DOE/MC/16230-152

RIVESVILLE MULTICELL FLUIDIZED-BED BOILER

Annual Technical Progress Report for the Period
July 1978—June 1979

Work Performed Under Contract No. AC21-81MC16230

Pope, Evans and Robbins, Incorporated
New York, New York

TECHNICAL INFORMATION CENTER
UNITED STATES DEPARTMENT OF ENERGY

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Prepared by

Pope, Evans and Robbins, Incorporated

Annual Technical Progress Report

For the period July 1978 - June 1979

Prepared for

UNITED STATES DEPARTMENT OF ENERGY
Morgantown Energy Technology Center
Morgantown, West Virginia

Under Contract No. DE-AC21-81MC16230

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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-ERDS/OCR Contract No. EX-76-C-01-1237 (E(49018)-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 Design Manual 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 1978 through June 1979 time period.

I. SUMMARY

1.1 Work Covered by This Report

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

- a) Operation of the boiler and auxiliary systems to evaluate performance and emission characteristics, establish operational reliability, and commercial operation feasibility,
- b) Preparation of Design Manual,
- c) Conduct test programs at the Alexandria, VA PDU Laboratory to support operation of the Rivesville MFB unit and to continue investigation of the atmospheric fluidized bed combustion process.

1.2 Review of Significant Events

The following significant events occurred in the period from July 1978 thru June 1979

Foster Wheeler Energy Corporation proposed new grid plate design utilizing "nut and bolt" design	July	1978
Continuous multicell steam generation with intermittent commercial operation	July - August	1978
Completion of series of 11 tests to measure unit performance	May - August	1978
Fire occurred at Rivesville MFB which damaged electrostatic precipitator, air preheater, I.D. Fan and inter-connecting ductwork. Reconstruction Plan (Project Phoenix) implemented to remove damaged equipment and restore MFB to operating capability.	August	1978
Project Phoenix outage utilized for MFB plant modifications and improvements.	August to March	1978 to 1979
Training Program initiated to provide detailed instruction to MFB Test Facility personnel	November	1978
Bed Drain system to ash cooler modified from pressure to vacuum system.	February	1979
Nut and Bolt grid plate installed in CBC.	March	1979
Flyash reinjection to CBC confirmed.	April	1979
Commercial operation established	April	1979
Two preliminary unit performance tests conducted to test run all data collection equipment and methods prior to upcoming test plan period	May	1979
Design and installation of a screw feeder for flyash transport to D cell; testing and modification performed	May	1979
Study of C and B-cells combustion by means of excess oxygen and superficial velocity measurements	May	1979
Initiation of isokinetic sampling.	May	1979

1.3 Summary of Activities

1.3.1 General

Operation of the Rivesville MFB unit continued and commercial operation was successful during the reporting period. Eleven performance tests were conducted in addition to two preliminary unit performance tests to test run data collection equipment prior to future testing.

Several problems were encountered which required modifications to the existing installation. Intermittent commercial operation followed until 9 August, when a fire occurred in the air pre-heater causing extensive damage and bringing about an outage lasting until early 1979. Project Phoenix was successfully completed and the unit put back in operation in March 1979. Following the outage, commercial operation was restored in April 1979.

1.3.2 Laboratory Work at Alexandria, Virginia

Work at the Alexandria Laboratory supported operation of the Rivesville 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

Hot model experiments and design verification tests were conducted using the process development unit (PDU) with a bed cross section of approximately nine square feet.

Tests were conducted to obtain the relative performance of cleaning modes for the baghouse i.e. reverse air and/or shake cleaning.

The perforated gridplate air distributor did not perform satisfactorily and various methods of insulating the gridplate from the operating bed led to the selection of drilled bolts projecting 1-3/4" above the plate. During operation, the bed material around the bolts did not fluidize and thus formed a 1-3/4" stagnant layer which insulated the base plate from the hot fluidized bed.

Tests were performed for the comparison of static Heat Transfer Bed Height (HTBH) and Effective Operating Bed Height (EOBH) bed level measurement. Four tests showed that the HTBH paralleled the static bed depth much more closely than the erratic EOBH. Not only did the sensor monitor the bed depth more closely, but the HTBH was much more sensitive to small changes in the bed depth compared to the EOBH.

The difficulties encountered in feeding flyash into the fluidized bed led to a series of hot and cold tests on the flyash reinjection system with and without solids flow to determine modifications required in the system.

Using data acquired during test, a new eductor-type oil nozzle was designed. This nozzle mixes the oil with air before it enters the bed. During these tests, the boiler was brought to operating conditions using coal and then oil feed was initiated. In one of the tests, a coal feed problem was simulated and the bed temperature was maintained, thus proving the efficacy of an oil backup system.

Process Analysis Results

It was established that a simple heat balance around a fluidized bed can be used to estimate many of the bed operating variables. With this premise, effort was directed toward the development of a general, simplified heat balance equation that could be used to predict operating conditions in a fluidized bed boiler.

In developing this heat balance equation, several key variables have been identified such as air feed rate (or superficial velocity), fuel feed rate and bed depth.

In general, solution of the heat balance equations involves specifying three of the four variables W_A , W_F , H and T_B and calculating the fourth. The two most useful variables to solve for are the fuel feed rate (W_f) and the bed temperature (T_B)

However, before the heat balance equations can be used, certain other key variables must be approximated. The key variables that affect the heat balance are: (e) combustion efficiency, (x) the fraction of inbed burning, (U_B) the inbed heat transfer coefficients between the bed and tube bundle, and (U_W) between the bed and waterwalls. Generally a few sets of operating data from a unit are needed to enable an estimation of these variables.

1.3.3 Rivesville MFB Plant Operations

During this reporting period, the Rivesville facility expanded its commercial operation despite an extensive outage and major modifications in equipment in the multicelled fluidized bed boiler. The outage was caused by a fire in the air preheater with subsequent extensive damage to the preheater and associated equipment. The equipment replacement and repairs were completed and commercial operation restored in April 1979 for the first time since the fire.

The August 1978 fire caused extensive damage to the facility and during the outage needed maintenance and repairs, in addition to the repairs needed as a result of the fire, were performed. A

new grid design was introduced during this period together with numerous other system improvements.

Continued cooperation from Monongahela Power Company, Stone & Webster Engineering Corporation, Union Boiler Company, Foster Wheeler Engineering Corporation and the Department of Energy enabled the Rivesville facility to accomplish many goals this year.

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 stress analysis of the 6 in and 10 in. duct runs for particulate sampling, design of additional access/egress facilities, studies involving removal of the coal dryer for relocation to a different site, installation of a new F.D. fan with a greater static head, relocation of Rotex screen, redesign of the bed material letdown system, an increase in the flue gas temperature from 730°F to 800°F.

Other studies and analyses included the following:

- Dust Collector NO. 1 Bypass
- Alexandria PDU Baghouse Filter
- Ash Cooler and Classifier Access Opening
- Investigate Limestone Storage and Quarry
- Silo Vent Baghouse Filter
- Baghouse Replacement
- Install Fire Protection System Extension
- Stack Silencer Installation
- Electrical Equipment Room Design
- Control Room Modification
- Coal Feed Eductor System
- Central Vacuum Cleaning System
- Instrumentation Shop Relocation
- Flyash Reinjection System
- Boiler Circulating Pumps
- Preventative Maintenance Program

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 include:

- Foster Wheeler Energy Corporation - MFB unit operative assistance and improvement
- Stone & Webster Engineering Corporation - plant advisory operation assistance

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 are conducted using a process development unit (PDU) with a bed cross section of approximately nine square feet. Heat transfer surfaces, including a submerged horizontal tube bundles, are capable of handling 10×10^6 Btu/hr heat release in the combustor 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 samples and on-line analysis of the flue gas.

Material handling systems to process $3\frac{1}{2}$ tons per hour of coal and limestone are located adjacent to the laboratory building. Processed material is stored and automatically transferred to weigh bins on the laboratory building roof.

Particulate removal from the flue gas is accomplished by a mechanical collector and by a baghouse that was recently added to the laboratory system for testing purposes.

2.2 Laboratory Systems

2.2.1 Process Development Unit (PDU)

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 and has a cross-sectional bed 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 installed in the lower rear manway and a 9 in diameter quartz window in the lower front manway permit viewing the hot fluidized bed. Figure 2.1 is a photograph of the hot bed taken through the quartz window. A second burner is located in the plenum to pre-heat 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 and four in the freeboard.

Figure 2.2 and 2.3 show the PDU after installation of insulation and piping. Figure 2.4 shows the dimensions of the PDU.

To light-off the boiler, the temperature of an eight inch static bed is raised to between 600°F to 700°F with the overbed and 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 superheat. Figure 2.5 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.

2.2.2 Fuel and Sorbent System

High sulfur bituminous coals from the Redstone seam and the Illinois #6 seam. Low sulfur bituminous coal from the Gauley Eagle #4 seam was also fired. Greer, Grove and Lowellville limestones were used as sorbents. Heating oil #2, 4 and high sulfur #6 oil were also fired.

Coal under ½ in is stored in a silo and pneumatically conveyed to a coal weigh bin with electronic load cells located on the roof of the combustion laboratory. The coal feed rate from the weigh bin is regulated by a variable speed SCR drive rotary feeder (RFCB, Rotary Feeder Coal Bin). Limestone under 1/8 in is stored in a silo and pneumatically conveyed to the limestone weigh bin with electronic load cells adjacent to the coal bin. The limestone feed from the limestone bin is also regulated by a variable speed rotary feeder (RFLB, Rotary Feeder Limestone Bin). Limestone and coal are gravity fed into a rotary airlock RFVFT and then into an eductor for pneumatic transport into the boiler through a submerged feed pipe.

2.2.3 Air Supply and Control System

A high pressure fan capable of providing 12,000 LBS/HR of air at 50"H₂O is used as the forced draft fan (FDF) to provide combustion/fluidizing air. The air flow rate is set by a

pneumatically operated F.D. damper (FCV-4). The air is preheated by a steam coil and metered by the main air orifice (F9). The freeboard space of the PDU is operated slightly below atmospheric pressure in accordance with conventional boiler practice, by means of an induced draft fan (IDF). A pneumatically operated I.D. damper (PCF-2) 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.

2.2.4 Feed Water and Steam Systems

Figure 2.6 shows the water and steam systems schematic for the PDU. Some boiler feedwater passes through the in-bed economizer bundle before going into the steam drum. The flow rate is controlled by a temperature controller (TIC-1, Temperature Indicating 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 three horizontal boiling bundles is withdrawn from the main lower headers and pumped through the bundles to the steam drum by a boiler circulation pump, (BCP). The steam produced, passes through the in-bed superheater bundle. The superheater bundle can also be used to measure the in-bed heat transfer coefficient.

Three methods of steam production control can be used. First, a constant flow controller can be used to maintain a constant flow of steam regardless of changes in boiler pressure. Second, a back pressure regulator can be used to maintain a constant pressure in the steam drum, thereby allowing the flow to fluctuate. The third method of steam flow control utilizes an air cooled steam condenser. The steam condenser can be used to vary the load on the boiler by changing the air flow over the condenser coils. 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 is used to preheat the main combustion air to 300°F.

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 an open tank which acts as a deaeration feedwater heater. The feedwater 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.

2.2.5 Fly Ash System

Fly ash is separated from the flue gas in two stages. The first stage is a multicone cyclone dust collector (DC-1). Fly ash from the bottom hopper can be reinjected into the boiler by means of an eductor. Any excess fly ash overflows into a rotary feeder (RFDC, Rotary Feeder Dust Collector) and is pneumatically transported to the fly ash storage silo for disposal.

Stack gas is drawn through an induced draft fan and goes into the second stage separation - a baghouse. Fly ash collected by the baghouse is batch weighed and then pneumatically transported to the fly ash silo for disposal.

The installation of the flue gas baghouse was completed in the third quarter of the reporting period. The baghouse was designed and procured from the Buell Emission Control Division of Environtech. The baghouse was designed to test the application of high temperature baghouse technology to fluidized bed coal combustion process. Figure 2.7 shows the installed baghouse and Figure 2.8 shows the flow diagram of the baghouse system. An opacity probe was installed at the baghouse outlet to obtain the relative performance of the different cleaning modes. To correlate the opacity probe readings, an isokinetic probe was installed at the baghouse outlet.

The baghouse which measures 8'11" x 4'2" x 35' long, was designed to handle 3,000 ACFM of flue gas at 400-500°F stack temperature using teflon finished fiberglass bags. Two air casing caps were provided. Figure 2.9 shows the two possible baghouse configurations which permit reverse air and/or shake cleaning mode and pulse-jet cleaning mode. Only reverse air and/or shake cleaning mode was used in the reporting period. In both configurations, the dirty flue gas enters the bottom hopper of the baghouse and exits from the top of the baghouse.

(1) Reverse Air/Shake Cleaning Mode

The shaker drive top works assembly is used and thirty-two bags (8"φ x 23', 9% teflon coated) are clamped to a bottom tube sheet. The dirty gas flows up the inside of the bags with the clean air on the outside during the filtration mode. The duration of the shake or reverse air cleaning period can be varied from 2 to 120 seconds at the baghouse control panel. The system can be operated with reverse air only, shake only or

combination of the two, via control and timing circuits. Figures 2.10A and 2.10B show the flow diagrams of reverse air and shake cleaning modes. In the reverse air mode, the flue gas bypasses the baghouse by the use of three pneumatically operated flow diverting valves (poppet valves). The gas in the baghouse is momentarily reversed as shown, causing the filter bags to puff out and knock off accumulated caked dust. In the shake mode, the dust is periodically knocked off the bags by shaking of the assembly.

(2) Pulse-Jet Mode

In the pulse-jet cleaning mode, the top assembly with air piping and valves will be used and the bottom tube sheet will be removed. The dirty gas flows up the outside of sixty six bags (6" ϕ x 10', 9% teflon coated) with the clean air on the inside during the filtration mode. The filter bags and air casing cap are periodically pulsed by high pressure air at automatically controlled time intervals, knocking off the accumulated dust. Figure 2.11 shows the pulse-jet cleaning gas flow diagram.

2.2.6 Bed Material Handling System

Bed material removal from the boiler is regulated by a stainless steel trim, air operated, knife-gate valve. Bed material passing through the valve is pneumatically transported by means of a rotary blower (RBBM, rotary blower bed material) and an eductor to the roof of the laboratory, where it is separated from the transport air by a cyclone dust collector (DC-2). Bed material can be dumped from the cyclone through dump valve (CDV) or dropped onto a vibrating screen (VS-2) through the cyclone block valve (CBV). Oversized bed material can be dumped, and undersized goes into a bed material storage bin. Bed material in the storage bin can be either gravity fed back to the PDU through a variable speed rotary feeder (RFBM), or dumped through a bin dump valve (BDV). Bed material dust in the cyclone air discharge is collected by a bin vent bag filter (BV-4), and discharged by (RFBMD, rotary feeder bed material, dust) to disposal.

2.2.7 Material Handling System

The system is designed to process 3.5 tons/hour of coal or limestone. An inlet moisture content of 7 percent can be reduced to an outlet moisture content of 3 percent. The feed of up to 1½ in top size can be reduced to a nominal 1/8 in top size.

Figure 2.12 shows the major components of the material handling system. The system can be operated in either "dry" or "no dry" modes. When in "dry" mode, coal or limestone is unloaded into materials loading hopper #1 and transported by TCD (tubular conveyor dryer) to the dryer. The output of the dryer is transported by TCC (tubular conveyor/crusher) to VS-1 (vibrating screen #1). When in the "no dry" mode, coal or limestone is loaded into materials loading hopper #2. The vibrating screen (VS-1) can process up to 6 tons/hour. The oversize material is dropped into RCLC (roll crusher coal/limestone) and after crushing is transported by TCV (tubular conveyor vibrating screen) back to VS-1. The materials passing through the screen are pneumatically transported by RBLC (rotary blower limestone/coal) to coal or limestone silos. Limestone is generally supplied in the required size and pneumatically transported directly to the storage silo.

The description of components follows:

Conveyors are six inch diameter tubular conveyors manufactured by Kalamazoo Conveyor Company, each with a Rotector zero speed switch, which is an interlock to shut down the system on conveyor failure. The motors on TCD and TCV are two speed motors with speed selection made at the materials handling control panel.

The dryer is a Rexnord Model QAD direct contact vibrating conveyor. Heated air passing through the air distributor conveyor pan, removes moisture and is cleaned in a pulse-jet baghouse to remove carryover fines before discharge.

Heat is supplied to an Eclipse Model 120, air heat duct burner unit. The burner is a fresh air system and has a capacity of 1,200,000 Btu/hr firing propane. Components are wired to a combustion control panel for safety start-stop and desired temperature controls. Forced and induced draft fans on the dryer, DFDF and DIDF, have dampers to regulate the dryer air flow.

The screen is a Rotex Model 402A with capability for double screening should undersize removal be added to the system.

2.2.8 Combustion Control System

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

Figure 2.13 shows a schematic 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 proportional-integral (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 O₂ in the stack. The third control loop is bed level and SO₂ control loop. This loop uses a P-I controller to maintain bed height and three proportional controllers to adjust the limestone feed and bed removal rates to maintain a set SO₂ concentration in the stack. Figure 2.14 shows the interface schematic between the PDU and the control system.

2.2.9 Chemical Analysis System

The fluidized-bed combustion laboratory includes a chemical laboratory which has the capability of handling nearly all of the required gas and solid chemical analysis. 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.

2.2.9.1 Solids Sampling

Bed material samples are withdrawn via a tube which penetrates the furnace wall, terminating just above the grid plate. Samples are collected in a nitrogen purged stainless steel container and cooled in order to preserve the carbon content and to reduce moisture deterioration of the sample.

Coal and limestone samples are taken at the base of the coal and limestone silos between fill cycles of the coal and limestone weigh bins. Samples of coal and limestone are taken once an hour during a test and usually combined for analysis.

Dust samples are collected from the stack using a Rader high volume isokinetic sampler. The sampling ports are located at the entrance and exit of the baghouse.

Fly ash samples of the cyclone catch are taken from the pneumatic transport line between the multicone dust collector (DC-1) and the fly ash silo by means of a diverter valve. A bag filter is screwed into one outlet of the diverter to collect a full flow sample.

2.2.9.2 Flue Gas Sampling

Samples are continuously taken from the freeboard and/or the upper bed regions of the PDU using a probe manifold consisting of three sintered stainless steel filters shown in Figure 2.15. The gas conditioning system removes any remaining dust and water from the gas sample before input to the flue gas analyzers. Components of the flue gas conditioning system are shown in Figure 2.16. The conditioned gas sample is pumped to the gas analysis system shown schematically in Figure 2.17.

2.2.9.3 Methods of Analysis

Table 2.1 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.

2.3 Test Results

During the reporting period the following tests were completed:

<u>Test No.</u>	<u>Date</u>	<u>Test Description</u>	<u>Duration, Hrs.</u>
701-1	7/11/78	Local fluidization	7
701-2	7/18/78	" "	7
701-3	7/20/78	" "	7
701-4	7/26/78	" "	7
701-5	7/31/78	" "	7
702-1	8/9/78	#2 Oil firing	6
702-2	8/15/78	" " "	6
700-24	8/24/78	Comparison PDU and FBM	16
702-3	9/8/78	#2 Oil firing	4½
702-4	9/13/78	" " "	4
702-5	9/14/78	Light-off trail on oil	7
702-6	9/19/78	#6 Oil firing	5
703-1	9/28/78	Large size coal	25
703-2	10/5/78	" " "	7
703-3	10/10/78	" " "	8
703-4	10/12/78	" " "	5
703-5	10/17/78	" " "	5
702-7	10/20/78	#6 Oil firing	6
700-25	10/25/78	Illinois #6 coal	5
702-8	10/27/78	#6 Oil firing	7
702-9	11/1/78	#6 Oil firing	16
700-26	11/13/78	Drilled bolt grid plate	5
700-27	11/14/78	" " " "	6

<u>Test No.</u>	<u>Date</u>	<u>Test Description</u>	<u>Duration, Hrs.</u>
700-28A	11/20/78	Drilled bolt grid plate and sulfur capture for Ca/S ratio of 1.	44
700-28B	12/5/78	Drilled bolt grid plate and sulfur capture for Ca/S ratio of 7.7	16
700-29	12/12/78	Drilled bolt grid plate	7
700-30	12/28/78	" " " "	7
700-32	1/9/79	" " " "	7
702-10	1/18/79	#4 Oil firing	4
702-11	1/19/79	" " " "	7
PDU-5	2/15/79	Shakedown coal feed system changes	6
700-33	2/28/79	Drilled bolt grid plate	6
700-34	3/2/79	" " " "	7
PDU-6	3/3/79	Shakedown	2
700-35	3/8/79	Drilled bolt grid plate	5
700-36	3/15/79	Lowellville limestone	27
700-37	3/21/79	" "	4
700-38	3/22/79	" "	31
700-39	4/6/79	" "	22
704-1	4/11/79	Georgetown coal and limestone	4
704-2	4/19/79	" " " "	8
704-3	4/24/79	Georgetown coal and limestone fly ash reinjection	5½
704-4	4/26/79	" " " "	6
704-5	5/3/79	Low sulfur coal and fly ash reinjection	30
704-6	5/22/79	" " " "	20
704-7	6/1/79	" " " "	5
704-8	6/5/79	Grove limestone and fly ash reinjection	5½
704-9	6/6/79	" " " "	8
704-10	6/18/79	Fly ash reinjection and preliminary baghouse	6
704-11	6/20-21/79	Fly ash reinjection and preliminary baghouse	32
704-12	6/27-29/79	Fly ash reinjection and baghouse	56

2.3.1 Air Distributor Testing

Over the past few years various air distributor designs have been tested at the Alexandria facility. Based on economic considerations and performance evaluations derived from these tests, a perforated plate was selected as the air distributor for the Rivesville MFB. The perforated plate air distributor has deteriorated and analysis concluded that buckling of the plate resulted from high temperature stress-oxidation cracking. This was believed to be the result of grid plate restriction during thermal expansion and exposure to excessive operating temperatures. Various

methods of insulating the grid plate from the operating bed have been tested, including layering rocks and refractory materials over the perforated grid plate (see Interim Report No. 4, 1978). As an alternative to these air distributors, a drilled bolt air distributor was tested in the PDU.

The drilled bolt air distributor is a prototype of the grid plate design now operating in D cell of the Rivesville MFB and in both cells of the Georgetown AFBB. The nine sq. ft. air distributor, contains 2784, 5/16" x 18 stainless steel bolts on 0.724" centers mounted on a 1/4" thick carbon steel base as shown in Figure 2.18. Each bolt projects 1-3/4" above the plate and is drilled out to 7/64" at the end projecting into the bed, yielding approximately 2% open area. During operation, the bed material around the bolts is not fluidized and forms a 1-3/4" stagnant layer which insulates the grid plate from the hot fluidized portion of the bed.

2.3.1.1 Test Procedures

Following installation of the air distributor, five Type K thermocouples with 1/8" diameter sheaths were mounted to monitor temperatures at various locations as shown in Figure 2.19. Thermocouples #1, #2, and #4 were mounted from underneath the distributor, 1/8" into the carbon steel base to measure metal temperatures. Thermocouple #3 was also mounted from below the grid plate but passed through the 1/4" thick base and extended 7/8" into the dead zone of bed material. Thermocouple #5 was located on the grid plate divider approximately one inch below the base. All thermocouples were peened into position. Preliminary tests were run to determine the quality of the air seal between the air distributor and boiler frame. Sulfated bed material was placed in half of the dead zone and air flow initiated. Any leaks would show as spouts through the bed material. Finding no leaks, an initial pressure drop across the grid plate at ambient conditions without bed material was recorded. Following this test, similar data was gathered at ambient conditions with 24 inches of static bed depth.

Test No. 700-26 was the first hot test run with the drilled bolt grid plate. To simulate Rivesville conditions, the bed was run at 24" static depth and 1550°F. The steam air preheater coil (designed for 300°F preheat) and the plenum propane burner were used during the entire test to achieve an air preheat near the Rivesville design value of 600°F. During this test, data was collected for plotting grid plate ΔP versus main air flow. An emergency shutdown was simulated by slumping the bed on the air distributor and monitoring the temperature as a function of time.

Test No. 700-27 was run at a bed temperature of 2000°F. This condition would exist in the Rivesville CBC and would be the most extreme temperature to which the grid plate would be exposed. In order to attain this temperature during the test run, it was necessary to operate with a reduced bed depth of 8" static. As in Test No. 700-26, the bed was slumped after being maintained at 2000°F for a period of time.

Following these tests conducted specifically for air distributor characterization, thermocouple #5 was removed from the plate divider and peened into the frame which supports the grid plate. After each hot test, the bed was drained and vacuumed clean and the grid examined for signs of buckling and bolt damage. To determine whether the bolts were plugging, a grid pressure drop test at ambient conditions without bed was performed after each hot test. If plugging occurred, the pressure drop at the same air flow rate would increase. When not in use, the grid plate was covered with 4" to 6" of bed material, and was occasionally rolled with steel bars. These steps were to induce plugging of bolt holes arising either from down time or maintenance on the boiler which would require walking on the grid as would be the case at Rivesville. The plenum was also examined for sifting of fine bed material through the grid. A summary of all grid plate test hours is given in Table 2.2.

2.3.1.2 Test Results

Air Distribution Pressure Drop

The change in grid pressure drop in inches of water with air flow is shown in Figure 2.20. Values are shown for 1550°F and ambient conditions with a 24" static bed as well as ambient conditions with a 24" static bed as well as ambient with no bed. Because of the severity of the 2000°F condition in Test No. 700-27, the air flow was not variable over a large enough range to both gather data and maintain temperature. The MFB design superficial velocity in the main cells at 1550°F bed temperature is 12 ft/sec with 600°F air preheat. The curve in Figure 2.20 shows that in order to attain 12 ft/sec at 1550°F, the pressure drop across the grid plate must be 16 in W.C. (7776 lb/hr through a PDU cross section of 9 sq ft). This result is lower than the design pressure drop of 20.0 in W.C. for the perforated plate type distributor.

Minimum Fluidization Velocity

During the ambient 24" static bed test, data was collected to determine the pressure drop across the grid plate needed to attain minimum fluidization. Following the onset of fluidization, the pressure drop across the bed is constant

over a wide range of fluidizing velocities since the excess air at higher velocities goes into bubble formation. This fact was used along with the pressure drop across the grid versus main air flow plot for the ambient 24" static bed case shown in Figure 2.21, to determine minimum fluidization. At minimum fluidization the drilled bolt grid plate pressure drop was 4.2" W.C. This was nearly identical to the perforated plate minimum fluidization pressure drop of 4.0" W.C.

Air Distributor Temperatures

Data for grid plate temperatures at normal operating conditions of 12 FPS, 1550°F and 24" static bed was collected during Test Nos. 700-28A, 700-29, 700-30, 700-33 and 700-36. For each test, the bed temperature was averaged as were the grid plate thermocouples #1, #2, and #4. Thermocouple #3 located midway in the dead zone was averaged separately. Thermocouple #5 located on the frame, read the same value as the grid plate thermocouples in all cases. The results are summarized in Table 2.3.

Test Nos. 700-26 and 700-27 simulated emergency slumping at 1550°F and 2000°F respectively. The results are shown in Table 2.4. Following slumping of the 1550°F bed in Test No. 700-26, the plate thermocouples were monitored to follow grid plate temperatures as a function of time. However, some thermocouples were influenced by the plenum burner. As a result, after slumping, these temperatures showed cooling trends since the burner was extinguished. Therefore, thermocouple #4, away from the burner yet still in the 600°F plenum air, was used as the grid plate temperature.

During Test No. 700-27, the plenum burner was not used resulting in the normal 300°F air preheat. This produced even temperatures across the grid plate. On slumping the 2000°F bed, temperatures were monitored every 30 seconds for 65 minutes. These temperatures versus time are shown in Figure 2.22. The maximum average grid temperature was 603°F for the three grid plate thermocouples.

Plugging and Sifting

Following each hot test the grid was examined for damage and signs of wear. As many as 50 to 100 nozzles were found to contain small amounts of bed material, most of which was easily removable. However, grid ΔP versus main air flow results at ambient no bed conditions after each hot test, were well within the limits of error when compared to the initial run at similar conditions prior to testing. This fact is shown in Figure 2.23 where the grid ΔP versus air flow rate curve after 184 hours of hot testing is nearly identical to the initial test, indicating no pluggage of bolts. Examination of the plenum after each test showed sifting to be minimal in all cases.

2.3.1.3 Conclusions

A total of 398 hours of hot testing including approximately 50 boiler cycles were completed during the reporting period. All results showed the prototype drilled bolt air distributor to be a satisfactory design for FBC units. Not only did the prototype show favorable pressure drop and fluidization characteristics, but the "dead zone" insulating concept thoroughly protected the grid plate from the severe operating temperatures of the bed. These conclusions have been borne out by the installation and successful operation of the drilled bolt air distributor at the Rivesville MFB cell D and the Georgetown AFBB facilities.

2.3.2 Bed Level Measurement

A probe was developed to measure the bed level using heat transfer principles as an alternative to the differential pressure tap means of determining bed height. Because the inbed heat transfer coefficient is 3 to 5 times the above bed coefficient, the steam flow rate can be changed by small changes in bed height in the region of the submerged tube bundle. Since the reliability and accuracy of the differential pressure tap level measurement method is limited by pluggage and pressure fluctuations, the development of a useful alternate is desirable.

2.3.2.1 Theory

The measuring technique currently employed is based on the large difference between heat transfer coefficients in the bed and in the gas above the bed. A fluid passed from the freeboard into the bed through a vertical tube should experience a more rapid heating rate in the bed region than in the freeboard. Thus, if the heating rate in the bed is greater, the temperature rise over a given distance in the bed will be greater than that over the same distance in the freeboard. Thermocouples located intermittently along the inside of the tube would give the temperature as a function of distance through the pipe. This temperature profile may be related to bed height by heat transfer principles in the following manner. Consider first, heat transfer to the inbed portion of the vertical tube as shown in Figure 2.24. Performing a steady state heat balance around the differential element dx ;

heat into element = heat out of element

$$\dot{m}C_p T + U_1 (T_B - T) = \dot{m}C_p (T + dT)$$

$$U_1 A_1 (T_B - T) = \dot{m}C_p dT$$

where the symbols are as follows:

C_p = Specific heat of fluid, Btu/lb/°F.

dT = Change in fluid temperature through differential element, °F.

dx = Differential element of height along tube, ft.

H = Height of tube from reference plane to top of bed, ft.

L = Overall height of tube above reference plane, ft.

$\dot{m}$ = Mass flow rate of fluid, lb/hr.

r = Radius of tube, ft.

T = Temperature of fluid entering the differential element, °F.

T_B = Temperature of bed, °F.

$T_H dT$ = Temperature of fluid leaving the differential element, °F.

U_2 = Overall heat transfer coefficient between fluid and freeboard Btu/hr/ft²/°F.

U_1 = Overall heat transfer coefficient between fluid and bed Btu/hr/ft²/°F.

x = Height of tube above reference plane.

A_1 = Area available for heat transfer, sq. ft.

The area available for heat transfer A_1 is $2\pi r dx$.

Substituting this quantity and rearranging gives the following expression:

$$\int \frac{dT}{T_B - T} = R_1 \int$$

where $R_1 = \frac{2\pi r U_1}{\dot{m} C_p}$

Integrating and substituting the following boundary conditions:

$$T = T_H \text{ @ } X = 0$$

$$T = T_C \text{ @ } X = H$$

one obtains

$$\ln (T_B - T) = R_1 (X - H) + \ln (T_B - T_C) \quad (1)$$

Thus, a plot of $\ln (T_B - T)$ vs X should yield a straight line of slope R_1 for all temperatures and distances in the bed.

A similar analysis of the freeboard region, assuming the temperature in the freeboard to be the same as that in bed, results in an equation similar to (1), i.e.

$$\ln (T_B - T) = R_2 (X - L) + \ln (T_B - T_1) \quad (2)$$

where $R_2 = \frac{2\pi r U_2}{\dot{m} C_p}$

A plot of $\ln (T_B - T)$ vs X for temperatures and distances above the bed should again yield a straight line of slope R_2 .

Thus, Equations 1 and 2 together give the temperatures-distance profiles in and out of the bed respectively. The intersection of the two lines is the heat transfer bed height (HTBH). Assuming Equations 1 and 2 to be of the form

$$\ln (T_B - T) = a_n X + b_n \quad (3)$$

The HTBH may be solved by Equation 3:

$$HTBH = \frac{\ln \frac{a_1}{a_2}}{b_2 - b_1} \quad (4)$$

2.3.2.2 Test Arrangement

To put this theory into practice, a test level probe was constructed and mounted in the PDU. The vertical tube was 3/4", 304 stainless steel piping 48 in. long. The fluid used was air for two reasons. First being a gas, it is very sensitive to temperature change relative to a liquid. Second was the fact that the C_p of air is about constant over the temperature range being considered (500°F - 1200°F). The air was supplied by a 20 CFM blower to keep the thermocouples below 1200°F. The thermocouples used were 1/16" chromel-alumel (type K) with inconel sheaths. The orientation of both the sensor and thermocouples in the bed are shown in Figure 2.25. The apparatus was mounted on the back wall of the PDU, with air flowing from the top and expelling into the bed.

2.3.2.3 Results and Discussion

In order to test the theory, the bed level sensor was operated during four tests, Test Nos. 702-7, 700-25, 700-28A and 700-28B. Normal pressure tap logs were kept during building of the bed to the 24" static level and the effective operating bed height (EOBH) from the pressure tap readings calculated as for every test. Figures 2.26 and 2.27 show typical plots of EOBH and HTBH respectively. As Figure 2.27 shows, the HTBH data falls on two straight lines as predicted by theory. To determine the intersection of the two lines, both sets of data points, those in and those above the bed were fitted to curves of the form of Equation 3 and solved simultaneously to obtain the HTBH by Equation 4.

Typical results of the HTBH method are shown in Figures 2.28 and 2.29. Since the first thermocouple was 19 in. off the distributor and a minimum of two points are needed to use Equations 1 or 2, data comparison was not attempted until the static bed height was at least 16 in., corresponding to an EOBH of about 25 in.

Comparison of EOBH and HTBH data indicates that for the same static bed heights, the two methods give different expanded bed depths. The reason for this lies in what each method actually measures. The EOBH from the differential pressure tap technique measures the mass of bed material above a point in the bed relative to a tap 1 in. above the grid. A pressure tap will only give a constant reading when covered by the bed. Therefore, by plotting the differential pressure versus tap location, the plot shown in Figure 2.26 is obtained to give the EOBH. The HTBH method on the other hand, measures the total height, including splashing. When the bed splashes due to bubbles breaking the surface, a portion of the bed is suspended and contributes nothing to the mass of the bed seen by the pressure taps. This splashing, however, is perceived by the bed level sensor since the heat contained by the hot bed is transferred to it by contact with the splashing material. Herein lies the difference in the EOBH and HTBH recordings.

The basic qualities to be examined with this new procedure were consistency, repeatability, sensitivity and endurance. Since the HTBH measures the splash height of the hot bed, at constant bed mass it is affected primarily by changes in superficial velocity. Because superficial velocity during bed height building is difficult to keep uniform from test to test, comparison between tests is difficult. However, trends repeatably show the HTBH to more closely parallel the static bed height than the EOBH (see Figures 2.28 through 2.31). This fact is clearly evident in Figure 2.29 where the bed building was very slow and extended to 28 in. static. The EOBH appears very erratic whereas the HTBH consistently parallels with static measurement.

The sensitivity of the HTBH technique is demonstrated in Tables 2.5 through 2.7. For Test No. 702-7 (Table 2.5), a change in the static bed from 16.5 in. to 19 in. produced no change in EOBH readings but was registered by a change in the HTBH. On the other hand when the static bed remained constant at 25.5 in., the EOBH showed a change of almost 3 in. whereas the HTBH remained essentially constant. Similar results may be seen in the 18 in., 21 in., and 24 in. static readings of Test No. 700-28A (Table 2.6) and the 25 in. reading of Test No. 700-28B (Table 2.7). Consistency is further demonstrated in Table 2.8. During Test No. 700-28B, data was taken during steady state running at 1550°F as bed depth was kept as close as possible to 24 in. It can be seen that this mean value of 48.90 in. agrees with the value of 48.50 in. for the same static reading during Test No. 700-28B. The probe held up well for the four runs for which it was used plus two preliminary test runs, a total of approximately 80 hours of hot testing.

2.3.2.4 Conclusions

Tests performed show the HTBH method to be a viable alternative to the differential pressure tap approach. Four tests showed that the HTBH paralleled the static bed depth much more closely than the erratic EOBH. Not only did the sensor monitor the static bed depth more closely, but the HTBH was much more sensitive to small changes in the bed depth compared to the EOBH. Results also show the HTBH method to be repeatable. In addition to these advantages, the HTBH may be a more useful control signal for bed level since it relates directly to heat output.

2.3.3 Flue Gas Baghouse Start-Up

In order to study the application of fabric filters for final particulate removal in the FBC process, a test baghouse was installed at the laboratory. The equipment is described in Section 2.2.5, Fly Ash System. Shakedown was completed and testing started during the reporting period. These activities are described below indicating some of the early problems.

Prior to the start of testing, the following instrumentation was installed to monitor the baghouse operation:

- (1) Pressure gages on the high and low side of the I.D. fan.
- (2) Differential pressure gages for measuring pressure drop across the bags.
- (3) Thermocouples in the inlet and outlet flue gas lines.
- (4) A Lear Siegler Model RM-41 transmissometer for measuring optical density of gas outlet stream.
- (5) Rader high volume automatic isokinetic samples in the inlet and outlet gas streams (sample simultaneously).
- (6) Load cell hopper to batch weigh fly ash collected.
- (7) Recorders for monitoring ΔP bags, optical density and dust weight continuously.

Having found all components of the baghouse operational during cold shakedown tests, two hot shakedown tests followed to familiarize laboratory personnel with the baghouse operation. The baghouse was heated with hot off-gas from the propane burners while heating a shallow bed of spent limestone prior to light-off. These tests showed that irrespective of the cleaning mode, dust hung up on the incline of the baghouse bottom hopper. To alleviate this problem, a manually operated vibrator was mounted on the hopper.

A performance test program was started with Test Nos. 704-10, 704-11, and 704-12. The shake only cleaning mode was selected for these tests. The goals of this test program were:

- (1) Determine the baghouse collection efficiency.
- (2) Determine the optimum cleaning mode and the cycle time (the time between the cleaning cycles).
- (3) Determine baghouse operating characteristics.
- (4) Correlate optical density of the gas outlet to dust loading.

Test No. 704-10 was a six hour run with a 60 minute cycle time and 30 second shake time. Prior to cleaning, ΔP bags was 5 in. w.c. After cleaning, ΔP bags increased from 1.5 in. w.c. to 2.0 in. w.c. after the fourth and final cleaning.

Test No. 704-11 was a 32 hour run with a 30 minute cycle time and the same 30 second shake time. Prior to cleaning, ΔP bags ranged from 4.2 in. to 5.4 in. w.c. After cleaning, ΔP bags ranged from 1.5 in. to 2.5 in. w.c. after the 47th and final cleaning. Half way through the test, the shake time was doubled to one minute. This increase showed no effect on the trend of increasing ΔP bags after each cleaning.

Test No. 704-12 was a 56 hour run with a 30 minute cycle time and one minute shake time. The baghouse hopper vibrator was tied into the cleaning cycle to operate simultaneously with the shaker. Seventy five cleaning cycles were completed and preliminary collection efficiency data were taken. For the first 24 hours of this test the cyclone overflow rotary feeder (RFDC) was not operational, causing the cyclone to overflow. The fly ash from the cyclone that overflowed to the baghouse increased the amount of fly ash deposited on the bags, causing ΔP bags prior to cleaning to be as high as 6 in. w.c. As in Test Nos. 704-10 and 704-11, ΔP bags after the cleaning steadily increased with the number of cycles but leveled off and never exceeded 215 in. w.c. recorded for Test No. 704-11. Due to operational problems with the automatic isokinetic samplers, a manual probe was used to sample the baghouse outlet and then the inlet. Care was taken to insure that the fly ash level in the cyclone was below the overflow so no fly ash collected in the cyclone was transported to the baghouse during isokinetic sampling. Collection efficiency was calculated to be 94 percent. A further check of these numbers was made by comparing the difference between input and output rates to the average collection rate determined by the weigh hopper. These two sets of

data agreed within nine percent. Due to the low collection efficiency, a check of the system was made and bypassing of dirty flue gas to the outlet through the reverse air poppet valve was discovered. Testing will be continued after readjustment of the valve.

2.3.4 Fly Ash Reinjection

A program of eductor development was begun to optimize the design of a low pressure fly ash reinjection eductor. This work was undertaken to provide total fly ash reinjection to the PDU and to provide performance data for fly ash eductor evaluation for the Rivesville MFB and the Georgetown AFBB projects. The previously developed design data for the coal feeder eductors, using low pressure air (less than 15 psig) as a motive force, was the basis for design work. This design was modified by the addition of a diffuser section on the outlet of the eductor to improve the performance.

Use of the eductor to feed fly ash into the fluidized bed proved to be more difficult than feeding coal. Part of the problem was caused by the fact that the fly ash in the cyclone is subject to the negative draft in the cyclone (usually 2 in. w.c.). The eductor must overcome this negative pressure while feeding a large amount of fly ash against the back pressure of the bed. Also since the fly ash is hot, expansion of air in the eductor causes additional pressure drop. The fly ash recycle rates encountered in the hot tests were estimated to be about 1000 lb/hr. At these rates the pressure losses caused by acceleration of the fly ash, transport friction losses in the pipe, and pressure drop through the fluidized bed resulted in the solids inlet pressure in the eductor increasing to the point where the pressure just equaled the cyclone pressure. At this point the fly ash feed rate could not increase and fly ash began to build up in the cyclone.

These problems indicated the need to develop lower solids inlet pressures for fly ash than were needed for coal. A series of hot and cold tests were run on the fly ash reinjection system with and without solids flow to determine modifications required in the system. The test program was conducted in the following sequence:

- (a) Tests on converging nozzle eductor with and without solids.
- (b) Tests on converging nozzle plus diffuser eductor with and without solids flow.
- (c) Tests on converging nozzle plus diffuser eductor with higher pressure blower.

The system used for the first tests, Nos. 704-3 through 704-8, is shown in Figure 2.32. The tests used a positive displacement rotary blower set up to deliver 67 scfm of air at up to 8 psig. The tests were run using a simple converging nozzle eductor. The throat diameter was determined using correlations developed during a previous coal feed eductor program, see monthly Progress Reports Nos. 66 and 67.

Since the purpose of the eductor is to minimize the solids inlet pressure at the design operating back pressure, the solids inlet pressure is considered a measure of the performance of the system. When all other variables are constant, the more negative the solids inlet pressure, the more effective the eductor. The solids inlet pressure, fly ash temperature, fly ash plus air mixture temperature and reinjection back pressure were all measured.

Problems developed during all of these tests due to a build-up of fly ash in the main dust collector. This buildup was the result of the negative draft in the cyclone, approximately 2 in. w.c., and the back pressure of the fluidized bed which has to be overcome by the eductor. When the equivalent static bed level was 10 in. or greater, the feed rates of fly ash dropped off from about 1200 pounds per hour to about 300 pounds per hour, measured using a heat balance around the eductor. This was due to the inadequate performance of the eductor.

After these tests, a diffuser section was added to the eductor to improve its performance at the same blower pressure, see Figure 2.33. The diffuser was installed so that the distance between the solids inlet and the diffuser section could be varied to obtain optimum performance.

The results of the cold tests for the eductor are summarized in Figure 2.34. There is a significant difference in the results of the solids flow and non-solids flow tests with the diffuser in place. This is due to the energy losses required for acceleration of the fly ash through the diffuser section. These losses resulted in a large increase in solids inlet pressure. Although the eductor performance dropped with solids flow resulting in minus 2 in. w.c. solids inlet pressure, this pressure could be maintained at a back pressure up to 40 in.

Test Nos. 704-9 and 704-10 were run on this eductor with diffuser and the data is shown in Figure 2.35. Problems again developed during these two tests with build-up of fly ash in the main dust collector. Even though the solids inlet pressure was sufficient to feed fly ash into the PDU, it was suspected that the fly ash recycle rate was not great enough. Because of unsatisfactory results from Test Nos. 704-9 and 704-10, two additional modifications were

made to the fly ash system; the reinjection blower was replaced with a blower which could operate at 12 psig and the rotary feeder at the bottom of the dust collector was removed and replaced with a four-inch slide gate. The rotary feeder was installed on an overflow dropout. These modifications are shown schematically in Figure 2.36.

After these modifications, there was no evidence of fly ash buildup in the main dust collector hence two longer tests, Nos. 704-11 (a 32 hour test) and 704-12 (a 56 hour test) were completed. The eductor performance data for the hot tests is shown in Figure 2.35.

2.3.5 Localized Fluidization of PDU

A series of partial fluidization tests were conducted, which was to investigate the potential of the partial fluidization technique as a means of improving both light-off and turndown procedures.

2.3.5.1 Equipment Modifications and Test Descriptions

For these tests, a slide gate and a divider plate were mounted in the plenum of the PDU, Figure 2.37. When the gate was open, complete fluidization of the bed could be obtained at normal air flow rates. The coal feed needle was also replaced by a design which could be rotated to feed only into the fluidized half of the bed. The in-bed portion of the new coal feed needle was angled so that it fed the middle of the entire grid area, or during partial fluidization, could be rotated so that the fuel was directed toward the center of the back half of the bed.

A series of cold tests were run on the boiler to check sealing around the slide gate. This check was important because no air should leak into the front half of the bed during slumping, otherwise the coal in the bed will continue to burn and hot spots may develop and result in the formation of clinkers. These cold tests also enabled the interface angle between the fluidized and slumped portions of the bed to be measured, Figure 2.38. The interface angle was measured to be approximately 51° . This angle is steeper than the normal angle of repose which is about 35° . During cold testing, it was observed that bed material was thrown onto the slumped portion of the bed. As this material rolled toward the slumped side, it formed the angle of repose. The actual interface had two regions, the first from near the air distributor to the top of the fluidized portion of 51° , and the second above this at 35° . Furthermore, bed material being transferred to the slumped side resulted in a decrease in the bed height on the fluidized side.

Test No. 701-1 was the first hot run using the slide gate. During this test both overbed and plenum propane burners and air preheater were used. When slumping occurred, the bed continued to burn in the front half of the boiler. After the test when the slide gate was checked, it was found that heat from the plenum burner and the air preheater had warped the divider plate, breaking the seal, and preventing isolation of the front half of the boiler.

For Test No. 701-2, furnace cement was used to seal the divider plate and gate and only the overbed burner was used during light-off. Conditions for Test No. 701-1 were repeated, but because of these changes, air leakage was minimized and the slumped portion of the bed began to cool off. After the front portion of the bed cooled, it was observed that the steam flow rates dropped off, see Table 2.9. This decrease in steam flow was due to the reduced superficial velocity through the upper tube bundles and to the reduced rate of coal feed to the fluidized portion of the bed.

Test No. 701-3 was the first test on the slide gate in which a complete set of samples was collected during the slumped and non-slumped periods of the test. The operating conditions of Test No. 701-3 are shown in Table 2.10.

During Test No. 701-3, a new method of light-off was developed using localized fluidization. With half the bed slumped, the active half of the bed was heated to 600°F and a ten pound charge of coal added. When the overbed propane burner began to ignite the coal, the fluidizing air was turned down to minimum fluidization and the bed temperature began to increase rapidly. Once the temperature of the active half of the bed came up to normal operating conditions, the coal feeder was rotated to the normal center position and the plenum slide gate opened. The hot material from the back of the boiler mixed with the previously slumped portion of bed and the entire bed reached operating conditions rapidly. This light-off procedure could be useful in situations where the amount of air available under cold conditions is limited.

Prior to Test No. 701-4, a 6 in. refractory wall was installed on the air distributor. The design of this wall was based on previous coal tests. Without the wall, the interface angle of the slumped bed was calculated to be 51°, see Figure 2.38. For this angle, none of the superheater tubes were in the slumped portion of the bed. The purpose of the wall was to raise the interface angle so that the desired number of superheater tubes would be immersed in the slumped portion of the bed during turndown. Because of the coal feeder location, only a 6 in. wall could be installed instead of the calculated optimum height of 18 in.

Test Nos. 701-4 and 701-5 were run to test the effect of the wall. The 6 in. wall did not immerse any tube in the slumped portion of the bed, and therefore the data collected is not as useful as desired. Test No. 701-5, see Table 2.10, was comparable to Test No. 701-3, see Table 2.11, with the exception that during Test No. 701-5, the refractory wall was in place. Samples were also collected during complete and partial fluidization periods of this test.

2.3.5.2 Heat Transfer Comparisons

As a means of evaluating the potential of partial fluidization for improving turndown procedure, comparative heat transfer coefficients and combustion efficiencies for partial versus complete fluidization were used. Heat transfer coefficients were calculated for Test Nos. 701-3 and 701-5 and were compared to Test No. 700-17 to evaluate the effects of localized fluidization versus superficial velocity change on steam production, Table 2-12. Heat transfer calculations were made using the data taken immediately after slumping while the bed temperature was still uniform. Calculations could not be made using the data taken long after slumping had occurred, because the slumped portion cooled off and the temperature distribution across the tube bundles was unknown.

The calculated heat transfer coefficients for the superheater, Table 2.12, show that the heat transfer rate decreases as the front half of the bed is slumped. The decrease in the overall coefficient during partial fluidization, 37 percent, is more significant than the decrease resulting from reducing the superficial velocities, 14 percent. This is important because using the calculated interface angle with and without the 6 in. wall, there were no superheater tubes in the slumped portion of the boiler. When the front half of the boiler is slumped, the back half is maintained at a superficial velocity of 12 feet per second near the air distributor with an area of 4.5 sq. ft. Due to the interface angle between the fluidized and slumped portions of the bed, the superficial velocity decreases with height above the air distributor. Because the superheater tubes are above the slumped portion of the bed, the increase in area results in a decrease in superficial velocity to about 6 feet per second. This decrease in superficial velocity accounts for some of the decrease in the overall heat transfer coefficient; the rest is due to the reduced EOBH which reduces contact area between the bed and the top of the tube bundles.

These heat transfer coefficients can be used to explain the decrease in bed temperatures observed during the slumped period of the test and a lower superficial velocities. In reducing the superficial velocity and slumping the bed, the amount of heat released is cut in half due to the fact

that the fuel and air feed rates have been cut in half. However, as shown in Figure 2.38, since the slumping action resulted in only a small amount of the in-bed tube bundle being covered by the slumped bed, too much heat was being removed by the tube bundle. Thus the bed temperature had to drop to maintain thermal equilibrium. This phenomenon was primarily the result of the test boiler geometry. Use of a larger bed would reduce this effect and result in a useful turndown technique.

2.3.5.3 Combustion Efficiencies Comparison

Combustion efficiencies of Test Nos. 701-3 and 701-5 are compared to two previous tests, Nos. 700-17 and 700-19 in Table 2.13. Test No. 700-17 was an eight hour test to determine the significance of superficial velocity on combustion efficiency and heat transfer. Test No. 700-19 was a sixteen hour test to obtain sulfur capture data and to determine the combustion efficiency. Tables 2.14 and 2.15 show the chemical analysis obtained on the bed material, fly ash and dust samples taken during Test Nos. 701-3 and 701-5.

Comparing the calculated combustion efficiencies before partial fluidization with the results of Test Nos. 700-17 and 700-19 at 12 feet per second, the results were found to be compatible. The difference between the values were due to changes in data collection techniques. The discrepancy among the values was due in part to a new fly ash sampling technique whereby a diverter valve in Test Nos. 701-3 and 701-5 replaced the ball valve assembly of Test Nos. 701-17 and 700-19. The results show that the modifications made on the boiler in these tests did not affect operation.

After half the bed was slumped, the calculated combustion efficiencies were less than those during the nonslumped portions of the test. Part of the reduction in combustion efficiency could be due to the overcooled condition in the bed. At 6 ft per second, the inbed tube bundles remove too much heat to maintain 1500°F at 3% O₂. This can be seen by the decreased bed temperature obtained during Test No. 700-17 at 6.5 ft per second. When half of the bed is slumped, the top portion of the bed is still fully fluidized due to the particular geometry of the PDU, Figure 2.38. Thus while the superficial velocity at the air distributor is 12 ft per second, the superficial velocity at the top of the bed (in the tube bundle region) is only 6 ft per second and the bed is overcooled. Since overcooling the bed results in bed temperatures and percent excess oxygen concentrations lower than optimum, the partial slumping action was expected to give lower combustion efficiencies.

However as shown in Table 2.13 the combustion efficiencies obtained were considerably lower than the combustion efficiency obtained during Test No. 700-17 at 6.5 ft per second. This would indicate that there is some mechanism other than overcooling contributing to the decreased combustion efficiency. It is possible that poor mixing in the region of the slumped-fluidized interface is responsible.

The results of the partial fluidization test program can be summarized as follows. First, a new light-off procedure was successfully demonstrated to be of value in cases of limited amounts of air under cold conditions. Secondly, heat transfer and combustion efficiency analyses of slumped and completely fluidized regimes showed partial fluidization to be a promising means of improving turndown in the PDU. Care, however, must be taken in extrapolating the capabilities of this procedure since the geometry of the PDU made it impossible to predict the effects of partial fluidization on larger beds.

2.3.6 Large Size Coal Feed

The objective of the 703 test series was combustion of large size coal in the PDU and the comparison of its behavior and performance to that of typical $1/4"$ x 0 coal under equivalent conditions. Firing larger size coal can affect the carbon content of the bed, NO_x production, SO_2 capture, combustion efficiency and heat transfer. These factors are of prime importance in the design and operation of an FBC.

A total of five large size coal feed tests were run. Test Nos. 703-1 through 703-3 used $1/2"$ x 16 mesh coal feed. A combination of equipment and sampling problems encountered during Test Nos. 703-1 and 703-2 caused early termination of these tests. As the result of the first two tests, minor modifications for large size coal feed data and sampling collection were made. Thereafter, Test No. 703-3 was a successful 8 hour run, see Table 2.16 for operating conditions.

Anticipation of the next larger size coal feed, $3/4"$ x $1/2"$, led to modification of the feed system. Whereas the $1/2"$ x 16 mesh coal was fed through the normal $1-1/2"$ feed needles into the bed, it was felt that $3/4"$ x $1/2"$ feed would plug this line. Therefore, a 2 in. overbed feed line was arranged through the middle front door of the boiler so that the pneumatic transport could still be utilized. With this modification, Test Nos. 703-4 and 703-5 with $3/4"$ x $1/2"$ coal were run. As in the case of Test Nos. 703-1, 2 and 3, no problem was encountered in establishing and maintaining steady state conditions while combusting large coal. However, difficulty was encountered in both bed sampling and draining.

Examination of the bed after testing revealed that numerous large ash particles had caused clogging of the bed drain system. Modification of the bed sampling and drain systems would be required in order to fire larger coal in a test program. Test work was consequently shifted to other priorities.

2.3.7 Oil Firing

2.3.7.1 Equipment Modifications and Initial Tests

The PDU was modified to feed oil into the bed by piping the oil through the air distributor to nozzles mounted on the grid plate. A manifold was constructed, Figure 2.39, to pipe the oil into the plenum using four independent feed points. Each feed point is independently controlled so that the boiler can be run with all or any combination of nozzles. The first two attempts at oil feed involved the use of a simple orifice nozzle, Figure 2.40. The oil pump forced the oil through a 1/16 in. orifice drilled into a stainless steel pipe cap. This orifice directed a fine stream of oil parallel to the air distributor and perpendicular to the other nozzles, Figure 2.40. The oil was injected in this manner in an attempt to maximize distribution in the bed and to increase the surface area of the vapor bubble with respect to the main fluidization air. It is important to accomplish complete burning of the oil in the bed where the heat removal is located, otherwise there will be a considerable amount of freeboard burning and difficulty in maintaining bed temperature.

Two tests, Nos. 702-1 and 702-2, were run using #2 fuel oil and coal combination. Both these tests had normal light-off using the propane burners. Oil firing for both tests was not initiated until after the boiler was up to operating conditions on coal. For operating conditions, see Table 2.17. Once the boiler was up to temperature, oil was fed to the boiler until the excess oxygen level in the flue gas dropped off, then the coal feed rate was cut back to bring the oxygen level back to 3 percent. This process of increasing the oil feed rate and then decreasing the coal feed rate to maintain the oxygen level was tried for both tests.

The method of varying oil feed and coal feed to maintain flue gas oxygen was found to be unsatisfactory in maintaining bed temperature during the two tests. The causes of low bed temperature were threefold. First, the burning characteristics of #2 oil result in increased freeboard combustion. Secondly, some of the nozzles plugged with bed material reducing the fuel input. Finally, the oil flow control valve was found to be defective which prevented accurate feed rates to the PDU.

The two tests which were run were not completely successful in fulfilling the test objectives. Although the boiler could not be run solely on light oil pumped directly into the bed, sufficient information was gained to design the feed nozzles for the following tests. It was surmised that in order to get complete combustion of the light oil in the bed, air has to be mixed with the oil before the oil enters the bed. It is believed that the oil can be vaporized more rapidly in this manner and have a higher probability of burning as a vapor in the bed.

2.3.7.2 Tests With Modified Nozzle Design

Using the data acquired during Test Nos. 702-1 and 702-2, a new eductor type oil nozzle was designed, Figure 2.41. This nozzle mixes the oil with air before it enters the bed. Also the flow control valve was rebuilt and satisfactory flow control was established.

Three tests were run using these new oil feed nozzles. The first test, No. 702-3, was unsuccessful because of a failure in the oil feed air blower. The second test, No. 702-4, was successful, with the boiler being completely switched over to oil. This test was a preliminary run to find the optimum conditions to enable an eight hour test to calculate heat transfer and combustion efficiencies for a third test, Test No. 702-5.

During these tests the boiler was brought to operating conditions using coal and then oil feed was initiated. The stack oxygen level was used as an indicator when switching from coal to oil. This procedure, which was not successful with the first feed nozzles, was effective with the new nozzles. When the oxygen level dropped below 3 percent with the coal and oil feed, the coal feed rate was reduced to bring the oxygen level back to 3 percent. This method continued until the coal feed was completely cutoff. Operating conditions for Test No. 702-5 are shown in Table 2.18.

During Test No. 702-5, spent bed material, stored in a bin on the roof, was used to build the static bed height to 24 in. prior to oil feed. At this bed height, a considerable amount of oil was burning in the freeboard as seen by increasing freeboard temperatures. Since this heat was not released in the bed, bed temperatures dropped. To maintain bed temperature at 1500°F, more oil had to be added to the bed, resulting in a decrease in stack oxygen concentrations. Because of greater freeboard burning and the excess fuel added, the stack oxygen levels, which are usually maintained around 3 percent for coal, dropped to below 1 percent, see Table 2.18. As the static bed height was built to 30 in., it was found that the fraction of oil that burned in the bed increased. Therefore, the oil feed rate could be cut back

resulting in increased stack oxygen at the same bed temperature. A heat balance around the bed showed that approximately 40 percent of the oil was burning in the bed at a static height of 24 in. At 30 in. static, the fraction of inbed combustion increased to approximately 60 percent. Combustion efficiencies were calculated at both levels and are shown in Table 2.19. Within the accuracy of the measurements, the combustion efficiencies are the same.

One of the problems during this test was maintaining bed height. Because the #2 oil is low in sulfur, about .4 percent, a large fraction of the limestone is not sulfated and attrits more quickly than sulfated limestone. This prevented reuse of limestone to make up bed losses due to attrition. Therefore, during the test, spent bed material was used to maintain and build bed height.

2.3.7.3 Results of #2 Oil Tests

The flue gas compositions obtained during Test No. 702-5, see Table 2.20, are compared when the boiler was fired on coal, coal and oil, and oil only to values obtained during Test No. 5, 700-24 and 700-19. These two tests were long term tests in which the PDU was operated at steady state conditions for over ten hours.

The hydrocarbon values fluctuated during the oil portion of Test No. 702-5. This was probably due to the amount of freeboard burning and to fluctuations in the stack oxygen. The magnitude of the hydrocarbon values during Test No. 702-5 were not significantly different from values obtained during Test No. 700-24 at stack oxygen less than 3 percent.

Comparing the NO_x emissions, the values were lower when the boiler was running on oil. The values observed are very close to the expected thermal equilibrium values, 20 to 30 ppm at 1500°F, and 3% O_2 . This range of values is probably due to the nitrogen content of the fuel, about .5 percent. The SO_2 emissions were also low. This was expected because the #2 oil is a low sulfur fuel.

The heat transfer results for Test No. 702-5, Table 2.21, are compared to values from Test Nos. 700-17 and 700-19. Test No. 700-17 was a heat transfer test where the superficial velocity was varied. The overall heat transfer coefficient is larger during the oil test than for either coal test. This is probably due to the burning characteristics in the bed at the tube bundle surfaces. When the boiler is operated on oil, a greater amount of burning of the fuel occurs at the inbed bundles. This is because the oil vapors cannot back mix to the bottom of the bed as burning coal does. This may produce greater heat transfer in the region of the tube bundle because the burning oil

vapors completely surround the tubes. A greater surface contact of the burning oil would also explain the larger values for the outside coefficients, while the average wall temperature for the tubes remained constant with coal and or with oil. The inside coefficient which is a function only of the fluid in the tube, remained relatively unchanged during the oil test.

2.3.7.4 Light-off With #2 Fuel Oil, Test No. 702-6

Tests had shown that #2 fuel oil could maintain bed temperature after the PDU was brought to operating conditions with coal. The objective of Test No. 702-6 was to start up and come to operating conditions without coal by feeding #2 fuel oil.

Initially, a six to eight inch bed fluidized at 4 ft per second was brought to a steady state temperature of 650°F with an overbed propane burner. Holding this condition for a short period of time, a small oil flow was introduced through nozzles Nos. 1 and 2 located below the burner inlet, see Figure 2.39. With a two nozzle feed, the bed and freeboard temperatures steadied at 1050°F. At this point, nozzles Nos. 3 and 4 were started in sequence causing the bed temperature to rise to 1200°F. Simultaneously, oil flow to all four nozzles was increased while the bed was built to 15 in., then, the bed temperature was increased to 1450°F by decreasing the superficial velocity, increasing the oil feed rate and initiating air preheat. During this period the stack was observed to be clear with no sign of smoke.

With the bed glowing orange, the oil feed was maintained while the overbed burner was shutdown. Immediately, combustion of oil in the freeboard increased, as indicated by rising freeboard temperatures and above bed flashing and the bed temperature steadily declined. A considerable amount of smoke was also observed at the stack. Next the burner was again turned on resulting in the restoration of bed temperature and a reduction of freeboard burning. A second attempt to shutdown the overbed burner was made when the bed temperature reached 1500°F with the same results as before.

From previous oil firing tests, it was recognized that with #2 fuel oil, a deeper bed was needed to increase the fraction of the fuel burned in the bed in order to maintain temperature. Thus, the following contradictory effects were noted. A deeper bed was needed to maintain combustion, yet increasing the bed depth any further would extinguish the propane burner, a situation already shown to cause the bed temperature to decline. At this point, the test was terminated. The overbed burner used to light-off the PDU was not designed to operate submerged in the bed, hence deep light-off techniques could not be attempted.

In conclusion, it appears that start-up and operation with #2 fuel oil can be readily accomplished provided the light-off burners work at maximum bed depth.

2.3.7.5 Preliminary #6 Fuel Oil, Test Nos. 702-7 and 702-8

Two preliminary tests, Nos. 702-7 and 702-8 were run using #6 fuel oil in combination with coal. These two tests were run to determine if any changes were required to the oil nozzle design and to characterize boiler operation on heavier fuel oil. Both tests had normal light-off using the propane burners and coal. The fuel oil was fed to the PDU when the boiler was at operating conditions on coal. Oil feed was controlled using the method developed for the #2 fuel oil. Stack oxygen level was monitored and used as an indicator to vary the oil feed.

Test No. 702-8 was a four hour test using #6 oil. The operating conditions for this test are shown in Table 2.22 where oil feed was initiated at 1130 hours.

The combustion efficiency for the test was 98 percent. Using this efficiency in a heat balance program, it was calculated that only 64 percent of the fuel was burning in the bed. This is comparable to values calculated for #2 oil with the same static bed heights and is probably due to insufficient mixing of the oil in the bed. Because of the amount of freeboard burning, an excess of fuel oil had to be fed to the boiler to maintain bed temperatures at 1550°F. Therefore, the stack oxygen levels were very low, see Table 2.22.

Heat transfer results from Test No. 702-8 were calculated, see Table 2.23. They are compared to Test No. 702-5, a #2 fuel oil test and Test No. 700-17, a coal test to determine the effects of superficial velocity on heat transfer; and Test No. 700-19, a sixteen hour coal run for sulfur capture. From the results, it was found that the overall heat transfer coefficients are larger when the boiler is firing oil. This may be due to the amount of oil burning in the tube bundle region; the calculated values are lower for #6 oil than for #2 oil; the difference is within the accuracy of the measurements.

One of the most significant demonstrations during these preliminary tests was a rapid switch to oil firing during a coal feed interruption. During Test No. 702-8 a coal feed problem was simulated; immediately oil feed was initiated and the bed temperature was maintained.

2.3.7.6 Test With The Modified Oil Air Feed System, Test No. 702-9

The major problem encountered during the preliminary tests was the amount of overbed burning. Prior to Test No. 702-9, the oil feed air piping was replaced to permit an increase in air flow to the oil nozzles in an attempt to reduce overbed burning. During Test No. 702,9, an eleven hour run with #6 oil, combustion efficiency and sulfur capture were measured with this improvement.

An immediate result of the improved oil-air feed system was that 1550°F bed temperature could be easily maintained with 4 percent oxygen in the stack. A visual inspection of the surface of the bed through the viewport showed no above bed burning. For operating conditions, see Table 2.24. With increased combustion in the bed, the heat transfer coefficients, shown in Table 2.25, are typical of the results expected when firing coal. At the end of the run, the PDU was switched from fuel oil to coal and the heat transfer coefficient remained essentially the same.

The overall heat transfer coefficient in Test No. 702-9 was slightly lower than in Test Nos. 702-5 and 702-8. During these earlier oil tests, the amount of oil burning at the upper portion of the expanded bed was high as evidenced by the flue gas analysis which showed higher levels of hydrocarbons and lower values of stack oxygen. It is suspected that the magnitude of the outside coefficient decreased somewhat during Test No. 702-9 because of more uniform burning throughout the bed compared to earlier tests.

Test No. 702-9 was the longest (11 hours) and the most efficient oil firing test and sufficient data was collected to measure combustion efficiency and sulfur capture. Using the measured 98% combustion efficiency, a heat balance around the bed indicated that about 95% of the fuel input was burned in the bed. This high rate of inbed burning was confirmed by visual observation during the test since very little above bed burning could be seen.

Sulfur capture was found to be 99%. This high rate of sulfur capture was apparent from the extremely low SO₂ emission levels during the test. Although an analyzer malfunction was suspected, every effort to find a problem confirmed the gas analysis. As a final confirmation of the flue gas analysis, coal firing was resumed at the end of the oil firing test to validate the flue gas readings.

Due to the low SO₂ emission levels, the fuel oil was sent to two independent laboratories for analysis. The results of these analyses are shown in Table 2.26. As can be seen, the average fuel oil sulfur content was 2.35% and the vanadium content was significant. For this reason, the

bed material was also analyzed for vanadium and the result is also shown in Table 2.26. Since the fuel oil and bed material contained a significant amount of vanadium, it is possible that the high sulfur capture rate was due to the influence of vanadium on the sulfur capture reaction.

Another interesting observation from the test was that when coal firing was resumed at the end of the test, the SO_2 concentration in the stack increased from 10 ppm to only about 330 ppm after about an hour of coal feed with a reduced Ca/S feed ratio. Since the SO_2 emission did not immediately increase on the switch to coal firing, and since the level of emission was lower than expected, activation of the limestone by the oil was suspected.

To further examine the unusually high sulfur capture, scanning electron micrographs (SEM) of samples of sulfated bed material were made, to examine physical changes in the material. Due to the time delay between sampling and the SEM work, some decomposition of the bed material obscured these changes. Figures 2.42 and 2.43 show some of the electron micrographs. Microcrystals of calcium carbonate show up as characteristic plates on the surface of the material and are well defined in all of these illustrations. The pictures do not show any smooth round surfaces that would be indicative of a "salt effect". They do not appear different from micrographs of bed material formed by burning coal. Although activation of the limestone particles by an element in the fuel is a possible explanation for the high sulfur capture, it was not evident from the micrographs.

2.3.8 PDU Performance Characterization

Characterization of the PDU continued with emphasis on the chemical analysis of fly ash and bed material as a function of size. By following changes in these components with boiler operating conditions, a better understanding of process mechanisms will result. This type of data can be useful in future AFBC plant designs where the sizing and selection of stack gas cleaning units and fly ash reinjection systems will depend on factors such as the ones studied here.

Normally the chemical analyses performed on solid samples from FBC tests are elemental only. However, these results do not reveal how the elements are distributed among the various components. For instance, an elemental analysis of bed material may reveal its total calcium content, but does not reveal how much calcium is in the form of CaCO_3 versus CaO or CaSO_4 . The method used to analyze test data was the simultaneous component balance technique previously presented in detail in Interim Report No. 3.

In this method, a simple elemental mass balance results in a system of linear equations which, when solved, yield a prediction of the component composition of desired process streams. The data required for this analysis is the elemental chemical analyses of the coal, its resulting coke and ash, as well as the limestone and the stream of interest.

This technique was applied to the results of various fly ash and bed material streams from Test Nos. 700-19, 700-24, 700-38 and 700-39.

2.3.8.1 Test Descriptions

During the sixteen hour testing period of Test No. 700-19, samples of the coal and limestone feed streams were taken every four hours. These samples were composited and analyzed for size distribution, carbon, hydrogen, sulfur, calcium, ash and Loss on Ignition (LOI). A sample of the coal feed was coked and ashed with both materials subject to the same analysis as above. Table 2.27 shows the summary of the operating conditions for Test No. 700-19. Near the end of the test period, after steady state conditions had been maintained for ten hours, two samples of fly ash, bed material and dust were taken one hour apart at 1430 and 1530 hrs. These samples were sized and analyses similar to those performed on the composited limestone and coal were completed on each individual size cut. The amount of dust collected in the three size cuts +200, -200+400, and -400 mesh was added to the similar size cuts of the fly ash and combined size distributions and chemical analyses reported.

Test No. 700-24 was run with nearly identical conditions as Test No. 700-19. Again samples were taken after ten hours of steady state run time and subjected to the same processing as those for Test No. 700-19. A summary of the operating conditions is given in Table 2.28.

Test Nos. 700-38 and 700-39 were limestone evaluation tests run with Lowellville limestone. As part of the elutriation evaluation procedure, the fly ash was examined using the simultaneous component balance. A summary of test conditions for Test Nos. 700-38 and 700-29 are given in Tables 2.29 and 2.30. Again, sample collection and processing were the same as Test Nos. 700-19 and 700-24.

2.3.8.2 Fly Ash Characterization

Analysis of the fly ash from PDU Test Nos. 700-19, 700-24, 700-38, and 700-39 was conducted to characterize the fly ash from the PDU. Comparatively, each test was run at near the same conditions, as seen in Tables 2.27 through 2.30. The only difference was that Test Nos. 700-38 and 700-39 used Lowellville limestone as the sorbent whereas

Test Nos. 700-19 and 700-24 used Greer limestone. The first comparison examined was the size distribution of the fly ash for each test which is summarized tabularly in Table 2.31 and graphically in Figure 2.44. As can be seen, Test Nos. 700-24, 700-38, and 700-39 gave similar distributions. Very distinctively, the major portion of the fly ash, 30% or more by weight repeatedly appeared in the -40+80 mesh size cut (300 μ) with some small variations in the sizes above and below this cut. Atypical of these results was Test No. 700-19 which showed two, nearly equal peaks occurring at -40+80 mesh (300 μ) and -140+200 mesh (90 μ). An obvious explanation for this difference was the sampling technique employed. During Test No. 700-19, two hand operated ball valves were used to divert the flow of the fly ash from the transport line to a sample bag. Due to the geometry of the sampling arrangement, fly ash often packed the sample line causing sampling difficulties. In some cases, samples were taken before entirely free sample flow had resumed which could have biased the material sampled. Prior to Test No. 700-24, and for all tests following, a diverter valve was installed in the fly ash transport line. With this improved sampling arrangement, all of the fly ash was diverted to the sample bag, with no pluggage of the line, giving a more representative sample. Therefore, fly ash results from Test No. 700-19 are not to be considered with those of Test Nos. 700-24, 700-38 and 700-39.

Establishing that a large portion of the fly ash occurs in the 300 μ range, explanations were inconclusively sought. Feed stream size distributions apparently had no effect. Even though the average limestone feed size was 2000 μ $\pm$ 100 μ for each test and the average coal feed varied from 876 μ in Test No. 700-24 to 1379 μ in Test 700-39 to 1510 μ in Test No. 700-38, the fly ash size distribution remained nearly the same. A review of operating conditions for these tests, Table 2.28 - 2.30, reveals no significant differences. Further investigations are planned to clarify mechanisms responsible for this PDU operational characteristic.

The second aspect of fly ash characterization examined, was the carbon content of the fly ash as a function of size. Since the fly ash rate varied from test to test, a basis of 100 grams of fly ash sampled was used to compare results. Figure 2.45 shows a plot of the grams of carbon per 100 grams of fly ash for the various size cuts. Test No. 700-19 has been omitted for reasons explained above. Again, the 300 μ peak is evident along with weak, infrequent peaks in the lower sizes. Not only does the -40+80 mesh size range represent the largest amount of fly ash leaving the PDU, but also the largest amount of carbon in the fly ash leaving the boiler. Theoretically, by reducing the grams of carbon/100g fly ash of this size cut by a factor of two, everything else remaining the same, the

combustion efficiency could rise 3-5%. For example, Test No. 700-39 with 10 grams carbon/100 grams fly ash, had an 84% combustion efficiency while Test No. 700-38 with 16 grams carbon/100 grams fly ash had an 82% efficiency.

Various combustion theories suggest that, in order to decrease the amount of carbon in a size cut, conditions must be such that the time for burnout of the particles must be brought closer to its residence time. Theoretically, this may be done for all size fractions by increasing both temperature and %O₂ in the freeboard. Test plans for the upcoming period include such runs.

The fly ash from Test Nos. 700-38 and 700-39 was analyzed as to the percent of limestone, calcined limestone, sulfated limestone, coal derived ash, and coal plus coke using the simultaneous component balance technique. The results of this calculation are shown in Figures 2.46 and 2.47 for Test Nos. 700-38 and 700-39 respectively. The height of each bar in the graph shows the weight percent of fly ash contained in each screen cut. Each bar is divided into smaller blocks to show the distribution of the various components within each screen cut. These figures show that for particles larger than 80 mesh (177 μ) larger amounts of fly ash are limestone derived while for particles under 80 mesh, the fly ash is largely coal derived. Here the predominance of the -40+80 size fraction in both mass and carbon contributions to the fly ash are more readily seen. The greater amounts of limestone derived material seen in Test No. 700-39 at 0900 hrs. versus 1300 hrs. of Test No. 700-38 may be explained by the fact that Test No. 700-39 had a higher limestone attrition than Test No. 700-38 as seen by the composite calcium analyses of the fly ash in both tests. Test No. 700-38 had a composite value of 10.73 percent calcium whereas Test No. 700-39 had a value of 20.48 percent, nearly twice as much.

2.3.8.3 Bed Material Characterization

The PDU bed material was also characterized using the same test periods as the fly ash. The bed material size distributions are shown tabularly in Table 2.32 and graphically in Figure 2.48. As can be seen, the bed material like the fly ash, has a prominent single peak in the size distribution. This single peak repeatedly appeared in the 8 mesh (2380 μ) range in both the Greer and Lowellville limestone tests.

The bed material was evaluated in terms of grams of sulfur per 100 grams of bed as a function of particle size. Chemical analyses for Test Nos. 700-19 and 700-24 were done on the entire range of nine sieve cuts while those for Test Nos. 700-38 and 700-39 were done on only the middle five cuts.

For ease of comparison, all figures are shown in terms of the middle five cuts, see Figure 2.49. As indicated, the -8+16 mesh (1785 μ) contained the largest fraction of total sulfur in the bed and is near the 8 mesh (2380 μ) range which was the predominant size fraction. However, more informative than Figure 2.49 for process characterization purposes, is the table of % sulfation, Table 2.33. This table shows the amount of Ca as CaSO_4 compared to the total Ca in the bed fraction i.e. Ca as CaSO_4 , CaO and CaCO_3 . Theoretically, if all the Ca in the feed were to capture sulfur to saturation, the resulting CaSO_4 material would have a sulfur to calcium mole ratio of 1.0. Since the sample fraction in the +1/4 and -30 size cuts was small, the possible error in the Ca analysis which involves a digestion and titration technique, could be large. Therefore, only the middle three cuts are shown. It is seen that as the particles grow smaller, the % sulfation or calcium utilization increases. Thus, even though more total sulfur was found in the -8+16 mesh cut, the best use of available calcium is in the -16+30 mesh cut. Reasons for difference in % sulfation values between the Greer and Lowellville limestone tests are speculated to be due to two factors. First, Greer Test Nos. 700-19 and 700-24 were run at calcium to sulfur molar feed ratios of approximately 2.0 while Lowellville Test Nos. 700-38 and 700-39 were run at values of 4.0 or more. Secondly, the average sulfur content of the coal for the Greer tests was 5 percent or more while for the Lowellville tests was 2.5 percent.

Finally, the simultaneous component balance for bed material of Test Nos. 700-19 and 700-39 are shown in Figures 2.50 and 2.51 respectively. Note the small amount of coal and coke found in the bed, usually less than one percent. Note also the amount of CaCO_3 and CaO available in the bed, not yet utilized. The larger amounts of CaSO_4 in Figure 2.50 versus Figure 2.51 are consistent with the % sulfation values of Table 2.33 and grams of sulfur per 100 grams of bed material peaks of Figure 2.49.

2.3.9 Limestone Evaluation Tests

Three sorbents were tested in the PDU: Greer limestone, Lowellville limestone, and Grove limestone. All the stones tested were used in larger AFBC applications. Greer and Lowellville limestones have been used at the Rivesville, West Virginia 30 MW_e MFB while Grove limestone has been used at the 100,000 lb/hr AFBB at Georgetown University. The purpose of the tests at the PDU laboratory was to provide baseline sulfur capture performance data and to anticipate any problems that might arise from using these stones in larger units.

Two criteria of the limestones were evaluated, sulfur capture and attrition. Sulfur capture performance was obtained during long tests that were usually conducted for other purposes (such as baghouse tests and PDU characterization tests). Attrition, while not necessarily detrimental to sulfur capture, could effect other operating conditions such as maintaining constant bed level. Specific tests were conducted to evaluate attrition and transient behavior.

2.3.9.1 Limestone Analyses

Table 2.34 shows the chemical compositions of the three limestones tested. Of the three stones, Lowellville has the highest CaCO_3 content (98.63%) while Greer has the lowest (87.33% CaCO_3). The size distributions of the three stones are plotted in Figure 2.52. The Lowellville stone showed the widest variation in size distribution. Although the larger size cuts are similar for the Lowellville stone, there was a marked difference in the minus 12 mesh material. The Grove samples showed more consistency in size. Only one size analysis was available for the Greer stone.

2.3.9.2 Steady State Sulfur Capture Tests

Table 2.35 lists the operating conditions obtained during tests where sulfur capture performance was evaluated. Generally more than 5 hours of steady operation was obtained prior to initial data taking although this was not always possible due to difficulties in coal and limestone feeding. The Lowellville tests were particularly troublesome due to the variations in fines. This sometimes resulted in erratic feed rates.

The Grove limestone tests were conducted to simulate the Georgetown plant operation, including the use of fly ash recycle. In Test No. 704-11, all of the fly ash caught in the cyclone was recycled to the PDU. In Test No. 704-12, the fly ash was split by means of an overflow system and some collected fly ash was not reinjected into the boiler. Table 2.35 shows the chemical analysis of the coal used during the tests shown in Table 2.35.

Figure 2.53 shows a plot of the sulfur capture as a function of Ca/S mole feed ratios. Included in Figure 2.53 are the points obtained on Greer limestone in the FBM (see Interim Report No. 5). Sulfur capture of the Grove limestone appears to be better than the Greer or Lowellville. However, the Grove limestone tests included fly ash recycle and this is known to significantly improve sulfur capture.

A comparison of the Greer and the Lowellville tests shows that while both sorbents perform nearly the same at Ca/S ratios near 2.0, the Greer limestone showed better sulfur

capture at higher Ca/S ratios. To achieve 85% sulfur capture, about 16% less Greer limestone would be required than Lowellville limestone. The scatter in the Grove limestone data, particularly at the high Ca/S ratios, could be due to either fluctuations in the amount of limestone fines or to fluctuations in the amount of coal fines. Another possible explanation could be that Test No. 700-38 (~82% sulfur capture at Ca/S ~5) was conducted at 1500-1550°F while Test No. 700-39 (~94% sulfur capture at Ca/S ~5) was conducted at 1440°F. This would suggest that the Lowellville limestone is more reactive at the lower temperature, although there is no other data to substantiate this possibility.

2.3.9.3 Transient Sulfur Capture Tests

Two tests were conducted to observe transient sulfur capture during start-up with a fresh limestone bed. Test No. 700-28A (shown in Figure 2.54) was conducted at an average molar Ca/S ratio of 1.5 after the first two hours, at an average superficial velocity of 11.8 ft/second. During the first two hours, the bed level was built with fresh limestone from 10 to 24 in. (static). Test No. 700-30 (shown in Figure 2.55) was conducted at an average molar feed Ca/S ratio of 1.9 at an average superficial velocity of 10.5 ft/second. During this test the bed level was not built as rapidly during the initial time periods.

In both tests, the SO₂ concentration in the stack was in the 100 to 200 ppm range shortly after light off. This is significant since during this time the bed is almost entirely fresh calcining limestone and should be at its most reactive state. This suggests that there is some upper limit of sulfur capture for the PDU that is somewhat below 100%. The 100 ppm SO₂ for these tests corresponds to about 96 or 97% sulfur capture. Freeboard combustion or bed geometry is probably responsible for the final 100 ppm and thus limits the sulfur capture to this value.

2.3.9.4 Attrition

The attrition characteristics of limestone are an important consideration in sorbent selection. In most FBC applications the limestone serves two functions: to remove SO₂ from the flue gas, and to remain in the combustion zone as the major constituent of the bed material. There are some instances where a limestone can show good sulfur capture potential but, due to a high rate of attrition, may require excessive feed rates to maintain bed inventory. The attrition rate of the limestone is dependent upon the rate at which the stone sulfates. This is because fully sulfated bed material is less susceptible to attrition than calcined limestone. Hence the sulfur of the coal is an important factor.

One measure of limestone attrition is the calcium fraction in the fly ash. Since coal has little calcium, most of the calcium in the fly ash will come from the limestone. Figure 2.56 shows the rate of calcium in lbs/hr leaving the PDU in the fly ash for Greer and Lowellville limestones. Since fly ash recycle was employed during the Grove limestone tests, these tests will be discussed separately. The line drawn in Figure 2.56 is a least squares fit of the Greer data. In spite of the fact that the fly ash collection rates were nearly the same for the two tests, the Lowellville stone seems to have attrited faster than the Greer.

A comparison of test results using Grove limestone was made to determine the effect of coal sulfur content on the rate at which the bed depth can be increased. It has been observed that highly sulfated limestone is more resistant to abrasion than fresh limestone. Thus there was some concern that use of a low sulfur coal could result in a situation where limestone feed rates in excess of those required for proper sulfur capture would be required to maintain proper bed depth. Furthermore, it was feared that when low sulfur coal is being burned, the bed level could not be increased with fresh limestone due to excessive attrition. For these reasons two tests, Nos. 704-5 and 704-9, were conducted to examine the bed building rates when using both low and high sulfur coals.

One objective of Test No. 704-5 was to determine the bed building rate for Grove limestone when using a one percent sulfur coal. The operating variables for the first eight hours of Test No. 704-5 are shown in Figure 2.57. Average bed temperature during the bed building portion of the test was 1540°F. During the first five hours of the test, the limestone feed rate was rather erratic. However, the area under the limestone feed rate curve shows that from the first to the fifth hour, 1950 pounds of limestone were fed to the PDU and the bed level increased 13.5 in. Thus, 144 pounds of limestone was required to build one inch of bed.

Test No. 704-9 was conducted to examine the bed building rate of Grove limestone using a three percent sulfur coal. The operating conditions for this test are shown in Table 2.37. The operating variables for the eight hours of operation are shown in Figure 2.58. The average bed temperature during this time was 1580°F. Again the area under the limestone curve shows that 1260 pounds of limestone was added to the PDU between the first and eighth hours of the test. The bed level increased 15 in. during that time. Thus only 84 pounds of limestone was needed to build one inch of bed. Thus it would appear that the use of high sulfur coal results in less limestone being needed to raise the bed level.

The major difference between these two tests that could affect the results was the average limestone feed rate during the bed building portion of the test. During the bed building portion of Test No. 704-5, the average limestone feed rate was 487 lb/hr while during Test No. 704-11 the average limestone feed rate was only 180 lb/hr. In order to check on the effect of the feed rate, data from Test No. 704-11 was examined.

Figure 2.59 shows the limestone feed rate and the static bed depth during the first eight hours of Test No. 704-11. This test was conducted using the same three percent sulfur coal as Test No. 704-9. During the first three hours of this test, 1080 pounds of limestone was added to the PDU and the bed level increased 15 in. The average bed temperature during this time was 1620°F and the average limestone feed rate was 432 lb/hr. This means that 72 pounds of limestone was needed to build one inch of bed which agrees with the 84 pounds of limestone needed to build one inch in Test No. 704-9. Hence the limestone feed rate did not significantly alter the comparison of results from these tests.

It appears that the effect of increased sulfur in the coal is to require less limestone to build the bed as expected. Test No. 704-5 showed that even with a low sulfur coal, Grove limestone could be used to build and maintain bed level. The combination of Grove limestone and low sulfur coal appears to be feasible from an operations standpoint.

2.3.10 Minimum Fluidization

Recent trends in FBC technology and cold light-off difficulties have led to reexamination of methods for calculating minimum fluidization velocity. Specifically, the use of large top size coal with overbed stoker feeding and the trend toward lower design superficial velocities to improve sulfur capture will result in operation near the minimum fluidization velocity of the bed. Also, cold light-off conditions set the design volume requirements for fluidizing air since mixing by fluidization of the bed is required. Occasionally, due to design limitations, change in limestone supply, or mixing of large coal particles in the bed, the bed mean particle size may become too large to be fluidized under start-up conditions. This situation has occurred in the Rivesville MFB and illustrates the need for accurate prediction of the minimum fluidization velocity of the bed.

The equation generally used to predict the minimum fluidization velocity is a modified form of the Ergun fluidization equation. In simplified form, the Ergun equation can be written as:

$$U_{mf} = \frac{\mu}{\rho_g D_p} \left\{ A^2 + B (G_A)^{\frac{1}{2}} \right\} (-A) \quad (1)$$

where: U_{mf} = minimum fluidization velocity
 μ = gas viscosity
 ρ_g = gas density
 D_p = particle diameter
 G_A = Galileo number = $\frac{D_p \rho_g (\rho_s - \rho_g) g}{\mu^2}$
 g = gravitational constant
 ρ_s = solid density
 $A = \frac{42.86 (1 - \epsilon_{mf})}{\phi_s}$
 $B = \frac{\phi_s \epsilon_{mf}^3}{1.75}$
 ϵ_{mf} = voidage at minimum fluidization
 ϕ_s = shape factor

Wen and Yu¹ have suggested using an approximation for ϕ_s and ϵ_{mf} such that:

$$\frac{1}{\phi_s \epsilon_{mf}^3} \approx 14 \text{ and } \frac{1 - \epsilon_{mf}}{\phi_s^2 \epsilon_{mf}^3} \approx 11 \quad (2)$$

This gives values of A and B in equation (1) of 33.7 and .0408 respectively. The Institute of Gas Technology², in developing a minimum fluidization velocity correlation for material normally found in an FBC, recently correlated available data to develop the following empirical values:

$$\frac{1}{\phi_s \epsilon_{mf}^3} = 8.81 \text{ and } \frac{1 - \epsilon_{mf}}{\phi_s^2 \epsilon_{mf}^3} = 5.19 \quad (3)$$

This gives values for the quantities A and B of 25.25 and 0.0649 respectively.

¹Wen, C.Y. and Yu, Y.H., AIChE Journal 12, 610 (1966).

²Institute of Gas Technology Coal Conversion System
 Technical Data Book 1978. 2.41

Figure 2.60 shows a plot of the calculated minimum fluidization velocities of limestone as a function of particle size at 80°F and at 1550°F for both above correlations. The curves show a rather large difference in the predicted minimum fluidization velocities. In previous cold minimum fluidization tests, the PDU and FBM yielded results in agreement with those predicted by the correlation of Wen and Yu. However, due to operational difficulties, the minimum fluidization velocity could not be measured under hot conditions. To test the correlations under hot conditions, a four inch diameter, propane fueled, hot cell was used to obtain the minimum fluidization velocity at 1550°F from pressure drop across the bed data. The tests were also run at 80°F in the same cell.

Two series of tests were run in the four inch hot cell using Lowellville limestone in the first and Greer limestone in the second. In each series, the minimum fluidization velocity was measured first at 80°F and then after one hour of operation at 1550°F. After each minimum fluidization velocity measurement, a sample was withdrawn for density and mean particle size determinations. Three trials were conducted in each series using fresh limestone for each trial. Minimum fluidization was determined in a conventional manner by plotting bed pressure drop versus superficial velocity. The results of the tests are shown in Table 2.38.

Table 2.39 shows a comparison of measured minimum fluidization velocities to the values predicted by the two correlations. In calculating the predicted values, the measured mean particle sizes and particle densities were used. In this way, effects due to calcination of the limestone were considered. The results in the table show that the prediction of Wen and Yu, Equation (2), gave better results under the cold conditions. This is consistent with earlier measurements made in the PDU and the FBM under cold conditions. However, at 1550°F, the IGT correlation, Equation (3), gave better results.

Solution of Equation (2) for ϕ_s and ϵ_{mf} gives values of $\phi_s = .57$ and $\epsilon_{mf} = .50$. The IGT correlation gives values of $\phi_s = .82$ and $\epsilon_{mf} = .52$. The major difference in the two correlations lies in the choice of shape factors. The results of these tests seem to indicate that the shape factor changes between cold and hot tests.

The shape factor is a measure of the roundness of the particle. Perfect spheres have a shape factor equal to one. The lower the value of ϕ , the more irregular the shape. Since the calcined limestone is softer than raw limestone, abrasion of the calcined limestone particles could result in smoother particles thereby changing their shape factor.

The results of these tests show that correlations should be developed and verified for the particular bed material in an FBC system.

2.4 Process Analysis Results

2.4.1 Inbed Heat Balance

The operating parameters of a fluidized-bed boiler are related to one another by the thermodynamic equilibrium between the fluidized bed and its surroundings. A change in an operating variable that affects the thermodynamic equilibrium will result in changes to other operating conditions in order to achieve a new equilibrium state. For example, an increase in the coal feed rate will result in an increase in the total heat input to the bed. This will force the bed temperature to rise until the heat output from the bed equals the new heat input. Furthermore, the increased coal feed results in more oxygen consumption, thereby lowering the percentage of oxygen in the flue gas.

A simple heat balance around a fluidized bed can be used to estimate many of the bed operating variables. The purpose of this effort was to develop a general, simplified heat balance equation that could be used to predict operating conditions in a fluidized-bed boiler. Figure 2.61 shows the general layout of a fluidized-bed boiler and the boundary for the heat balances. In developing this heat balance equation, several key variables have been identified that generally must be estimated based on experience. These variables will be discussed as the equations are developed.

(1) Heat Inputs

The heat inputs to the fluidized bed are the chemical heat released by the burning fuel and the sensible heat inputs from the various feed streams. The chemical heat released by the burning fuel is only the amount of heat released in the bed itself. Generally when combustion efficiency (E) is measured, it includes any combustion that may have taken place in the freeboard as well as in the bed. In order to use a measured combustion efficiency in the heat balance equation, the combustion efficiency must be multiplied by the fraction of combustion that takes place in the bed (X). Therefore, the chemical heat released by the burning fuel is given by:

$$Q_F = \Delta H_F W_F EX \quad (1)$$

where ΔH_F = as fired heat of combustion of fuel,
Btu/lb

W_F = fuel feed rate, lb/hr

E = overall unit combustion efficiency

X = fraction of inbed burning

Equation (1) will be the same for the case of fly ash reinjection to a coal fired bed since the term E is the total measured combustion efficiency. The sensible heat inputs, while small compared to the chemical heat release, should also be included in the heat balance equation. The sensible heat inputs are all referenced to 80°F in this calculation and are given by:

$$\text{Fuel} \quad Q_{FS} = W_F C_{pF} (T_S - 80) \quad (2)$$

$$\text{Limestone} \quad Q_{LS} = W_L C_{pL} (T_S - 80) \quad (3)$$

$$\text{Air (main combustion)} \quad Q_A = W_A C_{pA} (T_A - 80) \quad (4)$$

$$\text{Transport Air (coal and/or fly ash)} \quad Q_{TA} = W_{TA} C_{pA} (T_{TA} - 80) \quad (5)$$

$$\text{Reinjected Fly Ash} \quad Q_{FAR} = W_{FAR} C_{pFA} (T_{FAR} - 80) \quad (6)$$

where

C_{pF} = heat capacity of the fuel, Btu/lb, °F

T_S = temperature of incoming solids, °F

W_L = limestone feed rate, lb/hr

C_{pL} = heat capacity of limestone, Btu/lb, °F

W_A = main air feed rate, lb/hr

C_{pA} = heat capacity of air, Btu/lb, °F

T_A = main air temperature, °F

W_{TA} = transport air feed rate, lb/hr
(only if injected into the bed)

T_{TA} = transport air temperature, °F

W_{FAR} = fly ash reinjection rate, lb/hr

C_{pFA} = heat capacity of fly ash, Btu/lb, °F

T_{FAR} = temperature of fly ash (solids temperature prior to mixing with transport air), °F

The total heat input equation is given by:

$$Q_{in} = Q_F + Q_{FS} + Q_{LS} + Q_A + Q_{TA} + Q_{FAR} \quad (7)$$

These heat inputs will equal the sum of the heat outputs from the bed when the bed is at equilibrium.

(2) Heat Outputs

The two major heat outputs from a fluidized-bed boiler are the heat removed by the flue gas and the heat removed by the generation of steam. Other heat outputs are the solids sensible heat losses, calcination and sulfation heat losses and radiation heat losses.

The heat removed by the flue gas is given by:

$$Q_{Fg} = W_{Fg} H_{Fg} \quad (8)$$

where W_{Fg} = flue gas flow rate, lb/hr

H_{Fg} = flue gas enthalpy, Btu/lb

The flue gas enthalpy is a function of flue gas moisture and temperature. It should be noted that the temperature of the flue gas as it leaves the bed is equal to the bed temperature. An empirical curve fit to the flue gas enthalpy data presented on Pages 6-8 and 6-9 in the Babcock and Wilcox book "Steam: Its Generation and Use" is given by:

$$H_{Fg} = (0.00245m + .288) (T_B - 195) \quad (9)$$

where m = weight % moisture in flue gas

T_B = bed temperature, °F

Equation (9) is good for the temperature range of 1000 to 2000°F. This enthalpy is referenced to 80°F.

The heat transferred to the steam is only the heat removed by submerged or wetted heat transfer surfaces. Any tube surface in the freeboard will cool down the flue gas but does not affect the inbed heat balance. This is also true for any portion of an inbed tube bundle that is not completely submerged due to a shallow bed. For the case of a saturated steam boiling bundle, the heat transferred to steam is given by:

$$Q_S = U_B A (T_B - T_S) + U_W A_W (T_B - T_S) \quad (10)$$

where U_B = horizontal tube bundle inbed heat transfer coefficient, Btu/hr, ft², °F.
 A = wetted surface area of the tube bundle, ft².
 T_S = temperature of the saturated steam, °F.
 U_W = waterwall inbed heat transfer coefficient, Btu/hr, ft², °F.
 A_W = wetted surface area of the waterwall, ft².

In the case of an inbed superheater bundle, the quantity $(T_B - T_S)$ in the first part of equation (10) must be replaced by a log mean temperature difference (T_{LM}) between the superheated steam and the bed. The quantities A and A_W can be related to the expanded bed depth (H) and fixed bed geometrics as follows:

$$A = 0 \text{ if } H < H_B \quad (11A)$$

$$A = A_T \left\{ \frac{H - H_B}{H_T - H_B} \right\} \text{ if } H_B \leq H \leq H_T \quad (11B)$$

$$A = A_T \text{ if } H > H_T \quad (11C)$$

and $A_W = PH \quad (12)$

where A_T = total tube bundle area, ft²

H_B = distance from grid plate to bottom of tube bundle, ft.

H_T = distance from grid plate to top of tube bundle, ft.

P = bed perimeter, ft.

The solids sensible heat losses include the sensible heat from the fly ash and the sensible heat from the bed drain. Note that the chemical heat from the fly ash and the bed drain are not included in the heat balance directly. These have already been accounted for in Equation (1) by the combustion efficiency term. The solids sensible heat losses referenced to 80°F are given by:

$$Q_{SS} = W_{FA} C_{pFA} (T_B - 80) + W_B C_p (T_B - 80) \quad (13)$$

where W_{FA} = fly ash rate, lb/hr.
 Cp_{FA} = heat capacity of fly ash, Btu/lb, °F.
 W_B = bed drain rate, lb/hr.
 Cp_B = heat capacity of bed material, Btu/lb, °F.

When limestone calcines, it absorbs heat from the bed. The sulfation reaction releases heat to the bed. The heat removed from the bed will be the difference between the calcination reactions and the sulfation reaction given by:

$$Q_{C-S} = 1964.7W_L f_{Ca} + 1762.7W_L f_{Mg} - 6336.8W_F f_S S \quad (14)$$

where f_{Ca} = weight fraction of calcium in limestone.
 f_{Mg} = weight fraction of magnesium in limestone.
 f_S = weight fraction of sulfur in coal.
 S = fraction of sulfur capture.

Heat radiates from the top of the fluidized bed and is absorbed by surfaces in the freeboard. Since these surfaces see the entire top of the bed, the view factor is assumed equal to 1.0. The emissivity of the bed is also assumed equal to 1.0, thus the radiative heat loss is given by:

$$Q_R = \sigma A_S (T_{BR}^4 - T_{SR}^4) \quad (15)$$

where σ = $.1714 \times 10^{-8}$ Btu/hr, ft², °R⁴
 A_S = cross-sectional area of the bed.
 T_{BR} = bed temperature, °R.
 T_{SR} = steam temperature, °R.

Note also in equation (15) that the exposed metal surfaces are assumed to be at steam temperatures. Generally most freeboard surfaces are either natural convection waterwalls or forced convection boiling bundles.

The total heat output equation will be given by:

$$Q_{OUT} = Q_{Fg} + Q_S + Q_{SS} + Q_{C-S} + Q_R \quad (16)$$

The heat balance condition is given by:

$$Q_{IN} - Q_{OUT} = 0 \quad (17)$$

The solution of equations (1) through (17) can be used to predict the operating conditions in a fluidized-bed boiler.

(3) Application of the Heat Balance Equation

The primary operating variables for a fluidized-bed boiler are air feed rate (or superficial velocity), fuel feed rate and bed depth. These variables affect bed temperature and O_2 (or excess air). The excess air ratio (ϵ) is related to the fuel and air feed rates and the combustion efficiency by the following formula;

$$(1+\epsilon) = \frac{W_A}{W_F E(\overline{AIR})} \quad (18)$$

where $(\overline{AIR})$ = stoichiometric air/fuel ratio
(lb air/lb fuel)

In equation (18), the term W_A includes all air fed to the boiler. The $(\overline{AIR})$ term is given by the composition of the fuel. For most fossil fuels this term is given by:

$$(\overline{AIR}) = 11.53 f_C + 34.34 (f_H - \frac{f_O}{8}) + 4.29 f_S \quad (19)$$

where f_C = weight fraction of carbon in fuel.
 f_H = weight fraction of hydrogen in fuel.
 f_O = weight fraction of oxygen in fuel.
 f_S = weight fraction of sulfur in fuel.

Figure 2.62 shows the relationship between flue gas O_2 and percent excess air for a typical high sulfur eastern coal.

In general, solution of the heat balance equation involves specifying three of the four variables W_A , W_F , H and T_B and calculating the fourth. The two most useful variables to solve for are the fuel feed rate (W_F) and the bed temperature (T_B). The simplest way to solve the equations is to assume a value for the variable of interest, calculate the quantity $Q_{IN}-Q_{OUT}$, and then iterate on the value until the quantity $Q_{IN}-Q_{OUT}$ falls below a specified amount.

However, before the heat balance equations can be used, certain other key variables must be approximated. The key variables that affect the heat balance are combustion efficiency (E), the fraction of inbed burning (X), and the inbed heat transfer coefficients between the bed and the tube bundle (U_B) and between the bed and the waterwalls (U_W).

Generally a few sets of operating data from a unit are needed to enable estimation of these variables. Furthermore, the relationship between these variables and other operating conditions should be kept in mind.

Typical values for these key variables are shown in Table 2.40. Factors that affect the combustion efficiency are the type of fuel, fly ash recycle, bed temperature and freeboard geometry. Factors that affect the fraction of inbed burning are freeboard geometry, fuel type, bed depth and fly ash recycle. Bed mean particle size and superficial velocity affect the inbed heat transfer coefficients and the use of waterwall refractory obviously affects the waterwall coefficient.

Once acceptable values for the key variables have been chosen, the effect of operational changes can be studied. Figure 2.63 shows the effect of bed depth and superficial velocity on the flue gas oxygen concentration for the PDU operating at 1550°F. For this calculation the values of E , X , U_B and U_W were assumed to be constant regardless of bed depth or superficial velocity. Nevertheless, the predicted O_2 values are consistent with observed test data.

Table 2.41 shows a comparison of the calculated coal feed rates in the PDU versus the measured feed rates for Test No. 700-17. In these calculations, the measured combustion efficiency and measured heat transfer coefficients were used. Thus the only variable that needed to be specified was the fraction of inbed burning. For the first two time periods, $X = .80$ gave a reasonable approximation of the coal feed rate. For the time 1500, $X = .85$ gave a closer value of the coal feed rate. This would indicate that at the lower superficial velocities more of the total combustion takes place in the bed than at the higher velocities.

Another example of the usefulness of the heat balance equation was encountered during the oil firing test program. During these tests it was observed that a considerable amount of the oil was burning in the freeboard. The heat balance program showed that, while the overall combustion efficiency was 98 percent, the fraction of inbed burning was on the order of 50 percent. The extra freeboard combustion was consuming oxygen while not heating the bed and decreasing the overall combustion efficiency. This suggested that deeper beds could increase the inbed burning fraction and thus reduce the amount of fuel required to maintain bed temperatures. The result was higher combustion efficiency due to a higher excess air ratio.

Figure 2.63 shows some characteristics that are typical for most fluidized-bed boilers with an inbed tube bundle. First, when the tube bundle is completely submerged, an increase inbed depth has little effect on the flue gas

oxygen. However, when the tube bundle is not completely submerged, bed depth has a large effect on the O_2 . This is because the increased bed depth results in the bed contacting more heat transfer surface. This requires more heat to be supplied to the bed in the form of fuel to maintain bed temperature. A third characteristic is that, generally increasing the superficial velocity at constant bed depth and temperature results in a higher flue gas O_2 . The only time where this does not hold true is in the case of a tube bundle design that forces the bed to expand with increasing superficial velocity in such a way that the expansion removes enough heat so that the O_2 does not vary. The C-cell tube bundle of the Rivesville MFB appears to have such a design.

So far the heat balance calculations have been applied on the PDU, cells D and C of the Rivesville MFB and the Georgetown AFBB. The heat balance was also used during the oil firing activity of the PDU and is currently being used for the fly ash recycle program at Alexandria.

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 devolatilizing coal pieces burning. The dark background is the cooler bed material.

VIEW OF PDU COAL FEEDER AND BED DRAIN SIDE
2.52

Figure 2.2

VIEW OF PDU AIR INLET AND GAS OUTLET SIDE
2.53

Figure 2.3

DIMENSION DIAGRAM OF THE PDU

Figure 2.4

GAS AND SOLIDS SYSTEMS FLOW DIAGRAM

Figure 2.5

WATER SYSTEMS FLOW DIAGRAM
2.56

Figure 2.6

FLUE GAS BAGHOUSE AS INSTALLED AT ALEXANDRIA, VIRGINIA

2.57

Figure 2.7

STACK GAS BAGHOUSE FLOW DIAGRAM

Figure 2.8

BAGHOUSE CONFIGURATIONS

Figure 2.9

Figure 2.10A

Figure 2.10B

FLOW DIAGRAMS FOR BAGHOUSE CLEANING MODES

PULSE-JET CLEANING MODE FLOW DIAGRAM

Figure 2.11

PDU / CBC CONTROL SYSTEM
FLUIDIZED BED COMBUSTION LABORATORY
ALEXANDRIA, VIRGINIA

Figure 2.13

COMBUSTION CONTROL PANEL TO PDU INTERFACE SCHEMATIC

SAMPLE PROBE LOCATIONS

Figure 2.15

FLUE GAS SAMPLE CONDITIONING SYSTEM

GAS SAMPLE ANALYZERS AND HEADER SYSTEM

Figure 2.17

NOZZLE DETAILS

DETAILS OF DRILLED BOLT AIR DISTRIBUTOR

Figure 2.18

ENLARGED VIEW 'A-A'

THERMOCOUPLE LOCATIONS ON DRILLED BOLT AIR DISTRIBUTOR

Figure 2.19

PRESSURE DROP ACROSS DRILLED BOLT AIR DISTRIBUTOR
 AT VARIOUS CONDITIONS

Figure 2.20

DETERMINATION OF MINIMUM FLUIDIZATION WITH DRILLED
BOLT AIR DISTRIBUTOR

Figure 2.21

Figure 2.22

GRID PRESSURE VERSUS AIR FLOW

Figure 2.23

DIFFERENTIAL HEAT BALANCE ON THE INBED PORTION
OF BED LEVEL SENSOR

Figure 2.25

EXPERIMENTAL SETUP OF BED LEVEL SENSOR

TYPICAL DIFFERENTIAL PRESSURE PLOT OF EOBH

Figure 2.26

TYPICAL PLOT OF $\ln(T_B - T)$ VS. DISTANCE
FOR INBED AND FREEBOARD SECTIONS OF LEVEL SENSOR

Figure 2.27

COMPARISON OF EOBH AND HTBH METHODS FOR BED BUILDING IN TEST 700-25

COMPARISON OF EOBH AND HTBH METHODS
DURING BED BUILDING OF TEST 702-7

Figure 2.29

BED LEVEL MEASUREMENTS DURING BED BUILDING OF TEST NO. 700-28 A

Figure 2.30

BED LEVEL MEASUREMENTS DURING BED BUILDING OF TEST NO. 700-28B

Figure 2.31

FLY ASH EDUCTOR TEST FLOW DIAGRAM

Figure 2.32

CROSS SECTION OF SINGLE NOZZLE EDUCTOR AND DIFFUSER

Figure 2-33

EDUCTOR PERFORMANCE - COLD MODE TESTS

Figure 2.34

EDUCTOR PERFORMANCE - HOT MODE TESTS

Figure 2.35

Figure 2.36

MODIFIED FLY ASH HANDLING SYSTEM FOR THE PDU

PLANE VIEW OF PLENUM SHOWING
LOCATION OF DIVIDER AND SLIDE GATE

Figure 2.37

CUTAWAY SIDE VIEW OF PDU SHOWING RELATIONSHIP
BETWEEN SLUMPED BED INTERFACE AND TUBE BUNDLE

Figure 2.38

Figure 2.40

OIL FEED NOZZLE

Figure 2.41

Test 702-9, 11/1/78

Sample 2: Bed Material
(Magnified 10,000 times)

Sample 2: Bed Material
(Magnified 10,000 times)

ELECTROMICROGRAPHS OF SULFATED BED MATERIAL
2.92

Figure 2.42

Sample 1: Bed Material
(Magnified 10,000 times)

Sample 1: Bed Material (Magnified 10,000
times)

Figure 2.44

GRAMS OF CARBON PER 100 GRAMS OF FLYASH FOR TESTS
700-39, 700-38, AND 700-24

Figure 2.45

2.96

Figure 2.46

2.97

Figure 2.47

SIZE DISTRIBUTION OF BED MATERIAL FOR TESTS 700-19,
700-24, 700-38 AND 700-39

GRAMS OF SULFUR PER 100 GRAMS OF BED MATERIAL

Figure 2.50

LIMESTONE SIZE DISTRIBUTION FOR STONES TESTED

Figure 2.52

PERCENT SULFUR CAPTURE VERSUS CALCIUM TO SULFUR MOLAR FEED RATIO FOR STONES TESTED.

Figure 2.53

SO₂ EMISSIONS VERSUS TIME FOR TEST 700-28A

Figure 2.54

SO₂ EMISSIONS VERSUS TIME FOR TEST 700-30

Figure 2.55

CALCIUM LEAVING AS FLY ASH VERSUS C_a/S MOLAR FEED RATIO
FOR GREER AND LOWELLVILLE LIMESTONES

OPERATING VARIABLES FOR TEST 704-5
USING GROVE LIMESTONE AND LOW SULFUR COAL

Figure 2.57

OPERATING VARIABLES FOR TEST 704-9
USING GROVE LIMESTONE AND HIGH SULFUR COAL

Figure 2.58

OPERATING VARIABLES FOR TEST 704-11
USING GROVE LIMESTONE AND HIGH SULFUR COAL

MINIMUM FLUIDIZATION VELOCITY COMPARISON

Figure 2.60

INBED HEAT BALANCE BOUNDARY FOR A
TYPICAL FLUIDIZED BED BOILER

RELATIONSHIP BETWEEN FLUE GAS O_2 AND EXCESS AIR
FOR A TYPICAL HIGH SULFUR EASTERN COAL

Figure 2.62

PREDICTED OPERATING CONDITIONS OF THE ALEXANDRIA PDU

2.6 List of Tables (Section II)

Table 2.1

Methods of Analysis

<u>Analysis</u>	<u>Sample Type</u>	<u>Method</u>
Alkalinity	Boiler H ₂ O	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 ₂ O	Hack Model 7-P, Chloride Test Kit.
Conductivity	Boiler H ₂ O	Hack Portable Conductivity Meter Model 16300.
Hardness	Boiler Feed H ₂ O	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 2.1
Methods of Analysis (continued)

<u>Analysis</u>	<u>Sample Type</u>	<u>Method</u>
Hydrazine	Boiler H ₂ O	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 ₂ O	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 ₂ O	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 2.1

Methods of Analysis (continued)

<u>Analysis</u>	<u>Sample Type</u>	<u>Method</u>
Size distribution	Coal Limestone Bed material	ASTM Designation D410-38, (Reapproved 1969), Standard Method of Sieve Analysis of Coal.
Sulfite	Boiler H ₂ O	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.

Table 2.2

SUMMARY OF DRILLED BOLT AIR DISTRIBUTOR TESTS

<u>Test No.</u>	<u>Date</u>	<u>Duration</u>	<u>Purpose</u>
None assigned	10/30/78		Air Distributor delivered.
None assigned	11/08/78		Installation in PDU completed.
None assigned	11/09/78		Seals checked; initial ambient, no bed pressure drop tests.
None assigned	11/10/78		Ambient, 24" pressure drop tests.
700-26	11/13/78	5 hrs.	1500°F slump.
700-27	11/14/78	6 hrs.	2000°F slump.
700-28A	11/20/78	44 hrs.	Drilled bolt grid plate heat transfer and sulfur capture for Ca/S ratio of 1.
700-28B	12/05/78	16 hrs.	Drilled bolt grid plate heat transfer and combustion efficiency material balance for Ca/S ratio of 5.
700-29	12/12/78	7 hrs.	Drilled bolt grid plate, combustion efficiency low bed high temperature.
700-30	12/28/78	7 hrs.	Drilled bolt grid plate and sulfur capture fresh bed light-off, Greer limestone.
700-31	01/03/79	No run	Operational problems. Boiler never got up.
700-32	01/09/79	7 hrs.	Drilled bolt grid plate and bed material samples.
702-10	01/18/79	4 hrs.	Shakedown.
702-11	01/19/79	7 hrs.	Preliminary #4 Oil firing.
PDU-5	02/15/79	6 hrs.	Shakedown coal feed system changes.
700-33	02/28/79	6 hrs.	Drilled bolt grid plate.
700-34	03/02/79	7 hrs.	Drilled bolt grid plate.

Table 2.2 continued

<u>Test No.</u>	<u>Date</u>	<u>Duration</u>	<u>Purpose</u>
PDU-6	03/03/79	2 hrs.	Shakedown.
700-35	03/08/79	5 hrs.	Drilled bolt grid plate.
700-36	03/15/79	27 hrs.	Lowellville limestone.
700-37	03/21/79	4 hrs.	Lowellville limestone.
700-38	03/22/79	31 hrs.	Lowellville limestone.
700-39	04/06/79	22 hrs.	Lowellville limestone.
704-1	04/11/79	4 hrs.	Georgetown coal and limestone.
704-2	04/19/79	6 hrs.	Georgetown coal and limestone.
704-3	04/24/79	5½ hrs.	Georgetown coal and limestone, fly ash reinjection.
704-4	04/26/79	6 hrs.	Georgetown coal and limestone, fly ash reinjection.
704-5	05/03/79	30 hrs.	Low sulfur coal and fly ash reinjection.
704-6	05/22/79	20 hrs.	Low sulfur coal and fly ash reinjection.
704-7	06/01/79	6 hrs.	Low sulfur coal and fly ash reinjection.
704-8	06/05/79	5½ hrs.	Grove limestone and fly ash reinjection.
704-9	06/06/79	8 hrs.	Grove limestone and fly ash reinjection.
704-10	06/18/79	6 hrs.	Fly ash reinjection and pre- liminary baghouse.
704-11	06/20-21/79	32 hrs.	Fly ash reinjection and pre- liminary baghouse.
704-12	06/27-29/79	56 hrs.	Fly ash reinjection and bag- house.

Total Hot Test Hours: 398

Approximate # of Cycles: 50

Table 2.3

**Drilled Bolt Air Distributor Temperatures at
Normal Operating Conditions*****

Test No.	Avg. Bed Temp., °F	Avg. Air * Dist. Temp., °F	Avg. Mid** Dead Zone Temp., °F
700-28A	1528	351	466
	1513	346	463
	1558	352	489
700-29	1734	345	470
	1681	349	465
	1688	354	465
	1656	336	375
700-30	1476	326	417
	1472	332	404
	1489	327	391
	1513	331	410
700-33	1531	363	500
	1575	362	500
	1585	357	495
	1436	304	410
700-36	1560	378	502
	1543	361	475
	1481	351	448

* Avg. of TC #1, #2, #5 attached from the underside of the grid 1/8" into the base.

** TC mounted from underside of grid, passing through plate and extending 7/8" into dead zone of bed material.

*** Fluidizing Air Temperature 300°F.

Table 2.4

Summary of Drilled Bolt Grid Plate Slump Test Results

Test No.	Con- dition	Bed Temp.	Air Pre- heat	Max. Avg. Grid Temp.	Max. Dead Zone Temp.	Max. Divider Temp.
700-26 (11-13-78)	Run	1550	600*	653	620	903*
	Slump	1550	None	697	642	655
700-27 (11-14-78)	Run	2000	300	390	570	386
	Slump	2000	None	603	667	566

* Attained with below bed propane burner.

Table 2.5
Comparison of HTBH and EOBH Methods to Static Bed Depth
for Test 702-7*
10/20/78

<u>Static Bed</u>	<u>EOBH</u>	<u>HTBH</u>
16.5	31.0	37.85
19.0	31.0	41.59
20.0	32.5	44.39
22.0	36.0	47.42
24.0	36.5	48.35
25.0	38.5	49.62
25.5	38.0	49.41
25.5	41.5	49.10
26.5	43.0	50.30
27.5	43.0	50.88
28.0	44.0	51.38
28.0	44.0	51.11

* All measurements in inches.

HTBH = Heat Transfer Bed Height

EOBH = Effective Operating Bed Height

Table 2.6
Comparison of Static, HTBH and EOBH
Bed Level Measurements* for Test No. 700-28
(11-20-78)

<u>Static Bed Depth</u>	<u>HTBH</u>	<u>EOBH</u>
16.5	40.69	30.5
17	41.56	31.0
17.5	42.22	32.5
18	42.42	33.0
18	42.85	33.0
19	42.87	36.0
19.5	43.44	36.0
20	43.96	35.0
20	44.45	35.0
21	45.45	36.0
21	45.81	37.5
21.5	46.58	38.0
23	47.11	39.0
24	48.25	42.5
24	48.96	41.0

*All measurements in inches

HTBH = Heat Transfer Bed Height

EOBH - Effective Operating Bed Height

Table 2.7
Comparison of Static, HTBH and EOBH
Bed Level Measurements* for Test No. 700-28B
(12-5-78)

<u>Static Bed Depth</u>	<u>HTBH</u>	<u>EOBH</u>
15.5	37.14	30.0
17.5	38.54	30.5
19	41.68	34.0
20	44.18	36.5
21	45.80	41.0
23	47.56	42.0
24	48.18	40.0
24	48.82	41.5
25	49.48	41.0
25	49.48	42.0
25	49.53	41.0

*All measurements in inches

HTBH = Heat Transfer Bed Height

EOBH = Effective Operating Bed Height

Table 2.8
Consistency Test Results of HTBH Probe for
Test No. 700-28B
(11-20-78)

<u>Time (hr)</u>	<u>HTBH</u>
0300	48.93
0400	49.47
0500	49.37
0600	49.16
0700	48.19
0800	49.57
1200	49.03
1300	49.02
1400	48.33
1600	47.89

HTBH = Heat Transfer Bed Height

$$\text{HTBH} = 48.90''$$

$$s = 0.57''$$

Table 2.9
Steam Flow Rates, Test 701-2, July 18, 1978

Time/Hr.	12:00	12:30	13:00	13:30	14:00
Bed Temperature	1517	1520	1483	1698	1605
Superficial Velocity	11.4	11.6	11.2	13.8	13.6
Steam Flow Rates	3554	3369	3032	2266	2166

Table 2.10
Test 701-3, July 20, 1978

Time/Hrs.	10:30	11:30	12:00	12:30**	13:00	13:30	14:00	14:30
Bed Temp., °F	1478	1547	1573	1458	1470	1430	1415	1410
O ₂ in flue gas, %	4.3	4.7	4.3	.1	.5	.3	.2	.1
Coal feed rate, lb/hr	770	786	765	472	472	472	488	488
Fly ash collection rate, lb/hr	264	235	130	112	112	98	98	101
Limestone feed rate, lb/hr	420	279	208	249	304	246	212	12
Stack dust rate, lb/hr	*	*	24.97	*	*	17.22	*	12.87
Main air flow rate, lb/hr	7787	7571	7694	3407	4052	3935	3805	3404
Total air flow rate, lb/hr	8082	7867	7990	3708	4347	4231	4101	3700
Superficial velocity, ft/sec	11.6	11.7	12.0	10.5	12.4	11.8	11.4	10.2
Static bed depth, inches	25.0	27.0	25.0	25.0	25.0	27.5	28.0	30.0
EOBH (effective operating bed ht.), inches	36.0	39.0	37.0	31.0	32.0	35.0	34.0	31.0
SO ₂ , ppm	113	124	234	525	405	506	576	787
Steam flow	3929	4277	4417	2612	3165	3070	2988	2993

* Data not taken.

** Front half of bed slumped just prior to 12:30.

Table 2.11
Test 701-5, July 31, 1978

Time / Hrs.	11:30	12:00	12:30	13:30**	14:00	14:30
Bed Temp., °F	1456	1426	1553	1525	1470	1458
O ₂ in flue gas, %	2.0	1.9	1.0	.1	.7	.1
Coal feed rate, lb/hr	773	819	819	437	494	494
Fly ash collection rate, lb/hr	‡	‡	‡	‡	131	131
Limestone feed rate, lb/hr	258	249	249	120	114	114
Stack dust rate, lb/hr	23.71	*	26.13	17.71	*	19.74
Main air flow rate, lb/hr	7425	7418	7492	3842	3705	3699
Total air flow rate, lb/hr	7720	7714	7788	4138	4006	3995
Superficial velocity, ft/sec	11.0	10.7	11.6	12.2	11.5	11.4
Static bed depth, inches	27.0	25.0	25.0	22.5	23.0	22.0
EOBH (effective operating bed ht.), inches	41.0	35.0	36.5	30.0	31.0	30.5
SO ₂ , ppm	150	170	165	450	280	270
Steam flow	3788	3897	4307	2958	2958	2865

‡ Equipment problems.

* Data not reported.

** Front half of bed slumped just prior to 13:30

Table 2.12
Calculated Heat Transfer Coefficients

Test	700-17, 5/23/78					701-3, 7/20/78		701-5, 7/31/78	
						Before Slump	After Slump	Before Slump	After Slump
Bed Temp., °F	1535	1535	1521	1440	1383	1573	1458	1553	1525
Static Bed Depth, inches**	26.8	25.8	24.8	24.8	25.3	25.0	27.5	25.0	22.5
E.O.B.H., inches	40.0	42.5	35.0	35.0	35.0	37.0	35.0	36.5	30.0
Superficial Velocity	11.9	10.9	10.1	8.1	6.5	12.0	11.8 ‡	11.6	12.2 ‡
Economizer									
Overall Coefficient *	31.3	31.6	30.0	28.9	27.1	33.6	< 34.14	35.19 >	30.96
Inside Coefficient *	271.1	276.0	263.1	257.4	244.3	309.2	297.5	310.51	293.81
Outside Coefficient *	38.2	38.8	36.6	35.0	32.8	40.86	< 41.92	43.20 >	37.30
Calculated Ave. Wall Temp., °F	404	402	399	384	383	405.6	399.2	409.22	393.85
Superheater									
Overall Coefficient *	33.8	34.2	32.9	31.1	29.2	34.45	21.08	33.15	21.23
Inside Coefficient *	222.6	223.7	211.3	195.2	173.7	233.6	149.5	228.9	165.6
Outside Coefficient *	43.9	44.5	43.1	41.1	38.6	44.55	26.39	42.59	26.07
Calculated Ave. Wall Temp., °F	609	609	614	602	601	615.1	599.01	610.4	593.7

* All Heat Transfer Coefficients have units of Btu/hr/ft²/°F.

All Coefficients are uncorrected for bed temperature

** Difference in static bed depths are due to bed level control problems & not the results of partial slumping

‡ Based on 4.5 sq. ft. of active grid area

Table 2.13

Comparison of Localized Fluidization with Test
No. 700-17 (5/23/78) and No. 700-19 (6/8/78)

Test	700-17			700-19	701-3			701-5		
					Before Slump	After Slump		Before Slump	After Slump	
Bed Temp., °F	1535	1521	1383	1502	1573	1430	1410	1553	1525	1458
Static Bed Depth (in.)	26.8	24.8	25.3	25.5	25.0	27.5	30.0	25.0	22.5	22.0
E.O.B.H.* (inches)	40	35	35	39	37	35	31	36.5	30	30.5
Superficial Velocity	11.9	10.1	6.48	11.5	12.0	11.8	10.2	11.6	12.2	11.4
Combustion Efficiencies	92.30	86.06	86.80	89.05	88.81	83.64	81.90	89.60	84.12	80.43
Stack O ₂ %	3.3	1.4	2.5	3.4	4.3	.3	.1	2.0	.1	.1

*E.O.B.H. = Effective Operating Bed Height

Table 2.14

Test 701-3, 7/31/78

Composite Chemical Analysis of Bed Material, Fly Ash, and Dust

Material Analyzed	C%	H%	S%	L.O.I. %	Ash%	Ca%
Bed Material before Slumping	.49	.44	8.42	1.28	98.72	34.33
Bed Material after Slumping	.16	.21	7.67	3.58	96.42	38.07
Bed Material after Slumping	2.07	.30	7.78	6.52	93.48	32.55
Fly Ash before Slumping	39.19	.70	4.04	40.69	59.31	12.92
Fly Ash after Slumping	47.93	1.23	2.57	52.34	47.68	9.23
Fly Ash after Slumping	52.98	1.45	3.49	57.98	42.02	3.71
Dust before Slumping	15.73	.73	2.99	17.53	82.47	12.54
Dust after Slumping	27.89	.86	3.18	31.22	68.78	7.62
Dust after Slumping	38.13	1.24	2.48	42.10	57.90	3.84

Table 2.15

Test 701-5, 7/20/78

Composite Chemical Analysis of Bed Material, Fly Ash, and Dust

Material Analyzed	C%	H%	S%	L.O.I. %	Ash%	Ca%
Bed Material before Slumping	.67	.45	6.85	1.33	98.67	28.06
Bed Material after Slumping	1.88	.53	6.68	1.66	98.34	29.51
Bed Material after Slumping	.40	.42	7.31	1.56	98.44	31.13
Fly Ash before Slumping	20.89	1.03	5.93	25.89	74.11	6.10
Fly Ash after Slumping	36.97	1.08	2.85	40.54	59.46	7.99
Fly Ash after Slumping	49.77	1.13	1.75	56.75	48.25	3.44
Dust before Slumping	21.13	.86	2.49	24.16	75.84	6.52
Dust after Slumping	21.02	.86	2.49	26.37	73.63	7.34
Dust after Slumping	29.82	.95	1.87	37.72	62.28	5.74

Table 2.16
Operating Conditions, Test 703-3
10/10/78

Time-hr	12:00	13:00	14:45	16:20
Bed temp., °F	1610	1640	1590	1406
O ₂ in flue gas, %	1.3	1.0	3.0	5.0
Coal feed rate, lb/hr	808	714	638	629
Fly ash collection rate, lb/hr	276	258	228	224
Limestone feed rate, lb/hr	346	336	318	245
Bed drain rate, lb/hr	173	173	224	226
Stack dust rate, lb/hr	*	116.1	38.7	54.2
Main air flow rate, lb/hr	7383	7500	7682	7541
Total air flow rate, lb/hr	7678	7796	7977	7837
Superficial velocity, ft/sec	11.7	12.1	12.1	10.9
Static bed depth, inches	25.0	25.0	25.0	23.0
EOBH (effective operating bed ht.), inches	42.5	41.5	41.5	36.0

* Data not taken.

Table 2.17
Operating Conditions For Test 702-2
August 15, 1978

Time/Hr.	10:30	13:30	14:00	
Fuel Type	Coal	Coal Oil	Oil	
Bed Temp., °F	1504	1633	1452	
O ₂ in flue gas, %	3.0	1.4	3.8	
Coal feed rate, lb/hr	741	530	0	
Fly ash collection rate, lb/hr	236	220	244	
Limestone feed rate, lb/hr	294	0	0	
Bed drain rate, lb/hr	0	0	0	
Main air flow rate, lb/hr	7367	7012	6695	
Total air flow rate, lb/hr	7668	7307	6991	
Superficial velocity, ft/sec	11.1	11.3	9.7	
Static bed depth, inches	25	25	25	
EOBH (effective operating bed ht.), inches	38.5	36.0	32.5	
SO ₂ , ppm	120	958*	410	
Oil Flow	0	+	+	

* High SO₂ is probably due to combustion of coal in reducing regions caused by the burning oil.

+ Unknown due to Recorder malfunction.

Table 2.18
Operating Conditions for Test 702-5, 9/14/78

Time	11:00	11:30	12:30	13:00	14:00	14:30
Bed temperature, °F	1514	1513	1458	1479	1466	1469
O ₂ in flue gas, %	4.0	1.5	.5	.5	1.37	1.9
Coal feed rate, lb/hr	356	20	0	0	0	0
Fly ash collection, rate lb/hr	†	†	†	†	151	†
Limestone feed rate, lb/hr	255	0	0	0	0	0
Bed drain rate, lb/hr	0	0	0	0	0	0
Stack dust rate, lb/hr	†	†	†	7.064	10.16	†
Main air flow rate, lb/hr	7486	7487	7831	7817	7401	7935
Total air flow rate, lb/hr	7787	8033	8382	8578	8147	8661
Superficial velocity, ft/sec	11.3	12.4	13.6	13.2	12.6	13.4
Static bed depth, inches	26	25	25	24	30	31.5
EOBH (effective operating bed ht.), inches	38	36	36	34	43	45
SO ₂ , ppm	215	68	41	40	52	53
NO _x , ppm	446	138	29	33	34	34
Combustion efficiency,	†	†	†	98.5*	97.7	†
Oil feed air, lb/hr	235	245	250	250	250	250
Oil feed, gpm	0	1.02	1.2	1.2	1.19	1.06
Steam flow, lb/hr	4394	4811	4676	4676	4492	4569

† Data not taken

* Based on an assumed fly ash rate of 151 lb/hr

Table 2.19
Combustion Efficiencies for
Test No. 702-5, 9/14/78

Static Bed Height, inches	24	30
Combustion Efficiency	98.5	97.7
Heat Input from Fuel, Btu/hr	10.8	10.7
Fly Ash Heat Loss, Btu/hr	0.09*	.08
Stack Dust Heat Loss, Btu/hr	.013	.010
CO Heat Loss, Btu/hr	.032	.068
Hydrocarbon Loss, Btu/hr	.026	.086

*Based on an assumed fly ash rate of 151 lb/hr.

Table 2.20
Flue Gas Comparisons of Test Nos. 702-5, 700-19, and 700-24

Test	702-5			700-19			700-24
Date	9/14/78			6/8/78			8/24/78
Fuel	Coal	Coal +Oil	Oil	Oil	Oil	Coal	Coal
CO ₂ , %	13.7	11.2	10.7	10.6	11.5	14.2	15.4
CO, %	.175	.24	.165	.09	.20	.34	.25
Hydrocarbons, ppm	289	237	384	262	899	565	774
SO ₂ , ppm	215	68	41	40	52	987	263
NO _x , ppm	446	138	29	33	34	375	491
O ₂ , %	4.0	1.5	.5	.5	1.37	3.4	2.5

Table 2.21
Calculated Heat Transfer Coefficients

Test	700-17	700-19	702-5			
Date	5/23/78	6/8/78	9/14/78			
Fuel	Coal	Coal	Coal Oil	Oil	Oil	Oil
Bed temperature, °F	1535	1519	1513	1458	1479	1466
Static bed depth, inches	26.8	25.5	25	25	30	31.5
EOBH (effective operating bed ht.), inches	40.0	39.0	36	36	43	45
Superficial velocity	11.9	11.6	12.4	13.6	13.2	12.6
Economizer						
Overall coefficient*	31.3	32.21	38.9	37.4	36.6	37.3
Inside Coefficient*	271.1	312.2	304.2	292.6	298.6	303.7
Outside coefficient	38.2	38.8	49.2	47.1	45.6	46.6
Calculated average wall temperature, °F	404	387	419	409	406	407
Superheater						
Overall coefficient*	33.8	33.4	38.1	38.8	38.0	37.0
Inside coefficient*	222.6	223.6	250.7	243.6	243.6	235.3
Outside coefficient*	43.9	43.3	49.9	51.6	50.1	48.8
Calculated average wall temperature, °F	609	602	607	600	600	601

* All heat transfer coefficients have units of Btu/hr/ft²/°F.
All coefficients are uncorrected for bed temperature.

Table 2.22
Operating Conditions for Test No. 702-8
October 27, 1978

Time-hr	11:00	11:30	12:30	13:00	15:30
Bed Temp., °F	1520	1508	1462	1427	1441
O ₂ in flue gas, %	3.0	3.1	.1	.2	.3
Coal feed rate, lb/hr	696	573	0	0	0
Fly ash collection rate, lb/hr	326	264	80	62	42
Limestone feed rate, lb/hr	300	300	28	28	0
Bed drain rate, lb/hr	102	102	51	0	50
Stack dust rate, lb/hr	*	*	8.9	*	1.9
Main air flow rate, lb/hr	7539	7534	7445	6402	7369
Total air flow rate, lb/hr	8066	8065	7982	6939	7906
Superficial velocity, ft/sec	11.7	11.6	11.2	9.5	10.9
Static bed depth, inches	29.0	30.0	31.0	31.0	31.0
EOBH (effective operating bed ht.), inches	45.0	46.5	46.5	45.0	48.0
SO ₂ , ppm	423	383	68	91	85
Oil flow, gpm	0	0	1.02	.92	1.12

* Data not taken.

Table 2.23
Calculated Heat Transfer Coefficients

Test	700-17	700-19	702-5		702-8	
Date	5/23/78	6/8/78	9/14/78		10/27/78	
Fuel	Coal	Coal	#2 Fuel Oil	#2 Fuel Oil	#6 Fuel Oil	#6 Fuel Oil
Bed temp, °F	1535	1519	1479	1466	1427	1441
Static bed depth, inches	26.8	25.5	30.0	31.5	31.0	31.0
EOBH (effective operating bed ht.), inches	40.0	39.0	43.0	45.0	45.0	50.0
Superficial velocity	11.9	11.6	13.2	12.6	9.5	10.9
Economizer						
Overall coefficient*	31.3	32.21	36.6	37.3	36.4	36.6
Inside coefficient*	271.1	312.2	298.6	303.7	298.8	296.9
Outside coefficient*	38.2	38.8	45.6	46.6	45.4	45.7
Calculated average wall temp., °F	404	387	406	407	381.6	388.6
Superheater						
Overall coefficient*	33.8	33.4	38.0	37.0	35.1	37.3
Inside coefficient*	222.6	223.6	243.6	235.3	221.8	234.6
Outside coefficient*	43.9	43.3	50.1	48.8	46.3	49.4
Calculated average wall temp., °F	609	602	600	601	592.9	595.4

* All heat transfer coefficients have units of Btu/hr/ft²/°F.
All coefficients are uncorrected for bed temperature.

Table 2.24
Operating Conditions for Test 702-9, 11/1/78

Time, Hrs.	10:30	11:00	12:00	13:00	14:00	15:00
Fuel	#6 Oil	#6 Oil	#6 Oil	#6 Oil	#6 Oil	Coal
Bed Temp., °F	1523	1523	1515	1490	1540	1530
O ₂ in flue gas, %	4.1	4.0	4.5	4.5	3.2	4.0
Fuel feed rate, lb/hr	428.3	437.1	383.2	414.4	445.9	636
Fly ash collection rate, lb/hr	90	90	90	82	90	257
Limestone feed rate, lb/hr	206	197	200	186	221	174
Bed drain rate, lb/hr	1	1	42	2	2	*
Stack dust rate, lb/hr	15.79	16.10	*	22.91	27.87	*
Main air flow rate, lb/hr	7293	7293	7288	7126	7120	7120
Total air flow rate, lb/hr	7976	7976	7971	7807	7799	7811
Superficial velocity, ft/sec	11.6	11.6	11.5	11.1	11.4	11.4
Static bed depth, inches	30	30	29	31	31	29
EOBH (effective operating bed ht.), inches	46.5	44.5	45.0	48.5	45.0	43.5
SO ₂ , ppm	6	5	10	10	7.5	295
Steam flow, lb/hr	4486	4463	4341	4156	4620	4556
NO _x , ppm	204	206	217	218	189	560
Hydrocarbons, ppm	188	136	82	129	325	772
Combustion efficiency, %	--	98.44	--	98.36	98.11	--
Ca/S, molar feed ratio	--	5.01**	--	4.99**	5.5**	2.63

* Data not taken.

** Additional limestone fed to maintain bed level.

Table 2.25
Calculated Heat Transfer Coefficients for Test No. 702-9, 11/1/78

Time - hr	10:30	11:00	12:00	13:00	14:00	15:00
Bed temp, °F	1523	1523	1515	1490	1540	1530
Static Bed depth, inches	30	30	29	31	31	29
E.O.B.H., inches	46.5	44.5	45.0	48.5	45.0	43.5
Superficial velocity	11.6	11.6	11.5	11.1	11.4	11.4
Fuel	#6 oil	#6 oil	#6 oil	#6 oil	#6 oil	coal
Economizer						
Overall coefficient	34.81	34.54	35.25	33.82	34.40	34.77
Inside coefficient	301.55	299.7	305.8	302.88	324.8	328.5
Outside coefficient	42.84	42.47	43.39	41.32	41.69	42.08
Calculated Avg. wall temp, °F	387.1	387.3	387.2	387.0	390.4	387.5
Superheater						
Overall coefficient	34.79	33.86	34.68	31.06	33.32	33.75
Inside coefficient	234.9	233.6	228.2	219.2	240.8	237.2
Outside coefficient	45.05	43.57	45.19	39.57	42.38	43.22
Calculated Avg. wall temp, °F	607.1	604.6	608.5	597.9	602.7	600.6

Table 2.26

#6 Oil Analysis and Bed Material Analysis

Test 702-9, 11/1/78

	Lab I	Lab II
Carbon, %	85.22	85.76
Hydrogen, %	11.03	10.80
Oxygen, %	.53	1.04
Nitrogen, %	.19	.41
Sulfur, %	2.38	2.31
Ash, %	.09	.10
High Heating Value Btu/#	18,286	18,310
Trace Elements		
V, ppm	455	690 (as V_2O_5)
Ni, ppm	54	64 (as NiO)
Fe, ppm	6	61 (as Fe_2O_3)
Cl, ppm	238	590

Bed Material at 1400 hrs.

Sulfur, %	3.58
Iron, %	2.03
Vanadium, %	.07
Calcium, %	38.16

Table 2.27
Test 700-19, June 8, 1978

Time/Hr.	12:00	13:00	14:30	15:30	Avg.		
Bed Temp., °F	1502	1511	1519	1502	1509		
O ₂ in flue gas, %	4.1	3.6	3.0	3.4	3.5		
Coal feed rate, lb/hr	729	690	701	720	710		
Fly ash collection rate, lb/hr	218	206	216	216	214		
Limestone feed rate, lb/hr	366	352	346	348	353		
Bed drain rate, lb/hr	158	158	140	140	149		
Stack dust rate, lb/hr	--	27.6	17.9	21.5	24.3		
Main air flow rate, lb/hr	7624	7624	7638	7638	7631		
Total air flow rate, lb/hr	7920	7920	7934	7934	7927		
Superficial velocity, ft/sec	11.5	11.5	11.6	11.5	11.5		
Static bed depth, inches	26.0	30.0	26.5	25.5	27.0		
EOBH (effective operating bed ht.), inches	39.0	38.5	37.0	39.0	38.4		
SO ₂ , ppm	926*	879*	1035*	987*	957*		
Ca/S			2.48	2.43			
Combustion efficiency, %			82.21	89.05			

--No isokinetic sample taken.

* The SO₂ emissions are consistent with the low calcium sulfur feed ratios during this test.

Table 2.28
Operating Conditions, Test 700-24, August 24, 1978

Time	0400	0600	0800	1000	1200	1400	Avg.
Bed temp., °F	1474	1443	1453	1478	1519	1515	1480
O ₂ in flue gas, %	3.0	3.9	3.7	2.8	2.7	2.5	3.1
Coal feed rate, lb/hr	747	776	777	716	735	750	750
Fly ash collection rate, lb/hr	252	267	280	255	224	252	255
Limestone feed rate, lb/hr	321	339	345	345	321	350	337
Bed drain rate, lb/hr	189	130	227	155	155	150	168
Stack dust rate, lb/hr							
Main air flow rate, lb/hr	7412	7267	7184	7114	6911	6906	7132
Total air flow rate, lb/hr	7708	7563	7480	7409	7207	7201	7428
Superficial velocity, ft/sec	11.0	10.5	10.5	10.5	10.5	10.4	10.6
Static bed depth, inches	25.9	29	29	29	28	28	28.2
EOBH(effective operating bed ht.), in.	42.0	45.5	40.0	41.0	39.0	39.0	41.1
SO ₂ , ppm	--	--	100	244	259	263	217
Ca/S						1.65	
Combustion efficiency, %						75.3	

Table 2.29
Test 700-38
Operating Conditions

Date	3/23/79		3/24/79					
Time	1430	1900	0200	0500	0830	1000	1115	1300
Bed temp., °F	1507	1543	1571	1548	1509	1571	1553	1479
O ₂ in flue gas, %	3.5	3.7	3.1	3.5	3.2	3.0	3.1	3.5
Coal feed rate, lb/hr	735	708	752	728	734	774	772	708
Fly ash collection rate, lb/hr	*	246	238	252	339	240	220	*
Limestone feed rate, lb/hr	0	316	341	328	310	316	248	351
Bed drain rate, lb/hr	2	143	88	91	152	148	296	0
Stack dust rate, lb/hr	*	*	*	*	92.1	31.4	31.4	29.8
Main air flow rate, lb/hr	7401	7602	7485	7662	7250	7541	7507	7280
Total air flow rate, lb/hr	7662	7868	7782	7959	7545	7836	7801	7574
Superficial velocity, ft/sec	11.1	11.6	11.7	11.8	10.9	11.7	11.6	10.9
Static bed depth, inches	25.0	26.0	26.0	26.5	30.0	27.0	26.0	28.0
EOBH (effective operating bed ht.), in.	42.0	42.5	43.5	42.0	47.5	45.0	43.0	45.0
SO ₂ , ppm, (stack gas)	122	281	307	330	368	519	700	522

* Data not taken

Table 2.30
Operating Conditions for Test 700-39, 4/6/79

Time, Hrs.	0400	0700	0800	0900	1000	1100	1230	1300
Bed Temp., °F	1549	1493	1472	1438	1441	1457	1467	1443
O ₂ in flue gas, %	4.0	4.0	4.3	4.5	5.3	4.0	2.7	4.0
Coal feed rate, lb/hr	780	776	823	758	722	756	703	724
Fly ash collection rate, lb/hr	236	238	240	240	262	230	226	214
Limestone feed rate, lb/hr	366	358	350	325	0	342	346	328
Bed drain rate, lb/hr	*	*	*	*	*	*	*	*
Stack dust rate, lb/hr	*	*	27.5	56.5	*	*	*	34.8
Main air flow rate, lb/hr	7492	7401	7457	7256	7335	7335	7406	7249
Total air flow rate, lb/hr	7758	7667	7727	7523	7602	7597	7679	7511
Superficial velocity, ft/sec	11.5	11.0	11.0	10.5	10.6	10.7	10.9	10.5
Static bed depth, inches	26.0	26.0	26.0	28.0	27.0	27.0	26.0	28.0
EOBH (effective operating bed ht.) inches	43.0	43.0	42.0	44.5	45.0	43.5	43.0	45.0
SO ₂ , ppm	137	174	160	106	226	352	303	153
Ca/S,	*	*	4.74	4.80	*	*	*	*
Combustion efficiency, %	*	*	87.4	83.7	*	*	*	*

* Data not taken.

Table 2.31

Summary of Fly Ash Results for
Tests 700-19, 700-24, 700-38 and 700-39

BASIS OF 100 GRAMS OF FLY ASH PER TEST

Test 700-19: 1530 hrs

<u>U.S.S</u>	<u>MEAN</u>	<u>FRACT</u>	<u>g CARBON/ 100g FLY ASH</u>
+16	2275	.0244	.03
-16+30	890	.1548	.49
-30+40	505	.1260	.91
-40+80	299	.2194	4.25
-80+100	163	.0455	1.82
-100+140	127	.1097	4.17
-140+200	89	.1829	6.87
-200+400	56	.1178	4.29
-400	19	.0195	.56

Test 700-24: 1500 hrs

<u>U.S.S</u>	<u>MEAN</u>	<u>FRACT</u>	<u>g CARBON/ 100g FLY ASH</u>
+16	2275	.0109	.22
-16+30	890	.1148	2.51
-30+40	505	.1230	3.23
-40+80	299	.3743	16.43
-80+100	163	.0792	3.78
-100+140	127	.0847	3.87
-140+200	89	.0792	3.35
-200+400	56	.0984	3.52
-400	19	.0355	0.62

Test 700-38: 1300 hrs

<u>U.S.S</u>	<u>MEAN</u>	<u>FRACT</u>	<u>g CARBON/ 100g FLY ASH</u>
+16	2275	.009	.18
-16+30	890	.080	2.50
-30+40	505	.101	3.72
-40+80	299	.313	15.96
-80+100	163	.083	4.60
-100+140	127	.081	5.27
-140+200	89	.105	5.73
-200+400	56	.117	5.17
-400	19	.111	2.31

Table 2.31 Continued

Test 700-39: 0900 hrs

<u>U.S.S</u>	<u>MEAN</u>	<u>FRACT</u>	<u>g CARBON/ 100g FLY ASH</u>
+16	2275	.018	.22
-16+30	890	.119	2.71
-30+40	505	.139	4.29
-40+80	299	.301	9.95
-80+100	163	.072	3.71
-100+140	127	.068	2.98
-140+200	89	.062	2.41
-200+400	56	.125	4.35
-400	19	.096	1.62

Table 2.32

Bed Material Size Distributions for
Tests 700-19, 700-24, 700-38 and 700-39

U.S.S.	Mean Particle Diameter, μ	Size Fraction			
		Test 700-19	Test 700-24	Test 700-38	Test 700-39
		<u>1300 hrs</u>	<u>1530 hrs</u>	<u>1400 hrs</u>	<u>0900 hrs</u>
-.265"+.25"	6540	.0112	.0023	.0170	.0026
-.25"+6	4855	.1398	.0987	.0651	.00694
-6+8	2870	.3523	.2909	.2385	.2489
-8+12	2030	.2846	.3351	.3084	.3125
-12+16	1435	.1196	.1583	.2305	.2065
-16+20	1015	.0718	.0815	.1126	.1042
-20+30	715	.0176	.0310	.0256	.0289
-30+40	505	.0020	.0017	.0011	.0029
-40	210	<u>.0012</u>	<u>.0006</u>	<u>.0011</u>	<u>.0007</u>
Sample Mean Particle Size, μ		2075	1897	2581	1801

Table 2.33

% Sulfation of Bed Material As A Function of Size

U.S.S.	Mean Particle Diameter, μ	% Sulfation			
		Test 700-19	Test 700-24	Test 700-38	Test 700-39
- $\frac{1}{4}$ +8	4365	22.6	13.2	14.4	8.2
-8+16	1785	38.1	15.8	14.1	13.3
-16+30	893	46.7	29.4	23.6	22.6
Basis of 100 grams of bed material.					

Table 2.34
Analyses of Limestones Tested

	GREER	LOWELLVILLE	GROVE
Calcium	34.93	39.45	38.66
Ash	62.28	58.96	57.87
Loss on Ignition	37.72	41.04	42.14
Carbon	10.60	10.63	11.19
Hydrogen	0.17	0.19	0.03
Sulfur	0.02	0.18	0.02
CaCO ₃	87.33	98.63	96.65

TABLE 2.35

SUMMARY OF OPERATING CONDITIONS DURING SULFUR CAPTURE TESTS

LIMESTONE TYPE	GREER			LOWELLVILLE						GROVE					
Test No.	700-19	700-28A		700-36		700-38		700-39		704-11			704-12		
Date	06/08/79	11/21/78		03/15/79		03/16/79	03/23/79		04/06/79		06/20/79	06/21/79		06/28/79	06/29/79
Time	15:30	17:00	18:00	16:30	21:00	04:30	05:00	08:30	9:00	13:00	15:30	11:00	13:00	10:15	11:30
Bed Temp., °F	1502	1550	1521	1556	1531	1543	1548	1509	1438	1443	1588	1615	1600	1501	1604
O ₂ in flue gas, %	3.4	4.0	4.5	6.0	6.0	6.0	3.5	3.2	4.5	4.0	4.8	4.9	6.8	5.4	4.7
Fuel feed rate, lb/hr	720	684	700	929	946	948	728	734	758	724	828	848	819	840	814
Fly ash collection rate, lb/hr	216	168	175	226	247	208	252	339	240	214	**	**	**	90(+)	59(+)
Limestone feed rate, lb/hr	348	120	152	267	310	164	328	310	326	328	178	176	187	396	224
Bed drain rate, lb/hr	140	94	167	88	17	0	91	152	*	*	69	70	346	168	0
Stack dust rate, lb/hr	22	31	25	*	58	*	*	92	57	35	76	108	116	72	149
Main air flow rate, lb/hr	7638	7666	7597	7735	7780	7645	7662	7250	7256	7249	7394	7272	8207	7085	7269
Total air flow rate, lb/hr	7934	7967	7899	8007	8048	7924	7959	7545	7523	7511	8016	7895	8643	7658	7750
Superficial velocity, ft/sec	11.5	11.8	11.5	11.9	11.8	11.6	11.8	10.9	10.5	10.5	12.1	12.0	13.3	11.1	11.8
Static bed depth, inches	25.5	26.0	26.0	25.5	26.5	28.0	26.5	30.0	28.0	28.0	26.0	27.0	20.0	26.0	24.0
EOBH (effective operating bed ht.), inches	35.0	40.5	39.0	41.0	41.5	43.0	42.0	47.5	44.5	45.0	*	45.0	36.5	43.5	39.5
SO ₂ , ppm	587	1306	1101	562	549	1323	330	368	106	153	375	358	296	273	530
Ca/S feed ratio	2.4	1.5	1.9	3.4	3.9	2.1	5.4	5.1	4.8	5.1	2.8	2.6	2.9	4.5	2.6
% Sulfur capture	77.6	49.1	58.2	78.8	79.7	50.8	82.6	81.7	95.8	93.5	82.5	83.9	85.0	91.3	82.7
% Combustion efficiency	87.1	82.6	86.5	*	83.9	*	*	*	83.7	86.2	96.6	94.7	93.3	89.3	88.3

* Data not taken

** Fly ash reinjection

(+) Fly ash reinjection with overflow

Table 2.36

Analyses of Coal Used During Sulfur Capture Tests

Test No.	Carbon	Hydrogen	Sulfur	Ash	Calcium
700-19	68.4	4.7	5.2	14.6	0.4
700-28A	68.4	4.4	3.4	16.3	0.6
700-36	65.4	4.6	2.7	15.6	0.1
700-38	68.5	4.9	2.6	18.3	0.6
700-39	62.5	4.1	2.8	21.5	1.8
704-11	67.4	4.6	2.4	16.7	0.5
704-12	63.3	4.2	3.3	18.3	0.6

Table 2.37

Operating Conditions, Test 704-9, June 6, 1979

Time	1000	1100	1200	1300	1400	1500	1600
Bed Temp., °F	1626	1632	1610	1564	1548	1529	1553
O ₂ in flue gas, %	8.7	6.3	4.4	2.1	1.1	.8	1.3
Coal feed rate, lb/hr	330	376	446	574	612	692	720
Fly ash collection rate, lb/hr	0	0	0	0	0	0	0
Limestone feed rate, lb/hr	174	192	180	176	190	161	170
Bed drain rate, lb/hr	*	*	*	*	*	*	*
Stack dust rate, lb/hr	*	*	*	*	*	116.1	*
Main air flow rate, lb/hr	4460	4448	4320	4316	4622	4625	5504
Total air flow rate, lb/hr	4756	4745	4617	4608	4914	4912	5791
Superficial velocity, ft/sec	7.6	7.6	7.3	7.1	7.4	7.3	8.7
Static bed depth, inches	7.0	9.5	12.0	13.75	16.0	19.0	20.0
EOBH (effective operating bed ht.), inches	14.0	16.0	18.0	22.5	27.0	31.5	33.5
SO ₂ , ppm	7	16	40	117	502	813	453

* Data not taken

Table 2.38

Summary of Minimum Fluidization Tests
(Four Inch Diameter Bed)

Test	Material	Bed Temperature (°F)		Particle Size (microns)	Minimum Fluidization Velocities	
		Cold	Hot		80°F (fps)	1550°F* (fps)
Series I						
Trial 1	Lowellville LS	80	1565	1724	2.6	3.69
" 2	"	77	1540	1724	2.85	3.13
" 3	"	88	1570	1724	2.5	2.85**
Average		81	1558	1724	2.65	3.22
Series II						
Trial 1	Greer LS	82	1590	1785	2.65	3.0
" 2	"	78	1535	1785	2.50	2.6**
" 3	"	88	1560	1785	2.55	3.2
Average		82	1561	1785	2.57	2.95

* Superficial velocity under hot conditions were calculated using flue gas flow rates and densities.

** Scatter of data due to difficulty in measuring fluidization velocities.

Table 2.39

Summary of Predicted and Experimental
Minimum Fluidization Velocities*

Test	Bed Temperature	Equation 2 Prediction	Equation 3 Prediction	Experimental
Series I	80°F	3.25	4.52	2.65
Series II	80°F	3.34	4.62	2.57
Series I	1550°F	1.83	3.43	3.22
Series II	1550°F	1.95	3.61	2.95

*Ft/Second.

Table 2.40

Typical Values for Key Variables
for the Alexandria PDU

Variable	Range When Firing Coal	Range When Firing Oil
E	80-90%	95-99%
X	80-95%	50-100%
U_B	35-45 Btu/hr, ft ² , °F	35-45 Btu/hr, ft ² , °F
U_W^*	2.0 Btu/hr, ft ² , °F	2.0 Btu/hr, ft ² , °F

* Waterwalls of PDU are refractory covered.

Table 2.41

Comparison of Heat Balance Calculations
To Actual Test Data for Test 700-17

Time	V_S ft/sec	W_F measured lb/hr	W_F calcualted lb/hr	% difference
1300	11.9	697	739	5.72
1400	10.1	653	685	4.90
1500	6.5	454	476	4.85

III. RIVESVILLE, W. VA. MFB PLANT OPERATIONS

3.1 General

During the reporting period, the Rivesville MFB unit operated for 1750 cell hours including 164 hours of electrical power generation (commercial runs).

Eleven performance tests were conducted, in addition to two preliminary performance tests to check-out data collection equipment prior to testing.

Operation of the MFB unit was resumed in July after completion of extensive repairs to the west boiler circulating pump. An equipment fire occurred in August resulting in an eight-month delay in the program. During the reconstruction period, improvements and modifications to systems were incorporated to improve reliability and performance.

3.2 Operational Highlights

In July and until the equipment fire in August, the MFB unit was operated in continuous multicell steam generation mode with periods of commercial operation. To evaluate unit performance, a series of eleven tests were completed. Commercial operation of the unit was terminated at 2115 hours, 09 August due to inadequate coal supply and shortly thereafter at 2155 hours, the first visual indication of fire was reported. The "Project Phoenix" reconstruction program was implemented to remove damaged equipment and restore the plant to operating capability.

In the period prior to restart of MFB operations, an intensive operator training program was conducted to supplement previous operator training.

During the reconstruction outage, modifications were made to the Bed Material Transport System. The screens were removed from the classifier so that the bed material passes directly through. The rotary feeders at each end of the classifier were removed and the bed material is now transported by new steam jet ejectors.

Preliminary observations indicated that the replacement "nuts and bolts" air distributor in the CBC cell performed satisfactorily.

A screw feeder with a hollow shaft was installed to replace the eductor, which did not perform satisfactorily, to feed

ash into the CBC cell. The screw feeder was selected to provide a positive means of material feed. A diffuser was installed at the screw nozzle discharge to create negative pressure inside the screw because the GCA Corporation no longer subcontracts for isokinetic dust sampling work. In-house personnel were professionally trained to work after completing the week-long course and the certification test. Six successful candidates were inducted into the Results Department to handle this testing function.

Three tests were conducted on the MFB unit to determine optimum operating condition of B and C cells with respect to superficial velocity and excess oxygen.

3.3 Operations

3.3.1 Operational Summary

Operations were interrupted in June to perform extensive repairs to the west boiler forced circulation pump. Pump failure resulted in excessive shaft seal leakage caused by failure of the shaft sleeve o-ring. Shaft sleeve discoloration indicated operation under excessive temperatures. This could have been the result of pump operation prior to installation of the seal water cooler or possibly operation without sealwater flow.

During check-out operation of the reassembled pump, the throttle bushing again overheated and seized the shaft sleeve. The pump was removed and disassembled and a flexatalllic sealing gasket was found to be blocking sealwater supply to the shaft. This resulted from the bolts holding the spool piece to the motor hub being .030" too long. As a result, the lockwashers on this assembly could not completely compress. These bolts were shortened and assembled with additional lockwashers. The pump clearances and runout were then rechecked and found to be within manufacturer's recommendations. The pump was reassembled and a test run conducted.

With the pump in operation, an attempt was made to light-off the MFB with a single burner. This test was in anticipation of the possibility of installing an overbed coal feeder at some future date which would necessitate single burner light-offs. At 2145 hrs, 18 July, both burners were ignited at low fire rate and fifteen minutes later the north burner was valved out. The south burner was brought to maximum firing rate but the standard light-off procedure of alternately fluidizing and slumping a coal charged bed proved unsuccessful. Three hours maximum firing of the south burner produced a

maximum temperature of only 250°F. At that time, reverse flow conditions were established and a second coal charge admitted. Again, there was insufficient temperature to attain combustion. The CBC draft control was reopened and at 0510 hrs, 19 July, the north burner was ignited.

Four hours after lighting the north burner, coal fire was established in the CBC. While bed height was increasing in preparation for transfer to Cell C, MPC Unit #7 developed an economizer tube leak. MFB coal feed was discontinued at 1700 hrs and the unit cooled for inspection. The CBC was completely drained to permit inspection of the replacement gridplate section. Foster Wheeler Corporation proposed a gridplate design that employs 3/16-inch plates supporting a close array of two-inch bolts drilled length-wise to admit the fluidizing/combustion air. The two-inch extension above the support plate provides for an inert zone which should insulate the plate. (See detailed discussion later in this report.)

On 20 July, with Unit #7 in service, the MFB was prepared for light-off and performance testing in two cell operation. Bed height was slightly lower than normal when both burners were ignited at 2100 hrs. Coal combustion was established at 2240 hrs, but was immediately followed by a temperature excursion to 2300°F. The excursion was compensated for by increasing the combustion air flow which caused a high draft fan trip. The combination of high temperatures and sudden loss of combustion air agglomerated the bed. The unit was cooled and the clinkers removed via the CBC access door. During cleanout, the unit operating temperature limits were increased to accommodate anticipated higher bed temperatures required by the performance test program. Fuel feeder permissives were adjusted as follows:

<u>Alarm</u>	<u>Trip</u>
CBC - 2100°F	2150°F
C - 1600°F	1700°F
B - 1600°F	1700°F
A - 1600°F	1700°F

At 2100 hrs, 21 July, the unit was closed up and raw limestone added to build up bed height. At 2230 hrs, the Trane Burners were put in service but the CBC bed showed indications of only partial fluidization. Bed height was decreased by transferring material to C cell and combustion air pressure increased by closing off all but the CBC damper. Fan pressure was increased to maximum and still only partial fluidization

could be attained. The unit was cooled and a visual inspection of the bed provided no explanation for failure to fluidize fully. A complete inspection of fans, ductwork, gridplate and flues was conducted. Air flow tests were performed to confirm fluidizing air flow and analyses performed to determine bed material composition.

From the size distribution analysis of the bed, it was concluded that the mean particle size of the limestone created the fluidization difficulties. Refer to Table 3.1 for size analysis and mean particle size determination of the limestone that fluidized successfully.

Compare this to Table 3.2, which describes the limestone that would only partially fluidize. It should be noted that almost three quarters of this stone was between the -5 +10 screen sizes. All attempts to ignite coal in this partially fluidized bed resulted in low temperature agglomerates. After removing these soft clinkers, two tons of concrete sand (-10 +20 mesh) were added to the bed to reduce its mean particle density. This mixture fluidized and sustained coal fire at 2338 hrs, 27 July.

During the following day, operation of CBC cell was carefully monitored before C cell was ignited at 1930 hrs, 28 July. Two cell operation ended two hours later when a belt feeder standpipe plugged. Cell C was again ignited at 0130 hrs, 29 July, and operated for half an hour fed from the south side only when the north side belt feeder was shut down to reweld a supporting brace. This two cell operation lasted 15.5 hrs before a malfunction of the C cell bed temperature safety circuitry prematurely tripped out the fuel feed system. The temperature transmitter was reset and C cell was reignited at 2000 hrs. Again, standpipe pluggage curtailed two cell operation less than two hours later.

CBC continued operation until 0900 hrs, 30 July, when a faulty thermal overload magnetic coil tripped the boiler circulating pumps. CBC was slumped for approximately six hours for repair to the breaker and was reignited at 1645 hrs, 31 July. Two cell operation began at 2035 hrs and continued until 0600 hrs, 2 August. At that time, C cell was slumped to conserve condensate. Two cell testing was completed during a 15 hour period between 1 and 2 August.

Three cell operation was again attained 2136 hrs, 2 August, but was curtailed because of pluggage in the south B cell belt feeder. Cell C was again slumped to conserve condensate. Three cell operation was resumed 1310 hrs, 3 August. On stabilization and verification of steam conditions, steam

flow was routed to the MPC Turbine #5. Stable commercial operation and further testing continued from 3 August until 5 August when the unit was backed down to one cell because of debris plugging the south furnace feed system.

Three cell commercial operation resumed 6 August and continued for the next 36 hours before damage to the north B cell belt feeder forced a shutdown of cell B. Cell C operation was maintained from the south C cell belt feeder only while the north vibrating feeder was under repair. Cell B was reignited 1045 hrs, 8 August, and commercial operation was resumed 1.5 hours later. Commercial operation ended at 0406 hrs, 9 August, due to coal feed problems. Two cell operation continued until cell B was returned to service at 1330 hrs and steam flow to the turbine was restored at 1514 hrs. At 1530 hrs, a failure of the south coal scale created coal supply problems which were corrected without interruption to operation. Commercial operation was terminated at 2115 hrs, 9 August, due to inadequate coal supply to the furnace south side and pluggage on the furnace north side. At 2155 hrs, the first visual indication of fire was reported and one minute later, the CBC coal feed was manually tripped from the boiler safety circuit. A detailed report of the sequence of events leading to and the damage caused by a fire associated with the regenerative rotary air preheater (APH), follows the yearly operations summary.

Reconstruction of the fire damage was code-named Project Phoenix.

During the reconstruction outage, operations personnel performed the following functions:

1. Supplied the labor force for construction.
2. Expedited material for reconstruction effort.
3. Revised the MFB electrical and mechanical drawings to indicate as-built status.
4. Made design changes to the coal handling system, i.e., 75' belt and rotex modification.
5. Supplied personnel for maintenance activities.
6. Prepared the Phoenix Outage Inspection and Damage reports.

During November, operating personnel began a training program designed and taught by Stone and Webster Engineering Company.

The purpose of the Operations Training Program was to provide detailed instruction to Shift Supervisors, Control Room Operators and Equipment Operators on the MFB systems. This supplemented previous knowledge acquired in the MFB Power Plant Course. In addition, Shift Supervisors and Control Room Operators were given training in Supervisory skills. All shift Supervisors, Control Room Operators and Equipment Operators were also given training on prescribed responses to general and specific MFB emergency operating conditions.

The MFB was re-established in operational status during March 1979. Initial operational check-out resulted in several interruptions associated with equipment, design and control changes. Modifications to the bed material system and the Beckman gas analysis system during the outage proved to be significant improvements.

At 0520 hrs, 12 March, check-off began for CBC start-up. At 1031 hrs, both burners were ignited and a gradual increase in bed temperature began. After two 500 lb coal charges, at 1146 and 1305 hrs, bed temperatures increased rapidly. Coal feed was established at 1425 hrs. Good stack conditions were noted. The burners were backed off and the precipitator energized. At 1550 hrs, coal feed was discontinued due to a suspected clinker formation in the northwest corner of the cell. The unit was cooled and the clinkers removed from the CBC.

At 2200 hrs, 13 March, a failure of the auxiliary FD motor starter was reported and an investigation disclosed the main line contacts severely burned and the main coil fuses blown. Further study revealed that the curves of acceleration and thermal overload relay cross each other. While the motor acceleration time is 46 seconds, the thermal overload relay will trip in 15 to 20 seconds. Present operational procedure is to start the Aux. FD fan and when the thermal overload trips, to reset it while the fan is still rotating, and restart the motor. While this procedure permits the fan to be started, it has a damaging effect on the starting equipment and motor. The instant the motor is disconnected (tripped), it acts as a self excited generator that delivers substantial amounts of current because the magnetic flux within the air gap does not collapse immediately. The resulting residual voltage decreases in amplitude exponentially. The length of time it takes for the residual voltage to drop to zero depends on the motor's open circuit time constant. In addition to the residual voltage decreasing, the phase angle between the motor voltage and supply voltage changes. Reconnecting a motor to a power source while the motor residual voltage is high (the motor rotating at substantial

speed at the instant of reconnection) results in extremely high inrush currents which lead to breaker tripping, starter contact burning, motor damage and coupling damage.

The installation of magnetic overload relays eliminated the problem by preventing motor overload relay tripping.

At 2105 hrs, 14 March, the fuel oil burners were at high fire and preparations made to ignite D cell. After 32 hours of firing oil, D cell was ignited with coal. The lag time was due to problems with the Trane burners and difficulty fluidizing with cold air.

On 15 March at 1812 hrs, CBC coal fire was established. Bed height was increased to 30" (static) and held until 0344 hrs, 16 March, when C cell was ignited. At 0622 hrs, Monongahela Power Company ordered the MFB out of service due to low water levels in their deaerator and surge tanks. Later, with permission to relight CBC, coal fire was established at 0834 hrs. Preparations were made to relight C cell, but it was found that the DC-1 flyash transport system had malfunctioned and required repairing before transfer to C cell. At 2249 hrs, 16 March, coal fires were established in C cell and maintained until 0840 hrs the following day, when both cells were shut down to repair a broken spring and replace the rheostat on the CBC vibrating feeder.

At 1945 hrs, 17 March, repairs were completed and coal fires established in the CBC. At 2310 hrs, 17 March, C cell was ignited and fire maintained until 0404 hrs, 18 March. Coal fire was interrupted to C cell due to low auxiliary air flow on the fuel feed tubes. At 1108 hrs, 18 March, coal fire in C cell was re-established and maintained until 1250 hrs when coal flow was again lost to cell C. The problem appeared to have been coal jamming at the inlet of the vibrating feeders. At 1330 hrs, coal fire was extinguished and the unit cooled for inspection.

The MFB was placed back into service at 0645 hrs, 20 March, with coal ignition in the CBC. During the inspection, the general condition of the unit was found to be good with the exception of minor mechanical problems with auxiliary equipment. Additional sealing was done at the southeast corner of the CBC air distributor.

At 1220 hrs, the D-C slide gate was opened and a coal fire established in cell C. During the afternoon and evening, operation was hampered by flyash conveying problems. By 2200 hrs, these appeared solved. At 2000 hrs, an attempt was made to reinject flyash and from initial indications (oxygen,

bed temperature) there was some reinjection. A high hopper level in DC-1 prevented further testing and at 2030 hrs, the reinjection system was secured. The unit maintained steady state operation until 0405 hrs, 21 March, when the CBC vibrating feeder motor and rheostat failed. The problem was traced to the B phase of the circuit. C cell remained on line with the CBC slumped.

C cell remained in operation while D cell vibrating feeder motor and rheostat were repaired and tested for proper operation and the flyash reinjection system cleaned out.

Also during operation of C cell alone it was found that operating static bed heights greater than 25-inches contributed to coal furnace feed problems by causing bridging above belt feeders. Once the bed height was lowered to 24-inches, steady state operation was maintained with no problem. At 1427 hrs on 21 March, the fuel oil burners were lit. At 1452 hrs, the unit received a master fuel trip (MFT) which trips all running equipment. The alarm indication was FD and ID fan power loss.

As fan power had not been lost, equipment was restarted and coal fires in D and C cells established by 1506 hrs. The control room was notified that the cause of the trip was due to a loose wire on the FD and ID fan power loss relays.

A second MFT occurred at 1520 hrs while correcting wiring problems. Coal fires were re-established at 1526 hrs and the unit was brought back to stable condition.

The unit maintained steady state condition and flyash reinjection was attempted at 1840 hrs. At 2225 hrs, the reinjection system was shut down since it could not continuously transport material.

At 0720 hrs on 22 March, coal feed was cut to C cell and the cell taken out of operation due to low Mon Power condensate tank level. At the time, the MFB usage rate coupled with Monongahela Power's restriction on the raw water supply to the portable demineralizers, would not permit 24 hour 2-cell operation.

The CBC maintained steady operation until 0120 hrs, 23 March, when the vibrating feeder stopped operating. The cell was slumped and the problem found to be exposed wires in the motor. At 0528 hrs, a coal fire was re-established in D cell. Two hours later a coal fire was established in C cell.

Flyash reinjection was attempted at 0840 hrs. Coal feed was cut back and temperatures appeared good. Reinjection continued until 1006 hrs when DC #1 showed a high level alarm.

At 1315 hrs, a coal fire was established in B cell only and ended forty minutes later by a main fuel trip and a high steam trip. All cells were shut down. The fans were racked out at 0555 hrs, 24 March, and a weekend inspection of the unit began.

During the weekend a number of items were repaired and maintenance performed. C and D cell limestone rotary feeders were tested and calibrated. On 26 March, the unit was ready for service but start-up was delayed until 28 March due to a Monongahela Power tube leak in Boiler #7.

On 28 March, start-up was again delayed due to problems with the Monongahela Power's flyash system which threatened to shut down both their boilers.

At 1600 hrs, 28 March, release was given for MFB start-up. Boiler #7 developed a serious superheater leak and the unit was out of service for the remainder of the month.

Approval for MFB start-up was given at 1225 hrs on 1 April. Light-off procedures were implemented and equipment inspections conducted. At 1448 hrs, the north and south burners were lit and soon afterward the flame extinguished. Burner problems continued throughout the day and late into the night. The problems were corrected and both burners were at 100% fire rate by midnight.

At 0245 hrs, 2 April, a coal fire was established in the CBC and at 0704 hrs, combustion was transferred to C cell. Within thirty minutes coal feed problems developed in C cell. Three needle tubes became plugged and coal was not being fed by the south belt feeder. Coal feed was re-established by the equipment operators in fifteen minutes and smooth operation followed.

At 1356 hrs, flyash reinjection was begun at 10% with the coal feed hand/auto station at 80%. By 1436 hrs, flyash reinjection was at 100%. Coal feed was lowered to 42% with temperature maintained and flyash reinjection was confirmed. At 1450 hrs flyash reinjection was switched to by-pass mode.

At 1610 hrs on 2 April, coal feed to C cell was stopped; ten minutes later coal was halted to the CBC because of opacity of stack emissions.

Stack opacity was extremely high earlier in the day. At about 1300 hrs on 2 April, opacity was estimated at 60% and appeared orange-tan in color. At 1530 hrs, shortly before boiler shutdown, the stack cleared. After the unit was backed down, it was discovered that all ESP cabinets had tripped and all internals were shorted. Plans were made to cool the unit in order to perform a careful inspection and determine the cause of poor stack conditions and failure of the ESP. The following morning, 3 April, an internal inspection of the ESP was conducted.

Examination of the interior of the ESP revealed eleven broken wires in the inlet section and one each in the middle and outlet sections. The broken wires shorted out all ten emitting modules. It was also found that a considerable number of randomly located wires had stretched so that their weight hooks hung up on the support grid.

The broken emitting wires and their bottle weights were removed, marked as to their location and sent back to Buell for analysis. It is believed that the broken wires resulted from improper alignment of the precipitator collector plates, causing an electrical arc to be produced where emitting wires and collector plates were close to each other. The arc caused the wire to drop. The stretched and broken wires amounted to 12% of the total.

The remaining wires had approximately 1/16" uniform buildup of dark, fine ash, slightly oily to the touch. The inlet three modules exhibited a buildup of material that was hard and layered fine white ash. Some samples of the ash were taken from wires and plates and sent to Buell for analysis. Results will be reported when available.

Work continued on the ESP throughout 3 April and until 0300 hrs on 4 April. At that time the ESP was buttoned up and preparations made for light-off of the unit. At 0830 hrs, the burners were lit and brought up to 100%. By 1030 hrs, a coal fire was established in the CBC and by 1240 hrs, C cell was ignited. Coal feed problems developed at the belt feeder for C cell.

Precipitator performance during this period of operation was extremely poor. The opacity of the stack was almost 100% and the smoke nearly black. The control cabinets were found to be at low power or shorted. At 1337 hrs, shutdown of C cell started and at 1415 hrs the CBC was shut down. The shutdown was necessitated by excessive emissions.

On 5 April the entire unit was inspected. Following is a report of the inspection results.

ESP

Generally clean with no evidence of further build-up on either the plates or wires. All heavy dust deposits found earlier in the week were mostly gone and the hoppers were clean. One wire was broken in the D section and two bowed plates were discovered in the C and D sections, which explains why these cabinets had tripped at 600°F and restored on cooling down. These were removed.

APH Gas Inlet and Outlet

APH gas side was clean of any ash build-up or oil deposits. All gas side circumferential seals looked good and the ash deposits in the duct work were normal. The gas inlet straightening vane supports had bowed badly and showed evidence of carbon build-up. The gas outlet was virtually dust free but the expansion plate between the APH and ID Fan was bowed badly.

APH Air Inlet

This area was clean of ash and all probes and infrared detectors were clean. The FD outlet box expansion plate was badly bowed.

FD and ID Fans

Both fans looked good with no visible cracks on the blades or fan housings.

Furnace Uptake

Inspection of each cell uptake indicated that C cell was experiencing incomplete combustion on the northwest and southeast wall sections as noted by unburned coal deposits on the walls and generating tubes in these sections. A, B and C cell uptakes were clean and tube hangers were in good condition.

The uptake gas path had the usual light ash build-up but no smoldering was evident. The gas path entering DC-1 had approximately 12" carbon/coal smoldering on the isolation damper side where the material had no flow path.

Air Plenums

Each cell plenum looked clean with some material present on the floor of the plenums. The spare letdown grid section for each cell appeared to be leaking a considerable amount of material around the grid plate to the letdown seal.

DC-1 / DC-2

Both cyclones were clean with the exception of some ESP wire-hooks caught in three cyclone holes on DC-1.

D-Cell

The grid plate had about 8" of 1/2" rock on top; the bed was not empty as indicated on the control room chart. The bolt holes appeared to be clean but the bottom of the bolts, with the nuts attached, had bed material dust building up.

C-Cell

The grid plate had about 2" of material on it, located about 3" away from the top scallop bar. This indicates that peripheral leakage was greater around the edge. The plate was buckled and badly warped, with cracks and 1/4" to 1/2" gaps between the plate frames. The northwest and southeast corners of the cell were severely charred and the southeast tube support ladders were burned beyond recognition.

B-Cell

The top of the grid had 4" of material, much of which was low temperature clinkers. The grid plate had several holes but the general condition was good.

Because of the deteriorated condition of the C cell air distributor, it was decided to replace the grid with the spare. Removal of the worn grid began on 5 April and installation of the replacement was completed on April 14. A slightly different type of grid plate to scallop bar seal was utilized as shown on Figure 3.1.

During the unit outage, work was also accomplished on the ESP. Emitting wires that were broken, stretched or suspected to be out of line with the collector plates were removed. A total of 121 shrouded replacement wires were installed in Modules 2, 3, 4 and 8. The emitting wires that remained

undamaged were cleaned with brushes mounted on poles. The ESP was made ready for start-up on 16 April.

At 1043 hrs, 16 April, the fuel oil burners were both ignited and provisions made to light-off the CBC. Coal fire was established in the CBC at 1815 hrs and the transfer to C cell occurred at 2122 hrs. Operation continued smoothly until 0245 hrs, 17 April, when the south and then the north C cell belt feeders stopped running for a period of time. When the belt feeders were restarted, the temperature was too low for coal ignition. At 0633 hrs, C cell was reignited by transfer from the CBC. Three cell operation began at 0952 hrs despite a problem with the C-B transfer slidegate, which the control room was unable to close for quite some time. A half hour later, B cell underwent a high temperature fuel trip but was quickly recovered.

At 1208 hrs, the steam quality was approved and at 1457 hrs HPSV-1 was opened. The MFB began commercial operation for the first time since the Phoenix Outage. During the next four hours the unit developed coal feed problems twice, once due to master fuel trip on all coal feed. At 1842 hrs, coal flow was regained without interrupting turbine operation. The north B cell belt feeder was repaired and put back into service at 2003 hrs.

At 0520 hrs, 18 April, control of north and south B cell belt feeders was lost because of an electronic malfunction. The feed rate climbed to 80% and hand regulation was impossible. This terminated commercial operation after 5 hours 23 minutes.

The unit was backed down to D cell and held there until 1720 hrs, 19 April, when coal feed ceased to that cell. Operations began cooling the unit for inspection of the ESP hoppers for possible carryover of bed material. During the night these were inspected and a small amount was found in Hopper #1.

At 0400 hrs, 20 April, the burners were placed at high fire for light-off of the CBC. A coal fire was established at 0520 hrs in D cell and at 0955 hrs in C cell. On attempting to transfer combustion from C to B cell, it was discovered that the C-B transfer slidegate was open about 3" and would not entirely close. Three cell operation was aborted at this point. Preparations were begun for the weekend shutdown. Monongahela Power Company informed the control room that Boiler #7 would be shut down for the weekend. On Saturday, 21 April, Monongahela Power Company developed trouble with their flyash silo necessitating emptying the silo to perform maintenance. In the meantime, MFB flyash could not be discharged to the silo. Two days later this still hampered MFB operations.

Due to the silo problems, startup of the MFB was delayed until late on 23 April. The burners were lit at 2042 hrs, and a coal fire established at 2300 hrs in the CBC. An attempt was made to transfer to C cell at 0100 hrs, 24 April, but bed height appeared to be too great in C cell and combustion would not start. After draining material from C cell via transfer to B cell, and reinjection to the CBC from the bed material storage tank, the beds were at the proper height to attempt another transfer at 0513 hrs. A coal fire was established in C cell at 0515 hrs and in B cell at 0727 hrs. At approximately 1000 hrs, the steam quality was approved and at 1022 hrs, the steam valve was unseated manually to send steam to the turbine. Turbine operation continued until 1549 hrs, 24 April, when a faulty steam temperature limit forced Monongahela Power Company's interface steam valve to close at 885°F instead of 850°F.

Coal fires were maintained in all three cells despite a B cell high temperature fuel trip which occurred at 1606 hrs. Turbine operation began again at 1830 hrs and coal feed problems were present through midnight. At 0248 hrs on 25 April, high steam temperature caused pullout from the turbine followed by shutdown of B and then C cells at 0344 hrs.

Coal fire was discontinued in the CBC at 1154 hrs. Some routine repairs were made to the unit and preparations were made to light-off the following day. However, on 27 April, Monongahela Power Company advised they would not operate Boiler #7 over the weekend. The unavailability of feedwater delayed operation of the MFB until 30 April.

A coal fire was established in the CBC at 1100 hrs on 30 April; C cell was ignited at 1510 hrs, B cell transfer occurred at 1827 hrs, and commercial operation commenced at 2103 hrs. Some problems were encountered with fines plugging the north coal system, but these were overcome without interrupting the turbine run. At 1330 hrs, Monongahela Power Company reset the low temperature steam trip on HS-4 due to premature fuel trip during the past turbine run. At 1625 hrs, coal feed was lost on C cell north. The C-B and D-C slidegates were opened to sustain combustion. At 1637 hrs, B cell coal feed stopped and within five minutes, HS-4 and HPSV-1 were closed terminating the turbine run. The unit was backed down to D cell and held there until early the next morning. In the meantime, the belt feeders and rotary dump feeders were checked, cleaned and repaired, as required.

At 0055 hrs, 2 May, combustion was transferred to C cell and reinjection of spent bed material commenced to build bed

height for the B cell transfer. During the next three hours problems developed at the drain rotary feeders on all cells except the CBC. Because of bed height differentials, it is not possible to drain all beds through D cell drain rotary feeders. At 0400 hrs, coal feed was stopped to C cell and by 0430 hrs, all coal fires were out. Equipment was inspected and run to assure readiness. The drain rotary feeders were unplugged and operated. At 1307 hrs, 2 May, both burners were at 100% and at 1750 hrs, a coal fire was established in the CBC. Combustion was transferred to C cell at 2050 hrs. An attempt was made at 0056 hrs, 3 May, to light-off B cell but a defective silicon controlled rectifier in the north B cell belt feeder prevented the addition of a coal charge. A coal fire was established in B cell but was lost due to a feedwater flow permissive and further belt feeder problems. Another coal charge was added and B cell combustion was confirmed at 0256 hrs. The water chemist began superheated steam sampling at 0410 hrs and approved quality one-half hour later. Commercial operation began at 0455 hrs with coal feed in automatic throughout the afternoon. Routine operational problems were encountered but none serious until 2005 hrs when low steam temperature automatically closed HS-4. Two cell operation was maintained until 0045 hrs, 4 May, when it was decided to discontinue operation for routine maintenance and repairs.

Several outage-type items were accomplished and at 0756 hrs, 10 May, both burners were ignited in preparation to light-off D cell. At 1145 hrs, a coal fire was established in the CBC. After reinjection into C cell to build bed height, a coal charge was introduced. The combustion transfer attempt was unsuccessful due to faulty bed height indication on the control panel. The Hagan indicator showed twelve-inches while the height was only four-inches, thus preventing ignition. At 1645 hrs, C cell coal fire was established.

At 2035 hrs, the coal fire was transferred to B cell and commercial operation began at 2205 hrs, 10 May. The CBC temperature was boosted to 1800°F to commence flyash reinjection. At 2253 hrs, flyash was successfully reinjected for approximately one hour. The turbine run was maintained smoothly until 0304 hrs, 11 May. At that time, coal feed was lost to B cell and HS-4 closed. Cells C and B were shut down while the CBC was maintained until 0950 hrs. A complete unit inspection followed the shutdown and repairs made as required.

At 0755 hrs, 14 May, the burners were lit. A coal fire was established at 1110 hrs in the CBC. C cell combustion commenced at 1343 hrs but within three hours belt feeder

problems forced the shutdown of cell C. The problems were overcome and at 1420 hrs, 15 May, coal fire was transferred to B cell; steam entered the turbine at 1823 hrs. Commercial operation continued until 0356 hrs, 16 May, when the B cell south belt feeder plugged halting the turbine run. Equipment operators managed to continue B cell combustion for another two hours until five needle tubes plugged on B cell at 0505 hrs. The CBC and C cell operation was shut down about seven hours later.

Startup preparations began on the afternoon of 22 May. At 1615 hrs, the burners were firing at 100%. A coal fire was established at 1750 hrs in the CBC and the following day at 1235 hrs, 23 May, fire was transferred to C cell. At 1334 hrs, flyash reinjection began with coal feed to D cell cut back. D cell combustion continued but C cell was backed out at 2350 hrs, 24 May. D cell was extinguished at 1415 hrs, 24 May.

Routine maintenance and repairs were performed. Startup of the CBC occurred again on 30 May at 0825 hrs. Because of a serious leak in the flyash system, the unit was shut down thirty minutes later but came back on line at 0945 hrs. At 0611 hrs, 31 May, C cell was started up. Within an hour, a master fuel trip occurred but shutdown was prevented. At noon, coal fire was established in B cell. Belt feeder problems forced B cell shutdown at 1318 hrs. The entire unit was shut down at 1641 hrs to repair a circulation pump seal.

3.3.2 Forced Circulation Pump Failure

During the evening of 31 May, the MFB experienced a shutdown due to failure of the east boiler forced circulation pump seal. Disassembly and inspection of the pump commenced on the following day. Examination of the pump internals indicated more damage than a seal failure. The throttle bushing and shaft sleeve were worn on one side approximately 0.030". The upper throttle bushing and wear ring showed approximately 0.060" wear. Approximately 0.002" was machined off both the bushing holder and throttle bushing outside diameter.

Reassembly of the pump was completed on 8 June. When attempting to couple the pump shaft, it bound up inside the throttle bushing. A gasket was extruding just enough to allow the shaft sleeve to catch the inner spiral and draw it outward, binding the shaft.

On 9 June, the pump was again disassembled and throttle bushing removed. The inner gasket was machined and the casing pedestal and motor were assembled. All concentricities appeared normal. However, when alignments were made and the seal installed the following day, the pump was binding in the pump casing.

The pump was disassembled showing the impeller and casing wear ring scored slightly on one side. Upon removal, a severely scored impeller shaft was evident. Its bore was also scored in one area. A new shaft was installed. The runouts were measured after reassembly of the pump and an apparent high spot existed in the motor bearings causing the shaft to roll back in two spots. (This is uncommon in vertically mounted pumps.)

On 14 June the motor couplings were checked and found to have 0.002" runout. The pump shaft bore was found to have 0.007" runout but after final alignment this was reduced to 0.003". It was decided to have a new hub machined from 4340 aviation steel.

On 20 June it was discovered, in conversation with Mr. Baldo of United Centrifugal Pump, that the coupling and pump shaft assembly were not factory aligned. At that time, the motor was test run only to find severe vibration and noise from the bearing. The motor was disconnected and prepared for removal and transport.

As the pump was removed from the basement on 21 June, the motor was dropped, falling 12 feet to the concrete floor. Westinghouse was notified and requested to provide a full damage report. The damage was not extensive and repair began immediately.

The repaired east circulation pump motor arrived on site 28 June. By 1600 hrs that day, the motor had been placed on its pedestal. A new motor lead junction box was installed and the motor rewired.

On assembly of the pump coupling, the impeller bound up after 3/4 of a turn. The same difficulty resulted when the pump coupling was disassembled and reassembled several times.

An attempt was made to tighten the bolts on the side where high micrometer readings occurred, but the measurements remained the same. The assembly was raised and the pump coupling disconnected to assure that, upon reassembly, the wear ring did not interfere. Upon reassembly the wear ring runout was 0.014".

The impeller had to be removed and the wear ring machined to within required tolerances on 30 June.

<u>Impeller Wear Ring</u>	<u>Previous Runout</u>	<u>Final Runout</u>	<u>Pump Wear Ring Runout</u>
High	14.226	14.212	High 14.2515
Low	14.217	14.206	Low 14.249

A new pump cover casing gasket arrived and was installed. The seal was installed and readings taken as indicated:

	<u>Suspended</u>	<u>Cover Bolted</u>	<u>Impeller Coupling Unbolted</u>	<u>Impeller Coupling Bolted With Seal</u>
Motor Coupling	0.000"	0.002"	0.0005"	0.0002
Spacer				
<u>Top</u>	0.0005	0.002	0.002	0.002
<u>Bottom</u>	0.003	0.003		0.006
Impeller Coupling	0.0015	0.004		0.003
Impeller Shaft		0.005		0.003
Impeller Wear Ring	0.004			
Impeller Edge	0.0035			
Stuffing Box Bore to Stuffer	0.0015			0.003

From all indications an alignment problem exists between the cover and centerline of the case. This should not affect operations however.

3.3.3 Flyash Reinjection

The inability to feed flyash into the CBC with an eductor resulted in the design and installation of a screw feeder with a hollow shaft, see Figure 3.2. The screw was chosen for its ability to provide a positive means of material feed. The hollow shaft permits use of a transport medium for the material leaving the screw, which also cools the screw and prevents shaft distortion.

During the first hot test run, 10 May, the screw feeder ran material for 45 minutes. The screw jammed and material was found packed in the bearing. The screw feeder was disassembled and the bearing replaced with an air-injected bushing.

On the second hot test, 22 May, the screw jammed again as the result of flyash in the bushing. During this test, the reinjected material caused changes in bed temperature, O_2 , and reduction of coal feed to the CBC.

The unit was modified by replacing the bushing with a sealed bearing and leaving the injection air on the feeder side to prevent flyash from entering the bearing area. Cold tests were run under no load conditions to prove operation.

To improve feed, pipe was installed from the blower to the screw and a vent hopper and a 3" vent installed above the rotary feeder to prevent blow back. A diffuser was also installed at the screw nozzle discharge to insure a negative pressure inside the screw. As a result, the blower discharge pressure was increased from 10 psi to 15 psi. Additional cold tests were run to prove operation.

During the next operation run, the screw feeder will be tested under loaded (hot) conditions. If the bearing does not function and the feeder jams, the bearing will be moved outside the feeder housing and the inner bearing replaced with a mechanical seal. The nozzle pressure will also be increased to 60 psi by using a booster blower. As an alternate, the screw speed will be reduced so it will run full and provide a pressure seal.

3.4 Hollow Bolt Air Distributor

Based on successful tests at the Alexandria, Virginia facility, a decision was made to install a hollow bolt air distributor in the CBC of the MFB. The grid consists of twelve sections instead of the previous grid's nine. Figure 3.3 illustrates the hollow bolt and the hollow bolt spacing.

3.5 MFB Equipment Fire

This report describes the sequence of events and operations which led to a fire in the regenerative rotary air preheater on the MFB. The fire began about 2200 hrs on August 9, and was fought for 6 hours. The air preheater and induced draft fan were destroyed; the precipitator and some ductwork were damaged. Damage survey and removal of damaged equipment began at once as Project Phoenix.

The air preheater fire was similar to that which might occur in a conventional plant and was not directly related to AFB technology. Carbon rich deposits on the air preheater were ignited when the firing rate was reduced (lower air flow leads initially to higher metal temperatures). When air flow was increased to cool down the beds, combustion spread rapidly and either by flame front or detonation, involved the electrostatic precipitator.

3.5.1 Preincident Operation

The preincident plant operations described include the period of intermittent three cell commercial operation and testing from 31 July to 9 August.

Coal fires were established on 31 July in both cell C and cell D and the unit was stabilized in preparation for testing. The two cell testing was completed in 15 hours between August 1 and 2.

On 3 August, the unit was in three cell operation and steam exported to turbine number 5. Stable commercial operation and testing continued from 3 August till 5 August, when the unit was backed out of commercial operation because of debris clogging the south furnace feed system. Three cell commercial operation resumed on 6 August, and continued for 36 hours. At that time, cell B north belt feeder plugged and cell C north vibrating feeder strap failed. D and C cell operation continued while repairs were made. Cell B was reignited on 8 August and commercial operation resumed at a steam flow of 165,000 lb/hr.

On 9 August, the unit was backed out of commercial operation at 0406 hrs due to low pressure. At 0415 hrs, cell B was shut down because of coal feed problems. D and C cell operation continued and B cell was returned to service at 1330 hrs. Three cell operation to the turbine was restored at 1514 hrs. At 1530 hrs, the unit again experienced coal feed problems; both low level alarms on the coal bins tripped.

Coal feed was restored at 1600 hrs, and cell A was fluidized at 1700 hrs.

A check was made of the fans, APH, circulating pumps, and air compressor at 1800 hrs, and all were operating normally. The ID fan had been switched to manual control earlier in the day because of a draft swing, but it appeared to be working properly. At 2123 hrs, the unit was backed out of the turbine because of a bed temperature drop in cell C caused by coal feed problems. The south belt feeder was out of coal and the north belt feeder was functioning erratically. At 2210 hrs, operators first noticed flames coming from the MFB stack.

Unusual conditions noticed during this period of preincident operation were poor (ESP operation and a suspected boiler tube leak. For several days prior to the fire, the stack plume was heavy and had an unusual yellow tinge. On 9 August, a West Virginia Air Pollution Control Board Inspector threatened to shut down the plant because of this. It is believed that the stack condition was directly related to the fire.

● Salient Operating Factors

- No oil fire in the unit since July 31.
- Flyash temperature rose to 436°F at 1830 hrs on 9 August and remained there until 1915 hrs, 9 August (zero minus 3.5 hours to minus 2.75 hours).
- APH differential readings were taken by Foster Wheeler Service Engineers, Shift Supervisors and Operations Superintendent up to and including the day of the fire.
- APH was blown three times between 1445 hrs and 1707 hrs on 9 August (zero minus 7.25 hours to minus 4.9 hours). The recorded air temperature leaving APH indicated that the cleanliness factor was improved by blowing and that this temperature decreased by about 10°F from approximately 2000 to 2130 hours (zero minus 2.0 hours to minus 0.5 hours), then decreased later due to plugging.
- Flyash was not being reinjected. Analysis of flyash carbon content on 9 August showed 54.0%.
- Coal feed problems in C cell caused bed temperature in C cell to decrease and eventually caused the MFB to

be removed from turbine service at 2123 hrs (zero minus 0.61 hours) and reduced operation to firing coal in D cell only.

- Carbon content of bed material was 1.1%.
- MPC operator and nearby resident reported hearing an explosion about 2 to 3 minutes before flames stopped leaving stack.
- Some pluggage was noted in C cell economizer (about 5%).
- Fluidization was marginal at best due to grid plate leakage and heavier, larger limestone.
- C cell coal feed was heavier (17%) on the north side.
- Report from contractor hauling flyash from plant on 8 August said ash was hotter than usual.
- The West Virginia Air Pollution Control Commission complained of high opacity stack on 9 August.
- Rappers were placed in manual and rapped heavily by electrician between 2030 and 2130 hrs.
- Yellow stack was noted from 3 August until day of fire.

3.5.2 Cause of the Fire

When B and C cells shut down, the lessened cooling effect of the reduced air and gas flows caused ignition of the high carbon ash within the intermediate baskets of the APH.

When the B and C cell dampers were reopened, additional oxygen was supplied to the APH, increasing the quantity and intensity of the existing carbon ash fire to the point where metal ignition occurred. The intensity of the metal fire breached the APH air/gas seal interface, which resulted in a significant portion of the FD fan air passing through the combustion zone to the ID fan and out the stack. The thermal expansion of the gases coupled with the FD fan to ID fan short circuit, resulted in a reduction of the ID fan amperage and caused the high furnace pressure.

During the period before the B and C cell dampers were fully opened, choking and/or carbon monoxide formation occurred within C cell due to the carbon inventory which had not been

burned out. This combustible gas flowed with the air leakage from C cell dampers and primarily settled in ESP Hoppers 1 and 2, but was also carried to the APH by the draft from the APH after metal ignition and air/gas interface breaching. The combustible gas concentration was sufficient to support combustion but was not explosive. The combustion travelled back through the ducts and ESP to Hoppers 1 and 2, where the gas concentration was suitable for an explosion.

● Supporting Data

C cell bed temperature increased at a time consistent with the Control Room Operator's report that he fully opened the B and C cell dampers. This supports the occurrence of a coal/carbon monoxide formation and/or coking.

The APH vendor's data suggests that APH fires generally occur on load reductions, which result in a movement of the APH temperature gradient toward the cold end, causing ignition of condensed combustibles in the intermediate or cold end baskets. This occurrence does not require an outside source of ignition.

Inspection of the damaged APH indicated that most of the metal ignition occurred within the intermediate baskets, which are "shielded" from sparklers entering with the hot flue gas. No abnormal changes in APH hot air temperature or draft change occurred from the time that the sootblower was shut down until metal combustion was extensive. For this reason, it was concluded there was no rapid or extensive APH fouling.

ESP damage would suggest that detonation was low and that the detonation gas density was probably greater than the density of the flue gas. The damage also suggests that the heavier gas settled in the low velocity stream entering the ESP.

A low detonation would explain the bottle disengagement from the lower wire hooks, disengagement of the upper wire hooks from the top support, brittle fracture of the wires, and extruded fracture of the wires.

While the ignition source of the ESP detonation, i.e., C cell or the APH, cannot be conclusively established, the bowing of the inlet perforated plate would suggest a wave front movement counter to normal flow, or from the APH.

A low detonation would not necessarily bulge the ESP shell since the curtains would tend to dissipate the wave front prior to impact on the shell, but would not significantly reduce the wave front toward the inlet end or toward the center field.

- Refuting Data

If the ignition source was from the APH and an explosive gas mixture existed within the ESP, then one would expect more extensive damage to the west (outlet) field. If one assumes a combustible gas heavier than air and recognizes the low stream velocity through the ESP, one can picture Hoppers 1 and 2 acting as an inverted Y causing a different gas concentration within the hoppers than that in the general gas stream through the center and west (outlet) fields.

If one assumes that the general gas stream concentration was sufficient to support combustion but was not explosive, then in the 1-2 fps stream velocity a flame could propagate from the APH through the ESP to Hoppers 1 and 2 where the gas concentration is an explosive mixture. This would minimize damage through the outlet field and maximize damage in the inlet field.

3.5.3 Changes to Reduce Possibility of Future Fires and Aid Operations

The following modifications were accomplished after the fire to assure a similar situation does not occur again. These changes also permit more effective monitoring of boiler conditions.

- Inlet and outlet temperature indicators were installed for the ESP and APH in the Control Room.
- A furnace draft recorder was installed.
- Gai-tronic loudspeaker box locations were increased for better personnel communications.
- An APH deluge system was installed in a strategic location and is now operable.
- An infrared fire detection system was installed in the APH and the alarm located in the Control Room.

- The APH cleaning stream supply was modified to come from two sources, regardless of MFB load.
- Operator training was increased during November.
- Combustion air dampers were modified to enable tight shut off of idle cells.
- APH pressure differential gauges were placed in the Control Room.
- Temperature gauges for the ash silo were relocated to the Control Room.
- The alarm system was modified to include an APH high temperature alarm for both gas and air sides.
- The ESP cabinet breakers were modified by adding an alarm light to come on when breakers trip.
- A reading log for periodic ESP cabinet readings was installed.
- Rotary feeder jam alarms were installed.
- A carbon dioxide fire inerting system was installed in the coal bunkers.

3.5.4 Firefighting

The air preheater fire was first observed at about 2150 hrs on 9 August. The Rivesville Volunteer Fire Department (RVFD), PER management and Monongahela Power Company (MPC) were notified. At about 2225 hrs, the first hose was put into action to combat the fire, as employees and other fire departments arrived to help. The fire was controlled, and the last fire truck departed at approximately 0230 hrs on 10 August. The APH was hosed until the morning of 11 August to remove the residual heat.

After the first report of fire, the control room operator received a high furnace pressure alarm on Cells A, B, C and D.

The furnace pressure rose to +8 inches water column and the ID fan current dropped below 50 amperes (less than half normal operating current). The control room operator adjusted the ID fan inlet dampers, but with no effect. At this time, the equipment operator took control of the feedwater to the

boiler and at 2156 hrs, the control room operator tripped the coal feed to cell D. Both the FD and ID fans then tripped automatically, while the equipment operator continued to maintain drum level.

The Shift Supervisor told the control room operator to stop coal feed to cell D and went outside to inspect the stack. He told the control room operator to monitor the water level and to leave the fans on as long as possible. At 2158 hrs, the Shift Supervisor called MPC, the Marion County Fire and Rescue Squad, Project Manager, and the Operations Superintendent to notify them of the fire.

The RVFD truck arrived at 2215 hrs, and the Shift Supervisor went to meet the fire chief. Additional fire departments responding were the Grant Town Volunteer Fire Department, the Fairmont Fire Department and the Winfield Volunteer Fire Department. The Rivesville ambulance and the Marion County Rescue Squad also responded.

● Assessment of Fire

Firefighting continued through the night. The aerial truck and fire departments began to leave the plant at 0200 hrs. The RVFD was the last to leave, between 0230 and 0245 hrs. For the rest of the night, plant personnel rotated shifts on the roof, with two people spraying and one person watching at all times.

3.5.5 Equipment Damage

After the fire was contained, cool down and inspection began. Without fans, cooling of the boiler and all equipment took several days. The Rivesville staff was assigned to undertake inspection, dry out, cleanup and system checkout. To assure complete inspection, PER, New York, and Stone and Webster discipline specialists, as well as major equipment vendor representatives, also inspected fire and water damaged apparatus. Representatives of insurers, FW, MERC, DOE and Mitre/Metrek also reviewed damage.

● Air Preheater

The air preheater rotating parts were destroyed or damaged beyond repair. Fire was especially intense around the center post where combustion of metal reduced the heat exchange surface to slag.

Support bearings required replacement.

Casing maintained sufficient integrity to contain the fire but was badly distorted from overheating and required replacement.

Air preheater was completely disassembled, lowered to ground and irreparable components scrapped.

- Induced Draft Fan

The ID Fan was subjected to temperatures high enough to establish metal combustion. South side of rotor was melted and the inboard shroud completely consumed. The rotor shaft forging was severely cracked, north bearing housing exploded, and the south bearing failed. Anti-vortexing vanes and casing were badly warped and breached.

The motor was removed to a shop for disassembly and inspection. The cooling air screen was somewhat warped from overheating and shaft keyway was distorted.

Inlet dampers were scorched but appeared reusable.

- Electrostatic Precipitator

The ESP suffered major but indirect damage. The ESP is a conventional wire and plate design, located in the gas stream between the mechanical dust collectors and the air preheater.

Inspection revealed the precipitator to have suffered extensive internal damage. The inlet (East) section was ninety percent destroyed as most wires were broken and down, plates distorted and alignment frames (bottle guides) had collapsed. Approximately seventy percent of the collecting plates in the center section were buckled mainly toward the north end where nearly all the wires were down. Collecting plate damage was less severe toward the precipitator outlet (West). Hoppers were clean and there was no evidence of hopper fire or actual combustion within the precipitator.

It had previously been determined that the seam between the penthouse and the precipitator proper had been inadequately welded at original construction. This defect permitted rain water to enter the unit. Repair was scheduled for August/September.

- Ductwork

Ductwork around the air preheater suffered both directly and indirectly from combustion. The old gas duct (between air preheater and ID fan) and breeching (between ID fan and stack) were both severely buckled, had slag deposits and were burned through at several locations. Expansion joints and fairing plates were buckled and distorted.

Dampers between the air preheater and ID fan were warped and discolored but were reuseable. All other ducts, dampers and joints were undamaged.

- FD Fan

The FD fan showed no apparent damage. It was necessary to dismantle this fan and remove it from the roof in order to accomplish air preheater disassembly. The fan was inspected in great detail after it was on the ground. The motor was shop inspected.

- Stack

The stack was discolored but showed no physical damage or distortion. After a conference with U.S. Steel Metallurgists, we concluded the stack was not stressed beyond its yield point and required no repair. To confirm condition, coupons were cut for lab examination.

- Boiler

The boiler required repairs because of damage done by the weight of water accumulating in the combustion air plenum (estimated at seventy to one hundred tons). The plenum was cracked through in several locations and generally bowed. A water wall tube leak was detected and repaired in the web wall between A and B cells. This pinhole leak probably resulted from the weight of fire fighting water on the entire boiler structure.

- Electrical Gear

ID and FD fan motors were removed to a motor shop for disassembly and detailed inspection. Flooding from fire fighting water was concentrated at the northwest corner of the building. As this corner also contains the instrument room, control

room, lab and MCC area, wetting of electrical apparatus was extensive. Inspection revealed no major damage, but drying out and insulation checks took weeks.

- Instrumentation

Major water damage to instrumentation systems and laboratory equipment was averted by prompt protective measures. A laborious sequential checkout assured equipment integrity, particularly in the control room area.

Some apparatus, particularly sensors and panel at the top of the air preheater, (ID fan sensors) were damaged beyond repair by the fire and required replacement.

- Material Transport Systems

There was no known damage to coal, limestone, ash or bed transport systems. However, because of the soaking of electrical apparatus, equipment was subjected to an item by item checkout.

Six hundred and fifty tons of coal were in the bunkers at the time of the fire. Disposal of this coal was necessary to avoid the danger of a bunker fire. Coal handling system was therefore checked out first and coal disposed.

- Bunkers and Other Auxiliary Systems

Beyond destruction of compressed air and cooling water piping in the air preheater area, no damage to auxiliary systems occurred.

- Structure and Building

Asphalt roofing in both low roof and north high roof areas was badly damaged. Some flashing and miscellaneous roofing repairs were also damaged and replaced.

- Cleanup

Principle cleanup problems included removal of wetted bed material (forty to fifty tons) and demolition of damaged high roof equipment. Vacuum truck service was used for bed material removal. Roof demolition began immediately. A tower crane was set up 19 August.

- Fire Apparatus

No fire fighting apparatus was damaged, though a major reconsideration of equipment requirements was initiated.

3.5.6 Conclusions and Recommendations

The following conclusions have been drawn regarding the firefighting activities:

1. The 25 - 30 minute delay between the report of the fire and the arrival of personnel on the north low roof allowed the fire to progress and precluded application of water through the water wash system. Insufficient quantity and/or improper firefighting gear and inexperience of operating personnel may have contributed to the delay in undertaking firefighting activities.
2. The water wash system was inadequate to quench the metal fire alone but may have reduced or controlled the size of the fire. However, valve location prevented application of these firefighting resources.
3. The lack of readily available hose stations and the length of hose existing stations hampered firefighting activities.
4. Access/egress around the equipment, i.e., narrow platforms and vertical ladders, hampered firefighting activities.

The following changes and/or additions were made to aid in firefighting activities in the future:

1. The APH wash water valves and steam sootblowing valves were moved from under the APH.
2. The steam soot valve on top of the APH was eliminated.
3. Additional fire hose stations were installed on the high roof and the north low roof.
4. Quick-opening access doors were installed in the duct work to gain rapid access. Much time was lost in trying to remove access panel nuts with a wrench.

5. Additional personnel protection equipment has been provided including additional asbestos gloves, fire jackets and oxygen breathing apparatus.
6. A manually activated deluge system was installed on the APH with a control station in the Number 2 bunker area and the control room.
7. A new infrared APH fire detection system was installed with the alarm located in the control room.
8. Fifty feet of hose was added to the existing hose stations on the operating floor.
9. An additional hose station with at least 75' of hose was installed near column line D6/D7 on the operating floor to improve coverage on the north and east sides of the MFB.

3.6 Plant Reconstruction and Improvements

3.6.1 Major Tasks

Following the fire, a detailed schedule was developed for reconstruction of the plant and improvements.

Major tasks for reconstructing of the plant included:

1. Demolition and reconstruction of the rotary air preheater.
2. Demolition and reconstruction of the electrostatic precipitator internals.
3. Replacement of the induced draft fan, repair of the forced draft fan rotor.
4. Replacement of surrounding ductwork, re-insulation.

Plant improvement items included:

1. Fire protection system including an APH deluge system.
2. Improvement of the combustion air damper sealing.

3. Design changes to the ESP to increase collection efficiency.
4. Modifications to the instrument and control systems.

3.6.2 Plant Modifications/Improvements

Reconstruction of the fire damage provided the opportunity to implement pending plant modifications and improvements. The background of some of these improvements is described in the following sections.

Fuel and Limestone Handling Systems

A recurring interference with plant operations affecting both MPC and MFB, has been dust from several sources: limestone unloading and conveying, coal unloading and conveying, and flyash and bed material waste transport. Not only does this cause personnel discomfort and possible health hazard, but in the case of coal fines, produces a potentially explosive atmosphere. During the pneumatic unloading of limestone a large amount of very fine dust is produced, originating as very small (-20 mesh) limestone particles entrained on larger limestone particles. The standard procedure for capturing these dust particles was to utilize the intake of the induced draft fan to vacuum the area. The ID fan uptake is physically located close to the limestone bunker area. A vent baghouse was installed in September for the limestone bunker area. The area was also sealed more effectively using solid steel plate rather than deck grating. The baghouse, in conjunction with the ID fan, has eliminated dust problems associated with limestone unloading.

Coal dust problems were created by the 102' Redler conveyor which dumped into Bunker #2. The Redler was replaced by a non-attriting belt conveyor. By the nature of its operation, the Redler produced many fines which later caused problems downstream of Bunker #2. With the 75' belt conveyor, these problems were essentially eliminated.

To further reduce propagation of coal dust, the coal dryer, crusher, and their respective ductwork were removed from the plant. In December, the Rotex screen was moved from its position just above the operating floor on the east end of the boiler to the coal handling room. It is now positioned between the distribution hopper and the 75' belt conveyor redirecting any oversized coal back to the crusher. The volume of coal fines arriving at the plant site has also

diminished markedly as a result of several visits to the tipple point in Smithfield, Pennsylvania, to advise the vendor on truck loading and screening procedures. (See Exhibits 9 and 10.)

During May, access to the MPC coal unloading hopper generally used for the MFB was limited because of increased tonnage received by MPC. There are two additional MPC unloading stations which could be used for the MFB. Both stations include a hammer mill in the material handling path which would create fines unacceptable for MFB use. To overcome this condition and provide additional MFB coal unloading flexibility, a by-pass was fabricated around one of the hammer mills. This permits MFB coal to be unloaded at the inclined belt hopper and transported to Bunker #5 or #6 without creating additional fines. Coal can be unloaded at the rate of 260 tons per hour.

A potential fire hazard existed in the flyash system because of the level detectors which were located on the electrostatic precipitator bins. These detectors allowed air to enter the hot flyash stored in the hoppers. This air/combustibles mixture produced a possible flammable situation. Nuclear level detectors, installed during the reconstruction outage, have an airtight seal which prevents fresh air necessary for combustion from entering the hopper.

3.7 Emissions and Performance Testing

During May through August, a series of tests were conducted on the MFB to measure the performance of the unit. Prior to May, an instrumentation package had been installed to enable performance measurements to be made. Throughout the test period, malfunctions of various instrument components hindered testing efforts. Major malfunctions developed in the Beckman Gas Analyzer and the Autoweigh coal feeder belt scales.

Eleven tests were performed to study sulfur capture, combustion efficiency, boiler efficiency and gas emissions. Exhibit 12 lists the tests performed, dates, duration and number of cells in service.

Due to the malfunction, tests 10, 14 and 15, were sufficiently complete to represent unit performance.

Test data from runs 7, 11, 12 and 13 (all D cell tests), are incomplete due to a malfunction of the coal belt scale. Boiler blowdown is relatively high during single cell operation and, as this value is estimated, D cell coal feed rate could not be accurately calculated.

Tests 16 and 17 are not reported due to malfunctions with limestone and/or bed material dropout rotary feeders.

Test 9 was interrupted half-way through the test period due to high noise levels from steam vents.

Test 8 is not reported due to inconsistencies in the gas analysis test data.

Test Results

The results and test conditions for tests 10, 14 and 15 are shown in Tables 3.7 through 3.9. In all three tests, the carbon burnup cell was operated in a coal firing mode, which accounts for the combustion efficiencies.

Due to the malfunctioning Autoweigh belt feeders, reliable coal feed rates could not be measured. Therefore, the coal feed rates were calculated using the boiler duty, loss analysis, and heat balance. The heat balance gives the total coal feed rate. In order to determine the coal feed rate to the individual cells, a heat balance around the bed in D cell was conducted to estimate the coal feed rate to that cell (blowdown is zero during multi-cell testing).

With the calculated coal feed rates, excess air calculations were made to account for discrepancies between analyzers measuring percent O_2 in the flue gas. From the calculations, it was determined that the Beckman gas analysis system was leaking; therefore, gas fractions read during the test were diluted. Using a dilution factor, gas values were calculated (sample calculation shown later in this report).

In test 14 and 15, sulfur capture was running above levels reported for Greer Limestone by the Alexandria FBM. With higher Ca/S ratios and higher bed temperatures (1500 - 1550°F), sulfur capture will improve.

Sulfur capture in test 10 was inconsistent with results from tests 14 and 15. Possible explanations for this inconsistency are:

1. An undetermined malfunction of the gas analysis system.
2. Lower operating temperatures in all cells.

3. None of the three beds had reached a state of chemical equilibrium, resulting in overly sulfated beds.

NO_x emissions are well within EPA limits for the overall unit. D cell readings are high, but contribute only a small fraction of total gas flow from the unit.

Combustion efficiency results are reasonable and within the range of expected values. One noted highlight is D cell combustion efficiency. In test 10, D cell was operated with 8 coal feed points, while in tests 14 and 15, only 4 coal feed points were used. It appears that little change in combustion efficiency occurred, but D cell was also operated at higher temperatures in tests 14 and 15.

Boiler efficiency was calculated using ASME's loss method and is within the expected usage.

Conclusion

Although the test results show reasonable combustion efficiencies, operating bed temperatures were lower than desired. Due to equipment problems in the bed material drain system, static bed levels were difficult to control at the desired height. This resulted in additional heat removal from the beds. Limestone feed had to be continued in order to accomplish sulfur capture, leading to the increased bed heights.

In future tests, it may prove beneficial to operate the MFB at one set of conditions for combustion efficiency data and at another set for sulfur capture data. This would reduce the constraints on operational parameters and permit a full range of correlating data.

SO₂ monitoring and Ca/S feed ratios will also be improved to provide more reliable and acceptable results.

3.8 Test Plan Revision

The MFB test plan was reevaluated after the plant fire. Data collection and calculation procedures were re-examined. Problems which had arisen during previous test periods were analyzed and solutions sought. The result was a complete revision of test plan procedures, calculations and data collection methods. Table 3.12 shows a list of problems and solutions for unit testing.

A summary of test plans for the Rivesville MFB unit for the June through September 1979 period follows. Resumption of testing will represent the first formal testing at the facility following the fire.

A series of tests will be performed during stable operating runs to obtain steady state performance and emissions data. Conditions to be determined include:

- General operating characteristics
- Boiler efficiency
- Combustion efficiency
- Heat transfer rates
- SO₂ and NO_x emissions
- Sorbent effectiveness
- Solids elutriation

Reliability and accuracy of the data will be the primary objective. While longer duration in-turbine runs are desirable to demonstrate long term operability, reliability of test results will determine the actual run duration.

The overall objective of this testing period is to develop baseline performance and emissions characteristics of the MFB. The baseline data generated will allow comparisons to be made between the relatively large Rivesville beds with data from PDU's, giving a direct indication of scale-up effects on performance and emissions. Evaluation of scale-up effects will provide a better basis for future AFB designs and contribute to the commercialization of AFB combustion technology.

The baseline performance and emissions characteristics will be defined through a matrix of tests at nominal design conditions for the unit, with variations of process variables around the design point. The extent of variation of process variables will be restricted to conditions anticipated to provide desired performance and emissions and by the interrelation of variables imposed by design and operating restrictions. Two sets of process variables, one set for B and C cells and one for D cell, determine the operation and ultimately the performance and emission characteristics of the overall facility. In addition, performance and emission characteristics for the main cells, B and C, and the CBC are individually determined by the selected conditions.

The process variables which define a set of operating conditions for the main cells include coal feed rate, calcium/sulfur mole ratio, steam flow rate and conditions, bed temperature, superficial velocity, bed height, excess oxygen in flue gas, and bed drain rate. Of the process variables, only the Ca/S mole ratio is essentially independent. The remaining variables are interrelated and limited by the design and need to produce turbine quality steam. Venting of steam not within the turbine limits cannot be done for long periods because of limitations on boiler water feed. The turbine provides recycle of boiler water and is required during testing if equilibrium conditions are to be maintained. The interrelationships and limitations of the six process variables in the main cells are:

Coal Feed Rate: Through stoichiometry this variable is related to superficial velocity and excess O₂. Coal feed rate determines the heat input to the system.

Calcium/Sulfur Mole Ratio: The calcium/sulfur mole ratio defines the limestone feed rate relative to the coal feed rate and percent sulfur in the coal.

Steam Flow Rate and Conditions: Steam flow rate is determined through the heat transfer rates and overall heat balance.

Bed Temperature: Bed temperature is determined by a heat balance on the particular cell. Effects of heat input by coal feed rate, heat removal rates by total air and removal by the in-bed heat transfer surface determine bed temperature. In-bed surface is determined by bed height since varying bed height increases or decreases the number of in-bed tubes and the extent of the water wall contact with the bed. Bed temperature strongly influences sulfur capture and combustion efficiency.

Superficial Velocity: Superficial velocity also to a large extent determines carbon carryover, carbon combustion efficiency in the main cells and ultimately the heating value of the flyash injected into D cell.

Bed Height: Bed height affects the heat balance by determining heat transfer surface. By design, the MFB is limited to a maximum bed depth of 28-inches static. Maintaining acceptable sulfur capture requires running bed depth near the maximum.

Excess O₂: This variable is related to velocity and coal feed rate through stoichiometry. In addition, excess O₂ effects combustion efficiency and NO_x. Narrow limits on excess oxygen from about 3 to 5% are required to insure reducing conditions are not generated (lower limit) while the upper limit is set to limit heat removal from the bed by increased flue gas volume.

Bed Drain Rate: The rate at which bed solids are removed is set essentially by the limestone feed rate, i.e., the Ca/S mole ratio. Bed drain rate controls bed height. Operating conditions are controlled essentially by:

- Setting superficial velocity.
- Controlling excess oxygen to the desired level by varying coal feed rate.
- Controlling bed temperature to the desired level by adjusting bed height.

Effectively, calcium/sulfur, velocity, excess O₂ and bed temperature in the main cell can be independently set within limits while bed depth, coal feed rate and other variables are fixed by these choices.

Process variables in the CBC include those variables noted for the main cells, and in addition, a flyash to fuel ratio. These variables are interrelated just as the main cell variables with the following exceptions:

- Flyash to coal ratio is determined by heating value and quality of flyash available from the main cells.
- Bed temperature of the CBC is maintained at a higher level than the main cell to encourage carbon combustion. An upper limit of 2000°F is set by material and ash softening considerations. A lower limit of 1800°F is suggested by small scale tests of CBC's to affect fines combustion. The higher temperature will substantially reduce SO₂ capture and selection of the CBC operating temperature is effectively a tradeoff between combustion efficiency and sulfur capture in the CBC bed.

- Superficial Velocity: By design, the superficial velocity is lower in the CBC than in the main cells to encourage combustion of fines.
- Bed Height: To encourage combustion, bed heights are deeper.
- Excess Oxygen is maintained higher in the CBC than the main cells to encourage carbon combustion. As with the main cells, higher excess oxygen rates require lower bed height to maintain heat balance.

Effectively, with the CBC flyash to total fuel ratio, superficial velocity, excess oxygen and bed temperature can be set independently within operating limits. The remaining process variables are determined by these choices. Table 3.13 summarizes the range of the process variables obtainable in the facility.

To meet the total objective of defining baseline and emissions characteristics, a sequence of tests will be performed at and around the nominal design conditions for the facility. Testing will be divided into two phases.

Phase I will emphasize defining combustion characteristics of the main cells and CBC individually and the system as a whole. Tests in Phase I will utilize a high Ca/S mole ratio to allow rapid equilibration of SO₂ emissions. The primary variables to be explored in Phase I testing are the CBC temperature, CBC excess oxygen, the main cells temperature and the main cell excess oxygen. A matrix of test conditions to define these effects is given in Table 3.14.

Between Test Groups 1, 2 and 3, the CBC temperature is varied to determine its effect on combustion efficiency. Within each of these group tests are two levels of excess oxygen. Test Groups 4 and 5, along with Group 1, vary the main cell bed temperature with, again, a variation in excess oxygen. Because of the limitations on the range of achievable excess oxygen, the duplication of the change among the various tests will be used to determine the trend of the effect of this variable on performance. Variation of bed height is undesired, but necessary, to maintain operations. Existing PDU correlations of bed height versus residence time will be used to back up these effects in performance and emissions. Also, direct comparison of tests should allow separation from kinetic effects such as temperature and excess oxygen.

Phase II of the test program emphasizes characterizing emissions rates in the main cells, the CBC individually, and the system as a whole. In particular, the effect of varying Ca/S mole ratio on SO₂ emissions will be defined to establish a minimum Ca/S ratio needed to maintain various levels of sulfur capture. Several bed temperatures around the 1550°F optimum for limestone established by PDU experiments will be considered. During these tests, the temperature, excess oxygen levels and bed height in the CBC will be selected to optimize system combustion efficiency based on Phase I tests.

Table 3.15 shows a test matrix for Phase II testing. Main cell bed temperature is varied between Test Groups 1, 2 and 3. Within each of these groups Ca/S is varied at constant bed temperature. In these tests, excess oxygen level will be allowed to vary to maintain a constant bed level. Tests in Group 4 will attempt to vary velocity and bed height thereby varying residence time.

To carry out the test program, a plan has been developed involving 1250 hours of run time. Most of this operating time will be the 3-cell mode representing commercial operation with integrated data acquisition periods. The test plan requirements will dictate actual operating conditions. Review of data obtained during testing may result in modifying the test plan and schedule to repeat tests or to explore unexpected test results.

The overall schedule for operating the plant is shown in Table 3.16 which shows a total of 15 weeks of testing. Assuming no unscheduled outages, this schedule will allow for 39 data acquisition periods. Twenty-four periods are required by the test matrices. Six periods are allocated for duplication or repeat tests and the remaining nine may be consumed in a 500 hour test, if this option is chosen.

The planned sequence of events for a typical operation run is indicated in Table 3.17, showing operating steps for preparation, ignition, building inventory and testing. A typical schedule showing the operations on a week time frame is shown in Table 3.18. The schedule shows a midnight Sunday start, with Monday devoted to bringing the MFB on line, followed by establishing test conditions through Tuesday morning. Data acquisition periods would occur from Tuesday through Friday, with one period per day.

The details and procedures for collection of raw data and subsequent calculations are described in the manual "Rivesville Performance Test Guide". These procedures will generate a

summary of results which tabulates operating conditions along with sulfur capture, SO₂ emissions, NO_x emissions, combustion efficiency, boiler efficiency and material balances. Performance and emissions results will be determined for the carbon burnup cell and the main cells separately, along with overall system characteristics.

Results from Phase I testing will allow combustion efficiency and sulfur capture tradeoff for the CBC cell under various process conditions to be examined to determine optimum conditions for CBC operations. Phase II (along with Phase I results) will define emissions responses.

Figure 3.7 illustrates the steps needed to process the data taken during data acquisition periods. Preliminary test results are expected to be available two days following data collection. These preliminary results will utilize expected chemical analysis of solids samples. These preliminary results summaries will be used to flag data problems early and allow rescheduling of tests early. Final test results summaries will require 35 days to complete.

3.8.1 Airflow Calibrations

During January and February, calibration tests were performed on all airflow measuring devices in the combustion air and auxiliary air systems. These calibrations were necessitated from previous performance tests, where airflow measurements were in doubt. Test conditions for the Auxiliary FD Fan tests are shown in Table 3.19 and for the FD-ID Fan tests in Table 3.20. In both sets of airflow tests, a pitot traverse of the appropriate fan discharge was performed. Sketches of pitot traverse locations and points are shown in Table 3.20 and Figure 3.12.

Prior to testing, all sensing lines were leak checked and blown out. All instruments were zeroed and checked. In all cases, where applicable, instruments were placed as close as possible to the measuring device.

In all tests, conditions were allowed to stabilize before data collection began. Data was taken at five minute intervals and two complete pitot traverses were performed for each condition.

In order to maintain similar conditions for various test runs, duct pressure was closely monitored and restored to same pressure for all conditions.

Duct temperatures were monitored by a Doric Datalogger every five minutes.

Equations used for data reduction and test results are as follows:

$$\rho = .09658 [P_D + (P_B) (13.595)] / (T_D + 460) \quad (1)$$

where:

ρ = air density (lb/ft³)

.09658 = density constant @ 60% RH (lb/ft³)

P_D = duct pressure ("Hg)

P_B = barometric pressure ("Hg)

T_D = duct temperature (°F)

$$V_p = .1096.2 \sqrt{\Delta P_p / \rho} \quad (2)$$

Where:

V_p = pitot velocity (fpm) (from mfg's equation)

P_p = pitot differential pressure ("WC)

Auxiliary FD Fan Test

Aux FD Fan Pitot - Venturi

$$M_{AFD} = 49044.7 \sqrt{\rho \Delta P_{AFD}} \quad (3)$$

where:

M_{AFD} = Auxiliary FD Fan airflow (lb/hr)

ρ = Aux FD air density (lb/ft³)

T_D = Aux FD air temperature (°F)

ΔP_D = Aux FD fan press ("WC)

P_{AFD} = Aux FD fan pitot - venturi differential ("WC)

Vibrating Feeder Annubars

$$M_{VF} = 9470.96 \sqrt{\rho \Delta P_{VF}} \quad (4)$$

Where:

M_{BRN} = burner airflow

$\rho \Delta P_{BN}$ = burner annubar differential ("WC)

ρ = density of air at burner (lb/ft³)

Seal air in belt feeder leakage air could not be determined. For testing, seal air will be valved out.

FD - ID Test

Airfoil Airflow

$$M_{AF} = 1,333,256 \rho \Delta P_{AF} \quad (5)$$

M_{AF} = airfoil airflow (lb/hr)

$\rho \Delta P_{AF}$ = airfoil differential ("WC)

ρ = density of air at air foil (lb/ft³)

Cell Annubars

$$\text{A Cell} \quad M_{AANN} = 237,903 [\sqrt{\rho \Delta P_{AN}} + \sqrt{\rho \Delta P_{AS}}] \quad (6)$$

$$\text{B Cell} \quad M_{BANN} = 213,142 [\sqrt{\rho \Delta P_{BN}} + \sqrt{\rho \Delta P_{BS}}] \quad (7)$$

$$\text{C Cell} \quad M_{CANN} = 225,226 [\sqrt{\rho \Delta P_{CN}} + \sqrt{\rho \Delta P_{CS}}] \quad (8)$$

$$\text{D Cell} \quad M_{DANN} = 222,497 [\sqrt{\rho \Delta P_{DA}}] \quad (9)$$

where:

M_{XANN} = cell airflow from annubar (lb/hr)

ΔP_{xN} = cell north annubar differential ("WC)

ΔP_{xS} = cell south annubar differential ("WC)

ΔP_{DA} = D cell annubar differential ("WC)

ρ = density of air at annubar (lb/ft³)

Gridplate ΔP 's vs airflow

$$A \text{ Cell} \quad M_{\text{AGRID}} = 119,417 \sqrt{\rho \Delta P_{\text{AG}}} \quad (10)$$

$$B \text{ Cell} \quad M_{\text{BGRID}} = 119,417 \sqrt{\rho \Delta P_{\text{BC}}} \quad (11)$$

$$C \text{ Cell} \quad M_{\text{CGRID}} = 225,266 \sqrt{\rho \Delta P_{\text{CG}}} \quad (12)$$

$$D \text{ Cell} \quad M_{\text{xGRID}} = 42,998 \sqrt{\rho \Delta P_{\text{DG}}} \quad (13)$$

where:

M_{xGRID} = cell airflow from gridplate differential (lb/hr)

ρ = air density based on undergrid press ("WC)

ΔP_{xG} = cell grid plate differential ("WC)

Main Cells Annubar

$$M_{\text{MN}} = 2,603,922 \sqrt{\rho \Delta P_{\text{MN}}} \quad (14)$$

where:

M_{MN} = main cells annubar airflow (lb/hr)

ΔP_{MN} = main cells annubar differential ("WC)

ρ = density of air at annubar (lb/ft³)

D Cell Pitot - Venturies

$$M_{\text{DP}} = 67,716.73 \sqrt{\rho \Delta P_{\text{DE}}} + \sqrt{\rho \Delta P_{\text{DW}}} \quad (15)$$

where:

M = D cell airflow from pitot - venturies

ΔP_{DE} = D cell east pitot - venturi differential ("WC)

ΔP_{DW} = D cell west pitot - venturi differential ("WC)

ρ = density of air at pitot - venturi (lb/ft³)

Based on observations made during performance tests, airflow tests and Alexandria experience, the following recommendations are made:

1. Until further testing can be performed on gridplate ΔP vs airflow, this measurement should not be used with a "hot" cell. Based on data accumulated at Alexandria, this function changes greatly between hot and cold conditions. Further tests will be performed on gridplate ΔP as the unit is placed in service.
2. Due to leakage rates through A cell, A cell dampers should be blanked prior to testing.
3. Aux FD fan seal air should remain valved out during testing. This will not inhibit unit operation and will allow less error in total airflow.

Based on a review of all data, the expected error is $\pm 5\%$.

Damper leakages for each cell are as follows:

<u>Cell</u>	<u>Leakage (lb/hr)</u>
A	30,000
B	29,500
C	23,000
D	12,400

3.9 ESP and Dust Collector Performance

During April and May, several tests were conducted to determine the performance of the electrostatic precipitator (ESP) and the two mechanical collectors. A total of eight tests, seven by Buell personnel and one by GCA/PER personnel, were conducted on the ESP. Dust Collectors 1 and 2 were each tested twice by GCA/PER personnel. The two dust collector tests were run in conjunction with five ESP tests.

Performance testing of the ESP and dust collectors consisted of obtaining isokinetic samples at the inlet and outlet of each collector while maintaining the unit at constant operating conditions. Figure 3.13 illustrates the sampling locations. Due to physical limitations, the ESP outlet was sampled

downstream of the ID fan. The outlet conditions for Dust Collector No. 1 (DC-1) were calculated by difference from the ESP inlet and Dust Collector No. 2 (DC-2) outlet.

Test conditions for the ESP were determined by the manufacturer, Buell Envirotech. Buell had requested maximum firing rate for three cell operation. Test conditions are shown in Table 3.22. Collector efficiency is defined as weight of dust collected divided by weight of dust entering.

Sampling methodology was identical for both Buell and GCA. Both groups used modified EPA Method 5 trains to obtain samples. GCA sampled using a 48 point grid network at three minutes per point at DC-1 inlet, DC-2 inlet and DC-2 outlet. Buell used a 20 point grid network at three minutes per point at the ESP inlet and stack.

Test results are shown in Table 3.23. Disregarding the test labeled GCA #1, the average efficiency of the ESP was 88.6%. Since a visible dust plume was emitted from the stack during the tests, a low ESP efficiency was expected. The dust plume was not just characteristic of the tests, but is visible whenever the unit is in operation. Buell is aware of this situation and is investigating the problem.

Performance of the dust collectors had not been measured prior to these tests and they were assumed to be performing at or near design conditions. Design conditions and test conditions are shown in Table 3.24. The calculated pressure drops were derived from DC-1 inlet and ESP inlet pressures.

As shown in Table 3.24, neither dust collector is performing at design efficiency, however, the MFB was not operated at the specified design conditions. DC-1 was designed for A, B and C cells in operation, whereas, these tests were conducted with B and C in operation. DC-2 was designed for flyash reinjection into D cell, which was not in operation. The decreased DC-1 loading flow rate and pressure drop associated with these deviations from design conditions would be expected to lead to a decreased efficiency.

Figures 3.14 and 3.15 show manufacturer's efficiency curves for Dust Collectors 1 and 2. The dotted line in Figure 3.14 represents a predicted 10" WC drop line. In Figure 3.14, Pre-test 3 value was too low to fit on the curve. Pre-test 4 value on Figure 3.15 also exceeded curve limits.

Based on Figures 3.14 and 3.15, the low observed efficiencies of the dust collectors could be the result of small dust particle size. Both dust collectors were rated at 10% of

the solids flow less than 10 microns and DC-2 may be receiving up to 50% under 10 microns. Particulate size distribution tests could not be performed on the samples collected. However, Buell performed particle size distribution determinations on material entering the ESP. These results are shown in Table 3.25. By means of mass balance, it was estimated that the amount of -10 micron material entering the dust collectors was approximately 20-30% for these tests. This value is significantly above design specifications.

If the origin of the excessive flyash fines could be determined, some insight as to the cause of this problem would be obtained. The chemical analysis of the flyash was used with iron content as a tracer and ash content as a basis to determine relative contributions of coal derived ash and limestone derived ash in the ash fraction of the flyash. This is possible because the iron/ash ratio is high in coal and very low and constant in limestone. Table 3.26 shows details of this flyash component analysis including a sample calculation. Table 3.27 shows the overall and component flowrates and efficiencies. It can be seen that the wide variation in overall efficiency is largely due to unexplained wide fluctuations in the amount of limestone derived flyash entering each dust collector. Not only was the amount of coal ash entering each dust collector substantially constant, but its collection efficiency was also substantially constant. The outlet rates for collector remained about constant. Since the -10u fines entering the ESP were about identical for the two tests at about 60% of the total mass, it is suggested that most of these ultrafine particles were ultimately derived from the coal. It is noted that the volatiles, consisting primarily of carbon, were collected at a relatively high efficiency in all cases. The coal derived ash was always collected at a lower efficiency. This could be explained if the coal derived ash of completely burned out flyash was smaller in particle size or much less dense than carbon containing coal derived material. It is likely that the completely burned out coal ash is smaller in size which explains its lower collection efficiency.

Another problem with DC-1 is the low pressure drop. If pressure drop can be increased, DC-1 efficiency should also increase. One method of increasing the pressure drop across DC-1 is to operate with the DC-1 isolation damper closed. This damper was installed to isolate half of DC-1 during C cell operation. With DC-1 isolation damper closed, 4" WC pressure drop would occur bringing DC-1 efficiency to approximately 87%.

Conclusions from these tests are:

1. Both dust collectors appear to be receiving high quantities of -10 micron dust. Size distribution analysis of future samples should be performed in order to confirm this.
2. The unit should operate in 3 cell operation with the DC-1 isolation damper closed.
3. Further testing should be performed to confirm these results.

3.10 Characterization Tests of C-Cell (Venting Tests)

The Rivesville MFB provides an opportunity to study three different designs of fluidized bed combustors at the largest available scale. These are represented by D cell, C cell and B cell. The scales of these units are 72, 120 and 144 ft², respectively. Heat transfer surfaces are wall only, evaporating bundle and superheating bundle. Freeboards range from as high as 10 feet to as low as structurally feasible. A program is proceeding to individually characterize the performance of each cell to determine the ranges of conditions to be studied in the formal test program. Also, peculiarities of beds of this scale will be identified and a performance data base for large fluidized bed combustion systems of these designs will be developed. Parameters to be studied are bed expansion versus bed height, heat transfer to submerged surfaces, combustion chemistry, freeboard chemistry, steam production variation (turndown), and gaseous emissions. Both experimental and theoretical studies are in progress. On May 23 and 31, two test sequences were conducted on C cell which were the initial tests to characterize the performance of C cell. The performance of C cell was monitored while superficial velocity, bed depth or temperature were varied. Gaseous emissions, steam production and bed operating parameters were noted. Also visual inspection of the hot bed with an observation of expanded bed height was made at each point. Oxygen levels sampled at the top of the economizer bundle and measured by the Beckman gas analysis system were a primary dependent variable.

These test conditions are shown in Table 3.28. Specific data points are numbered in the order they were taken. Bed temperatures ranged between 1450°F and 1550°F. Static bed height, determined by pressure drop across the fluid bed, ranged from 18 to 29 inches. Superficial velocities varied from 7.2 to 12.7 ft/sec. These velocities are calculated as

the sum of C cell annubar and auxiliary air. Excess oxygen in the C cell economizer ranged from 3.0% to 8.5%. A preliminary analysis of this data has been completed while a more comprehensive analysis continues. The following discussion indicates the direction of this analysis and some tentative conclusions.

The theoretical comparison of actual with predicted performance is based on an In-Bed Heat Balance. This is represented by one of several computer programs currently under development, and is described in detail elsewhere in this report. In this method all heat inputs and outputs are calculated as well as excess air, calculated bed temperature and calculated coal feed rate. Inputs and outputs required are flow rates, chemistries and temperatures as well as bed geometry, bed chemistry, bed expansion and heat transfer parameters. Any of these parameters can be varied in successive computer runs to give insights into their relative importance as a variable. Thus, by comparison of actual experimental performance variation with calculated variation, insights into fluidized bed combustion can be obtained which can be useful, not only in performance prediction, but also in FBC design.

An example of this type of calculation is shown in Figure 3.16. This represents, based upon data obtained at the Alexandria Laboratory, a first guess theoretical estimate of C cell performance at 1500°F and superficial velocities of 8, 10 and 12 feet/second. The theoretical volumetric oxygen level in the cell exit gases are presented as a function of static bed height. The heat balance dictates that when none of the tube bundle is submerged (i.e., a static bed height of less than about 10-inches) and the temperature of the flame is maintained low, at 1550°F, large amounts of excess air must be used to carry away the heat generated. As the tube bundle becomes more submerged on increased bed height, the heat load is increasingly shifted to the tube bundle, less excess air is required and the flue O₂ decreases. On submergence of the tube bundle, further increases in bed height do not add significant amounts of additional heat transfer surface and, therefore, the exit O₂ becomes approximately constant with further increases of bed height. Figure 3.17 shows the calculated C cell exit oxygen as a function of superficial velocity and bed temperature. It should be noted that these are preliminary calculations yet to be correlated to experimental data. C cell exit oxygen is shown to increase approximately as a straight line with superficial velocity. An increase in bed temperature decreases the exit oxygen level significantly.

The excess oxygen versus superficial velocity data measured experimentally is shown in Figure 3.18 & 3.19. The dotted line in both figures represents full submergence of the in-bed tube bundle. Visual observations were made of effective operating bed height for all points except numbers 1 and 2 on 23 May. The tube bundle was observed to be totally submerged for points 3 - 8 on 23 May, and points 4 and 5 on 31 May. Results confirm that once the in-bed tube bundle is completely submerged, any increase in bed depth has little effect on oxygen consumption. This is in agreement with theoretical calculations based on an in-bed balance.

Again referring to Figures 3.18 & 3.19 it can be seen that submerged tube bundle lines are offset from each other. This agrees with heat balance calculations which indicate that the exit oxygen level will decrease about one percent per each 50°F increase compared to 23 May, accounting for the offset in the limiting lines.

These lines on Figures 3.18 & 3.19 can be used as the basis of a correlation. Since it is known that points on this line represent a completely submerged tube bundle, referring to Figure 3.18, it is noted that the O_2 level remains approximately constant for all bed heights where the bundle is totally submerged. Therefore, these data lie on a horizontal line on an O_2 versus bed height curve. This is plotted on Figure 3.20. The curves generated from this data differ markedly from those of Figure 3.16. Whereas the small scale test unit bed expansion data upon which the theoretical calculations were based predicted a gradual bed expansion through the tube bundle up to perhaps 28-inches static bed height, in practice it is found that at 20-inches static bed height the tube bundle is completely submerged except at the lowest superficial velocity (7.2 ft/sec). This rapid expansion through the bed was not expected and may be a phenomenon resulting from circulation and turbulence of large scale beds which will be clarified in further tests. Another significant result of this can be applied to achieve turndown. It can be seen that in the region of 18 to 20 inches bed height, the velocity can be decreased from 12 ft/sec to 8 ft/sec without changing the exit O_2 . The bed subsides through the tube bundle at this height, thereby maintaining the heat balance upon turndown.

Also, preliminary indications are that the heat transfer coefficient indicated by these data is lower than that used for design of the unit. The design outside coefficient for submerged tube surface was 40 Btu/M ft² °F. Preliminary calculations show that the experimental heat transfer coefficient may be somewhat less than this for the large scale

tube bundle in C cell. This difference may be due to operating conditions or other parameters and will continue to be studied.

The major factor discovered by this analysis is that the design conditions of 12 ft/sec superficial velocity, 1550°F and 3% O₂ in the exit gases cannot be met simultaneously. This is due to the lower than expected heat transfer coefficient and the resulting undersize of the submerged tube bundle. Two of the three conditions can be met with deviation of the third. These combinations are (1) 12 ft/sec, 1550°F, 7% O₂; (2) 12 ft/sec, 1650°F, 3% O₂; and (3) 9 ft/sec, 1550°F, 3% O₂. These and other possible combinations will be considered in specifying the desired test conditions in the upcoming formal test program.

During these tests, feedwater flow (steam production), was measured at each condition. Figure 3.21 shows the relationship of feedwater flow and superficial velocity. Results show that with a fully submerged tube bundle, the largest effect on steam production is bed temperature. Superficial velocity affects steam flow only when the tube bundle is not completely submerged.

Gaseous emissions from C cell appear to be mainly affected by the percentage of excess oxygen in the flue gas. Generally, CO and hydrocarbons increased as oxygen decreased as can be seen in Figure 3.22. In the 3% to 5% range of excess oxygen, hydrocarbons averaged 1000 ppm and CO averaged 2500 ppm. These values are approximately the ranges found in the Alexandria FBM and PDU studies. However, compared to these studies, higher levels of CO and hydrocarbons were exhibited in the main uptake and air preheater inlet. Three possible explanations for this condition are:

1. Poor sample point location;
2. Poor feed distribution in the bed; or
3. Flue gas combustion or other reactions.

Investigations are being made to determine the cause of this discrepancy.

Poor feed distribution could result from the present feed tube configuration. The 16-1/2° feed needles and 33° feed needles terminate within one foot of each other and about two feet from the north and south water walls. This causes more coal feed to the north and south walls and not enough to the center of the bed. Locating the feed needle discharges

closer to the center of the bed and air distributor may improve fuel distribution and lower CO and hydrocarbon emissions.

In summary:

1. Individual cell fluidized bed combustion parameters are being characterized to aid in future operations and future design.
2. In-Bed Heat Balance computer programs are useful as theoretical guides in the interpretation of this data.
3. Bed expansion is more rapid through the C cell tube bundle than expected from pilot scale tests.
4. Bed heights in the region of 18 - 20 inches can be used for turndown.
5. The heat transfer capabilities of the C cell tube bundle are less than design resulting in an under-sized tube bundle. This has led to identifying appropriate ranges to operate the boiler.
6. Steam flow can be affected by turndown of velocity or changing temperature.
7. Carbon monoxide and hydrocarbon emissions decrease with oxygen level and approach design levels at design operating conditions.

As a result of the work, the following actions will be taken:

1. Further tests will be performed on each cell. Particular emphasis will be on testing those conditions expected to be near the optimum.
2. A traverse will be made at C cell uptake to determine if the gas analysis sampling ports are located properly.
3. The needles in C cell will be lengthened to provide better coal distribution.

3.11 Results From Pre-Tests 3 and 4

On 2 and 3 May, two preliminary unit performance tests, designated Pre-Tests 3 and 4, were conducted. These tests were designated "pre-tests" because the MFB plan was not approved at the time of their performance. The primary objective of these pre-tests was to check out all data collection equipment and methodologies to be used in the upcoming test plan. Data reduction procedures would also be checked using data produced from these tests. The results of these two data acquisition dry runs have provided much useful information on test instrument performance and/or availability. The following deficiencies were identified by Pre-tests 3 and 4.

1. The final superheater outlet pressure gauge readings were unreliable; therefore, the control room indications were used.
2. The FD fan and Aux. FD fan inlet and outlet thermocouples were not operable, which led to an evaluation of all Doric thermocouples. The values reported are from previous tests at similar conditions.
3. Because of thermocouple breakage and in-bundle bed temperatures, bed temperatures could not be computer averaged.
4. The backup gas analysis system was not available due to instrument failure. All gas analysis was performed by the Beckman gas analysis system.
5. Limestone and bed material rotation counters were not available. Flowrates were derived from control room settings.
6. C-north and B-south limestone rotary feeders were not operable.
7. Bed drain system was not fully operational.
8. Belt feeder integrators were not fully operational for both pre-tests.

All deficiencies have been noted and corrected.

Pre-Tests 3 and 4 data was entered into the computer to check program continuity and preliminary unit performance. Test results and conditons are shown in Tables 3.29 and 3.30. These tables are the "final calculation" printout of the

data reduction computer program. Both pre-tests were run under Phase I conditions, emphasizing combustion efficiency and sulfur capture is not reported.

Results from Pre-Test 3 indicate high combustion efficiency, 87% in the main cells. This can be attributed to the higher bed temperatures of B and C cells. The combustion efficiency in Pre-Test 4 was 77%, due to an increased flyash rate of 2-1/2 times that of Pre-Test 3. This is being investigated.

Hydrocarbons and CO in the gas stream showed a steady increase at sampling points throughout the system from the cell to the air preheater inlet. This is described in the operations section of this report.

The results of Pre-Test 3 are based on calculated coal feedrates as opposed to measured rates due to C cell integrators malfunction. All but one integrator was operational during Pre-Test 4. Measured coal rates were within 1% of the calculated values. It should be emphasized that Pre-Tests 3 and 4 were not originally intended for data reduction and the results should be treated accordingly.

3.12 List of Figures (Section III)

"C" CELL GRID PLATE TO SCALLOP BAR SEAL

D Cell Hollow Bolt Spacing
3.57

Figure 3.3

PREVIOUS COAL HANDLING SYSTEM

Figure 3.5

Figure 3.6

DATA PROCESSING FLOW DIAGRAM

AUX AIR SAMPLING LOCATION

Circular Duct - ID = 19.25"
Number of Sampling Points = 20

Aux Air Duct Sampling Location

Traverse Point Number on Diameter	Traverse Point Location on Diameter (Inches)
1	1.0
2	1.6
3	2.8
4	4.4
5	6.6
6	12.7
7	14.9
8	16.4
9	17.7
10	18.2

- 0. PD Fan Inlet Damper
- 1. PD Fan
- 2. PD Fan Discharge Damper
- 3. Air Preheater
- 4. Airfoil
- 5. Pitot Traverse
- 6. D Cell Annubar
- 7. D Cell Pitot - Venturi (2)
- 8. D Cell Damper
- 9. D Cell Gridplate
- 10. Main Cells Annubar
- 11. C Cell Annubar (2)
- 12. C Cell Damper
- 13. C Cell Gridplate
- 14. B Cell Annubar (2)
- 15. B Cell Damper
- 16. B Cell Gridplate
- 17. A Cell Annubar (2)
- 18. A Cell Damper
- 19. A Cell Gridplate

FORCED DRAFT FAN CIRCUIT

MAIN AIR SAMPLING LOCATION

Duct Dimensions - 7' X 7'6" (84" X 90")

Diameters (Equivalent) - 86.9"

Number of Sampling Points - 30

MAIN AIR DUCT SAMPLING CONFIGURATION

- 1. DC-2 Inlet
- 2. DC-2 Outlet
- 3. DC- Inlet
- 4. ESP Inlet
- 5. Stack

ISOKINETIC SAMPLING LOCATIONS

DC-1 Collector Efficiency *

Overall
Collector
Efficiency
% By
Weight

* Curves provided by manufacturer: JOY MANUFACTURING CO., INC.

Note: Curve is based on a minimum inlet dust concentration of .8 gr/acf

○ Value from pre-test 4

DC-2 Collector Efficiency *

% Material Under 10 Microns

Note: Curve is based on a minimum inlet dust concentration of .8 gr/acf

* Curves provided by manufacturers: JOY MANUFACTURING CO., INC.

○ Value from pre-test 3

Figure 3.15

Theoretical Oxygen Concentration
versus Bed Height Relationship
for C cell at 1550°F

First approximation based upon
small scale combustor experience
and "In-Bed Heat Balance"
computer program.

Figure 3.16

PRELIMINARY CALCULATION OF C CELL EXIT
OXYGEN AS A FUNCTION OF SUPERFICIAL VELOCITY

Figure 3.17

C Cell
Exit
Oxygen %

Oxygen levels at C cell exit as a function
of C cell superficial velocity
Data of 5-23-79

C Cell
Exit
Oxygen

Note: Points 1, 2 & 3 were
observed to be below the
top of the in bed tube bundle.

Oxygen level at C cell exit as a function
of C cell superficial velocity
Data of 5-31-79

Figure 3.19

Experimented C Cell Oxygen Versus Bed Height Data and Correlation

Feedwater
Flow
 10^3 lb/hr

Feedwater flow as a function of C cell
superficial velocity

Figure 3.21

Experimental C Cell CO and Hydrocarbon Levels

Figure 3.22

SIZE DISTRIBUTION ANALYSIS

3.13 List of Tables (Section III)

Test No. _____

Date Completed: 04-18-77

Test Date: 04-14-77

Analysis By: Kuhl

Sample Type: Limestone

Sample No. Truck 3

Table 3.1

U.S.S. SIEVE	WEIGHT (grms)	(X_i) FRACTION	FRACTION SMALLER THAN	(d_p) MEAN PARTICLE DIAMETER (μ)	X_i/d_p
+6 407.0	9.7	0.008			
-6 +8 432.3	35.0	0.030			
-8 +10 498.8	101.5	0.088			
-10 +12 575.0	177.7	0.154			
-12 +14 616.1	218.8	0.189			
-14 +16 531.9	134.6	0.116			
-16 +20 531.3	134.0	0.116			
-20 +40 530.7	133.4	0.115			
-40 609.0	211.7	0.183			
TOTALS	1156.4	1.0000			

Sample Wt. (before screening) 1160.3 grms

% Diff. _____

Sample Mean Particle Size 654.46 μ

3.77

SIZE DISTRIBUTION ANALYSIS

Test No. _____

Date Completed: 07-26-78

Test Date: 07-26-79

Analysis By: _____

Sample Type: Limestone

Sample No. #7

Table 3.2

U.S.S. SIEVE	WEIGHT (grms)	(Xi) FRACTION	FRACTION SMALLER THAN	(d _p) MEAN PARTICLE DIAMETER (u)	Xi / d _p
+½	0	0.00			
+½ -½	6.90	0.99			
-½ +5	69.20	9.94			
-5 +10	514.80	73.95	Would not fluidize in		
-10 +14	87.40	12.56	D-cell w/80" pressure.		
-14 +20	11.10	1.59			
-20	6.70	0.96			
TOTALS	696.10	99.99 1.0000			

Sample Wt. (before screening) 698.50 grms % Diff. _____

Sample Mean Particle Size 2630.01 u

POPE, EVANS AND ROBBINS, INC.

RIVESVILLE, WEST VIRGINIA

MFB YEARLY OPERATING SUMMARY

Table 3.3

Period July 78 - June 79

OPERATING HOURS

	<u>Month</u>	<u>To Date</u>
Oil Fired - CBC	<u>166.18</u> hrs.	<u>1637.93</u> hrs.
Coal Fired - CBC	<u>786.03</u> hrs.	<u>2745.86</u> hrs.
- C Cell	<u>450.65</u> hrs.	<u>980.35</u> hrs.
- B Cell	<u>204.85</u> hrs.	<u>444.97</u> hrs.
- A Cell	<u>0.00</u> hrs.	<u>5.75</u> hrs.
Commercial Operation	<u>172.50</u> hrs.	<u>269.50</u> hrs.

Total Coal Consumption	<u>8,097.477</u> lbs.	<u>4,048.74</u> T.
Total Sorbent Consumption	<u>2,252,494</u> lbs.	<u>1,126.25</u> T.
Approximate Feedwater Consumption	<u>87,799,769</u> lbs.	gals.
Approximate Steam Vented	<u>46,318,880</u> lbs.	<u>5,555,088</u> gals.
Approximate Blow Down	<u>12,367,000</u> lbs.	<u>1,483,411</u> gals.
Commercial Steam Generation	<u>28,849,500</u> lbs.	<u>3,459,434</u> gals.
Total Fuel Oil Consumption		<u>34,159</u> gals.
Demineralized Water Credit		<u>6,925,662</u> gals.

Number of Starts

CBC	<u>22</u>
C Cell	<u>28</u>
B Cell	<u>13</u>
A Cell	<u>0</u>

Multicell Operation

D	<u>346.02</u> hrs.
DC	<u>235.16</u> hrs.
DCB	<u>204.85</u> hrs.
DCBA	<u>0.00</u> hrs.

COAL AND LIMESTONE ANALYSIS FOR

JULY 1978 - JUNE 1979

Table 3.4

	Jul	Aug	Sep	Oct	Nov	Dec	Jan	Feb	Mar	Apr	May	Jun	Average	TOTAL
COAL														
% Carbon	—	70.62	—	—	—	—	72.11	72.09	69.18	68.52	68.41	67.62	69.79	—
% Hydrogen	—	4.82	—	—	—	—	5.25	5.55	4.38	4.89	4.91	4.81	4.94	—
% Sulfur	—	2.13	—	—	—	—	1.69	2.74	3.00	2.45	2.88	2.59	2.50	—
% Ash	—	13.81	—	—	—	—	12.40	14.17	15.5	15.9	17.16	17.16	15.28	—
% H ₂ O (Hyg)	—	0.84	—	—	—	—	1.03	0.75	0.71	0.9	0.84	0.90	0.85	—
% H ₂ O (Free)	—	1.78	—	—	—	—	3.2	0.9	1.54	1.53	0.86	0.78	1.51	—
Btu / lb.	—	12676	—	—	—	—	12843	12864	12426	12258	12244	12143	12493	—
Size (Microns)	—	4521	—	—	—	—	2502	4401	2421	3467	4507	4164	3712	—
Tons Delivered	0.0	1789	0.0	0.0	0.0	0.0	22	414	1236	3055	1146	191		6853
LIMESTONE														
% Calcium	—	33.0	—	—	—	—	38.41	38.3	37.02	37.49	37.49	37.67	37.05	—
% Sulfur	—	0.03	—	—	—	—	0.05	0.03	0.02	0.00	0.05	0.05	0.03	—
% LOI	—	37.98	—	—	—	—	41.83	41.59	40.21	41.09	40.72	40.86	40.61	—
% H ₂ O	—	0.15	—	—	—	—	0.30	0.28	0.19	0.17	0.18	0.14	0.20	—
% Magnesium	—	NA	—	—	—	—	0.49	0.49	NA	NA	NA	NA	0.49	—
Size (Microns)	—	2428	—	—	—	—	NA	2298	1712	1978	2181	2059	2109	—
Tons Delivered	0.0	330	0.0	0.0	0.0	0.0	18	78	305	196	159	22	—	1108

Table 3.5
Contributing Events and Major Evidence

	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>	<u>C</u>
Incomplete Combustion/Freeboard Burning	x		x		x	x
Rapping of ESP	x	x				
Grid plate air leakage	x				x	x
Uneven coal feed	x		x		x	
Opening of B & C cell dampers	x	x	x	x	x	x
ID fan breakdown				x		
<u>Major Evidence</u>						
Poor stack conditions		x	x		x	x
Carbon deposits on economizer tubes	x		x		x	x
Furnace uptake H ₂ O readings	x		x		x	
C cell analyzer readings	x		x		x	x
Economizer steaming	x		x		x	
ESP roof heat					x	
High APH pressure differential readings			x	x		
Uneven ESP damage			x		x	x
ID fan bearing explosion	x				x	
Char and buildup on ESP plates & wires					x	
<div style="display: flex; justify-content: space-between; margin-top: 20px;"> <div style="width: 30%;"> <p>-Incomplete Combustion in C Cell. causing ignition in ESP and APH</p> <p>-Ignition of hot ash in APH</p> <p>-Independent APH & ESP fires caused by glowing ash</p> <p>-Ignition of APH deposits with flashback detonation ESP</p> <p>-Repeated combustion in ESP glowing ash carryover to APH</p> <p>-Spontaneous Ignition of ash in APH, flashback into ESP</p> </div> <div style="width: 65%;"></div> </div>						

C = Consensus

Table 3.6

<u>Test</u>	<u>Date</u>	<u>Duration (hrs)</u>	<u>Cells In Service</u>
7	05/16/78	4	D
8	05/22/78	4	D & C
9	05/23/78	2.5	D, C, B & A
10	05/25/78	4	D, C & B
11	06/13/79	4	D
12	06/14/78	4	D
13	06/15/78	4	D
14	08/01/78	4	D & C
15	08/02/78	4	D & C
16	08/04/78	4	D, C & B
17	08/06/78	4	D, C & B
<u>Pretest</u>			
1*	04/17/79	1.5	D, C & B
2*	04/24/79	1.75	D, C & B
3	05/01/79	3.25	D, C & B
4	05/03/79	4	D, C & B
<u>O₂ C Cell</u>			
1	05/24/79	1.70	D, C
2	05/31/79	3.5	D, C
<u>Buell</u>			
1	04/24/79	1	D, C & B
2	04/26/79	1	D
3	05/01/79	1	D, C & B
4	05/01/79	1	D, C & B
5	05/03/79	1	D, C & B
6	05/03/79	1	D, C & B
7	05/03/79	1	D, C & B

*Pretests 1 and 2 were data training periods.

Tests Performed, Dates, Duration and Number of Cells in Operation.

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

Table 3.7

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	98,000	5
Steam Flow, lb/hr	164,000			6
Steam Conditions, psig/°F	1324/921			
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	3,489		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			

TEST REPORT October 30, 1978
 TEST NUMBER 14
 TEST DATE August 1, 1978

Table 3.8

OPERATING CONDITIONS	CELL		NOTES
	D	C	
Coal Feed, lb/hr	1,787	11,366	1
Limestone Feed, lb/hr	200	1,398	2
Coal % S	3.29	3.29	3
Ca/S (mol ratio)	1.0	1.1	4
Main Air Flow, lb/hr	30,400	127,100	5
Steam Flow, lb/hr	123,400		6
Steam Conditions, psig/°F	1218/567		
Bed Temperature, °F	1,556	1,423	7
Vs, fps	5.6	13.2	8
Static Bed Height, inches	28.6	28.6	9
Oxygen in Flue Gas, %	9.9	4.6	10
Bed Drain, lb/hr	450	750	11
Flyash, lb/hr	679	3,338	12
TEST RESULTS			13
Sulfur Capture, %	47	63	
SO ₂ , lb/10 ⁶ Btu	3.00	2.99	
NO _x , lb/10 ⁶ Btu	1.27	0.46	
Combustion Efficiency, %	86	79	
Boiler Efficiency, %	71		

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

TEST REPORT October 30, 1978
TEST NUMBER 15
TEST DATE August 2, 1978

Table 3.9

OPERATING CONDITIONS	CELL		NOTES
	D	C	
Coal Feed, lb/hr	1,657	10,925	1
Limestone Feed, lb/hr	200	878	2
Coal % S	3.86	3.86	3
Ca/S (mol ratio)	0.9	0.6	4
Main Air Flow, lb/hr	31,400	129,100	5
Steam Flow, lb/hr	112,700		6
Steam Conditions, psig/°F	1233/569		
Bed Temperature, °F	1,550	1,440	8
Vs, fps	5.9	13.6	
Static Bed Height, inches	25.2	27.8	9
Oxygen in Flue Gas, %	9.7	4.0	10
Bed Drain, lb/hr	125	250	11
Flyash, lb/hr	540	3,018	12
TEST RESULTS			13
Sulfur Capture, %	31	32	
SO ₂ , lb/10 ⁶ Btu	4.48	4.45	
NO _x , lb/10 ⁶ Btu	1.18	0.43	
Combustion Efficiency, %	89	77	
Boiler Efficiency, %	69		

Table 3.10

Test No:	10	14	15
M BTU/HR			
Boiler Duty	187.93	112.89	100.38
Unburned Carbon Loss	39.53	30.76	29.71
Dry Gas Loss	15.92	12.83	11.78
Fuel Moisture Loss	.02	.05	.04
Hydrogen Moisture Loss	1.04	.55	.53
Air Moisture Loss	.43	.40	.35
Radiant Heat Loss	1.06	1.03	1.03
Dust Heat Loss	.34	.37	.35
Calcination Loss	.96	.45	.29
Sulfurization Loss	-1.40	-.84	-.48
Bed Drain Loss	<u>.07</u>	<u>.25</u>	<u>.08</u>
Total Losses	58.04	45.85	43.68
Boiler Efficiency, %	76	71	69

Unit Losses

October 30, 1978

CALCULATION OF COAL FEED

I. Equations Used

1. Total Heat of Coal = Boiler Duty + Heat Losses - Additional Heat Inputs
2. Knowing the total heat of the coal, the coal feed rate is calculated by the following equation:

$$Q_v = WF \times (1-H_2O) H_v + C_p (T_c-T_a)$$

Where, H_v = Heating Value of Coal
 H_2O = Total Moisture Content of Coal
 T_c = Temperature of Incoming Coal
 T_a = Ambient Temperature
 C_p = Specific Heat of Coal
 WF = Coal Feed Rate
 Q_c = Total Heat of Coal

The following is a sample calculation of test #14:

Data:

Fluegas Rate = 320727 lb/hr (from GCA data)
 Enthalpy of Flue Gas = 40 BTU/lb (from correlation)
 Boiler Duty = 111.0 M BTU/hr
 Total Air Feed = 273226 lb/hr
 Mass Fraction Moisture of Air = 0.0154
 Air Temp = 145°F
 Mass Fraction Coal Moisture = 0.0205
 Heating Value of Coal = 11882 BR/lb
 Coal Temp = 89°F

I. QG = Heat Loss From Fluegas

$$QG = (320727 \text{ lbs/hr}) (40 \text{ BTU/lb}) = 12.83 \text{ M BTU/hr}$$

II. Heat Loss Analysis: All Units Are M BTU/hr

Unburned Carbon Loss	=	30.76
Gass Heat Loss	=	12.83
Fuel Moisture Loss	=	0.05
Hydrogen Moisture Loss	=	0.55
Air Moisture Loss	=	0.40
Radiant Heat Loss	=	1.03
Dust Sensible Heat Loss	=	0.37
Calcination Heat Loss	=	0.45
Sulfurization Heat Loss	=	-0.84
Bed Drain Losses	=	0.25
Total Heat Losses	=	45.85 M BTU/hr

Table 3.11 (con'd)

$$\begin{aligned} \text{III. Air Heat In (Dry Basis)} &= M C_p T \\ &= 273226 (1-.0154) (.25) (145-83) \\ &= 4.17 \text{ M BTU/hr} \end{aligned}$$

IV. Additional Heat Inputs All In M BTU/hr

$$\begin{aligned} \text{Air Heat} &= 4.17 \\ \text{Auxilliary Air Heat} &= 0.14 \\ \text{Air Moisture Heat} &= 0.58 \\ \text{Boiler Circulation Pump Heat} &= 0.75 \\ \hline \text{Total Additional Heat Inputs} &= 5.64 \end{aligned}$$

$$\text{V. Boiler Duty} = 111.0 \text{ M BTU/hr (calculated)}$$

$$\begin{aligned} Q_c &= 111.0 + 45.85 - 5.64 \\ &= 151.21 \text{ M BTU/hr} \end{aligned}$$

$$SO, \quad 151.21 \times 10^6 \text{ BTU/hr} = WF (1-.0205) 11882 + 0.30 (80-83)$$

Solving For WF,

$$WF = 12990 \text{ lb/hr}$$

CALCULATION OF FLUE GAS CONSTITUENTS

I. Equations used

$$1. \quad 21L + (H)(V) / (L + V) = B$$

Where:

L - Leakage
H - Hagan O₂ volumetric %
V - Sample volume
B - Beckman O₂ volumetric %

$$2. \quad (X)(V) / (L + V) = B_x$$

Where:

X - Calculated gas constituent
V - Sample Volume
L - Leakage
B_x - Beckman gas constituent

Table 3.11 (con'd)

II. Sample Calculation from Test 10

Hagan O₂ = 1.07%

Beckman O₂ = 12.5%

$$(21L + 1.07V) / (L + V) = 12.5$$

$$12.5L + 12.5V = 21L + 1.07V$$

$$8.5L = 11.43V$$

$$L = 1.345V$$

$$V = .744L$$

$$L/V = 1.345$$

Beckman SO₂ = 394 ppm

$$(XV)/(1+V) = 394$$

$$394L + 394V = XV$$

$$(1.345V)(394) + 394 = XV$$

$$(1.345)(394) + 394 = X$$

$$X = 924 \text{ ppm}$$

Table 3.12

PROBLEM	SOLUTION
Inaccurate or missing readings on critical data	Installation of new and/or redundant instrumentation on critical readings
Questionable instrument calibration	Extensive calibration of <u>ALL</u> instruments
Uncertainty over T/C location	Extensive mapping of T/C locations
Coal belt feeders losing calibration	Recalibration of all belt feeders with five correlations to assure coal flow accuracy
Improper coal sampling	Design and install new coal samplers
Limestone and bed material flow rates inaccurate	Installation of limestone and bed material mechanical counters to determine revolutions also bag calibration tests
Limestone and bed material feeders jamming at low RPM	Installation of new feeder and/or motor sprockets to increase motor speed. Also installation of high torque motors
Poor bed material sampling from cells A, B and C	Design and install bed material samplers
Inaccuracies in DC-2 inlet flow caused by duct configuration	Possible installation of load cells on DC-2 will eliminate two isokinetics
Inaccurate airflow measuring	Installation of annubars in A, B and C cells with inclined manometer indication. Also, extensive calibration of all airflow measuring devices in main air and auxiliary air systems
Inability to control belt feeders from control room	Modification of feeder control circuitry

Table 3.12 (con'd)

PROBLEM	SOLUTION
<p>Inability to monitor Beckman & Doric outputs from control room</p> <p>Inaccuracies in gas analysis</p>	<p>Installation of Beckman and Doric repeaters in control room</p> <ol style="list-style-type: none"> 1. Replace damage probes and filters from the fire. 2. Replace all plastic tubing with stainless steel. 3. Replace all 3-way valves with ball valves and check valves. 4. Extensive system and instrument calibration of Beckman system. 5. Modifying TECO system to provide a full backup for Beckman system. 6. Installation of a 4th)₂ analyzer

Table 3.13

	UNITS	B or C Cell		D Cell		Reason- able Incre- ment
		Minimum Value	Maximum Value	Minimum Value	Maximum Value	
Temperature	F	1500	1650	1800	2000	25
Superficial Velocity	fps	10	14	6	10	1
Flue Gas Oxygen	%	3	5	4	6	1
Calcium / Sulfur Ratio	Molar Ratio	1	5	NA	NA	1
Flyash to Total Fuel Ratio	%	NA	NA	60	100	1
Dependent Bed Height (Static)	Inches	18	28	28	36	2
Coal Feed Rate	Tons/hr	11	15	--	--	0.5

SUMMARY OF RANGE OF OPERATING CONDITIONS

Table 3.14

Group Test		Main Cell						CNC					
		Independent				Dependent Process Vnr		Independent				Dependent	
		Velocity fps	Ca/S	Temperature °F	Oxygen %	Bed ht inches	Conl Rnte ton/hr	Velocity tps	Plynth/ fuel %	Temp. °F	Oxygen %	Bed ht inches	
1	a	12	5	1550	3	25	12.3	9	80	1800	4	36	Vary O ₂ — 1
	b	12	5	1550	5	21	10.9	9	80	1800	6	28	
2	a	12	5	1550	3	25	12.3	9	80	1900	4	34	Vary CNC Temp.
	b	12	5	1550	5	21	10.9	9	80	1900	6	26	
3	a	12	5	1550	3	25	12.3	9	80	2000	4	32	
	b	12	5	1550	5	21	10.9	9	80	2000	6	24	
4	a	12	5	1500	3	27	12.3	9	80	1800	4	36	Vary Main Cell temp.
	b	12	5	1500	5	25	10.9	9	80	1800	6	28	
5	a	12	5	1600	3	23	12.3	9	80	1800	4	36	
	b	12	5	1600	5	19	10.9	9	80	1800	6	20	

PHASE I TEST CONDITIONS

TABLE 3.15

		Main Cells					CRC Cell										
Group	Test	Independent				Dependent		Independent				Dependent					
		Velocity fps	Ca/S	Temperature °F	Oxygen %	Bed ht inches	Coal Rate tons/hr	Velocity fps	Flyash/ fuel%	Temp. °F	Oxygen %	Bed ht inches					
1	a*	12	5	1550	4	23	11.6	Condition Selected to Optimize Combustion Efficiency based on Phase I results.									
	b	12	4	1550	4	23	11.6										
	c	12	3	1550	4	23	11.6										
	d	12	2	1550	4	23	11.6										
2	a*	12	5	1500	5	23	10.9										
	b	12	4	1500	5	23	10.9										
	c	12	3	1500	5	23	10.9										
	d	12	2	1500	5	23	10.9										
3	a*	12	5	1600	3	23	12.3										
	b	12	4	1600	3	23	12.3										
	c	12	3	1600	3	23	12.3										
	d	12	2	1600	3	23	12.3										
4	a	10	3	1550	3	22	10.3										
	b	10	3	1550	5	18	9.0										

Vary
Ca/S

Vary Red
Temp.

Vary velocity
bed height

* Optional Tests which duplicates some of Phase I Tests. If Phase I emissions results adequate, these will be deleted.

PHASE II TEST CONDITIONS

TABLE 3.16

WEEK	1	2	3	4	5	6	7	8	9	10	11	12	13
NUMBER OF STARTS	1	1	1	1	1	1	1	1	1	1	1	1	1
CELLS IN SERVICE	2	3	3	3	3	3	3	3	3	3	3	3	3
NUMBER OF HOURS COAL FIRED	50	100	100	100	100	100	100	100	100	100	100	100	100
TEST PHASE	I	I	I	I	I	II	II	II	II	II	II	II	II
NUMBER OF DATA ACQUISITION PERIODS	2	2	4	4	3	3	3	3	3	3	3	3	3
OPTIONAL 500 HOUR RUN

NOTE: It is anticipated that two weeks during this period
will be required for equipment maintenance.

OPERATING SCHEDULE DURING FIVE MONTH TEST PERIOD

TABLE 3.17

PLANNED SEQUENCE OF EVENTS IN A TYPICAL OPERATING RUN

Table 3.18

Table 3.1b												
IMP	S	M	T	W	T	F	S					
0	M & R Weekend Maintenance and Repair	BB Build Bed	Establish PC	Establish New Pre- Test	Establish New Pre- Test	Establish New Pre- Test						
1												
2												
3		BI Ignition	Data Acquisition									
4												
5		CID Coal-Cell D BB Build Bed										
6												
7		CIC Coal-Cell C BB Build Bed										
8												
9		CIB Coal-Cell B FAR Flyash Reinject										
10												
11	Stack Sample
12												
13												
14		STT Steam To Turbine										
15												
16		PC Pretest Conditions										
17												
18												
19												
20												
21												
22												
23												

TYPICAL OPERATING SCHEDULE FOR WEEK NO. 3 OR 4
DURING FIVE MONTH TEST PERIOD

Table 3.19

Date	Run	Condition
01/16/79	1	D w/seal air
01/16/79	1	C-N w/seal air
01/16/79	1	B-N w/seal air
01/16/79	1	A-N W/seal air
01/16/70	1	A-S w/seal air
01/16/79	1	B-S w/seal air
01/16/79	1	C-S w/seal air
01/16/79	2	D, C-S w/seal air
01/16/79	3	D, C, B-N w/seal air
01/17/79	4	A w/o seal air
01/17/79	5	A, B w/o seal air
01/17/79	6	A, B, C w/o seal air
01/17/79	7	A, B, C + D w/o seal air
01/17/79	8	A, B, C + D w/seal air
01/17/79	9	C, D and burners

Auxiliary F.D. Test Conditions

Table 3.20

Date	Run No.	Conditions
02/08/79	1	All Cells Open
02/08/79	2	A, B, C Open
02/08/79	3	A + C Open
02/08/79	4	C Open
02/11/79	5	A Open, B, C + D Capped
02/11/79	6	A Open, B, C + D Capped
02/13/79	7	A 80%, B 80%, C+D Capped
02/13/79	8	A 80%, B 40%, C+D Capped
02/13/79	9	A 80%, B 20%, C+D Capped
02/13/79	10	A, B 80%, C 80%, D Capped
02/13/79	11	A, B 80%, C 40%, D Capped
02/13/79	12	A, B 80%, C 20%, D Capped
02/15/79	13	A, B, D 80%, C 0%
02/15/79	14	A, B 80%, C 0%, D 100%
02/15/79	15	A, B 80%, C 0%, D 100%
02/15/79	16	A 80%, B, C, D 0%

FD - ID Fan Test Conditions

TABLE 3.21

- 0. Aux FD Inlet Damper
- 1. Aux FD Fan
- 2. Aux. FD Discharge Damper
- 3. Aux FD Pitot - Venturi
- 4. Pitot Traverse
- 5. Belt Feeder Bypass Equalisation Line (Closed)
- 6., 8., 10., 12., 14., 16., 18., Vibrating Feeder Annubar.
- 7., 9., 11., 13., 15., 17., 19., Vibrating Feeders
- 20. N. Burner Annubar
- 21. N. Burner
- 22. S. Burner Annubar
- 23. S. Burner
- 24. Seal Air

Aux FD Fan Circuit

Isokinetic Test Conditions

April - MAY 1979 ✓

Table 3.22

5-24-79

Test No.	GCA #1	Pretest 2	Buell #2	Pretest 3		Pretest 4		
		Buell #1		Buell #3	Buell #4	Buell #5	Buell #6	Buell #7
Test Date	4-18-79	4-24-79	4-26-79	5-1-79	5-1-79	5-3-79	5-3-79	5-3-79
No cells operating	1	3	1	3	3	3	3	3
D CELL								
Vs ft/sec	6.66	6.60	5.75	9.0		8.6		
Bed temp. °F	1587	1560	1512	1663		1642		
Coal lb/hr	2200	1900	2300	2200		1802		
Limestone lb/hr	NA	NA	NA	1009		600		
C CELL								
Vs ft/sec	NA	10.4	NA	11.0		11.2		
Bed temp °F	NA	1500	NA	1525		1446		
Coal lb/hr	NA	NA	NA	7719		9813		
Limestone lb/hr	NA	NA	NA	829		1200		
B CELL								
Vs ft/sec	NA	10.5	NA	11.2		12.1		
Bed temp °F	NA	1557	NA	1653		1575		
Coal lb/hr	NA	NA	NA	9138		10143		
Limestone lb/hr	NA	NA	NA	501		1000		

SUMMARY OF ISOKINETIC TESTING

APRIL - MAY 1979

Table 3.23

-24-79

Test No.	GCA #1	Pretest 2	Buell#2	Pretest 3		Pretest 4		
		Buell #1		Buell #3	Buell #4	Buell #5	Buell #6	Buell #7
Test Date	4-18-79	4-24-79	4-26-79	5-1-79	5-1-79	5-3-79	5-3-79	5-3-79
No. Cells Operating	1	3	1	3	3	3	3	3
DC-1 INLET								
% H ₂ O	NA	NA	NA	5.93		6.75		
Temp °F	NA	NA	NA	609		618		
ACFM	NA	NA	NA	111,352		136,457		
GR/ACF	NA	NA	NA	3.83		7.11		
LB/HR	NA	NA	NA	3,652		8,316		
DC-1 OUTLET								
% H ₂ O	NA	NA	NA	5.93		6.75		
Temp °F	NA	NA	NA	609		618		
ACFM	NA	NA	NA	111,352		136,457		
GR/ACF	NA	NA	NA	1.69		1.58		
LB/HR	NA	NA	NA	1614		1851		
Collector Eff %	NA	NA	NA	55.8		77.7		
DC-2 INLET								
% H ₂ O	NA	NA	NA	3.48		4.30		
Temp °F	NA	NA	NA	646		628		
ACFM	NA	NA	NA	28,787		27,584		
GR/ACF	NA	NA	NA	3.73		1.53		
LB/HR	NA	NA	NA	921		362		
DC-2 OUTLET								
% H ₂ O	NA	NA	NA	4.04		5.44		
Temp °F	NA	NA	NA	646		649		
ACFM	NA	NA	NA	26,297		25,602		
GR/ACF	NA	NA	NA	0.560		0.420		
LB/HR	NA	NA	NA	126		93		
Collector Eff. %	NA	NA	NA	86.2		74.4		

SUMMARY OF ISOKINETIC TESTING

APRIL - MAY 1979
Table 3.23 (con't)

Test No.	GCA #1	Pretest 2	Buell #2	Pretest 3		Pretest 4		
		Buell #1		Buell #3	Buell #4	Buell #5	Buell #6	Buell #7
Test Date	4-18-79	4-24-79	4-26-79	5-1-79	5-1-79	5-3-79	5-3-79	5-3-79
No Cells Operating	1	3	1	3	3	3	3	3
ESP INLET								
% H ₂ O	3.39	6.76	1.41	4.01	5.24	7.14	6.16	6.30
Temp	484	600	562	586	588	605	607	606
ACFM	69,568	170,056	78,638	145,884	149,865	156,634	154,480	153,660
GR/ACF	0.220	1.71	0.927	1.41	1.34	1.47	1.46	1.46
LB/HR	132	2486	625	1759	1722	1978	1929	1924
STACK								
H ₂ O	² 1.34	² 1.72	² 1.81	3.44	5.37	² 1.54	² 1.48	² 1.37
Temp	219	306	237	310	287	299	312	340
ACFM	92,556	150,934	89,262	161,148	163,996	170,358	170,109	173,380
GR/ACF	0.070	0.326	0.050	0.182	0.124	0.150	0.128	0.147
LB/HR	53	421	38	252	174	219	186	219
Collector Eff %	59.7	83.1	93.9	85.7	89.9	88.9	90.3	88.6

1 DC-1 Outlet Numbers Are Calculated Values

2 Preliminary Moisture

Table 3.24

Dust Collector Design Conditions and Test Conditions

	Design Conditions	Pre-test 3	Pre-test 4
DC-1			
flow, acfm	180,000	111,352	136,457
Δp , "WC	3.6	2.5	2.8
calc. Δp	---	1.5	2.25
% eff.	95	55.8	77.7
DC-2			
flow, acfm	19,800	28,787	27,584
Δp , "WC	3.4	5.0	4.5
% eff.	95	86.2	74.4

Table 3.25

MICRONS	TEST 1	TEST 2	TEST 3	TEST 4	TEST 5	TEST 6	TEST 7
60	.10	.22	.04	.045	.045	.02	.05
50	.01	.03	.01	.005	.005	.01	.01
40	.03	.06	.01	.005	.005	.01	.01
30	.05	.08	.04	.025	.045	.04	.04
20	.10	.11	.10	.08	.075	.09	.10
10	.18	.15	.20	.19	.205	.19	.20
8	.08	.04	.11	.08	.10	.09	.09
6	.09	.06	.12	.16	.11	.11	.10
5	.05	.04	.06	.10	.05	.07	.05
4	.07	.035	.06	.09	.09	.09	.08
3	.06	.055	.07	.07	.07	.07	.05
2	.05	.05	.07	.06	.09	.09	.09
1	.06	.05	.08	.06	.09	.08	.09
.6	.03	.01	.02	.01	.03	.03	.03
.3	.02	.01	.02	.02	.01	.01	.01
Size, u	15.08	24.54	10.49	10.26	10.54	9.58	11.43
4-10u	51	35	51	65	64	64	59

ESP INLET SIZE DISTRIBUTION
as Performed By BUELL ENVIROTECH, LEBANON, PA

Mass fraction in each size cut.

Table 3.26

Details of Flyash Ash Component Analysis Calculations

	Pre-test 3	Pre-test 4
Coal Fe/Ash	.0991	.1228
Limestone Fe/Ash	.0182	.0210
DC-1 in FA Fe/Ash	.1058	.0992
% Coal in DC-1 FA	100%	76.60
% ls in DC-1 _i FA	0%	23.40
DC-1 out FA Fe/Ash	.0706	.0696
% coal in DC-1 _o FA	64.58	64.14
% ls in DC-1 _o FA	35.43	35.869
DC-2 in FA Fe/Ash	.0528	.1196
% Coal in DC-2 _i FA	42.45%	92.14%
% ls in DC-2 _i FA	57.55%	2.86%
DC-2 out FA Fe/Ash	.0604	.0944
% Coal in DC-2 _o FA	51.9%	75.62%
% ls in DC-2 _o FA	48.31%	25.38%

$$\text{Equation: } (\text{Fe/Ash})_{\text{FA}} = X [(\text{Fe/Ash})_{\text{C}}] + (1-X) [(\text{Fe/Ash})_{\text{LS}}]$$

where: $(\text{Fe/Ash})_{\text{FA}}$ - iron to ash ratio of flyash
 $(\text{Fe/Ash})_{\text{C}}$ - iron to ash ratio of coal
 $(\text{Fe/Ash})_{\text{LS}}$ - iron to ash ratio of limestone
 X - fraction of coal derived flyash ash

Table 3.27

OVERALL AND COMPONENT MATERIAL FLOW RATES AND EFFICIENCIES

	Dust Collector No. 1				Dust Collector No. 2			
	Pretest 3		Pretest 4		Pretest 3		Pretest 4	
	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet	Inlet	Outlet
Total Flyash Flowrate, lb/hr	3652	1614	8316	1851	921	126	362	93
Flyash Volatiles Flowrate, lb/hr	1826	348	3933	505	156	18	123	23
Flyash Ash Flowrate, lb/hr	1826	1266	4383	1346	765	108	239	70
Coal Derived Ash Flowrate, lb/hr	1826	818	3357	863	325	56	232	52
Limestone Derived Ash Flowrate lb/hr	0	449	1026	483	440	52	7	18
Overall Collection Efficiency, %	55.8		77.7		86.2		74.4	
Volatiles Collection Efficiency, %	80.9		87.2		88.5		81.3	
Coal Derived Ash Collection Eff'y %	55.2		74.3		82.8		77.6	
L.S. Derived Ash Collection Eff'y %	Negative		52.9		88.2		Negative	

C CELL CHARACTERIZATION TESTS

Table 3.28

Point No.	Bed Temperature	Bed Depth	Superficial Velocity	C cell exit O ₂
	°F	"WC	ft/sec	%
Date:5/23/79				
1	1520	20	7.2	3.0
2	1510	21	9.7	3.5
3	1500	22,5	13.0	7.2
4	1500	24	12.7	7.8
5	1500	26	11.8	6.7
6	1510	27	10.6	5.9
7	1510	28	9.8	4.5
8	1450	29	8.8	3.0
Date:5/31/79				
1	1550	18	10.4	8.5
2	1555	18	8.0	7.5
3	1540	19	8.3	6.7
4	1550	20	10.6	4.3
5	1530	21	11.9	5.1

Table 3.29

POPE, EVANS AND ROBBINS, INC
 RIVESVILLE, W.V.
 TEST NO. 3
 DATE 5/ 1/79
 3 CELL(S) IN OPERATIONS
 FLYASH REINJECTION NOT IN SERVICE
 STEAM TO TURBINE

OPERATING CONDITIONS	CELL			
	D	C	B	A
COAL FEED, LB/HR	2200	7719	9138	0
LIMESTONE FEED, LB/HR	1009	829	501	0
COAL % S	2.59	2.59	2.59	2.59
CA/S (MOLE RATIO)	5.40	1.27	.65	00
MAIN AIR FLOW, LB/HR	36882	91560	105326	32880
AUX AIR FLOW, LB/HR	7071	4474	4975	0
STEAM FLOW, LB/HR	171756			
STEAM CONDITIONS, PSIG/F	1265/ 894			
BED TEMPERATURE, F	1663	1525	1653	526
VS, FPS	9.0	11.0	11.2	1.7
STATIC BED HEIGHT, IN.	18	21	19	2
OXYGEN IN FLUE GAS, %	14.9	8.6	9.6	+++++
BED DRAIN, LB/HR	600	950	500	0
FLYASH, LB/HR	921	3652		
TEST RESULTS	D	MAIN CELLS		OVERALL
SO2, LB/M BTU	1.3	2.9		4.0
NOX, LB/M BTU	1.3	.7		.9
COMBUSTION EFFICIENCY, %	95	87		88
BOILER EFF., %(LOSS)	78			

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MAY 30 1979

DATE OF TEST 5/ 1/79
3 CELL(S) IN OPERATION

Table 3.29 (Con'd)

TOTAL MASS BALANCE CALCULATIONS			
COMPONENT	D	MAIN CELLS	UNIT
COAL IN (LB/HR)	2200	16857	19057
LIMESTONE IN (LB/HR)	1009	1330	2339
COMBUSTION AIR IN (LB/HR)	36882	229765	266647
FEEDER AIR IN (LB/HR)	7071	9449	16520
FLYASH IN (LB/HR)	0	+++++	+++++
TOTAL MASS IN (LB/HR)	47162	257402	304564
FLYASH OUT (LB/HR)	921	3652	4573
BED(S) OUT (LB/HR)	600	1449	2050
H2O GAS OUT (LB/HR)	1263	8963	10226
O2 GAS OUT (LB/HR)	9249	29893	39142
NOX GAS OUT (LB/HR)	35	146	182
HC GAS OUT (LB/HR)	0	87	87
SO2 GAS OUT (LB/HR)	34	625	660
CO2 GAS OUT (LB/HR)	4534	25359	29893
CO GAS OUT (LB/HR)	3	303	306
N2 GAS OUT (LB/HR)	43773	179979	223752
TOTAL MASS OUT (LB/HR)	60416	250460	310876
MASS BALANCE CLOSURE %	128	97	102

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MAY 30 1979

DATE OF TEST 5/ 1/79
3 CELL(S) IN OPERATION

Table 3.29 (Con'd)

COMBUSTION EFFICIENCY CALCULATIONS

D CELL

UNBURNED COMBUSTIBLES IN FLYASH	1.41 (MBTU/HR)
UNBURNED COMBUSTIBLES IN BED MATERIAL	.11 (MBTU/HR)
CO HEAT LOSS	.02 (MBTU/HR)
HC HEAT LOSS	.01 (MBTU/HR)
TOTAL UNBURNED COMBUSTIBLES	1.54 (MBTU/HR)
COMBUSTION EFFICIENCY	95 (%)

CELLS A, B & C

UNBURNED COMBUSTIBLES IN FLYASH	23.80 (MBTU/HR)
UNBURNED COMBUSTIBLES IN BED MATERIAL	.40 (MBTU/HR)
CO HEAT LOSS	1.32 (MBTU/HR)
HC HEAT LOSS	2.24 (MBTU/HR)
TOTAL UNBURNED COMBUSTIBLES	27.76 (MBTU/HR)
COMBUSTION EFFICIENCY	87 (%)

OVERALL

UNBURNED COMBUSTIBLES IN FLYASH	25.21 (MBTU/HR)
UNBURNED COMBUSTIBLES IN BED MATERIAL	.51 (MBTU/HR)
CO HEAT LOSS	1.33 (MBTU/HR)
HC HEAT LOSS	2.25 (MBTU/HR)
TOTAL UNBURNED COMBUSTIBLES	29.31 (MBTU/HR)
COMBUSTION EFFICIENCY	88 (%)

LOSS ANALYSIS CALCULATIONS

UNBURNED COMBUSTIBLE LOSS	29.31 (MBTU/HR)
GAS HEAT LOSS	11.20 (MBTU/HR)
TOTAL MOISTURE LOSS	6.41 (MBTU/HR)
RADIANT HEAT LOSS	2.06 (MBTU/HR)
DUST SENSIBLE HEAT LOSS	.29 (MBTU/HR)
SULFATION HEAT LOSS	3.03 (MBTU/HR)
BED DRAIN LOSSES	.75 (MBTU/HR)
TOTAL LOSSES	53.05 (MBTU/HR)
BOILER DUTY	191.39 (MBTU/HR)
INDIRECT BOILER EFFICIENCY	78 %
CALCULATED COAL FLOW	18923 (LB/HR)

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Table 3.30

PDPE, EVANS AND ROBBINS, INC
RIVESVILLE, W.V.
TEST NO. 4
DATE 5/ 3/79
3 CELL(S) IN OPERATIONS
FLYASH REINJECTION NOT IN SERVICE
STEAM TO TURBINE

OPERATING CONDITIONS	CELL			
	D	C	B	A
COAL FEED, LB/HR	1805	9813	10143	0
LIMESTONE FEED, LB/HR	600	1200	1000	0
COAL % S	3.03	3.03	3.03	3.03
CA/S (MOLE RATIO)	3.31	1.22	.99	.00
MAIN AIR FLOW, LB/HR	34101	96545	118776	28921
AUX AIR FLOW, LB/HR	8263	4751	4708	0
STEAM FLOW, LB/HR	171604			
STEAM CONDITIONS, PSIG/F	1320/ 896			
BED TEMPERATURE, F	1642	1446	1575	527
VS, FPS	8.6	11.2	12.1	1.5
STATIC BED HEIGHT, IN.	18	21	21	8
OXYGEN IN FLUE GAS, %	15.3	9.1	9.3	+++++
BED DRAIN, LB/HR	800	0	0	0
FLYASH, LB/HR	362	8316		
TEST RESULTS	D	MAIN CELLS		OVERALL
SO2, LB/M BTU	4.1	3.5		3.5
NOX, LB/M BTU	1.2	.7		.7
COMBUSTION EFFICIENCY, %	93	77		79
BOILER EFF., %(Loss)	69			

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MAY 3 / 1979

TEST NO. 4
 DATE OF TEST 5/ 3/79
 3 CELL(S) IN OPERATION

Table 3.30 (Con'd)

TOTAL MASS BALANCE CALCULATIONS			
COMPONENT	D	MAIN CELLS	UNIT
COAL IN (LB/HR)	1805	19955	21761
LIMESTONE IN (LB/HR)	599	2200	2800
COMBUSTION AIR IN (LB/HR)	34101	244242	278343
FEEDER AIR IN (LB/HR)	8263	9459	17722
FLYASH IN (LB/HR)	0	+++++	+++++
TOTAL MASS IN (LB/HR)	44770	275857	320627
FLYASH OUT (LB/HR)	361	8315	8677
BED(S) OUT (LB/HR)	799	0	799
H2O GAS OUT (LB/HR)	1502	12201	13703
O2 GAS OUT (LB/HR)	9076	32998	42074
NOX GAS OUT (LB/HR)	26	158	184
HC GAS OUT (LB/HR)	0	122	122
SO2 GAS OUT (LB/HR)	91	852	944
CO2 GAS OUT (LB/HR)	3926	33726	37653
CO GAS OUT (LB/HR)	4	412	416
N2 GAS OUT (LB/HR)	41794	213130	254924
TOTAL MASS OUT (LB/HR)	57583	301918	359501
MASS BALANCE CLOSURE %	128	109	112

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3 CELL(S) IN OPERATION

Table 3.30 (Con'd)

COMBUSTION EFFICIENCY CALCULATIONS

D CELL

UNBURNED COMBUSTIBLES IN FLYASH	1.43 (MBTU/HR)
UNBURNED COMBUSTIBLES IN BED MATERIAL	.15 (MBTU/HR)
CO HEAT LOSS	.02 (MBTU/HR)
HC HEAT LOSS	.00 (MBTU/HR)
TOTAL UNBURNED COMBUSTIBLES	1.60 (MBTU/HR)
COMBUSTION EFFICIENCY	93(%)

CELLS A, B & C

UNBURNED COMBUSTIBLES IN FLYASH	50.76 (MBTU/HR)
UNBURNED COMBUSTIBLES IN BED MATERIAL	0. (MBTU/HR)
CO HEAT LOSS	1.79 (MBTU/HR)
HC HEAT LOSS	3.12 (MBTU/HR)
TOTAL UNBURNED COMBUSTIBLES	55.68 (MBTU/HR)
COMBUSTION EFFICIENCY	77(%)

OVERALL

UNBURNED COMBUSTIBLES IN FLYASH	52.20 (MBTU/HR)
UNBURNED COMBUSTIBLES IN BED MATERIAL	.15 (MBTU/HR)
CO HEAT LOSS	1.81 (MBTU/HR)
HC HEAT LOSS	3.12 (MBTU/HR)
TOTAL UNBURNED COMBUSTIBLES	57.28 (MBTU/HR)
COMBUSTION EFFICIENCY	79(%)

LOSS ANALYSIS CALCULATIONS

UNBURNED COMBUSTIBLE LOSS	57.28 (MBTU/HR)
GAS HEAT LOSS	12.02 (MBTU/HR)
TOTAL MOISTURE LOSS	8.24 (MBTU/HR)
RADIANT HEAT LOSS	2.06 (MBTU/HR)
DUST SENSIBLE HEAT LOSS	.26 (MBTU/HR)
SULFATION HEAT LOSS	2.98 (MBTU/HR)
BED DRAIN LOSSES	.30 (MBTU/HR)
TOTAL LOSSES	83.14 (MBTU/HR)
BOILER DUTY	186.60 (MBTU/HR)
INDIRECT BOILER EFFICIENCY	69 %
CALCULATED COAL FLOW	21534 (LB/HR)

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IV. NEW YORK OFFICE 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 EPA Particulate Control Test

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 (6 in and 10 in.) from the flue gas duct at the electrostatic precipitator to the EPA mobile equipment.

Stress analyses for the 6 in. and 10 in. duct runs with and without expansion joints were completed. Results indicated that expansion joints were required to avoid overstress. Piping arrangement drawings were prepared which included expansion joints, guides and anchors to reduce stresses to allowance limits, and released for construction.

4.3 Access/Egress

In order to facilitate maintenance and operation of the Rivesville MFB systems and equipment, the design of additional access/egress facilities was undertaken. Drawings of platforms were forwarded to Rivesville Operations for installation of the following:

- Access to platform above coal dryer distribution hopper (coal dryer now removed),
- Existing platform extension at top of limestone bunker
- Additional access to air preheater platform and preheater service platform west side extension, and
- Stairs to connect high roof to south low roof.

4.4 Coal Dryer Removal and Reinstallation

After a field trip to the site to investigate the feasibility of removing the existing coal fryer, drawings were prepared detailing a step-by-step removal procedure. Low clearance

hoists were investigated and selected to accomodate available space. A study was initiated on relocating the dryer to the Smithfield Mine Tipple to insure an available supply of dry coal. A final report was prepared and submitted to DOE which included a budget cost estimate to provide for relocating the dryer and the addition of a combustor heat source, dryer building, a coal storage shed, and covered conveyors.

4.5 F.D. Fan Relocation

A study was initiated to investigate the possibility of relocating the F.D. fan to another area inside the plant in order that a second fan could be provided as back-up. Various alternate locations were evaluated during a field trip to the site. The conclusions of the study were presented in a final report on redundant equipment which included the F.D. fans, I.D. fans and a baghouse filter to augment the ESP. The study included cost figures and drawings for four alternate arrangements with recommendations for lowest cost scheme.

4.6 Duct Temperature Study

To improve the efficiency of the electrostatic precipitator, it was proposed that the flue gas temperature be increased from 730°F to 800°F. An investigation was conducted to determine if the existing duct system and expansion joints could withstand this increase in addition to the pressure increase of 20 in. to 25 in. W.C. resulting from the new I.D. fan. The study concluded that it was permissible to operate at the increased temperature and pressure with periodic checks of the ducts and joints for signs of distortion.

4.7 Dust Collector No. 1 Bypass

The efficiency of Dust Collector No. 1 is low when the MFB is in one or two cell operation due to the reduced quantity of flue gas. A shop drawing of the D.C. No. 1 by-pass damper was prepared for diversion of the flue gas to dust collector No. 4, originally designated for use with the eliminated coal dryer. The duct modification and damper installation drawings prepared by S&W were reviewed before the by-pass installation.

4.8 Alexander PDU Baghouse Filter

During excavation for the baghouse foundation, water was encountered. On inspection, proper drainage and backfilling procedures were recommended.

4.9 New I.D. Fan

The fire-damaged I.D. fan was replaced with a fan having a greater static heat in order to overcome additional pressure drops in the system resulting from modifications to the electrostatic precipitator and possible future replacement

with a baghouse filter. Design drawings for the revised ducts and supports systems prepared by S&W were reviewed and existing structural support system checked for adequacy before proceeding with the I.D. fan installation.

4.10 Ash Cooler and Classifier Access Opening

Advised Rivesville Operations that because of structural integrity, the proposed access opening in the existing concrete wall between ash cooler and classifier should not be made.

4.11 Investigate Limestone Storage at Quarry

Budget cost estimates were prepared for several sizes of storage silos for on-site storage of limestone at the quarry to insure continuous supply during MFB operation.

4.12 Rotex Screen Relocation

With the replacement of the horizontal Redler conveyor by a 24" belt conveyor to minimize coal degradation, the Rotex screen was relocated to the head of the belt conveyor and the crusher moved adjacent to the screen. As a result of equipment relocation, the existing support steel was checked for adequacy and new members added where required.

4.13 Silo Vent Baghouse Filter

To eliminate discharge of transport air to the flue gas duct which significantly lowers the flue gas temperature, it was proposed that a direct vent to atmosphere be provided on top of the 80 T silo. The vent would discharge into a small baghouse filter similar to that in use at the Alexandria facility. To minimize down time, arrangements were made to ship the Alexandria baghouse to Rivesville and a unit ordered for replacement. Installation drawings were sent to Rivesville Operations for implementation.

4.14 Baghouse Replacement

Quotations for a baghouse filter proposed to augment the electrostatic precipitator for MFB flue gas discharge, were received from Standard Haven, Combustion Engineering and Zurn Industries. Recommendations for award were submitted on completion of evaluation, which included the impact on the existing structure and additional structural work that would be required to accomodate the offered baghouses.

4.15 Bed Material Letdown System Modernization

The existing pressurized material transport system for classified bed material was converted to a vacuum system and additional required equipment ordered. Installation drawings for system modification were prepared and forwarded to Rivesville Operations for implementation.

4.16 Install Fire Protection System Extensions

The existing fire protection system in the vicinity of the northwest roof area was extended by the addition of three more stations with provision for a fourth. Drawing and bill of materials were prepared and forwarded to Rivesville Operations for implementation.

4.17 Stack Silencer Installation

Based on a review of calculations prepared by S&W on the stability of the existing stack with the new silencer, it was determined that the critical wind velocity (for instability) was greater than the 50 year maximum recorded in the Rivesville area. Accordingly, Rivesville Operations was advised that structural reinforcing of the stack due to addition of the silencer was not required.

4.18 Electrical Equipment Room

Reviewed and checked support members for the Electrical Equipment Room on a field inspection and apprised Rivesville Operations of maximum permissible loadings in that area.

4.19 Control Room Modification

Structural information regarding the modifications required for extension of the present control room was provided to S&W for two alternate locations under consideration.

4.20 Coal Feed Eductor System

An analysis of available pneumatic blowers which could provide the required volume of air at 10 psi for the fuel eductor system to be installed in Cell "C", led to the selection of the Sundstrand Centrifugal Blower. In order to prove the effectiveness and reliability of the eductor system, it was decided to only equip Cell "C" initially.

Layouts were prepared showing the installation of the blowers, needles, control and monitoring equipment for the north and south bank of coal feed needles for Cell "C".

4.21 Central Vacuum Cleaning System

Preliminary designs for a central vacuum cleaning system for the MFB plant were prepared along with a budgetary cost estimate and submitted to Rivesville.

4.22 Instrumentation Shop

During a trip to the site, it was decided that the Instrumentation Shop would be best located in an area above the demineralizer installation. Sketches and designs for the structural framing were prepared and issued to Rivesville Operations for implementation.

4.23 Preventative Maintenance

A Preventative Maintenance Program utilizing the VisiRecord System was developed and introduced to Rivesville Operations. At the right side of the VisiRecord card, the frequency code is defined by numbers for weekly, monthly etc, action. Small metal tabs attached to the card side identify the period that maintenance is required for the subject equipment. Upon completion of the equipment maintenance, the tabs are moved down to the next scheduled period.

4.24 Fly Ash Reinjection Piping

A thermal stress analysis for the fly ash reinjection piping for Rivesville was conducted and it was determined that the piping would be overstressed if flexible connections were not provided.

4.25 Boiler Circulating Pumps

An investigation was conducted into the possibility of obtaining better accessibility to the boiler circulating pumps for more ready maintenance. The study also included a determination of the feasibility of installing a redundant boiler circulating pump. Alternate arrangements, costs, and schedules were prepared and are currently being evaluated.

V. SUBCONTRACTOR ACTIVITIES

5.1 General

To assist in operation of the MFB unit and evaluation of performance, the services of the Foster Wheeler Energy Corporation (FWEC) and Stone & Webster Engineering Corporation (S&W) were subcontracted for. A description of their activities during the reporting period follows:

5.2 Foster Wheeler Energy Corporation

MFB Steam generation development, MFB operations and maintenance assistance

5.2.1 Activities

Foster Wheeler activities included support of MFB unit operation and assistance in performance improvement. During the reporting period, the boiler design manual was issued for comment and problems associated with the Beckman gas analyzer equipment resolved. Replacement air distributors were designed, fabricated and delivered based on utilizing the previously proposed hollow bolt configuration. FWEC also completed:

- Evaluation of C-Cell inbed heat transfer surface based on operating test data,
- Evaluation of reduced design fluidizing velocity for C-cell,
- Evaluation of techniques for increasing the flue gas exit temperature from cells, A, B and C
- Evaluation of C-cell ignition with a single ignitor,
- Evaluation of techniques for reducing air leakage through the control dampers for each cell.

Test Instrumentation

Continued use and evaluation of results obtained from the Beckman gas analyzer equipment indicated that considerable air in-leakage was occurring. It was determined that the leakage was due to the use of nylon tubing, particularly at the end connections. The nylon was replaced with stainless steel tubing and subsequent operation was satisfactory. Modifications were also made to the gas analyzer cabinet to replace leaking gas sequencing valves with an improved design valve.

Systems Evaluation

A study of the air control dampers for A, B, C and D-cells of the MFB unit to determine the cause of excessive leakage was performed. Based on the results of the study, equipment modifications and modified operating techniques were recommended. Replacement air distributors for B, C and D-cells based on utilizing the hollow bolt design which was listed in the Alexandria PDU, were designed, fabricated and delivered for installation in the MFB unit.

5.3 Stone & Webster Engineering Corporation - MFB Operations and Assistance

5.3.1 Activities

Stone & Webster activities concentrated on providing headquarters and onsite engineering and design support for the testing program, operations and specified investigations. Following the 9 August fire, S&W focused primarily in assisting in reconstruction related items and specified tasks assigned by PER. Support was provided in those areas critical to the scheduled January, 1979 MFB startup. Particular attention was devoted to new ID fan procurement, fabrication of associated ductwork, stack and steam vent silencers, combustion air damper repairs, general arrangement drawings, site plans and instrument lists. Early in 1979, the bulk of S&W work focused on modernization studies to be incorporated during the summer outage. At the end of this reporting period, reports had been prepared and submitted on the Control Building Study, Evaluation of Noise Control Equipment, Coal and Limestone Handling, Fan Vibrating Analysis, Turbine Bypass System and Draft Equipment Modernization (Baghouse).

Boston Office Activities

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

- EDR- 2 Evaluation of Noise and Installed Noise Control Equipment
- EDR- 5 Precipitator Operations at Low MFB Load, and Vibrating Analysis of the Forced Draft and Induced Draft Fans
- EDR- 8 Coal and Limestone Delivery, Distribution and Injection System
- EDR-10 Estimate for MFB Turbine Bypass System
- EDR-14 Control Room Modernization

EDR-23 Particulate Emission Control and Draft
 Equipment Modernization

Field Activities

The following Field Reports were Completed:

FR-12	ESP Cleanout Ports
FR-14	Instrument Air Compressor Cooling Water Booster Pump
FR-15	Water Requirements for Ecolochem Demineralizer (may be re-opened)
FR-16	Precipitator Hopper Access Platform
FR-17	Combustion Air Dampers
FR-18	Georgetown D.D. Fan (cancelled)
FR-19	Estimate for Dryer and Storage Facilities for Coal and Limestone
FR-20	Water Supply for APH Deluge System
FR-21	MS-4 Circuit Review
FR-22	Review of Boiler Safety Circuit
FR-23	Relocation of Rotex Screen (S&W involvement)
FR-24	Air Duct and Plenum Drain Lines
FR-25	DC-1 Bypass
FR-26	Instrument Schedule
FR-27	New I.D. Fans
FR-28	Automated Coal Sampler
FR-29	Limestone L-Valve
FR-30	Damages (in shipment) to APH Deluge Pump
FR-31	PER Ash Silo Unloading Chute
FR-32	MPS System Review
FR-33	F.D. Fan Ground Faults

- FR-34 Limestone Unloading Baghouse
- FR-35 MFB Cell CO Monitors
- FR-36 Relocation of BFWV-9
- FR-37 MFB Air Preheater (File)
- FR-38 Auxiliary F.D. Fan System

Results

Reviewed entire results effort including instrumentation, test procedures, reporting and formatting and data collection. Evaluated direction of future results effort. Verification analysis of performance calculation equations was completed. Test Plan was updated and reformatted and computer program reviewed and revised.

Safety

Throughout the reporting period, daily safety tours were conducted in the MFB plant, safety meetings were held, recommendations made, and accident investigations conducted. Weekly "tool box meetings" were held with maintenance, operations and construction personnel. Safety memos were issued on unsafe conditions and general plant safety.