreign representatives, and the Under Secretary of the Department of Energy. Visitors from 17 countries were on-site including, The United Kingdom, Brazil, Japan, Canada, Sweden, USSR, Australia, France, India, Ireland, Germany, Norway, Taiwan, Finland, the Netherlands and Peru.

Prominent visitors during this period were Pennsylvania State Representatives B. DeWeese, R. Roberts, M. Fisher, E. Burns, J. Wright, D. Sioat, R. Gamble, B. O'Grien; Representatives R. Mollohan (WV), Senator R. Byrd (WV), Senator J. Randolph (WV), and Under Secretary D. Meyers.

Several visitors attended two important occasions. The first was the dedication ceremonies of the MFB which took place on August 26, 1977. A dedicatory address was delivered by Senated Majority Leader, Robert Byrd, D-WV. An example of local coverage is illustrated on the following page.

The second occasion was the visitation by attendees of the Fifth International Conference on Fluidized Bed Combustion. Sixty-one visitors from the conference toured the site. The conference was held on December 12, 1977 in Washington, D.C. Nine members of the Rivesville staff attended.

$23 million coal plant dedicated at Rivesville

RIVESVILLE—An innovative new method of burning coal to produce electrical energy—the fluidized bed converter—was formally dedicated here yesterday by Senate Majority Leader Robert Byrd, D-W.Va.

Also present at the ceremonies were Sen. Jennings Randolph and Rep. Robert O. Mollohan, both D-W.Va.; Rep. Austin Murphy of western Pennsylvania; Attorney General Chauncey Browning, representing Gov. Jay Rockefeller; Dr. Philip White, assistant Administrator of fossil energy with the U.S. Energy Research and Development Administration (ERDA).

The plant was named for the late John W. Bishop, an employe of one of the firms working on the facility, who developed the fluidized bed process in use here.

Byrd, as the keynote speaker, traced for the crowd present the history of coal, a known energy source for millenniums which only developed its full power when the Industrial Revolution began.

He described the energy crisis as "one of the greatest challenges facing us in this century. And I believe that the rich coal resources of West Virginia will provide us with an enormous energy alternative in coming decades."

The pilot project here is an example of cooperation between the federal government and private industry at its best, he believes.

The four-burner bed was completed at a cost of almost $23 million. It offers both economic and environmental advantages.

The limestone base in which coal is burned acts like a magnet to attract sulfur, thus reducing emmissions. This allows the burning of coal at a lower temperature, permitting more efficient use of the fuel.

Byrd, chairman of the Interior Appropriations Subcommittee, noted, "Coal has played an indispensable role in the history and development of civilization. But in an increasingly energy-hungry world, the destiny of coal lies not in the past but in the future.

"The foresight, research, skill and scientific expertise that produced the fluidized-bed boiler here will make it possible for us to bequeath to future generations of Americans a rich energy legacy—the renewed legacy of the 'black diamond.' "

Energy officials speaking also included Don Hollen, vice president of Monongahela Power Co.; Richard Valentine, Rivesville mayor; Michael Pope, president of Pope, Evans and Robbins; William Stevens, vice president of Foster-Wheeler Energy Corp., and John Hamilton, plant manager.

Also present for the meeting were members of the Fossil Energy Advisory Committee, which held one of its three annual meetings this week at Lakeview. Prior to the dedication, Byrd and others toured the Morgantown Energy Research Center.

Also speaking at the dedication was Dr. White, who focused his comments on discussion of the new power generating facility and how it has paved the way for larger facilities to be built using the same principles.

The use of coal presents a dilemma in that it is abundant but clearing away the technical, environmental and economic roadblocks to utilizing that abundant energy source is a difficult process.

"Because projects such as this one at Rivesville help clear away such roadblocks, they are essential to unlocking the potential of coal," he said. "We in ERDA see it as another important and exciting step in our efforts to increase the use of coal, the nation's most abundant energy source."

He continued, "This installation, a first-of-its-kind boiler, is designed to verify the technical and economic feasibiltiy of the fluidized-bed boiler concept and help develop it to the commercial stage in the early 1980's.

"When compared to conventional coal-fired boilers, with scrubbers, fluidized-bed systems offer higher power generation efficiencies and cleaner exhaust gases, when burning the high sulfur coals that predominate in the East."



NEW ENERGY PRODUCERS were discussed during this after-the-dedication get-together between Sen. Jennings Randolph, D.-W.Va.; Rep. Austin Murphy, western Pennsylvania; and Senate Majority Leader Robert C. Byrd, D-W.Va. They were among those dedicating the new fluidized-bed boiler system at Rivesville yesterday. Also pictured is William Stevens, vice president of Foster-Wheeler Energy Corp., partially visible behind Murphy.

The conference addressed the worldwide development and programs in fluidized bed combustion.

Three papers concerning the Rivesville installation were presented:

1. "First Performance Results from the Rivesville Multicell Fluidized Bed Boiler", G. Claypoole, D. Hill, R. Mineo.

   Initial results regarding combustion efficiency, sulfur capture, dust losses and loadings presented and reviewed.

2. "Start-Up and Initial Operation of the Rivesville 30 MW Fluid Bed Boiler", T. Stringfellow.

   This paper discussed the start-up, testing and initial operation of the multicell fluidized bed boiler at Rivesville, West Virginia. It was composed of four sections:

   1. A brief history of fluidized bed boilers.

   2. Equipment check-out, initial running and initial preoperational tests such as air flow calibration.

   3. A summary of the MFB operating history, including problems encountered and solutions attempted.

   4. Conclusions derived from the experience gained at Rivesville which will be helpful to future designers and operators of fluid bed boilers.

3. "Light-Off of Multicell Fluidized Bed Boilers by Hot Bed Transfer", L. L. Gasner, D. Turek, Department of Chemical Engineering, University of Maryland, Callixtus Aulisio and Duane Hill, Pope, Evans and Robbins Incorporated.

ABSTRACT

Light-off by hot bed transfer has been demonstrated to be a successful light-off procedure for multicell atmospheric fluidized bed boiler operation. The routine successive cell light-off in the 30 MW_e Rivesville Multicell AFBC boiler has been very successful. Raw data from this effort has been used to develop a correlation believed useful to predict successive cell light-off by hot bed transfer. Bed height,

pressure drop, temperature and time to coal ignition calculation methods are presented along with correlations to large scale data. The above correlation has also been extended to the simultaneous light-off of two successive cells in series.

IV. DESIGN AND ENGINEERING

4.1 General

The PER New York Office design and engineering staff has been engaged in various tasks in support of the Rivesville MFB unit operation. These include studies and evaluations of systems and equipment and modifications to the MFB plant to improve performance and increase reliability based on experience gained from continued operation. Co-ordination of subcontract work and test programs was also conducted.

Work on several such tasks was initiated and was completed or continued. A brief description of each follows.

4.2 Steam Silencers

With continued operation of the Rivesville MFB unit, steam vented through the silencers to atmosphere created excessive noise. Stone & Webster Engineering conducted a study recommending that the existing Maxim silencers be replaced by American Air Filter units that would provide additional attenuation. A study indicated that the existing structure could take the increase in load of the larger silencers. After installation of the new silencers and start of operations, the reduction in noise level was less than anticipated and still objectionable to the local citizenry.

Stone & Webster proposed adding a 12 ft extension to each silencer for increased attenuation. The existing structure was checked for loads imposed by the addition of 3500 lbs to each silencer. The structure was also checked for sound insulation to be added to both the I.D. and F.D. fan duct work for additional sound attenuation. The additional sound insulation for each duct would be approximately 6000 lbs. A report and sketches were prepared indicating the structural modifications required to support the additional sound insulation.

4.3 Instrument Air Compressors

As the result of unanticipated additional demand, the existing instrument air system at Rivesville had insufficient capacity for operation of controls and instrumentation. The design and engineering staff designed a replacement system to supply higher capacity improved quality air. Competitive quotations were solicited and a purchase order issued for a skid mounted system consisting of two Ingersoll Rand model ESV instrument air compressors, 200 ACFM 100 psi each, complete with receiver, after cooler, and desiccant dryer with prefilter and afterfilter.

Vendor drawings were reviewed and approved and assembly of the system authorized. The system is located in the basement of the MPC Plant on the existing concrete mat. A drawing was prepared providing the necessary construction and installation details.

An inspection was conducted at the vendor plant to check the equipment for compliance with the specifications before shipment to Rivesville. The package was received in Rivesville in late June and installed in July.

4.4 Waste Material Sample Storage System

Design and engineering of the 80 ton capacity waste sample storage system was completed and competitive construction bids solicited. A contract was awarded to Union Boiler Corp. the low bidder. Shop drawings were reviewed and released for construction. To accommodate an increase in weight for handling bed material only, a larger capacity rotary ash conditioner recommended by the supplier was approved. At the close of the reporting period, construction was approximately 70 percent complete.

4.5 Limestone Unloading System Vent Filters

At the time limestone was delivered to the Rivesville MFB plant, venting of the limestone unloading system cyclone created a visible plume of particulates discharging from the stack. The system was redesigned to allow the cyclone collector to be vented into a baghouse filter to capture the fines carried over. An installation drawing was prepared along with a budget cost estimate and released to the field for construction.

4.6 Boiler Feedwater Supply

Studies and evaluations were conducted for alternate means of providing a permanent, independent feedwater supply for the MFB. Among the alternates considered was barging, trucking and rail transporting of treated water from MPC's Fort Martin Station. Further study indicated that the most practical and feasible solution was the use of city water, demineralized to obtain the purity required for use as boiler feedwater.

Two portable demineralizer units were leased from Ecolochen resulting in an independent feedwater supply for the MFB permitting continuous operation during the period that steam is vented to atmosphere. The units were initially mounted on trucks and stored outside. However, because of severe winter weather, the units will be relocated inside the plant to prevent freezing and permit operator accessibility for maintenance and adjustments.

The units will be located on the operating floor between column lines E1-E5 and F1-F5. A demolition specification was prepared for removal of existing obsolete MPC equipment and associated piping and released to the field for solicitation of competitive bids for the work required.

4.7 EPA Particulate Sampling Program

The U.S. Environmental Protection Agency (EPA) will conduct a particulate sampling and evaluation program at the Rivesville MFB to establish performance standards for AFB emission control equipment. In connection with this program, installation drawings were prepared for sample ducts from the flue gas duct at the electrostatic precipitator to the EPA equipment.

The EPA equipment is mobile trailer mounted, and consists of a baghouse filter, wet scrubber and electrostatic precipitator. A fourth trailer houses the auxiliary equipment for the electrostatic precipitator. Two slip-streams are required for these emission control devices. A 6 in uninsulated duct to serve the baghouse and a 10 in heat traced duct to serve the scrubber and electrostatic precipitator. Approximately 300 ft of duct run is required including provision for expansion, anchors, guides, etc. Based on the prepared design drawings, an oder was placed for the long lead delivery thin wall, 6 in and 10 in stainless steel ducts.

4.8 Overbed Coal Feeder-Test Coal Bin

Provisions have been included in the Rivesville MFB for future installation and testing of an overbed coal feeder to demonstrate its applicability to feed fluidized bed boilers. An independent test coal bunker is required to feed this system. Existing Bunker #2 was investigated to determine the optimum location for the new bunker and associated coal supply chute. An investigation of existing structural members was conducted to determine their capacity to support additional and redistributed loads. The test coal bin was designed, fabricated and installed.

4.9 Flash Tank

Surface blowdown is maintained at a high level during start-up of the Rivesville MFB unit to reduce the time required to produce steam. During this start-up period, the surface blowdown is discharged into a flash tank causing it to vibrate considerably. An investigation concluded that the most probable cause of

vibration was high water level in the flash tank which submerged the blowdown pipe inlet nozzle. Disassembly of the tank drain revealed that the float valve of the 40 year old steam trap was inoperative. The float and valve seat of the trap were replaced and the existing drain line replaced by a larger line capable of serving the existing and a second available flash tank. Should the repaired steam trap prove to be deficient in capacity, it is intended to provide an additional steam trap for the second tank, placing both tanks in parallel operation.

4.10 Bed Drain System

Rocks in the fluidized bed and clinkers which form in the fluidized bed during operation collect in and plug the bed drains preventing continuous drainage of spent bed material. Plugging of the six inch drain lines was reviewed with Rivesville Operating Personnel and to alleviate this problem, the solution adopted was installation of a 3/4 in air injection line in the drain line immediately above the rotary air lock. Injection of 10 cfm, 60 psi air assists the flow of solids and encourages more complete burn-out of remaining carbon thereby reducing the tendency to form clinkers. Additional improvements included provision for more convenient service clean-outs at several bends in the pneumatic system used for transport of hot bed material.

4.11 F.D. and I.D. Fan Motor Pedestals

A complete inspection of the MFB was performed during the outage period and cracks on both the F.D. and I.D. fan motor pedestals were noted. The grout at the edge of the F.D. and I.D. fan base plates had cracked as the result of ice forming between the grout and edge of the plate. The problem was reviewed and a sketch prepared with recommendations for repairs and forwarded to the field for follow-up action.

4.12 Coal Feed System

Difficulties encountered with the coal feed system led to recommendations that eductors be utilized to transport the 1/2 in coal into the boiler combustion zone. As a result of eductor tests conducted at the Alexandria Laboratory, it was determined that a 10 psig air supply would be required to serve 52 eductors, one for each feed needle. An investigation of three alternate schemes to provide the necessary air supply led to the selection of a central system utilizing three positive displacement blowers.

4.13 Boiler Structural Support Framing

With the numerous modifications to the boiler, including splitters in the supply air plenum, observation doors in various cells, new ignition burners and proposed overbed feeders, there was a marked increase in the load on the structural support framing. An analysis of the framing was conducted to evaluate the effect of the additional loads imposed on the structure. As a result, it was determined that the new loads did not significantly affect safety factors provided in the original design.

4.14 Access Platforms and Stairs

In order to facilitate maintenance and operation of MFB equipment, additonal access platforms and ladders were designed. Sketches were prepared and forwarded to Rivesville Operations for fabrication and erection. The following items were provided;

- Access platform to Dust Collector No. 1,
- Access platform to coal vibrating conveyor, south side,
- Access stair from MPC walkway to precipitator air lock platform,
- Access stair from basement area to classifier and silo area outside building,
- Extension to coal vibrating conveyor platform,
- Access to north limestone vibrating conveyors.

4.15 Weigh Belt Feeders

The design and engineering staff assisted FWEC in a program to replace the existing rotary feeders with weigh belt feeders by investigating the structural support requirements for the feeders and designing access platforms and ladders. A structural design drawing indicating the modifications and additions to the existing structural framing system was completed, checked and released for construction. The steel fabricators shop drawings were reviewed and released to FWEC for fabrication and erection.

4.16 Stack Gas Sampling

In order to analyze and obtain samples of the MFB flue gas in support of the EPA emission evaluation program, it was necessary to provide an access platform to permit gas sample withdrawal at a point 12 ft below the top of the stack. An analysis was conducted and design drawings prepared for an access platform, ladders and gas probe supports. The drawing was released to the field along with a budget cost estimate to obtain competitive quotations and issue a purchase order for fabrication and installation.

V. SUBCONTRACTOR ACTIVITIES

5.1 General

In support of development, operation and improvements to the Rivesville MFB unit, the expertise of several major organizations was subcontracted for. The more significant subcontracts during this reporting period were with the Foster Wheeler Energy Corporation (FWEC), Stone & Webster Engineering Corporation (S & W) and Zurn Industries. A brief description of their activities follows.

5.2 Foster Wheeler Energy Corporation - MFB Steam Generator Development, MFB Operations and Maintenance Assistance

5.2.1 Activities

FWEC efforts during the past year on the Rivesville MFB boiler were divided into two general roles. One was that of consultant, providing maintenance and operational advice by full-time, on-site service engineers, supported by engineering and laboratory services from the Livingston, New Jersey Headquarters. The second role was that of engineer/constructor on various plant modifications and additions, the majority of which were accomplished during the winter outage. Following is a list of these items:

Belt Feeder Installation

The rotary feeders were removed from the coal feed system and replaced by belt feeders which have improved availability.

Large Start-Up Burners

The original 4 million Btu/hr burners were replaced with 7 million Btu/hr burners, minimizing the unit start-up time.

Test Plan Installation

A gas analyzing system, an extensive thermocouple system and a multitude of pressure and flow monitoring devices were installed to monitor all significant aspects of boiler operation. Data generated by this test plan is required to establish design and operational parameters for future fluid bed combustion boilers.

Metallurigical Examination

Metal samples were removed from various boiler tubes and supports and subjected to detailed metallurgical examination.

Miscellaneous

Smaller projects included installation of automatic auxiliary air controllers, installation of wide angle observation doors in all four boiler cells and access doors installed in the air distribution grid plate in A, B, and C cells.

5.3 Stone & Webster Engineering Corporation - MFB Operations Assistance

5.3.1 Activities

During the 14 month period, May 1977 through June 1978, S & W provided on-site and headquarters engineering and design support in areas involving the testing program, operations, and alternative investigations.

Boston Office Activities

The following Engineering/Design Reports (EDR) were completed;

EDR-1	Demineralized Water Availability (Superseded by EDR-10)
EDR-2	Off-site Noise During Atmospheric Venting
EDR-3	Truck Unloading Hopper Design (Ref. EDR-8)
EDR-4	Coal Dryer and Material Handling System Review
EDR-5	Precipitator Review
EDR-7	Ash Sluice System
EDR-9	CBC Carbon Reinjection System
EDR-10	Air Cooled Condenser
EDR-11	High Carbon Content Ash Cooling
EDR-12	RF-3 Air Leakage Review

Status of EDR's in process at the close of the reporting period was as follows:

EDR-6 Coal Bin Ventilation

Held pending proposed modifications to Coal Handling System.

EDR-8 Coal Delivery and Injection System

Work continued on this item

EDR-13 Bed Material Letdown System Review

Work continued on this item.

Field Activities

The following Field Reports (FR) were completed:

FR-1	Support Hangers for 2 in Needle Tubes
FR-2	Space Heater Oil Filling Stations
FR-3	Plant Vacuum System
FR-4	Coal Dryer Removal
FR-5	Recirculation Pump Seal Water Cooler
FR-6	Floating Grid Plate
FR-7	Draft Fans Cooling Water Fouling
FR-9	Fly Ash Reinjection Nozzle
FR-10	Fire in A-Cell T/C Plenum Penetration
FR-12	ESP Cleanout Ports
FR-13	Single Circulation Pump Operation

Status of FR's in process at the close of the reporting period was as follows:

FR-8 Feedwater Deaerator Estimate

Estimate prepared and submitted.

FR-11 Main Steam Safety Valve

Modification analyses continuing.

FR-14 Instrument Air Compressor Cooling Water Pump

Effort will commence in July 1978.

FR-15 Water Supply to Ecolochem Demineralizers

Effort will commence in July 1978.

FR-16 ESP and BF Access/Maintenance Platforms

Effort will commence in July 1978.

RES-1 Performance Tests

Continued acting as test supervisor.

Safety

Maintained responsibility for on-site safety; participated in training programs and attended Management Advisory Group (MAG) meetings.

5.4 Zurn Industries - Conceptual Design of Packaged Industrial Fluidized Bed Boiler

5.4.1 Activities

In October 1977, Zurn Industries was commissioned to prepare conceptual designs for a packaged (prefabricated) industrial atmospheric fluidized bed boiler of 100,000 lb/hr saturated dry steam capacity. Based on operating experience at both the Alexandria Laboratory and Rivesville, design parameters were established which are listed below.

5.4.1.1 Design Conditions

Steam Output	100,000#/hr.
Design Pressure	200
Operating Pressure	150
Saturated Steam Temperature	366
Feedwater Temperature	225
Air Temperature	80
Blowdown	5
Elevation	Sea Level
Excess Air	18 percent

5.4.1.2 Fuel

Kind	Bituminous - Pittsburgh Seam No. 8
Size	-1/4" Crushed
High Heating Value	13,000 Btu/lb.
Fusion Temperature of Ash	2080
Ultimate Analysis Percent by Wt.	Ash - 8.5 S - 4.3 H_2 - 5.07 C - 71.2 H_2O - 3.3 N_2 - 1.3 O_2 - 6.33

5.4.1.3 Submerged Surface

Tube Size	1 1/2" O.D.
Overall Horizontal Heat Transfer Coefficient	40 Btu/hr. ft^2 (hor. tubes)
Distance above air distribution plate	18 inches

5.4.1.4 Bed Conditions

Superficial Gas Vel.	8 ft/sec.
Operating Temp	1550°F
Static Bed Depth	18 inches
Expanded Bed Depth	36 inches
Overall Vertical Heat Transfer Coefficient To 36" above Air Distribution Plate	40 Btu/Hr., $ft^2°F$ (hor. tubes) 50 Btu/Hr., $ft^2°F$ (vertical tubes)
Unburned Combustible Loss	1.335 percent
85 percent of Available Heat Released to Bed	
1t percent Heat Released to Freeboard	

5.4.1.5 Boiler Concept

Zurn concept of the industrial packaged boiler would not employ individual cells but would have an open bed influenced by local fluidization. A sectionalized plenum would provide air to portions of the bed and contribute to turndown capabilities. Air would pass through a distribution plate similar to that in the Rivesville boiler and fluidize the fuel bed. Combustion gases would flow upward from the bed and into a baffled convection zone running parallel to the furnace and exit the gas outlet on the side of the boiler. Hoppers under the first baffle would collect carryover for reinjection into the main bed and increase combustion efficiency.

At the close of the reporting period, Zurn had completed their designs which included detailed calculations and arrangement drawings. The material developed was in sufficient detail to permit a competant boiler manufacturer to prepare detailed fabrication drawings and construct a prototype unit. Zurn has started preparation of a budget cost estimate for fabrication.

A P P E N D I X A

Table I

Operating Conditions for Tests PDU-4, 700-1, 2, and 3

Test No.	PDU-4				700-2		700-3			
Date of Test	1-31-78				2-27-78		2-28-78			
Time	14:30	15:00	14:00	12:00	13:00	14:00	12:00	13:00	14:00	15:00
Bed temp., °F	1518	1505	1510	1495	1495	1475	1530	1585	1585	1575
O_2 in flue gas, %	1.2	0.8	6.5	3.3	3.2	4.3	2.9	2.2	2.9	3.0
Coal feed rate, lb/hr	700	765	768	680	712	750	765	720	819	810
Fly ash collection rate, lb/hr	212	212	225	--	--	--	--	--	--	--
Limestone feed rate, lb/hr	203	203	0	0	0	0	394	360	270	210
Bed drain rate, lb/hr	0	0	0	0	0	0	--	--	--	--
Stack dust rate, lb/hr	40.4	34.3	28.7	20.4	24.1	18.7	13.9	13.4	15.3	16.0
Main air flow rate, lb/hr	5110	4664	6411	7366	7097	6955	7869	7547	8032	8213
Total air flow rate, lb/hr	5344	4888	7381	7595	7326	7179	8103	7781	8266	8447
Superficial velocity, ft/sec	7.8	7.1	10.7	10.9	10.6	10.2	11.8	11.7	12.4	12.6
Static bed depth, inches	26.1	31.2	27.7	28.2	26.1	26.0	28.2	27.2	27.2	27.2
EOBH (effective operating bed ht.), inches	33.0	41.5	38.5	39.0	34.5	37.0	41.0	41.0	40.5	40.0
SO_2, ppm	1165	940	1581	5500	9900	9350	3076	3245	2226	2292
Ca/S	1.5	1.6	0	0	0	0	3.1	3.0	2.0	1.5
Combustion efficiency, %	84.7	81.6	80.8	f	f	f	f	f	f	f
Sulfur Capture	b	b	b	b	b	b	b	b	b	b
lb_2SO /10^6 Btu	b	b	2.71	b	b	b	b	b	b	b
Average grid plate temperature, °F	--	--	--	--	--	--	--	--	--	--

b Non-steady state bed composition.

f Data not taken.

Table II

Operating Conditions for Tests 700-4, 5, 6, 7 and 8

Test No.	700-4		700-5			700-6				700-7		700-8		
Date of Test	3-15-78		3-21-78			3-22-78				3-28-78		3-30-78		
Time, Hr	13:00	14:00	12:00	14:00	15:00	12:00	13:00	14:00	15:00	13:00	14:00	12:00	13:00	15:00
Bed Temp., °F	1535	1538	1519	1532	1565	1545	1543	1541	1511	1555	1552	1536	1552	1536
O_2 in flue gas, %	3.2	4.1	3.8	3.8	3.0	2.8	2.7	2.6	3.2	2.6	2.6	3.7	3.2	3.3
Coal feed rate, lb/hr	695	731	747	747	758	744	764	770	711	714	720	520	528	451
Fly ash collection rate, lb/hr	194	189	234	214	240	228	228	224	210	222	235	140	147	135
Limestone feed rate, lb/hr	428	424	375	350	350	348	348	348	336	412	412	363	363	363
Bed drain rate, lb/hr	115	115	221	221	221	210	210	210	210	256	256	167	167	167
Stack dust rate, lb/hr	15.5	16.9	21.3	19.6	20.8	22.1	23.0	25.5	25.4	20.8	26.1	16.5	19.6	22.3
Main air flow rate, lb/hr	7106	7736	7615	7607	7607	6908	7049	6901	7042	7740	7872	5699	5678	5575
Total air flow rate, lb/hr	7462	8091	7957	7939	7849	7250	7386	7242	7378	8087	8219	6040	6024	5922
Superficial velocity, ft/sec	11.0	12.0	11.6	11.7	11.7	10.7	10.9	10.7	10.7	12.0	12.2	9.0	9.1	8.3
Static bed depth, inches	24.9	24.5	23.6	24.6	26.4	26.5	25.5	26.5	25.1	26.1	26.2	19.0	19.1	19.7
EOBH (effective operating bed ht.), inches	37.5	39.0	38.3	40.5	41.2	38.3	38.0	39.4	36.2	40.9	39.3	31.3	31.0	32.7
SO_2, ppm	1618	1430	1040	900	570	422	636	544	538	1150	1133	648	760	750
Ca/S	4.00	3.74	3.16	2.95	2.91	2.52	2.46	2.44	2.51	a	a	a	a	a
Combustion efficiency, %	90.9	92.1	88.7	91.3	84.5	92.0	90.1	90.1	92.4	a	a	a	a	a
Sulfur Capture	b	b	b	b	b	89.8	84.8	87.3	86.3	a	a	a	a	a
lb SO_2/10 Btu	b	b	b	b	b	0.72	1.07	0.90	0.97	a	a	a	a	a
Average grid plate temperature, °F	658	612	903	967	984	972	979	992	981	1009	999	986	990	973

a Analysis incomplete.

b Non-steady state bed composition.

Table III

Operating Conditions for Tests 700-9, 10, 11 and 12

Test No.	700-9			700-10		700-11			700-12	
Date of Test	4-6-78			4-18-78		4-20-78			4-27-78	
Time	13:00	14:00	15:00	14:00	15:00	13:00	14:00	15:00	13:00	14:00
Bed temp., °F	1489	1498	1553	1519	1500	1547	1530	1601	1585	1561
O_2 in flue gas, %	2.8	2.6	2.6	3.0	3.1	3.0	3.1	3.2	3.0	3.4
Coal feed rate, lb/hr	397	397	397	666	662	420	429	429	736	736
Fly ash collection rate, lb/hr	88	88	95	204	210	103	104	117	276	252
Limestone feed rate, lb/hr	394	394	394	424	424	375	375	378	492	564
Bed drain rate, lb/hr	283	283	283	300	309	194	194	194	272	272
Stack dust rate, lb/hr	22.7	17.2	14.3	19.1	18.4	21.1	21.1	21.6	24.2	22.3
Main air flow rate, lb/hr	3938	4177	4059	6982	6719	4962	4840	4896	7921	7921
Total air flow rate, lb/hr	4250	4488	4371	7283	7020	5263	5141	5197	8222	8222
Superficial velocity, ft/sec	6.2	6.6	6.6	10.6	10.1	8.0	7.7	8.1	12.4	12.2
Static bed depth, inches	18.4	18.5	16.8	26.2	28.2	16.1	16.5	15.1	26.7	26.7
EOBH (effective operating bed ht.), inches	23.0	23.0	21.5	38.0	40.0	24.0	26.3	22.4	41.0	40.0
SO_2, ppm	457	469	816	b	850d	764	709	814	723	1091
Ca/S	b	b	b	3.11	3.13	6.10	5.97	6.05	4.61	5.28
Combustion efficiency, %	c	c	c	89.8	92.3	89.6	90.2	89.2	82.6	91.0
Sulfur capture	b	b	b	b	80.7d	a	a	a	a	a
lb $SO_2/10^6$ Btu	b	b	b	b	1.62d	1.72	1.53	1.77	1.44	2.19
Average grid plate temp. °F	947	963	1048	956	944	998	986	1045	277	273

a Analysis incomplete.

b Non-steady state bed composition.

c Not calculated due to an error in fly ash sampling.

d Analysis based on SO_2 reading taken at 23:00 hours.

Table IV

Operating Conditions for Tests 700-15, 16, 17 and 18

Test No.	700-15	700-16						700-17					700-18			
Date of Test	5-16-78	5-18-78						5-23-78					5-31-78			
Time	13:00	11:00	12:30	12:45	13:00	13:15	14:00	13:00	13:30	14:00	14:30	15:00	10:30	11:45	13:30	15:00
Bed temp., °F	1514	1558	1569	1533	1548	1555	1706	1535	1535	1521	1440	1383	1541	1565	1765	1680
O_2 in flue gas, %	2.2	2.2	3.0	2.5	2.6	3.4	1.8	3.3	2.0	1.4	2.1	2.5	2.0	5.3	0++	0++
Coal feed rate, lb/hr	496	324	462	462	624	624	583	697	715	653	555	+++	570	664	854	700
Fly ash collection rate, lb/hr	0	0	0	0	--	--	159	219	213	177	115	115	168	192	350	240
Limestone feed rate, lb/hr	297	187	326	272	420	420	426	431	390	390	348	358	322	422	460	387
Bed drain rate, lb/hr	f	187	187	f	f	f	f	310	310	310	208	208	0	480	0	112
Stack dust rate, lb/hr	f	f	f	f	f	f	f	15.6	f	18.4	f	22.8	f	f	f	f
Main air flow rate, lb/hr	5102	3159	4683	5464	6326	7081	7756	7779	7082	6543	5460	4572	6188	7974	7833	6166
Total air flow rate, lb/hr	5409	3461	4985	5765	6627	7382	8058	8080	7383	6844	5761	4873	6484	8270	8129	6462
Superficial velocity, ft/sec	8.0	5.3	7.6	8.7	10.1	11.2	13.2	11.9	10.9	10.1	8.1	6.5	9.8	12.7	13.7	10.4
Static bed depth, inches	17.0	17.5	16.5	17.0	16.6	16.9	16.7	26.8	26.8	24.8	24.8	25.3	15.5	14.0	15.0	15.7
EOBH (effective operating bed ht.), inches	26.0	26.0	25.0	25.5	25.5	25.5	25.5	40.0	40.5	35.0	35.0	35.0	27.00	27.00	29.0	29.5
SO_2, ppm	920	593	332	355	558	557	1321	491	645	779	663	717	f	528	2792	1917
Ca/S	a	a	a	a	a	a	a	3.28	a	3.41	a	3.51	a	a	a	a
Combustion efficiency, %	f	f	f	f	f	f	f	92.5	f	86.1	f	89.6	f	f	f	f
Sulfur capture	f	f	f	f	f	f	f	82.5	f	70.1	f	79.8	f	f	f	f
lb $SO_2/10^6$ Btu	f	f	f	f	f	f	f	1.00	f	1.72	f	1.16	f	f	f	f
Average grid plate temp. °F	241	338	283	248	314	287	281	947	930	944	876	293	313	311	291	293

a Analysis incomplete.

f Data not taken.

Table V

Operating Conditions for Test 700-19 and 20

Test No.	700-19					700-20					
Date of Test	6-8-78					6-13-78					
Time	12:00	13:00	14:30	15:00	Ave.	12:15	12:45	13:15	13:45	14:30	15:00
Bed temp., °F	1502	1511	1519	1502	1509	1535	1513	1531	1500	1349	1356
O_2 in flue gas, %	4.1	3.6	3.0	3.4	3.5	3.5	4.0	2.1	2.9	1.8	2.2
Coal feed rate, lb/hr	729	690	701	710	710	702	702	686	625	564	484
Fly ash collection rate, lb/hr	218	206	216	216	214	262	262	197	164	122	112
Limestone feed rate, lb/hr	366	352	346	348	353	330	362	390	380	196	312
Bed drain rate, lb/hr	158	158	140	140	149	238	238	238	238	68	68
Stack dust rate, lb/hr	f	27.6	17.9	21.5	24.3*	f	f	f	f	f	f
Main air flow rate, lb/hr	7624	7624	7638	7638	7631	7997	8996	7087	6496	5474	4473
Total air flow rate, lb/hr	7920	7920	7934	7934	7927	8298	9297	7388	6795	5775	4774
Superficial velocity, ft/sec	11.5	11.5	11.6	11.5	11.5	12.2	13.5	10.9	9.8	7.7	6.4
Static bed depth, inches	25.9	25.9	25.9	25.9	25.9	25.9	25.9	23.9	25.9	24.9	24.9
EOBH (effective operating bed height), inches	39.0	38.5	37.0	39.0	38.4	41.5	41.5	39.0	36.5	36.5	31.5
SO_2, ppm	926	879	1035	987	957	4500	5500	7500	7500	9000	8500
Ca/S	a	a	a	a	a	a	a	a	a	a	a
Combustion efficiency, %	a	a	a	a	a	f	f	f	f	f	f
Sulfur Capture	a	a	a	a	a	f	f	f	f	f	f
lb $SO_2/10^6$ Btu	a	a	a	a	a	f		f	f	f	f
Average perforated grid plate, temperature, °F	866	880	882	868	868	900	891	923	890	674	303

* Avg. of stack dust rate at 12:00, 13:00 and 14:00.

a Analysis incomplete.

f Data not taken.

Table VI

Operating Conditions for Tests 700-21 and 22

Test No.	700-21					700-22			
Date of Test	6-22-78					6-27-78			
Time	12:30	13:11	13:45	14:28	14:52	12:30	13:00	13:25	14:00
Bed temp., °F	1648	1610	1585	1668	1630	1609	1675	1701	1771
O_2 in flue gas, %	6.0	6.0	7.3	7.5	7.0	4.5	3.0	5.0	1.8
Coal feed rate, lb/hr	416	511	511	528	644	660	750	750	773
Fly ash collection rate, lb/hr	135	132	152	195	171	213	255	273	247
Limestone feed rate, lb/hr	189	171	195	180	180	268	217	217	270
Bed drain rate, lb/hr	156	156	156	156	156	--	--	--	--
Main air flow rate, lb/hr	6282	6731	7558	8443	8923	7545	7524	7526	7498
Total air flow rate, lb/hr	6578	7027	7854	8739	9219	7846	7825	7827	7800
Superficial velocity, ft/sec	10.5	11.4	12.3	14.1	14.6	12.2	12.6	12.2	13.1
Static bed depth, inches	9.9	9.9	9.9	8.9	9.4	12.4	11.9	11.4	11.4
EOBH (effective operating bed ht.), inches	20.5	21.5	20.5	20.5	21.0	30.0	30.0	30.0	30.0
Average grid plate temperature, °F	291	278	273	274	276	--	--	--	--
Expanded metal screen temperature, °F	--	--	--	--	--	978	1104	1120	1242
Plenum temp. (°F)	257	251	245	245	247	259	260	258	263
Air distributor ΔP ("H_2O)	10.0	11.0	13.0	15.0	17.0	12.0	12.0	12.5	12.5

A P P E N D I X B

cc: Mr. J. S. Maurer, Engineering
Mr. W. D. Stone, RESEARCH CENTER
Mr. A. J. Desmarais, RESEARCH CENTER
File-CM-77-92

Wilmington, Delaware
March 31, 1978

TO: Mr. F. N. Bent
Engineering Department

FROM: R. D. Crooks
Engineering Department

ERDA FLUIDIZED BED BOILER TUBES FINAL REPORT

Background

Pope, Evans and Robins, Alexandria, VA have operated an experimental fluidized bed boiler for a number of years for ERDA. Upon acquisition of a new boiler the old boiler was dismantled and arrangements made for Hercules to obtain some of the boiler tubes for metallurgical examination. Of concern was the possibility of various types of tube deterioration, such as erosion, oxidation, carburization and sulfidation.

Figure 1 illustrates the approximate tube arrangement in the fluidized bed boiler. Service records of PE&R indicate the unit was operated with fluid bed temperatures of 1400-1600°F over an 8 year period involving about 1000 start-stop cycles for a total operating time of about 8000 hours. Limestone was added to the fluidized coal bed during most of the operating time and on occasion either $NaCO_3$ or NaCl were also added.

Progress report #1, December 20, 1978, indicated that the boiler tubes were of plain carbon steel analysis equivalent to ASTM-A192. Hardness measurements made on the exposed tubes did not reveal significant differences between tubes or at different tube locations. Spectographic analysis of the tube fireside deposits revealed the presence of concentrations of Sulfur, Sodium, Lead and Vanadium.

The report which follows covers the results of our analysis of surface attack, loss of wall thickness and microstructural changes in the same tubes.

Summary

Examination of these boiler tubes has revealed no serious deterioration attributable to the service conditions, hence plain carbon steel ASTM-A-192 can be considered an adequate boiler tube material for this service.

Tube Sampling

Tube samples were obtained from various locations in the boiler (see Fig. 1) and these tubes were cut at three locations representing the top, exposed to combustion product gases, the middle, fluid bed-gas interface, and at the bottom of the fluid bed as shown in Figure 2. At each level there are two samples; the fireside sample which should be representative of the final tube condition after use, and outside of the tube at the same level which should be unchanged from its original condition. Tube samples are identified by Tube number + Location code + fireside or outside orientation (i.e., 5W-T-F = Tube 5W, Top fireside).

Sample Preparation

Small pieces were cut from each tube wall location and mounted in Bakelite. These mounts were ground, polished and etched in preparation for metallographic examination.

Surface Decarburization

After etching the mounted samples in 2% Nital, large amounts of surface decarburization and grain growth were visible to the naked eye on some of the samples. Some of the most pronounced cases are illustrated in Figures 3, 4 and 5. Note that in several instances there is more decarburization on the inside tube surface than on the outside or fireside. The extent of grain growth accompanying the decarburization is illustrated in Figure 6. Surface decarburization measurements made are summarized in Table I.

This anomalous condition cannot be explained by any reasonably expected operating conditions in the boiler since the fluid coal bed should produce a carburizing atmosphere. Moreover, only water or steam contacts the inner tube walls and these environments are not decarburizing.

It appears that the only reasonable explanation for the presence of this irregular decarburized surface is that the condition pre-existed in the tube material prior to use.

Tube Surface Condition

Examination of the interior and exterior tube surfaces at various locations revealed in general that the waterside was coated with an adherent scale layer which prevented corrosion, as illustrated in Figure 7. An exception to this is shown in Figure 4 sample 6E-M-0 wherein significant waterside corrosions has occurred, presumably as a result of loss of the protective scale layer.

The exposed fireside of the tubes showed some general roughening probably as a result of combined erosion and corrosion, however, this was seldom more than a few mils deep. A comparison of a typical fireside surface profile with a waterside profile is shown in Figure 8. At one location on Tube 14 a high temperature corrosion-oxidation cell was found about .006" deep. (See Fig. 9)

No evidence of carburization or sulfidation was observed on any of the fireside surfaces.

Tube Microstructure

The microstructure of all tubes examined was normal consisting of ferrite grains with a small excess of iron carbide as illustrated in Figures 8, 10 and 11. Matrix grain size varied from ASTM 8 to 10 except in heavily decarburized areas such as in Fig. 6 where grain size was ASTM #1 or larger. No evidence of carburization or sulfidation was seen on any of the fireside surfaces despite the presence of a sulfur rich carbonaceous layer. It appears that tube wall temperatures were simply not high enough for these diffusion reactions to occur.

A summary of the microstructural observations made on each sample is shown in Table II.

Tube Wall Thickness Losses

Boiler tube wall thicknesses may be reduced in operation by oxidation or erosion on the fireside or by corrosion or erosion on the waterside. Measurements of tube wall thickness were made microscopically at various locations and these data are shown in Table III.

To evaluate a change, the initial thickness must be known and of course no such measurements are available. Based on the thicknesses measured, it has been assumed that the tubes were originally 11 gage, nominal .120" thick and loss of thickness has been calculated from this base.

For clarity, the data of Table III are also plotted in Figure 12A and B against an original thickness of 11 gage (.120") and 12 gage (.109"). From these plots it is seen that little or no thickness change occurs on the air side wall, confirming the absence of waterside corrosion (Fig. 12A). Fig. 12B showing fireside thickness change suggests the loss is greatest at the mid tube location, i.e., the fluid bed-gas interface. This seems reasonable if it is assumed that the abrasive or erosive effect of the fluid coal bed is greatest at this location.

Observations

Examination and analysis of these tubes has shown remarkably little change in the surface or subsurface characteristics of the tube material as a result of exposure to the high temperature fluidized coal bed or its combustion products. The most probable reasons for the absence of change are the relatively low operating temperature of the bed coupled with the maintainence of low tube wall temperatures via adequate water flow.

Specifically it was found that:

a - Surface decarburization appeared to be a pre-existing condition in some tubes, and not present in others, leading to the conclusion that decarburization did not occur in service.

b - Despite the presence of a carbonaceous deposit layer on the fireside of the tubes, no carburization of the tube surfaces was observed.

c - Despite the presence of high sulfur levels in the tube fireside deposits, no evidence of tube surface sulfidation was observed.

d - The microstructure of the tubes showed no evidence of overheating nor was there any significant surface attack on the fireside.

e - Small losses of tube wall thickness were observed on the tube fireside concentrated at the mid tube location as a result of abrasive erosion.

These observations lead to the conclusion that for the service conditions in this fluidized bed boiler, no serious deterioration of the boiler tubes occurred hence it appears that plain carbon steel ASTM A192 is a satisfactorily serviceable material.

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Table I - Surface Decarburization of Boiler Tubes

Sample Location		Tube 5E Decarburization - in. Water Side	Tube 5E Decarburization - in. Out-Fire Side	Tube 6E Decarburization - in. Water Side	Tube 6E Decarburization - in. Out-Fire Side
Top-Out		.0401	.0147	.0171	.000
Top-Fire		.0195	.000	.0377	.0137
Mid-Out		.0285	.0137	.0411	.0171
Mid-Fire		.000	.000	.0171	.0034
Bot-Out		.0103	.0315	.0377	.0137
Bot-Fire		.0206	.000	.0223	.0034
	range	0-.0401	0-.0315	.0171-.0411	0-.0171
	mean	.0198	.0099	.0288	.0086

Sample Location		Tube 13 Decarburization - in. Water Side	Tube 13 Decarburization - in. Out-Fire Side	Tube 14 Decarburization - in. Water Side	Tube 14 Decarburization - in. Out-Fire Side
Top-Out		.0274	.0106	.0343	.0181
Top-Fire		.000	.000	.0205	.0069
Mid-Out		.0295	.0113	.0298	.0124
Mid-Fire		.0096	.000	.0206	.0034
Bot-Out		.0274	.0137	.0274	.0156
Bot-Fire		.000	.000	.0242	.0027
	range	0-.0295	0-.0137	.0205-.0343	.0027-.0156
	mean	.0156	.0059	.0261	.0096

Tube - 5W - No measurable Decarburization

Tube - 10 - No measureable Decarburization

Table II - Summary of Microstructural Observations

Tube No.	Location	Fire Side		Out Side	
		Fire	Water	Air	Water
5E	Top	BGH	AKGH	LGH	AG
	Middle	MNGH	AGH	LGH	AG
	Bottom	BNGH	AGH	LGH	AG
5W	Top	BDJ	AL	LD	ADJ
	Middle	BDJ	ANDJ	LDJ	ANJD
	Bottom	BFJ	AL	LFD	ALDJ
6W	Top	BDJ	AL	LD	ADJ
	Middle	BDJ	AND	LDJ	ANJD
	Bottom	BFJ	AL	LFJ	ALDJ
10	Top	BDJ	AFH	LDJ	LEJ
	Middle	BDJ	AE	LDJ	AEJ
	Bottom	LDJ	LE	LDJ	AEJ
11	Top	BDJ	LEJ	LDJ	AEJ
	Middle	BDJ	LFJ	LDJ	LFJ
	Bottom	LDJ	LFJ	LDJ	NEJ
13	Top	BD	E	LF	LGH
	Middle	BD	AF	LGH	LGH
	Bottom	BND	AEJ	LGH	LFH
14	Top	LFJ	AGH	LGH	LGH
	Middle	BND	AF	LGH	LGH
	Bottom	BFJ	AGH	LGH	LGH

Key to Symbols

A - normal water side scale
B - rough-erosion-corrosion
D - no decarburization
E - Patchy decarburization
F - light continuous decarb.
G - heavy continuous decarb.
H - coarse grain surface
J - fine grain matrix
K - scale nodules
L - no surface attack
M - high temp. scale
N - pitting

Table III - Summary of Tube Wall Thickness Loss

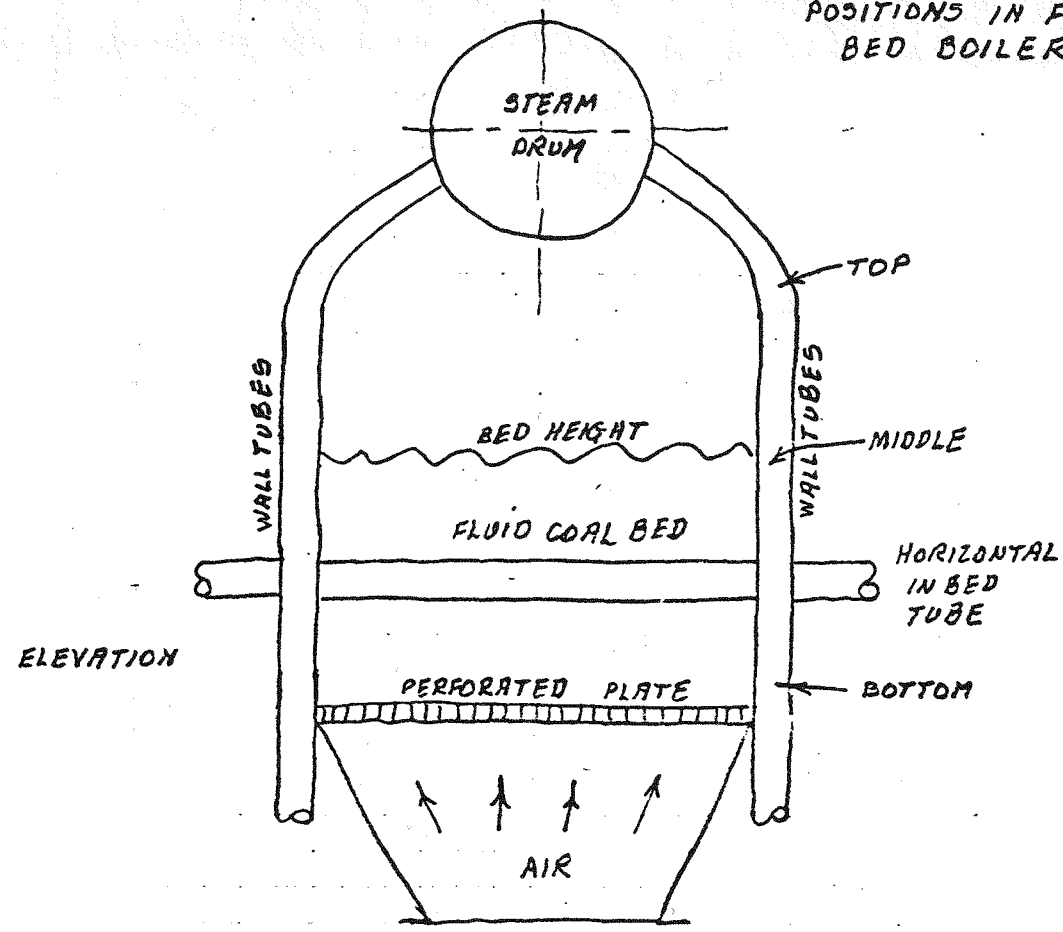
Sample Location	Tube 5E Measured Wall-in	Tube 5E Thickness* Loss From 11G	Tube 5W Measured Wall-in	Tube 5W Thickness* Loss From 11G
Top-Coat	.1104	.0096	.1113	.0087
Top-Fire	.1086	.0114	.1060	.0140
Mid-Out	.1096	.0104	.1084	.0116
Mid-Fire	.0599	.0601	.1039	.0161
Bot-Out	.1093	.0107	.1080	.0120
Bot-Fire	.1096	.0104	.1047	.0153

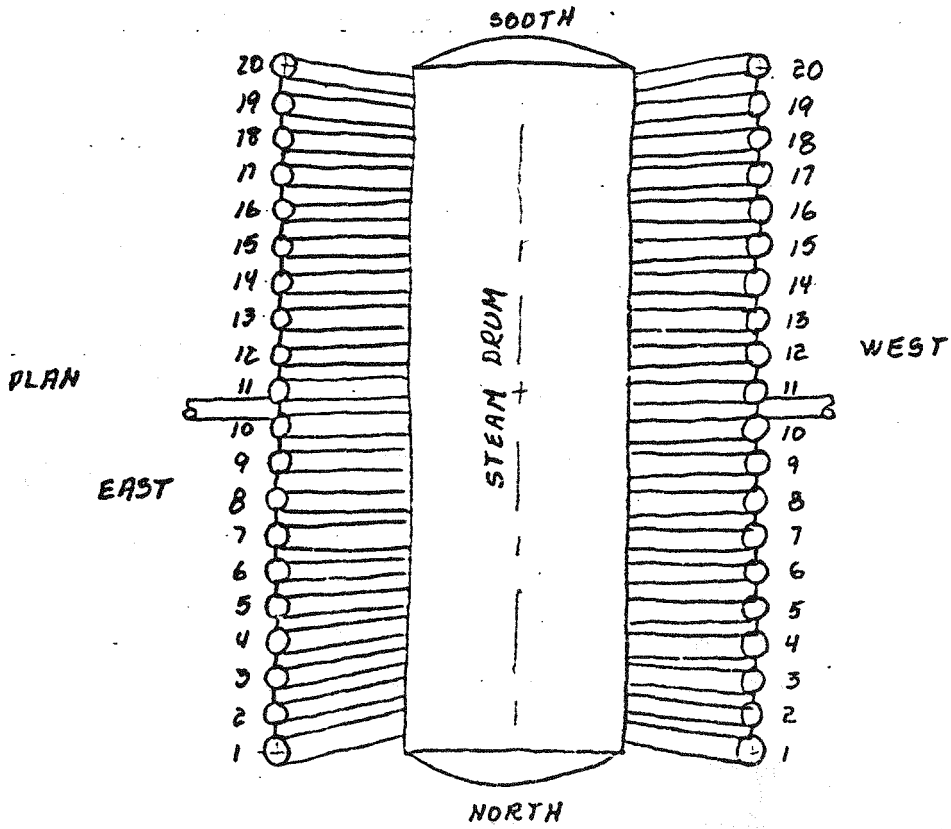
Sample Location	Tube 6E		Tube 10	
Top-Out	.1069	.0131	.1107	.0093
Top-Fire	.1103	.0097	.1081	.0119
Mid-Out	.1089	.0111	.1130	.0070
Mid-Fire	.1055	.0145	.0991	.0209
Bot-Out	.1096	.0104	.1042	.0158
Bot-Fire	.1125	.0075	.1123	.0077

Sample Location	Tube 13		Tube 14	
Top-Out	.1089	.0111	.1108	.0092
Top-Fire	.1050	.0150	.1061	.0139
Mid-Out	.1078	.0122	.1081	.0119
Mid-Fire	.1022	.0178	.1045	.0155
Bot-Out	.1082	.0118	.1058	.0142
Bot-Fire	.1028	.0172	.1065	.0135

*Note - Losses are computed from assumption that an 11 gage tube is originally .120" thick.

FIGURE 1- SKETCH OF TUBE POSITIONS IN FLUIDIZED BED BOILER

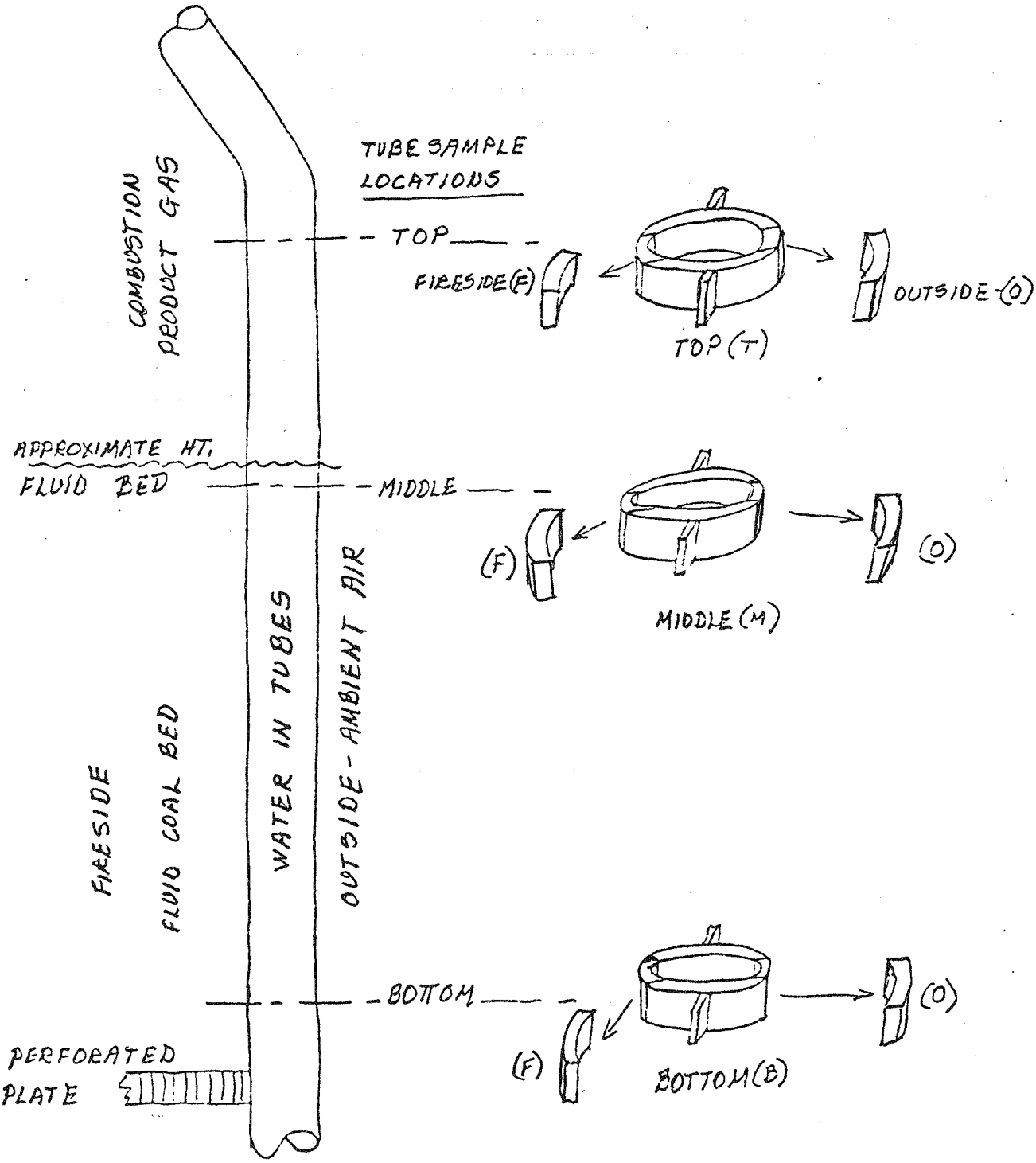




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FIGURE-2 - LOCATION OF SAMPLES CUT FROM BOILER TUBES AND METHOD OF IDENTIFYING

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AUTHOR R. D. CROOKS
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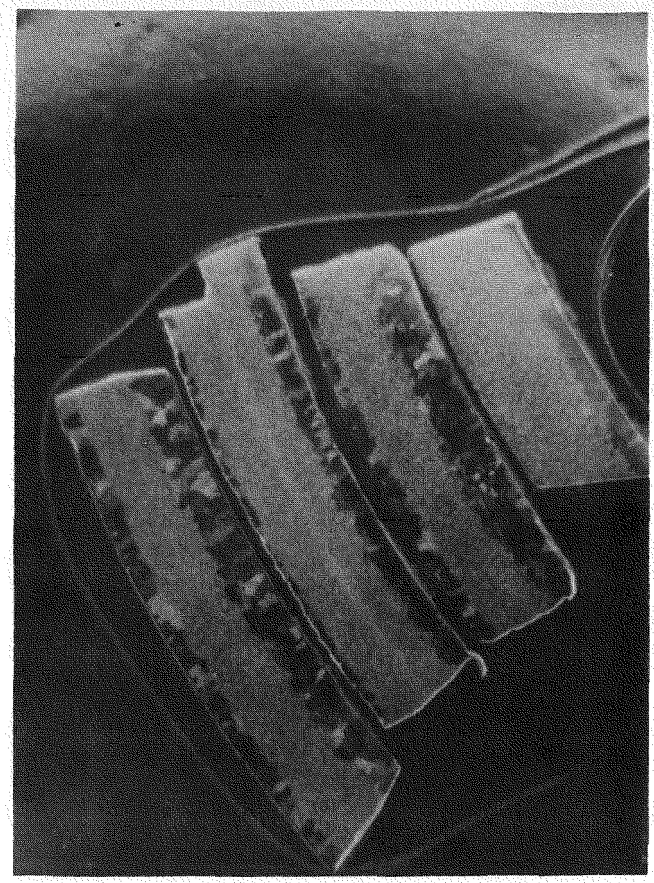


FIGURE 3

Section through top of Tubes 5E and 6E showing decarburization and grain growth on interior and exterior surfaces.
Mag. 6X - Etched 2% Nital

Orientation

Left - Tube 6E - Top - Outside
Tube 6E - Top - Fireside
Tube 5E - Top - Outside
Right - Tube 5E - Top - Fireside

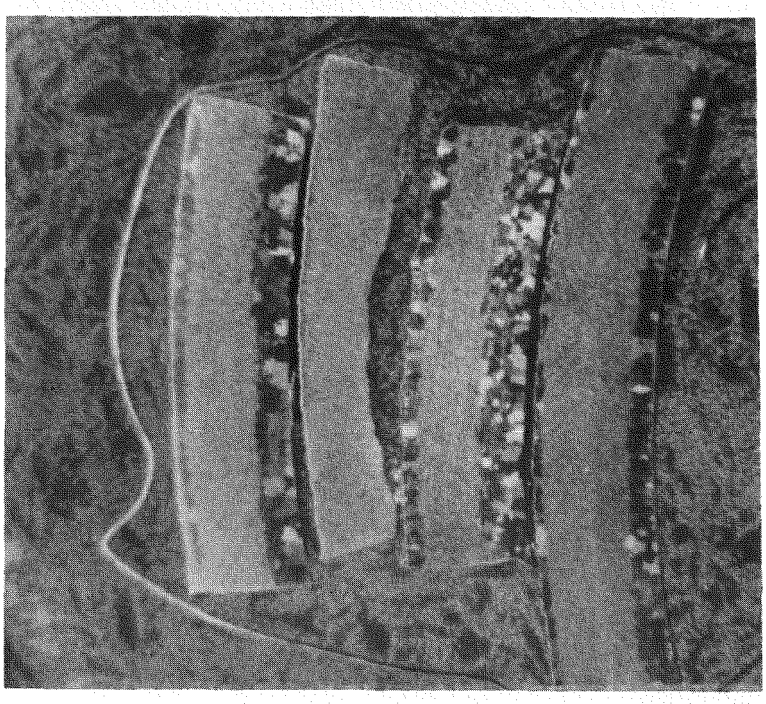


FIGURE 4

Section through middle of Tubes 5E and 6E showing decarburization and grain growth on interior and exterior surfaces. Note also waterside corrosion on 6EMO.
Mag. 6X - Etched 2% Nital

Orientation

Left
Tube 5E - Middle - Outside
Tube 5E - Middle - Fireside
Tube 6E - Middle - Outside
Right
Tube 6E - Middle - Fireside

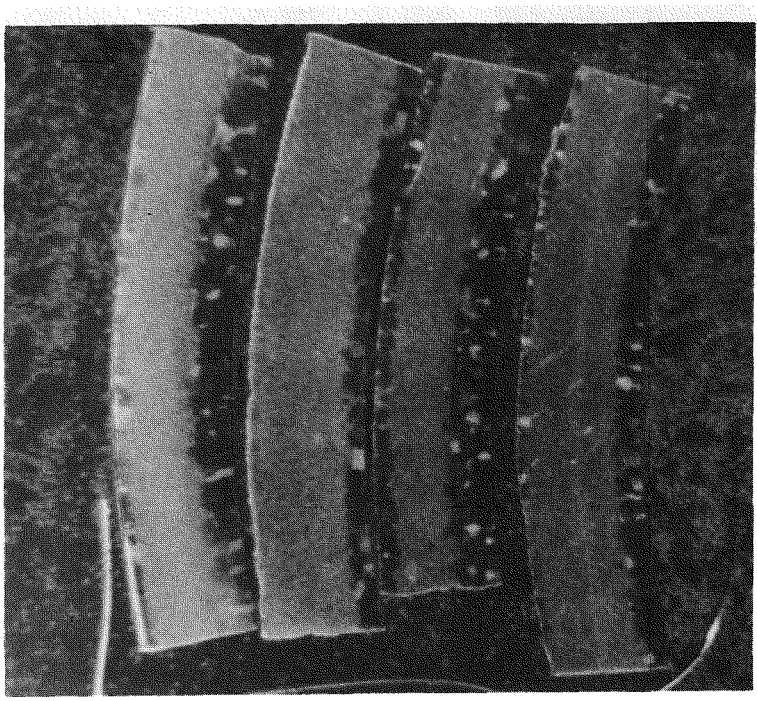


FIGURE 5

Section through bottom of Tubes 5E and 6E showing decarburization and grain growth on interior and exterior surfaces.
Mag. 6X - Etched 2% Nital

Orientation

Left
- Tube 5E - Bottom - Outside
- Tube 5E - Bottom - Fireside
- Tube 6E - Bottom - Outside
- Tube 6E - Bottom - Fireside

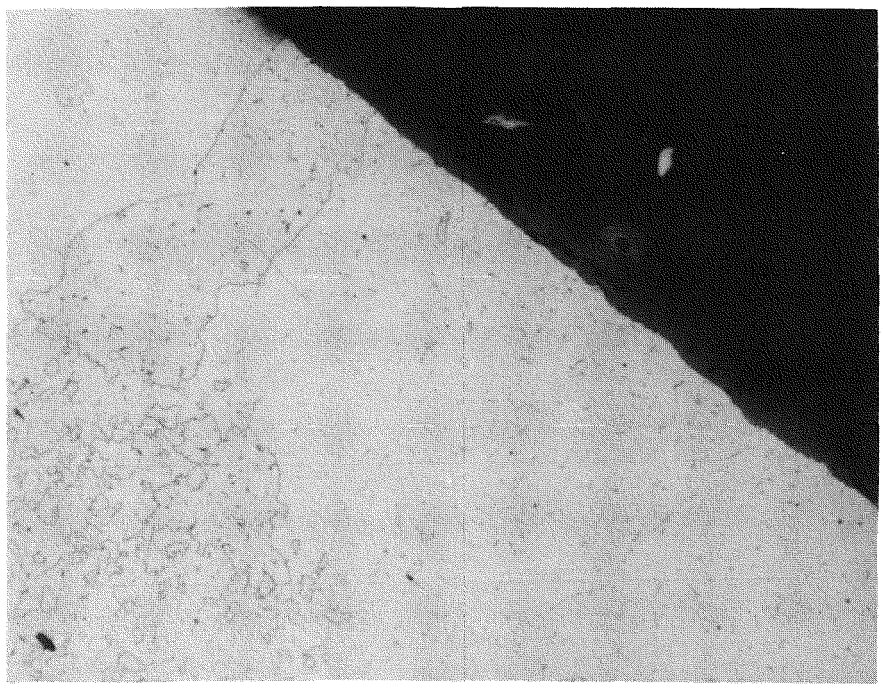


FIGURE 6 - Cross section showing .0175" deep decarburized layer under waterside scale of tube sample 5ETF.
Grain size in decarburized zone is ASTM #1.
Matrix grain size is ASTM #9.
Mag. 100X - Etched 2% Nital

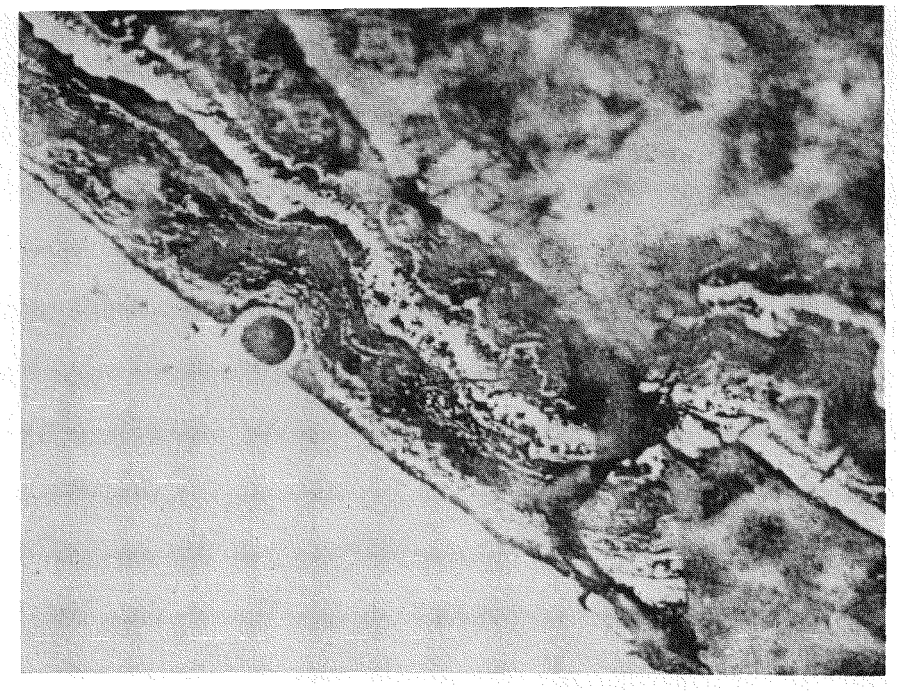


FIGURE 7 - Cross section through Tube sample 5ETF showing waterside scale formation. Adherent layer of Ca scale .0012" thick plus .005" layer of hydrated iron oxide. Mag. 200X - Unetched.

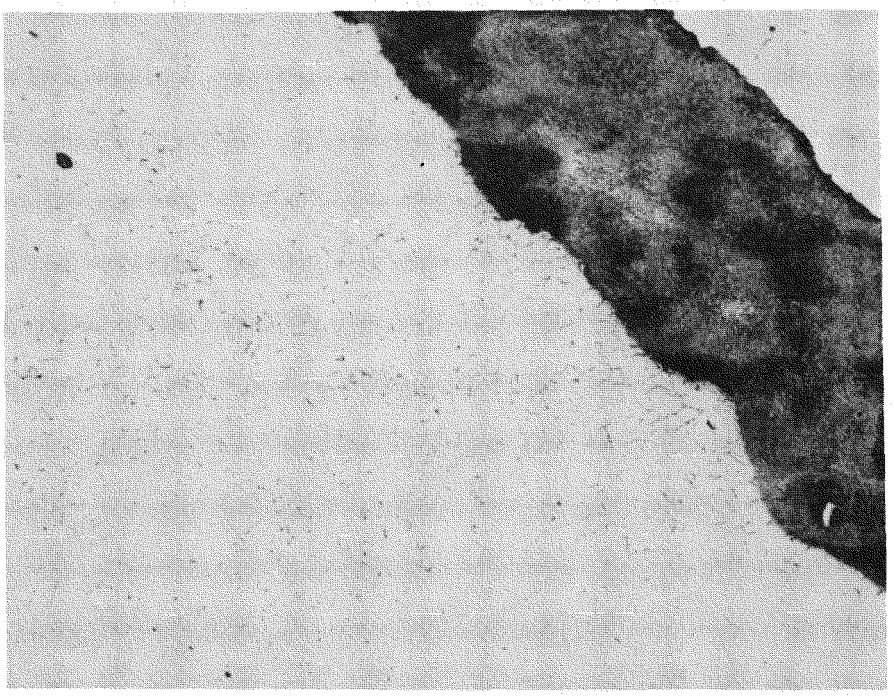


FIGURE 8 - Cross section through Tube 5ETF showing erosion and roughening of fireside tube surface. Matrix ferritic grain size is ASTM #8.
Mag. 100X - Etched 2% Nital

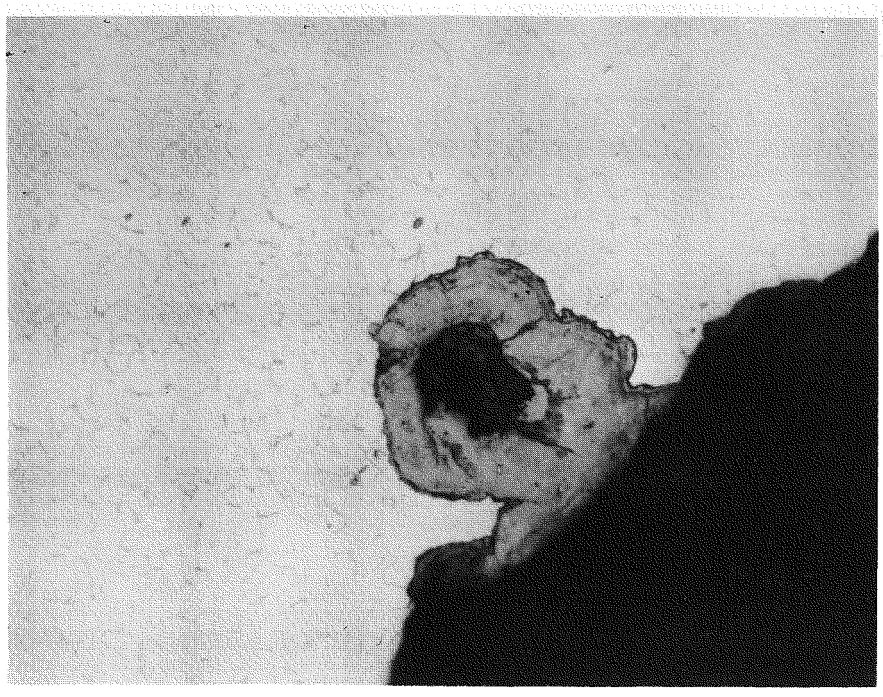


FIGURE 9 - Cross section through tube sample 14MF showing general erosive roughening and a high temperature corrosion-oxidation pit .006" deep.
Mag. 200X - Etched 2% Nital



FIGURE 10 - Cross section through tube sample 5WMO showing appearance of the normal internal microstructure, ferrite grains plus iron carbide. ASTM grain size 9-10. Mag. 100X - Etched 2% Nital

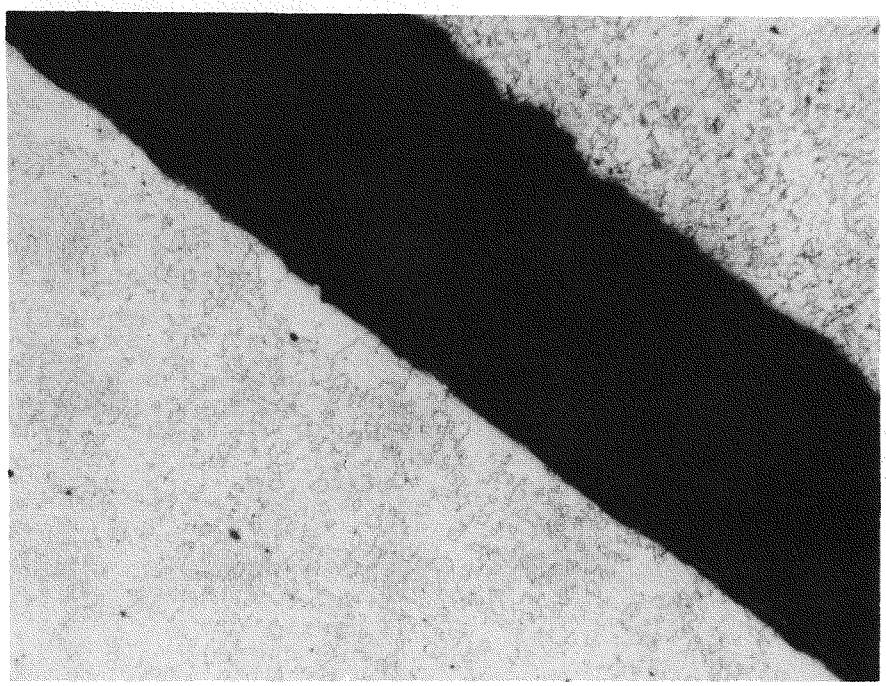


FIGURE 11 - Cross section through tube sample 10TO showing appearance of patchy decarburization on the water-side. Normal matrix structure is ferrite plus iron carbide. ASTM grain size 9-10.
Mag. 100X - Etched 2% Nital

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AUTHOR R.D. Cooper

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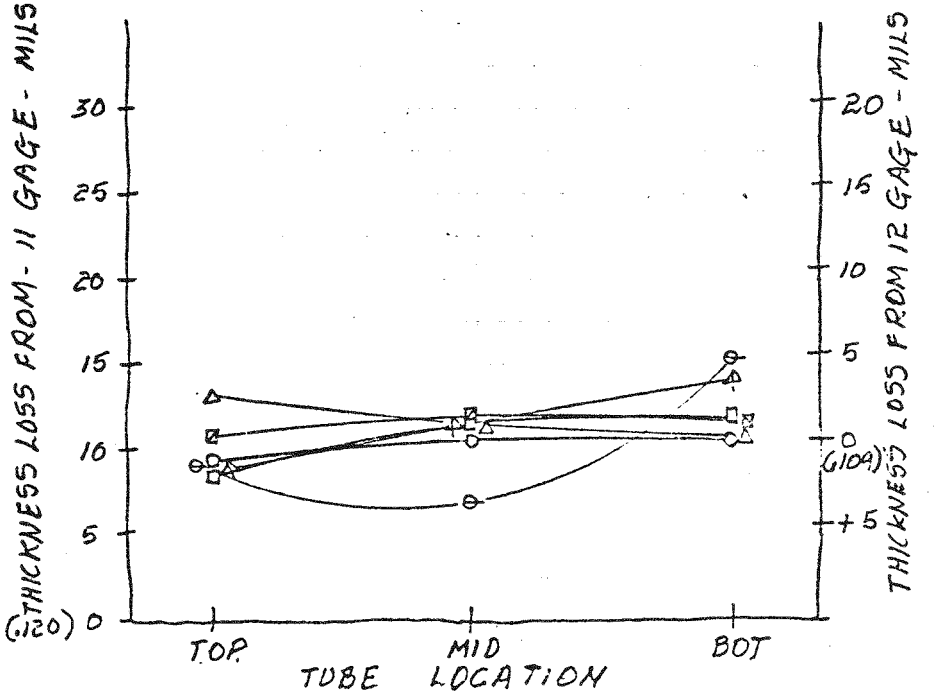


FIG-12A - OUTSIDE TUBE WALL DATA

KEY

TUBE	SYMBOL
5E	○
5W	□
6E	△
10	⊖
13	⊠
14	⊕

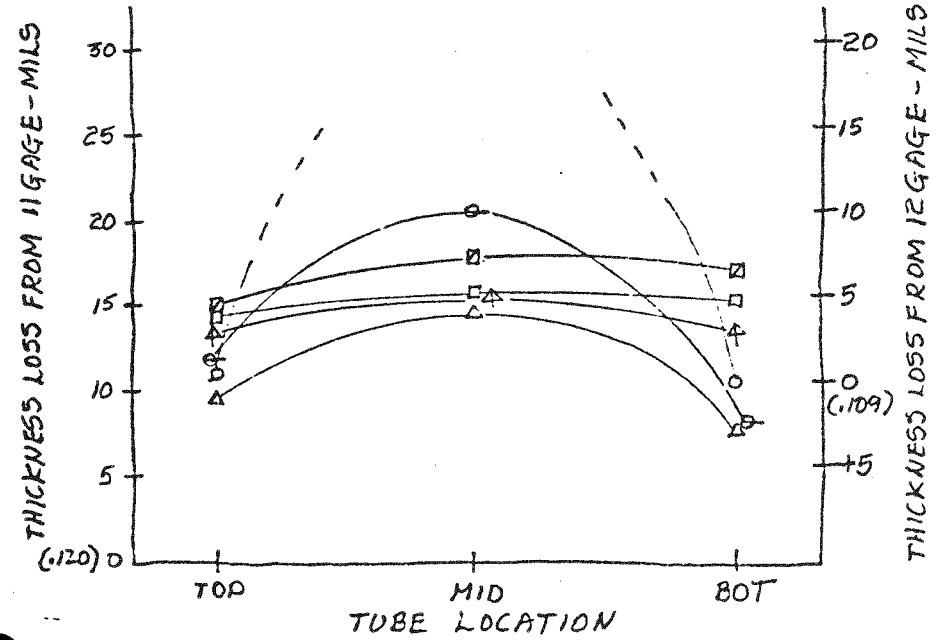


FIG-12B - FIRESIDE TUBE WALL DATA

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