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Multicell Fluidized Bed Boiler Design, Construction, and Test Program

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Interim Report for the Period July 1976-June 1977

July 1978

Prepared for
U.S. Department of Energy
Assistant Secretary for Energy Technology
Division of Fossil Fuel Utilization

MASTER

Under Contract No. Ex-76-C-01-1237

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

For
U.S. Department of Energy
Assistant Secretary for Energy Technology
Division of Fossil Fuel Utilization
Washington, DC 20545

Under Contract No. Ex-76-C-01-1237

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

This report presents information on start-up and sustained operation of the Rivesville MFB steam generating plant, identification and analysis of operational problems and corrective action to improve performance and reliability; studies and tests to develop and evaluate alternate MFB light-off techniques, fly ash reinjection and alternate fuel feed systems; construction completion and preparation of operation manuals.

17. Key Words and Document Analysis. 17a. Descriptors

Air distributor
Coal
Computer
Limestone
Sorbent
Sulfur

17b. Identifiers/Open-Ended Terms

Fluidized-Bed Module (FBM)
Multicell Fluidized-Bed Boiler (MFB)

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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, without excessive maintenance problems. The fluidized-bed boiler will produce 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 is being 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 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 ERDA/OCR Contract No. EX-76-C-01-1237 (E(49-18)-1237) is to test a multicell fluidized bed boiler as a pollution-free method of burning high-sulphur coal or highly corrosive coals without excessive maintenance.

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 Design and Operation Manuals and the Design of Utility Size Boilers

This Interim Report covers work performed under Phases II and IV of the Contract during the July 1976 and June 1977 time period.

SUMMARY

1.1 Work Covered by This Report

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

a) Administration and supervision of subcontracted work,

b) Evaluation of systems and equipment and development of modifications thereto to reflect advances in fluid bed technology and to improve performance and reliability,

c) Operation of the boiler and auxiliary systems in cold model shakedown status for acceptance and correction of operational differences,

d) Operation of the boiler and auxiliary systems in hot mode shakedown/demonstration status to prove operational performance, evaluate operational characteristics, identify areas for improvement and initiate necessary action to implement same,

e) Preparation of Design and Operation Manuals,

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

1.2 Review of Significant Events

Between the October 1972 and June 1977 contract period, the following significant events occurred:

Start of contract	October	1972
Site selection completed	February	1973
Rehabilitation of FBM complete at Alexandria, Virginia laboratory	March	1973
Site selection approved by OCR	March	1973
Evaluation of boiler manufacturers completed	March	1973
Selection of Foster Wheeler Corporation (FWC) and OCR approval	March	1973
Bureau of Mines laboratory tests completed at Alexandria	March	1973
PER boiler design completed	April	1973
FWC boiler evaluation, redesign and cold component testing contract signed	May	1973
Modernization of the Alexandria laboratory completed	June	1973
Monongahela Power Company agreement signed	June	1973
Design verification tests started in the Alexandria laboratory	June	1973
Fluidized bed boiler plant design started	June	1973
Boiler redesign completed and released for bidding	October	1973
Salt tests completed in the Alexandria laboratory	October	1973
Boiler bids submitted by two manufacturers	November	1973
Bid evaluation completed and approved by OCR	December	1973
Notice to Proceed given to FWC for boiler fabrication	December	1973
Cold component testing in the Livingston, New Jersey laboratory started	December	1973
Fluidized bed boiler plant design completed and released for bidding	January	1974
Air distribution plate pre-selection tests in the Livingston laboratory completed	January	1974
Air pollution control permit filed	February	1974
High N_2 content coal NO_x tests completed in Alexandria	February	1974
Plant construction bids received from three contractors	February	1974
Request for additional funding submitted to OCR	February	1974

Plant construction bids submitted to OCR for evaluation	February	1974
Air pollution control hearing	February	1974
Coal feeder hot tests completed in the Alexandria laboratory	March	1974
Preliminary grid plate tests completed in the Alexandria laboratory	April	1974
Construction Contract No. 1 for Rivesville boiler support steel work signed	May	1974
Air pollution permit granted by the State	May	1974
Rivesville boiler support steel work started	May	1974
Coal feeder tests started in the Rexnord, Inc., Louisville, Kentucky plant	May	1974
Furnace feed system design completed	June	1974
Combustion control system design completed	June	1974
Fluidized bed boiler plant design package, including furnace feed and combustion control systems released for re-bid	July	1974
Boiler drum and headers delivered to site	July	1974
Construction Contract No. 1 for Rivesville boiler support steel completed	July	1974
Preparation of Interim Report No. 1 started	July	1974
Design and fabrication of a horizontal tube bundle for the Alexandria FBM started	July	1974
Interim Report No. 1 completed and submitted to OCR	August	1974
Plant construction re-bids received from three contractors and submitted to OCR	August	1974
Request for additional funding resubmitted to OCR	August	1974
Erection of the 30 MW_e MFB at Rivesville started	September	1974
General Construction Subcontract signed with Champion Construction & Engineering Co., Inc.	September	1974
Construction of the large cold test model (6'x6') in the Livingston, N.J. laboratory completed	September	1974
Fabrication of the horizontal tube bundle for the Alexandria FBM completed	September	1974
Purchase of plant equipment by Champion Construction & Engineering Co., Inc. started	October	1974
Additional funding for the project approved by OCR	November	1974

General mobilization by Champion Construction & Engineering Co., Inc. completed	November	1974
Demolition Subcontract work at the Rivesville site started	November	1974
Purchase of major plant equipment by Champion Construction & Engineering Co., Inc. completed	December	1974
Testing of the large cold model with the Rivesville horizontal tube bundle arrangement started	December	1974
Fabrication and delivery of all steam generator parts from the FWEC Dansville, N.Y. and Mountaintop, Pa. plants completed	December	1974
Installation of the horizontal tube bundle in the Alexandria FBM completed	January	1975
Testing of the large cold model with the Rivesville horizontal tube bundle arrangement completed	February	1975
Initial test of the horizontal tube bundle in the Alexandria FBM completed	February	1975
Fuel injection tests in the large cold model started	February	1975
Regeneration Technology Workshop	March	1975
Phase I Erection Sequence of the 30 MW_e MFB at Rivesville completed	April	1975
Preservation work of the 30 MW_e MFB at Rivesville completed	April	1975
Developed computer program for heat and material balances for the Alexandria FBM	April	1975
Completed testing in the large cold model	May	1975
Preparation of operation and maintenance manual for the Rivesville, W. Va. MFB started	May	1975
Preparation of a test and instrument program for the Rivesville, W. Va. MFB started	May	1975
Successfully performed test and evaluated the burning of coke-breeze in the Alexandria, Va. FBM	May	1975
Installed pre-selected Rivesville MFB air distribution grid plate in the Alexandria, Va. FBM for final grid performance tests	June	1975
ERDA Technical Audit of PER - July 7 & 8, New York office, July 9 Livingston, N.J., July 10 Rivesville, July 11 Alexandria - completed	July	1975
Started final testing of the selected Rivesville MFB air distribution grid plate in the Alexandria, Va. FBM	July	1975

Preliminary review of the Rivesville, W. Va. MFB instrumentation and test program with ERDA and FWEC	July	1975
Issued Limestone Regenerator Design Report	July	1975
Completed final testing of the selected Rivesville MFB air distributor grid plate in the Alexandria, Va. FBM	August	1975
Issued Final Report for large cold model test work	September	1975
Started detailed horizontal tube bundle heat transfer testing in the Alexandria FBM	October	1975
Design for conversion of the Alexandria FBM to a Carbon Burnup Cell started	November	1975
Modification of the Alexandria FBM facility started	December	1975
Completed preparation of a test and instrumentation program for the Rivesville, W. Va. MFB	January	1976
Completed preparation of an operation and maintenance manual for the Rivesville, W. Va. MFB unit	January	1976
Preparation of an operation and maintenance manual for the Rivesville, W. Va. MFB Plant Auxiliary Systems started	February	1976
Issued Staffing and Training Program for MFB Plant Personnel	March	1976
ERDA review of the Rivesville, W. Va. MFB plant instrumentation and test program at the Plant site	April	1976
Issued Safety Manuals for MFB Plant Operating Personnel	April	1976
Completed design for conversion of the Alexandria FBM to a Carbon Burnup Cell	May	1976
Completed hydrostatic testing of 30 MW_e MFB and the high pressure steam piping system at Rivesville, W. Va.	June	1976
Completed horizontal tube bundle heat transfer testing in the Alexandria, Va. FBM	June	1976
Completed construction of the Alexandria, Va. FBM facility modifications/minimum equipment improvement	July	1976
Started cold mode operation of the Rivesville, W. Va. MFB Plant systems and equipment	July	1976
Started cold mode operation of the Alexandria, Va. laboratory new systems and equipment	July	1976
Completed chemical cleaning of the Rivesville, W. Va. MFB	July	1976

Completed cold mode operation of the Alexandria, Va. Laboratory new systems and equipment	August	1976
Started hot mode operation of the Alexandria, Va. Laboratory new systems and equipment	September	1976
Fluidized limestone bed in Cell D (CBC) of the Rivesville, W. Va. MFB	September	1976
Fired start-up burners in Cell D (CBC) of the Rivesville, W. Va. MFB	September	1976
Received first shipment of coal at the Rivesville MFB Plant, and placed in No. 5 Storage Bunker	October	1976
Started warm-up of the Rivesville, W. Va. MFB unit by continuous firing of the startup oil burners	November	1976
Accomplished first lightoff and operation with coal of the Rivesville, W. Va. MFB	December	1976
Operated one cell (CBC) of the Rivesville, W. Va. MFB unit in a continuous mode, burning coal	December	1976
Performed tests for bed material transfer between Cells D (CBC) and C of the Rivesville, W. Va. MFB	January	1977
Accomplished first light-off of Cell C and operation of two Cells (D & C) of the Rivesville, W. Va. MFB unit	February	1977
Initiated coal feeder test and evaluation program at Alexandria Laboratory	March	1977
Performed tests at Alexandria Laboratory to develop alternate light-off procedures for the Rivesville, W. Va. MFB	March	1977
Completed all Rivesville MFB plant auxiliary systems subcontract work	March	1977
Accomplished first light-off of Cell B and concurrent operation of three cells, Cells D, C and B of the Rivesville MFB unit	April	1977
Initiated modifications to MFB unit fuel feed systems to eliminate continuing operational problems	May	1977

1.3 Summary of Activities

1.3.1 General

Substantial progress was made during this reporting period in achieving overall MFB program objectives. The most significant accomplishment was successful initial light-off of the MFB unit in December 1976 and subsequent sustained coal firing operation. Continued operation of the unit brought to light several operational problems, the most troublesome which were lack of a continuous supply of boiler feedwater to support operation of the MFB during the period when steam is not exported and condensate not returned and difficulties with components of the fuel feed system which prevented continuous feeding of coal/limesonte mixture to the bed. Action was taken to correct these and other problems. Shutdown of the unit and prolonged outages was caused on several occasions by failure of commercial equipment and lack of fuel.

Additional areas of significant progress was the development of a comprehensive test instrumentation plan, development of auxiliary systems to improve the reliability and performance of the MFB and test programs carried on at the Alexandria Laboratory in support of MFB operation.

1.3.2 Laboratory Work at Alexandria, Va.

Work at the Alexandria fluidized-bed combustion laboratory, supported operation of the Rivesville MFB Unit and continued development of the atmospheric fluidized-bed combustion process.

Following is a summary of the work performed at the laboratory, including test results, process analysis results and support activities.

<u>Test Results</u>

The FBM was converted to operation as a Carbon Burn-Up Cell. A series of tests were conducted to verify the Rivesville CBC operating conditions, estimate the number of feed points necessary for good combustion, and to evaluate a prototype fly ash injection system. Tests using 3, 2 and 1 fly ash feeder(s) in the FBM indicated that one feed point was insufficient for reliable control of bed temperatures near the ash fusion point. Two feed points showed a higher degree of reliability and performed nearly as well as three feed points. The prototype fly ash injection system was found to be difficult to operate but satisfactory for large scale use.

An automatic combustion control system installed at the laboratory was set up and tuned. This system completely controls the fire side of the FBM. A boiler water level controller and a steam flow controller were also installed to control the waterside. With both units operational, the boiler operation is completely automatic.

A series of tests were conducted to study alternate coal feeding methods. Two types of underbed feeders, as well as overbed coal feeders were tested. The underbed feeders tested were mushroom cap, similar to the fly ash feeders, and a tee feeder.

Three alternate light-off procedures were studied. The first used an oil soaked bed to rapidly heat the bed to ignition temperature. The second used a partially blocked grid plate to allow ignition of only 1/3 of the bed. The third procedure used a duct burner located in the plenum to raise the bed temperature to ignition.

A study of the effect of simultaneous coal and fly ash combustion at 1600°F was begun. Tests using fly ash previously generated were conducted to determine the once through carbon burnup of fly ash. Other tests were conducted recycling the fly ash to the FBM as it was produced.

Results of the Corrosion/Erosion analysis begun in the last reporting period are reported. Two additional new racks, each containing fifteen different metals, were installed in the bed and in the freeboard of the FBM.

Process Analysis Results

A simple mathematical model was developed to correlate the results of the CBC test program. The model incorporates the effects of bed depth, superficial velocity, bed temperature, and percent excess oxygen to predict percent carbon burnup.

A mathematical model of the hot bed transfer process for light-off adjacent cells was developed. This model utilized data from Rivesville during a light-off test to calculate an orifice flow coefficient for hot bed material flowing through the transfer gate between Cells C and D.

A mathematical model was developed to calculate the maximum amount of oversize ash which will accumulate in a fluidized bed. The model was applied to size distribution of Sewickley coal and Greer Limestone used at the FBM.

The shrinking core model to predict sulfur capture in a fluidized bed was expanded to permit calculation of the effect of a classifier operation and the effect of various bed depths.

Support Activities

A computerized Data Acquisition System was installed at the Alexandria FBM Facility. The equipment, supplied by MITRE Corporation, includes a mini-computer and a tape drive to store test data on magnetic tape. The system automatically records most of the data previously taken by hand.

1.3.3 Rivesville MFB Plant Operations

Construction of the MFB plant was completed and the plant accepted from the construction subcontractor early in the reporting period. Equipment and systems were operated in the cold mode for checkout and calibration of instrumentation. The first successful coal firing of the unit was accomplished in December 1976. Subsequent sustained coal firing of Cells C, C and B was also accomplished during the remainder of the reporting period. Operation of the unit was curtailed due to lack of adequate boiler feedwater supply from MPC and problems with the fuel feed system. Action was taken to correct these conditions. Several equipment failures occurred which resulted in shutdown and prolonged outages. These included shearing off of the F.D. fan blades, breaking of the electrostatic precipitator structure, distortion of the MFB grid plates. To encourage continuous operation of the coal feed system on an interim basis, coal for combustion will be limited to dry, double screened coal in the -½" to +¼" range. Coal to this specification was difficult to obtain locally which contributed to sporatic operation of the MFB unit.

Repair and maintenance activities were carried on throughout the reporting period as required, to improve operations, repair equipment failures and modify systems as determined by operating experience.

Formal light-off procedures, including supporting charts and tables of MFB characteristics were prepared to standardize techniques for Cell D light-off and subsequent light-off of Cells C, B and A.

1.3.4 New York Office Design and Engineering Work

Design and engineering work was carried on during this reporting period at PER's New York office to improve the performance, increase the reliability and alleviate operational problems of the Rivesville MFB unit.

Work included development of a CO_2 Fire Inerting System to detect and extinguish fires in Bunker Nos. 2, 5 and 6, resulting from oxidation of stored high sulfur coals and development of a waste material sample storage system to transport, collect and unload MFB waste materials.

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

- improving limestone delivery,
- evaluating coal dryer operation,
- improving MFB light-off procedures,
- improving fuel feed systems,
- investigation of alternate feedwater sources,
- coordination of MFB test program requirements,
- improve coal handling systems.

1.3.5 MFB Steam Generator Design, System Development, Model Testing, Operating Instructions and Supplemental Construction

Activities under this heading, relating to the MFB unit, were accomplished by Foster Wheeler Energy Corporation under a subcontract to Pope, Evans and Robbins Incorporated. Much of the work was completed and reported in previous Interim Reports. Activities during this period included completion of boiler installation and chemical cleaning. A revised test instrumentation plan was developed, reflecting ERDA's comments on preliminary plan and implementation of the revised plan initiated. The MFB plant operation manual was completed and issued during this period. Miscellaneous modifications to the MFB installation for improved operation were completed. A study was conducted evaluating various alternate MFB light-off techniques to reduce light-off time and approval granted to proceed with recommended scheme. Approval also granted to modify coal feed system and provide automatic control of vibrating feeder pressurizing air.

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 full-scale boiler module, designated the FBM (fluidized boiler module) with a maximum bed cross section of approximately nine square feed. Heat transfer surfaces, including a submerged horizontal tube bundle, are capable of handling 10 x 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.

The Alexandria facility is operated by Pope, Evans and Robbins Incorporated (PER) with equipment provided by ERDA, to simulate design conditions of the multicell fluidized-bed boiler, verify operational and component performance parameters, and identify and find solutions to potential problem areas.

The overall objectives of the work are process development which encompasses hardware design, operational procedures, analysis and performance evaluation methods, application and review of basic technology, economic improvements, and instrumentation and control application.

The laboratory program attempts to utilize and combine the diverse technologies of combustion, fluidization, steam generation, chemical reaction and air pollution control in the conduct of the experimental work.

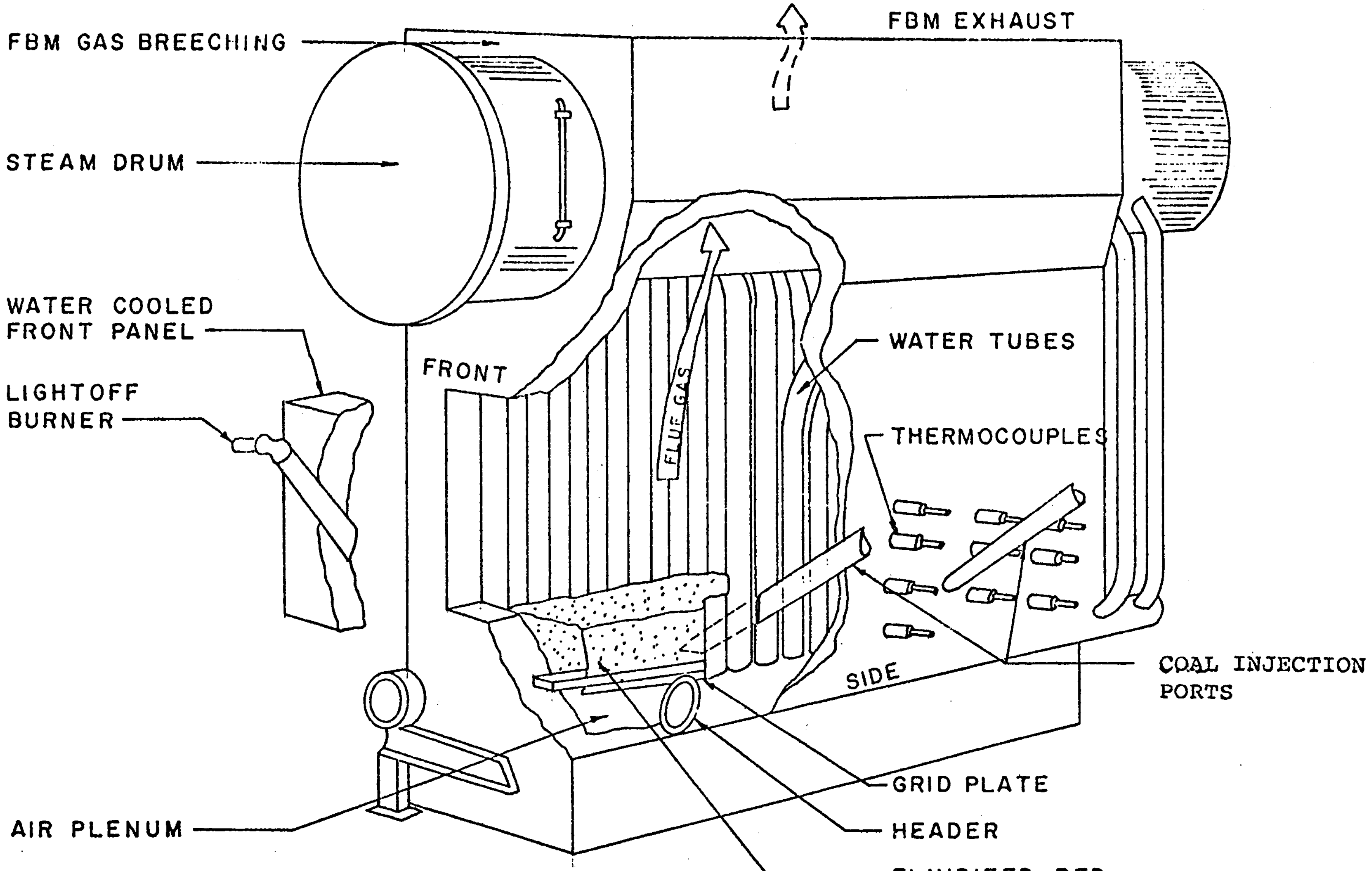
2.2 Laboratory Systems

2.2.1 Full-Scale Boiler Module

The FBM is an atmospheric pressure fluidized bed reactor capable of generating steam up to 300 psig. In this unit, the fluidized-bed is contained in a rectangular enclosure in which each side wall is a row of verticle boiler tubes seal-welded so as to form a gas-tight enclosure.

A cut-away view of the FBM, essentially the same as when first placed into service in 1967, is shown in Figure 2.0.

FIGURE 2.0



FLUIDIZED-BED MODULE (FBM) INTERNAL CONSTRUCTION

The FBM air flow cross section is approximately 18 x 72 inches, providing a maximum of 9 square feet for fluidization. The bed is surrounded by vertical boiler tubes which extend from two cross headers below the grid plate to the steam drum. The boiler tubes are joined together by welded fins and are backed by insulation. The combustion space is accessible through a water-cooled panel at the front of the unit. The panel contains a view port and a premix gas burner used to fire the bed during start-up. The burner directs a flame downward onto the front of the bed.

From the plenum at the base of the unit, air is directed upward through the air distribution grid to fluidize a limestone bed. The static bed depth may be varied from 6" to over 30", although the useful range is narrower. Bed sampling drop out pipes and valves are provided. Thermocouples are mounted throughout the bed.

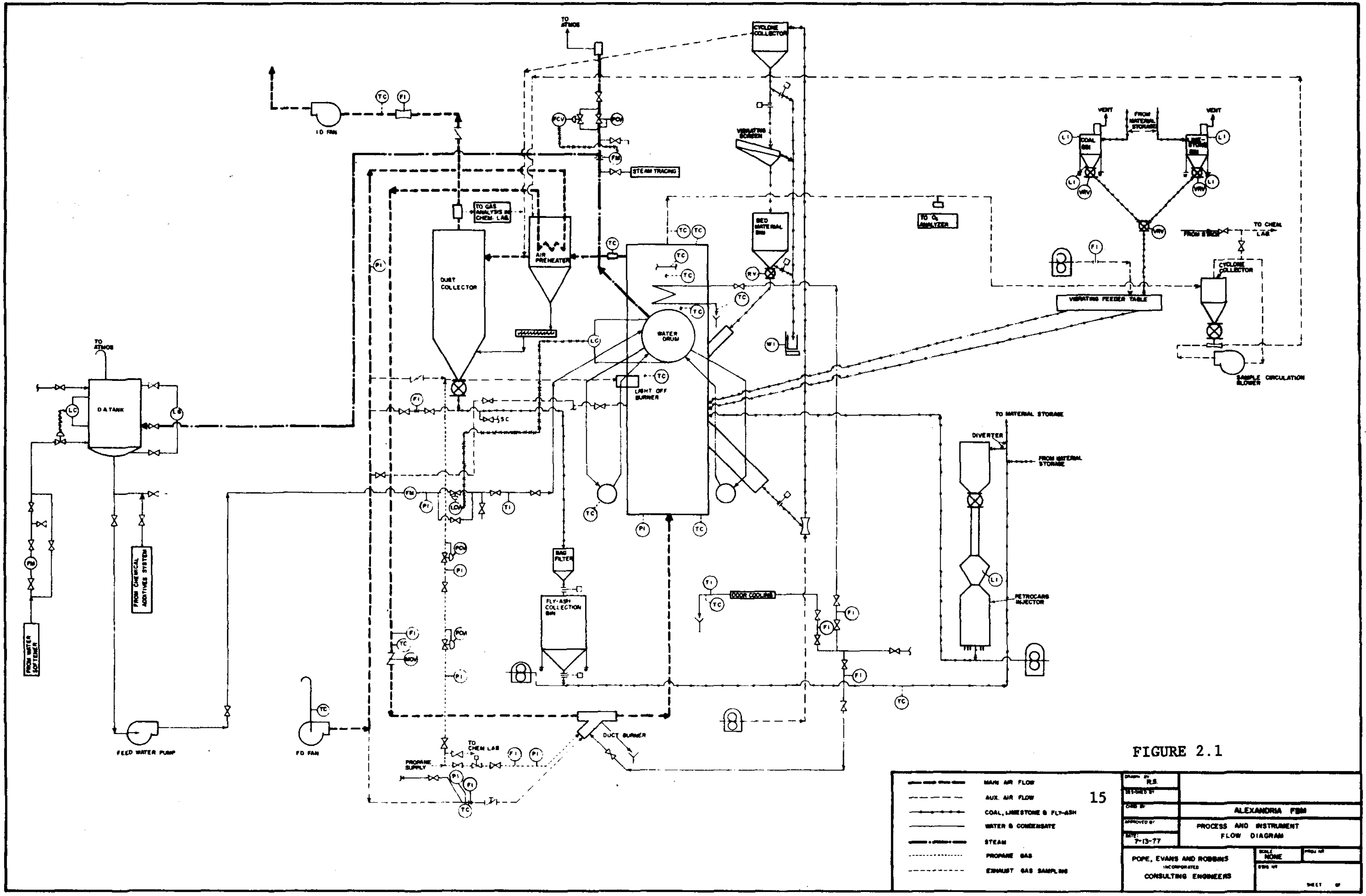
In operation, the bed is raised to the ignition point of coal by use of a propane burner mounted in the front panel. Combustion of the coal begins in the vicinity of the light-off burner flame and propagates rapidly throughout the bed. A water-cooled tube array just above the drum simulates the convection bank of a conventional boiler. The FBM produces about 3,500 lb/hr of steam at 200 psig when 600 lb/hr of coal is fed. Figure 2.1 shows the schematic flow diagram for the combustor and auxiliary systems utilized during the reporting period.

2.2.2 Fuel and Sorbent Systems

During the reporting period, high sulfur Sewickley coal and fly ash were fired with Greer limestone as the sorbent bed material. All solid fuels and sorbents are processed, stored, and delivered by the automated materials preparation and handling system which was completed during the first quarter of the reporting period. A detailed description of this system can be found in Interim Report No. FE/1237/Y-76/3.

Coal is dried in a fluid bed dryer, crushed in a roll crusher, screened on a vibrating screen with the +¼" size recycled. Coal under ¼" is stored in a silo and pneumatically conveyed to a weigh hopper located on the roof of the combustion laboratory. The coal feed rate from the weigh hopper is controlled by a variable speed rotary valve.

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


TO ATMOS
I D FAN
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CYCLONE COLLECTOR
VIBRATING SCREEN
BED MATERIAL BIN
TO O_2 ANALYZER
VENT
FROM MATERIAL STORAGE
COAL BIN
LIME-STONE BIN
LI
VRV
FROM STACK
TO CHEM. LAB
CYCLONE COLLECTOR
VIBRATING FEEDER TABLE
SAMPLE CIRCULATION BLOWER
TO GAS ANALYSIS IN CHEM. LAB.
DUST COLLECTOR
AIR PREHEATER
WATER DRUM
LIGHT OFF BURNER
LC
RV
WI
PI
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SC
LCV
D.A. TANK
LS
FROM WATER SOFTENER
FROM CHEMICAL ADDITIVES SYSTEM
FEED WATER PUMP
FD FAN
BAG FILTER
FLY-ASH COLLECTION BIN
DOOR COOLING
TO MATERIAL STORAGE
DIVERTER
FROM MATERIAL STORAGE
PETROCARB INJECTOR
MOV
DUCT BURNER
TO CHEM LAB
PROPANE SUPPLY
MAIN AIR FLOW
AUX. AIR FLOW
COAL, LIMESTONE & FLY-ASH
WATER & CONDENSATE
STEAM
PROPANE GAS
EXHAUST GAS SAMPLING
DRAWN BY R.S.
DESIGNED BY
CHKD BY
APPROVED BY
DATE: 7-13-77
ALEXANDRIA FBM
PROCESS AND INSTRUMENT FLOW DIAGRAM
POPE, EVANS AND ROBBINS INCORPORATED
CONSULTING ENGINEERS
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FIGURE 2.1

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

2.2.3 Gas Supply and Control System

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

The position of both the F.D. and the I.D. dampers can be controlled from the automatic combustion control panel. The I.D. damper can either be set manually or automatically to maintain a set furnace pressure. The F.D. damper can either be set manually, to deliver a constant air flow, or automatically, in response to system variables, through the automatic combustion control system which is described ~~elsewhere~~ in paragraph 2.2.7.

2.2.4 Feedwater and Steam System

The FBM uses water of a quality commensurate with the operating pressure of 200 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 deaerating feedwater heater. Steam produced in the FBM is vented to the atmosphere. A back pressure regulator is used to maintain steam pressure in the boiler. The operating pressure is normally set at 100 to 200 psig.

A new aspect of the boiler feedwater and steam system was the addition of automatic controls to the boiler feedwater valve and the addition of a constant steam flow control valve. These improvements were made in the third quarter of the reporting period. The feedwater controller is a P-I Controller that maintains the boiler water level in the steam drum at a constant set point. The constant steam flow controller overrides the back pressure regulator. It utilizes a flow signal from the main steam flow orifice.

2.2.5 Fly Ash Collection System

The products of combustion are separated from the gas by a multi-cyclone dust collector. Heavy duty vibrators are installed on the hoppers under the preheater and cyclone collector to prevent fly ash from accumulating on the hopper walls. The ash is screw-fed from the air preheater hopper to the mechanical collector hopper and pneumatically conveyed from the cyclone hopper to a baghouse and weigh station outside the laboratory.

2.2.6 Fly Ash Reinjection System

During the reporting period a system to reinject fly ash into the FBM was evaluated for use in the Rivesville 30 MW_e MFB. A Petrocarb Incorporated fly ash reinjection system was installed during the first quarter of the reporting period and has been in operation since that time. The injector is a proprietary design of Petrocarb Incorporated. Since its installation, the injector has been used to feed both fly ash and 1/8" coal to the FBM in various test programs.

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

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

The Petrocarb fly ash injector was equipped with an automatic control system. This system was designed so that the unit would deliver all fly ash fed to it. Figure 2.2 shows a schematic of the automatic control system. The capacitance level probe is converted to a signal which acts as the set point for the automatic controller. The controller maintains the pressure in the primary injector at the set point. In this way the level in the surge vessel will reach a stable level which will give the proper primary injector pressure for the fly ash feed rate.

Considerable time was expended calibrating the fly ash injector and solving operational problems. Problems with holdup of fly ash or coal in the seal leg and with the automatic controls prevented normal operation of the unit during most of the testing programs. Operation of the unit during these tests is discussed in the test results section. However, with elimination of the seal leg, the Petrocarb injector system appears to be operating satisfactorily.

2.2.7 Automatic Control System

During the reporting period a pneumatic automatic control system was installed at the laboratory. The control system was designed to operate the FBM based upon the control logic used in the Rivesville 30 MW_e MFB system.

Figure 2.3 shows a schematic of the FBM/CBC control system at Alexandria. 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. damper to maintain a constant furnace pressure. Loop two, the bed temperature controller, consists of a P-I controller to maintain bed temperature, and eight proportional controllers to adjust the coal and combustion air feed rates to compensate for fluctuations in steam pressure, air temperature, furnace pressure, and percent O_2 in the stack. The third control loop is the bed level and SO_2 control loop. This control loop uses a P-I controller to maintain bed height and three proportional controllers to adjust the limestone feed and bed removal rates to maintain a set SO_2 concentration in the stack. Figure 2.4 shows the interface schmatic between the FBM and the control system.

FLYASH FEED CONTROL SYSTEM

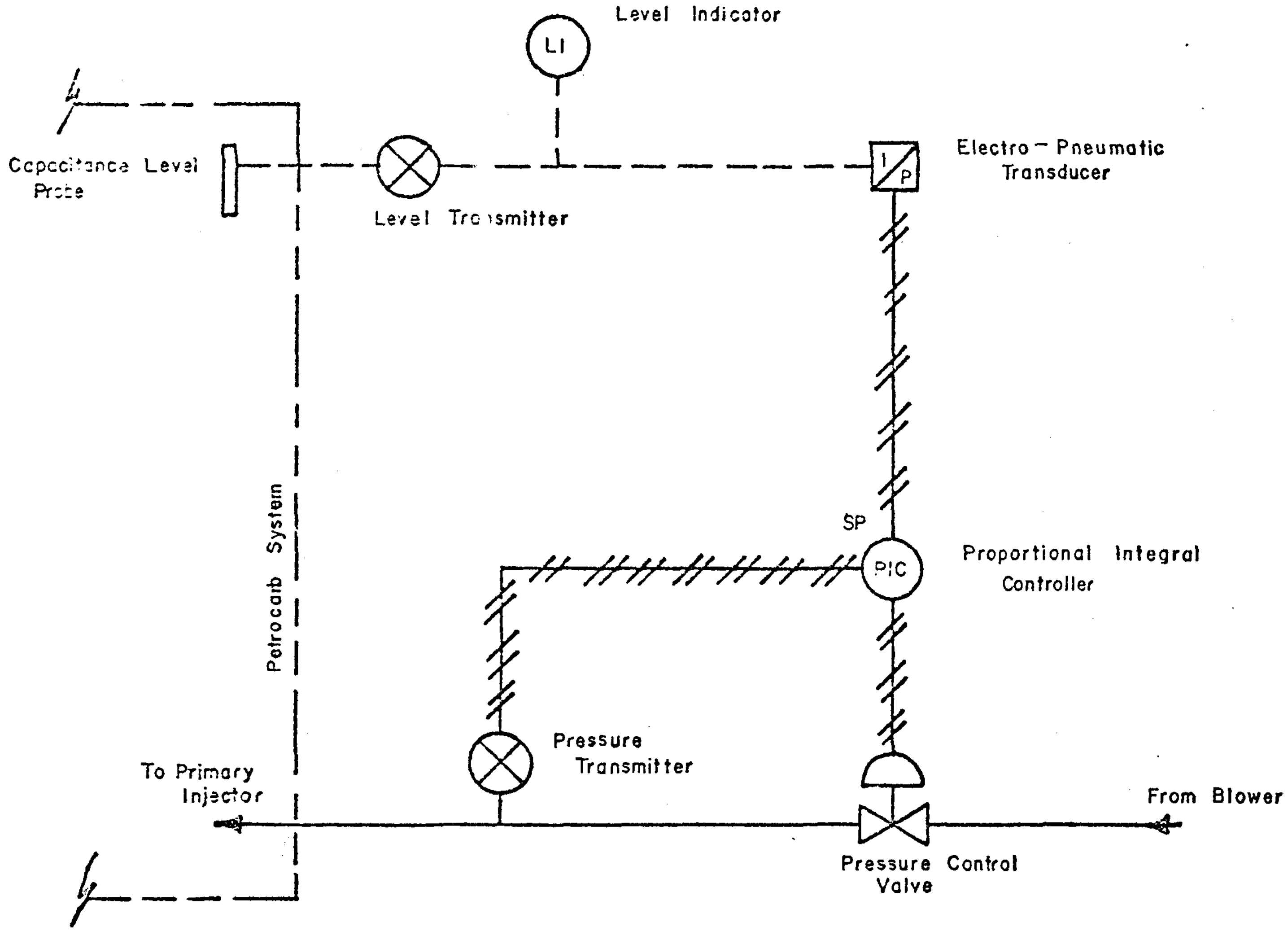
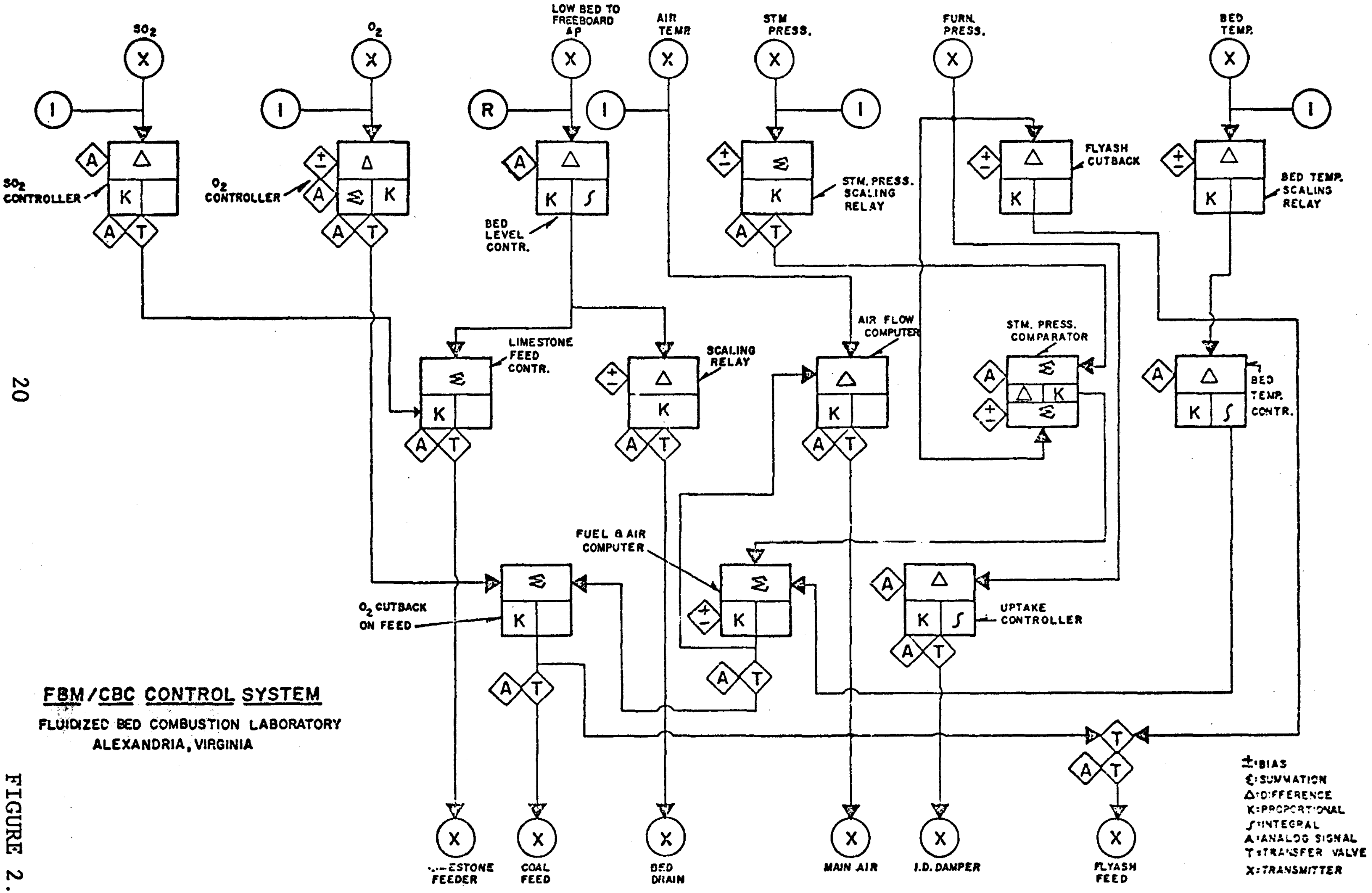


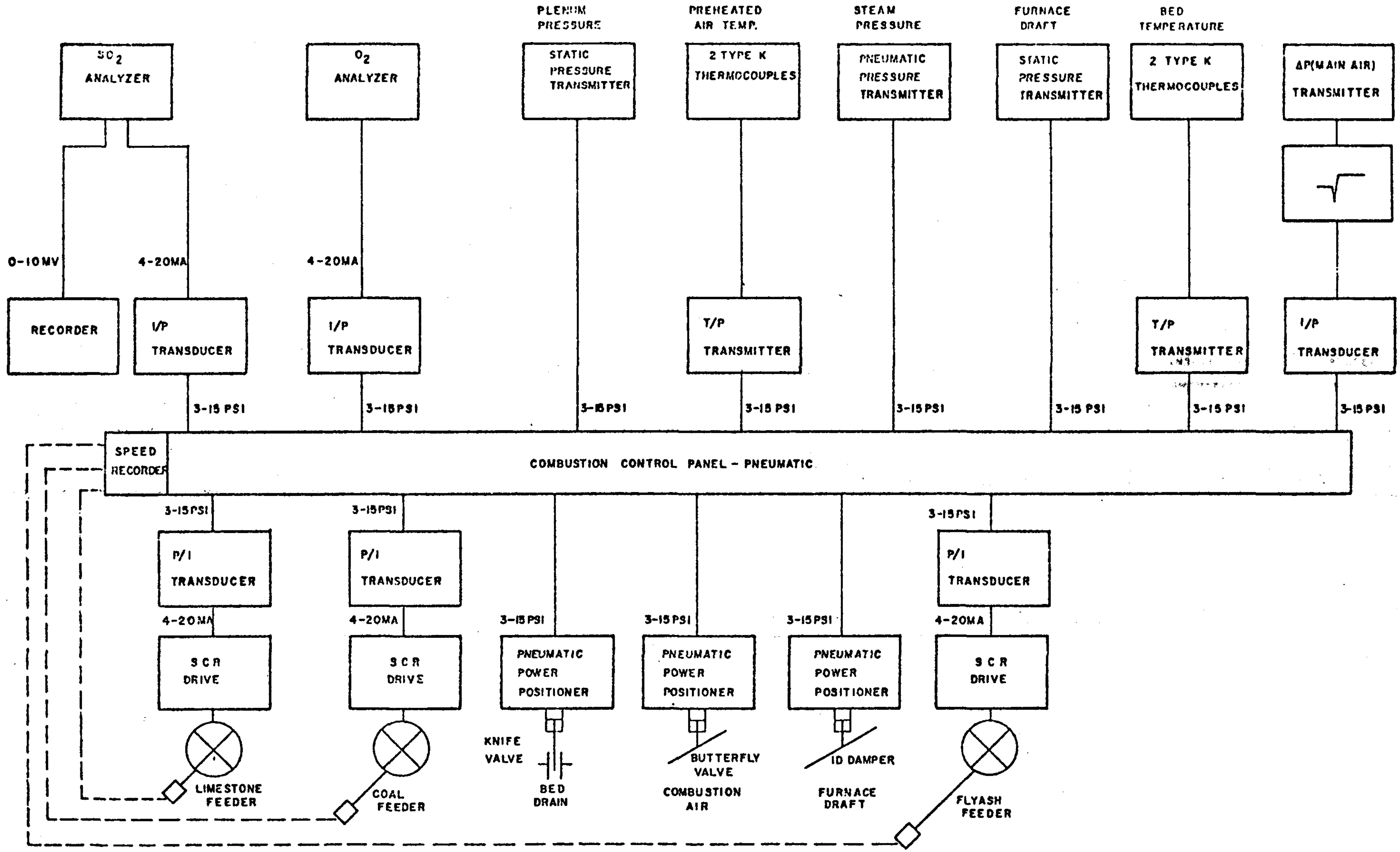
FIGURE 2.2



FBM/CBC CONTROL SYSTEM

FLUIDIZED BED COMBUSTION LABORATORY
ALEXANDRIA, VIRGINIA

FIGURE 2.3



FBM CONTROL PANEL TO PROCESS INTERFACE SCHEMATIC

FIGURE 2.4

2.2.8 Chemical Analysis Systems

The fluidized-bed combustion laboratory includes a chemical laboratory which has the capability of handling nearly all of the required gas and solid chemical analyses. On line gas analyses are performed and recorded during the test by automatic flue gas monitoring equipment. Chemical analyses of solid samples taken periodically during 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.

Flue gas is continually sampled by the gas sampling and conditioning system described in (Report No. FE/1237/Y-76/3. ~~Interim No. 3~~. After passing through the sample conditioning system which permits determination of the moisture content of the sample, the sample is continuously supplied to instruments which analyze the gas for carbon dioxide, carbon monoxide, oxygen, nitric oxide, NO_x, sulfur dioxide, and total hydrocarbons.

Table 2.1 lists the gas analysis equipment, as well as methods used to analyze solid samples.

Routine analysis are performed on samples of coal, limestone, bed material, fly ash, dust, flue gas, and boiler water. Samples are collected according to the following methods:

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

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

ASTM Designation D2928-71, Standard Method of Sampling Stacks for Particulate Matter.

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

2.2.9 Data Acquisition and Reduction Systems

Instrumentation and associated data acquisition equipment includes: thermocouples with multipoint strip chart recording and digital data logging; main air, steam and tube bundle water flow orifices: Bailey, Westinghouse, Taylor, and Thermox brand oxygen analyzers with recorders; and flue gas analyzers listed previously with strip chart recorders. Solids feed rates are determined from load cell readings recorded on a digital data logger. The auxiliary air and gas flows, cooling water rates and operating pressure are observed on manometers and logged on data sheets by laboratory personnel as required. During the reporting period, a time

TABLE 2.1

Methods of Analysis

Analysis	Sample Type	Method
Carbon Hydrogen	Coal Limestone Bed material Fly ash Dust	Coleman Model 33 Carbon-Hydrogen Analyzer.
Sulfur	Coal Limestone Bed material Fly ash Dust	LECO Sulfur Determinator Model 517.
Calcium	Coal Limestone Bed material Fly ash Dust	ASTM Designation D2795-69 (Reapproved 1974), Standard Method of Analysis of Coal Ash and Coke Ash.
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.
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

Analysis	Sample Type	Method
Moisture	Coal Limestone	ASTM Designation D3173-73, Standard Method of Test for Moisture in the Analysis Sample of Coal and Coke.
Size distribution	Coal Limestone Bed material	ASTM Designation D410-38 (Reapproved 1969), Standard Method of Sieve Analysis of Coal.
Bulk density	Coal Limestone Bed material	Volume and mass unpacked sample is measured in a graduated cylinder.
Particle density	Coal Limestone Bed material	Mass is determined by weighing in a graduated cylinder. Liquid displacement is used for volume measurement.
Hardness	Boiler water	Taylor Chemical Inc. total hardness drop test kit.
pH	Boiler water	Orion Model 801 digital pH/MV meter
Phosphate Sulfite	Boiler water	Water Services Laboratories Water Test Kit.
Carbon dioxide Carbon monoxide Nitric Oxide	Flue gas	Beckman NDIR
Sulfur dioxide	Flue gas	Thermo Electron Corp. SO_2 Pulsed Fluorescent Analyzer and Theta Sensor SO_2 Emission Monitor.

Sheet 3 of 3

TABLE 2.1
Methods of Analysis Continued

Analysis	Sample Type	Method
Nitric oxide and NO_x	Flue gas	Thermo Electron Corp. Chemiluminescent NO-NO_x Gas Analyzer.
Oxygen	Flue gas	Taylor Servomax Oxygen Analyzer.
Hydrocarbons	Flue gas	Beckman ID8A Hydrocarbon Analyzer.
Heat of Combustion	Coal Fly ash	Plain Jacket Parr Oxygen Bomb Calorimeter.

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

sharing computer and an HP-97 programmable calculator have been used to reduce the time required for data handling and reduction after data has been collected.

2.2.10 New Alexandria PDU

During the last half of the reporting period a program was initiated to design and produce a new fluidized bed boiler to replace the existing Process Development Unit (PDU). The new unit will have a square cross-section as compared to the present rectangular section. The unit will have a nominal cross-sectional area of 12 ft^2. Steam generation will take place in the waterwalls by natural convection and by forced circulation in boiling tube bundles located inside the bed and above the bed. Two other smaller tube bundles in the bed will act as a superheater and an economizer. Steam production will be about 4,000 lb/hr at 300 psig and 150°F superheat. The boiler is being manufactured by Cleaver Brooks.

Figure 2.5 shows the proposed water and steam flow schematic for the new unit. Some boiler feedwater will pass through the economizer tubes located within the bed. The remainder of the required feedwater will pass through a liquid level control system and into the steam drum. Water for the boiling bundles will be withdrawn from the lower headers by a recirculation pump. The steam produced will be passed through the bed superheater bundle, cooled to saturation in a desuperheater station, and condensed in air cooled heat exchanger. The combination economizer-superheater in-bed bundle can be removed, in which case steam flow can be controlled either by a steam flow controller or a back pressure regulator. Some of the desuperheated steam will be used to preheat the air.

2.3 Laboratory Results

During the reporting period, test programs included carbon burnup cell (CBC) testing, automatic control tests, coal feeder tests, alternative light-off procedures, simultaneous combustion of coal and fly ash, and corrosion/erosion tests. No hot tests were conducted from July 1, 1976 to September 19, 1976 due to the construction program for the automated materials handling system. Hot tests conducted during the reporting were as follows:

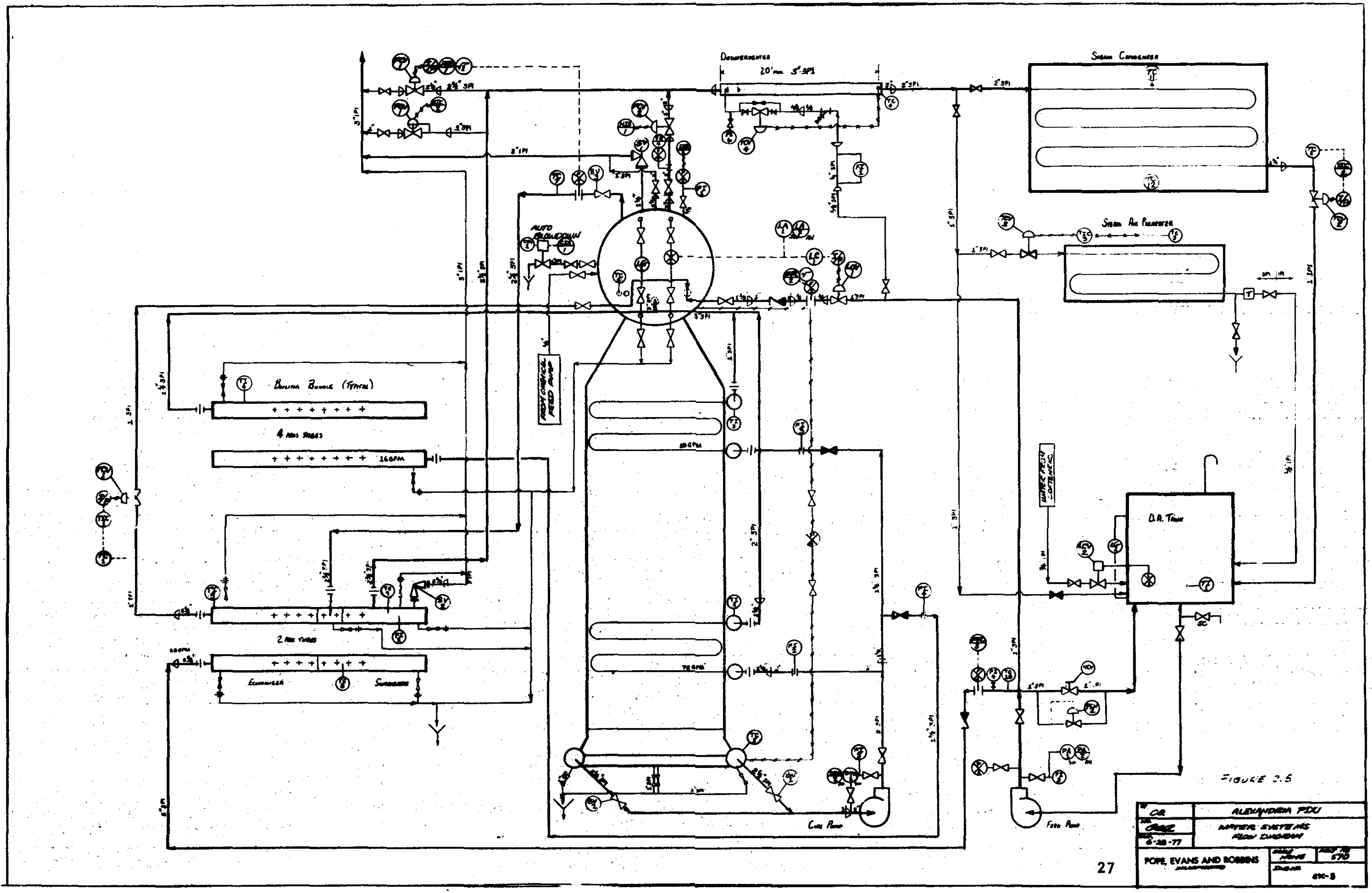

Desuperheater
20' min. 5"-3PI
Steam Condenser
Steam Air Preheater
Auto Blowdown
From Chemical Feed Pump
Boiling Bundle (Typical)
4 rows tubes
16GPM
2 row tubes
Economizer
Superheater
Water from Softener
D.A. Tank
Circ Pump
Feed Pump
CR
6-28-77
ALEXANDRIA PDU
WATER SYSTEMS FLOW DIAGRAM
POPE, EVANS AND ROBBINS INCORPORATED
NONE
570
SK-3

FIGURE 2.5

Test No.	Date	Test Description
	9/20/76	Shake down run on new systems and production of spent bed material for ERDA Projects
	9/27/76	Same as above.
	9/28/76	Same as above.
	10/1/76	Same as above.
	10/11/76	Same as above.
	10/14/76	Same as above.
644-1	10/25/76	Baseline Test Run for CBC Tests
644-2	10/27/76	Shake down fly ash feed system.
	11/10/76	Shake down after insulating upper water walls.
644-4	11/16/76	CBC Test, 3 feeders
644-5	11/19/76	CBC Test, 3 feeders
644-6	11/23/76	CBC Test, 3 feeders
644-7	12/1/76	CBC Test, 2 feeders
644-8	12/2/76	CBC Test, 1 feeder
644-9	12/15/76	CBC Test, 2 feeders
644-10	12/16/76	CBC Test, 1 feeder
645-1	1/12/77	Automatic control test
645-2	1/18/77	Automatic control test
644-11	1/27/77	CBC Test, 2 and 1 feeder
644-12	2/3/77	CBC Test, 1 feeder
644-13	2/8/77	CBC Test, 1 feeder
644-14	2/10/77	CBC Test, 1 feeder
648-1	2-15 through 2/18/77	77 hour continuous run/ automatic control test
648-2	2-1 through 2/25/77	86 hour continuous run/ automatic control test
648-3	3/1/77	Automatic control test
648-4	3/2/77	Automatic control test
647-1	3/22/77	Light-off test/oil assisted
647-2	3/23/77	Light-off test/oil assisted
646-1	3/17/77	Shake down coal feeder test
646-2	3/18/77	Shake down coal feeder test
646-3	3/24/77	Coal feed test
646-4	3/30/77	Coal feed test
647-3	3/31/77	Light-off test/reduced air flow
646-5	4/14/77	Coal feed test
646-6	4/18/77	Coal feed test
646-7	4/20/77	Coal feed test
646-8	4/25/77	Coal feed test
646-9A	4/27/77	Coal feed test
646-9	4/28/77	Coal feed test
647-4	4/29/77	Duct burner light-off test
647-5	5/3/77	Duct burner light-off test
646-10	5/16/77	Coal feed test
650-1	5/19/77	Simultaneous coal and fly ash test/duct burner light-off
650-2	5/23/77	Simultaneous coal and fly ash test, open loop

Test No.	Date	Test Description
650-3	5/25/77	Simultaneous coal and fly ash test, open loop
650-4	5/26/77	Same as above, open loop
650-5	6/2/77	Same as above, open loop
650-6	6/10/77	Same as above, closed loop
650-7	6/15/77	Same as above, closed loop
650-8	6/17/77	Same as above, closed loop
650-9	6/21/77	Same as above, open loop
650-10	6/23/77	Same as above, open loop
650-11	6/28/77	Same as above, open loop

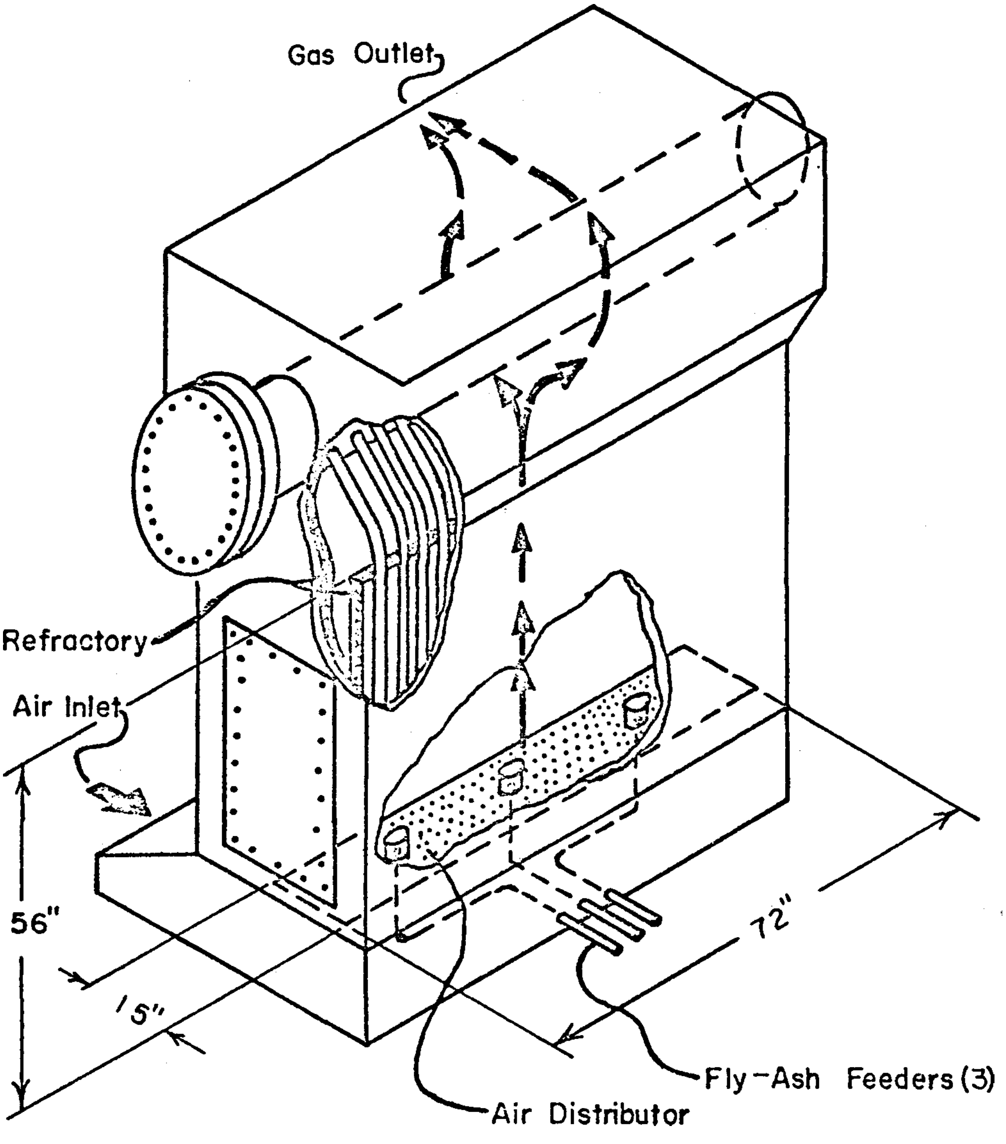
A summary of the operating conditions for the hot tests, other than shakedown tests, is given in Appendix A.

2.3.1 Carbon Burn-Up Cell Tests and Results

During the reporting period a test program was conducted operating the FBM as a carbon burn-up cell. The purpose of the test program was to verify design operating conditions for the Rivesville CBC and to evaluate the Petrocarb Inc. fly ash injection system for use at Rivesville.

The test program was based on modification of the fluidized bed module (MFB) to combust fly ash at 2000°F, 5 percent oxygen in the flue gas, 24 inch static bed depth, and 9 ft/sec. superficial velocity. When lined with 1½ inch thick refractory, to permit increased operating temperatures, the FBM has a cross section of 7.5 square feet as can be seen in the general arrangement of the combustion chamber, Figure 2.6. Three fly ash feeders of the type shown in Figure 2.7 were installed through the air distributor and mainifolded to allow feeding either one, two, or three feed points. Two coal and limestone feeders entering the side of the chamber from above were utilized as required during the test program for light-off and to continue operation between periods of fly ash feeding.

The fly ash fuel for the test program was produced by burning high sulfur Sewickley coal in a bed of Greer limestone during several previous test programs. This fly ash was collected in a multicyclone collector, approximately 80 percent efficient. Table 2.2 indicates the size distribution and composition of the fly ash fuel. The carbon content of the fly ash varied due to the different operating conditions during which the fly ash was produced. Since a fluidized bed will elutriate material below a certain size, depending on the superficial velocity, fly ash produced in the FBM and burned in the test program should be representative of fly ash produced in the larger MFB cells. Two differences between test conditions and MFB operation are the greater freeboard height of the MFB cells and the higher collection efficiency of the Rivesville mechanical collectors. Based on these differences the overall



FLUIDIZED BED MODULE

FIGURE 2.6

FLY-ASH FEED NOZZLE CROSS SECTION

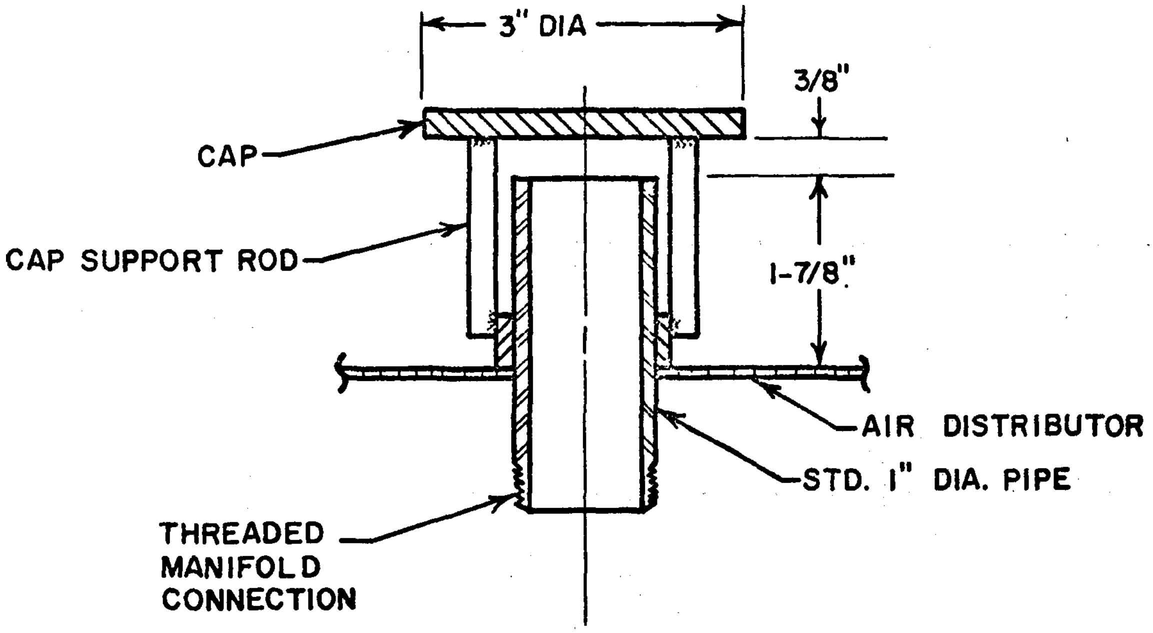


FIGURE 2.7

TABLE 2.2

Typical Fly Ash Fuel

SIZE DISBRIBUTION		COMPOSITION	
USS Sieve Size	Mass Fraction	Component	% Range
20	0.001	Carbon	32-45
- 20 + 30	0.006	Calcium	8-13
- 30 + 40	0.029	Sulfur	2-3.5
- 40 + 48	0.068	Hydrogen	0.5-1.0
- 70 + 80	0.120	Balance*	57.5-37.5
- 80 +100	0.101		
-100 +140	0.122		
-140 +170	0.117		
-170 +200	0.073		
-200 +270	0.036		
-270 +325	0.097		
-325 +400	0.033		
-400	0.156		

* Coal and limestone derived ash.

carbon burnup in the multicell design may be better than in the test system.

Bed material for the fly ash combustion tests was limestone which had been calcined and sulfated while burning high sulfur coal. No attempt was made to capture sulfur or replenish bed material during the test runs. Bed material was withdrawn to maintain a constant bed level.

A major limitation to the scope of the test program was the availability of only 30 tons of fly ash. This meant that when unforseen problems were encountered, fly ash combustion had to be stopped and solutions attempted by cold testing or by firing coal.

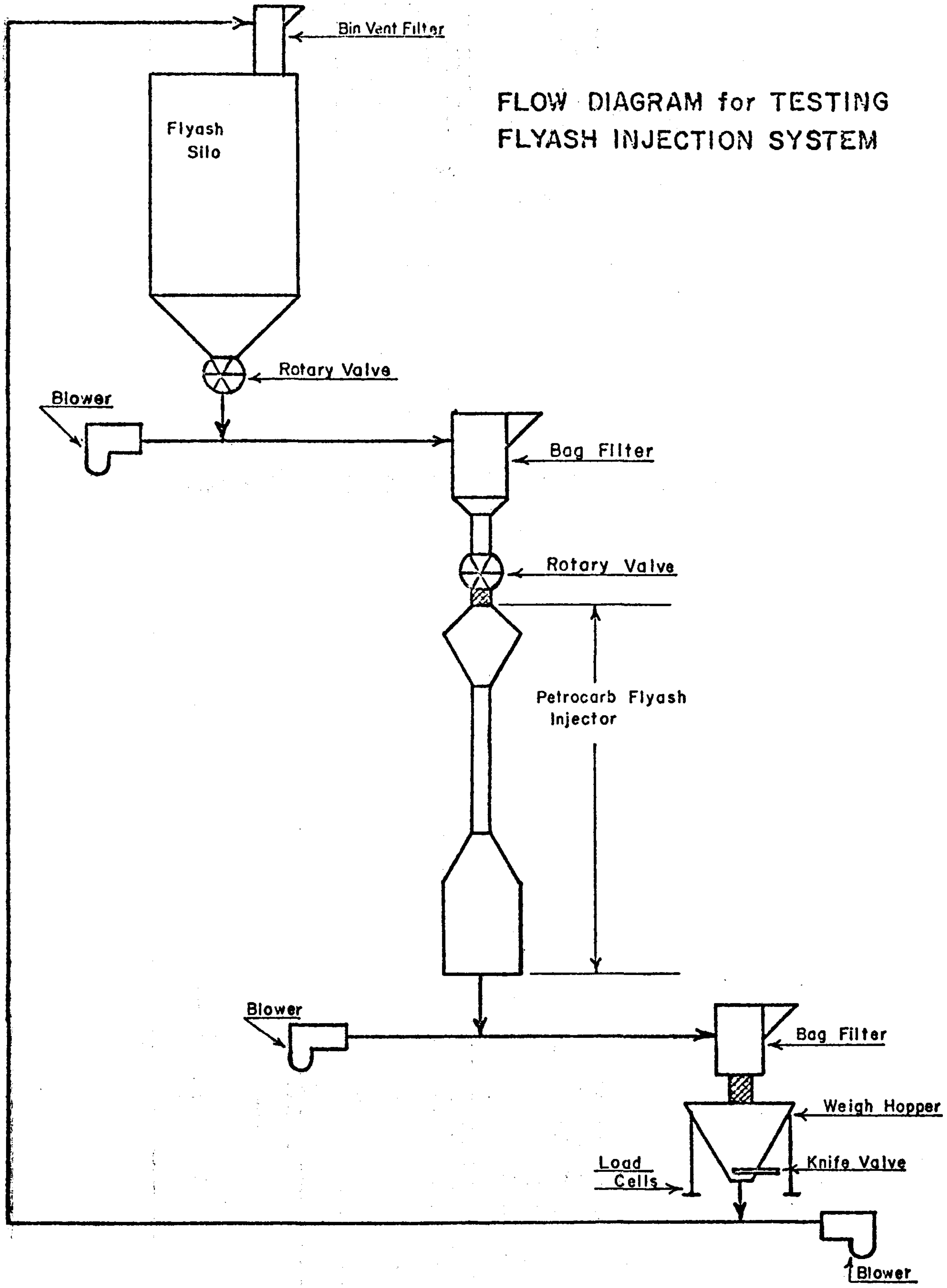
TEST APPARATUS DEVELOPMENT

Two major development problems of the test program were (1) operation and control of the fly ash feed system, and (2) inability to operate the FBM at 2000°F bed temperature over a wide range of bed depths and excess oxygen.

Fly Ash Feed System

The prototype fly ash feed system designed and furnished by Petrocarb, Inc. was based on adaptation of their proprietary design to the low pressure Rivesville application. The fly ash feed system has two primary functions; dividing the fuel into multiple streams with equal solids flow rate and delivering the fuel to the combustor. A thirty foot column of fly ash was used to seal the system and thus permit application of low pressures for overcoming line losses and the back pressure at the bottom of the fluidized bed. Adjustment of the feed system and evaluation of feed splitting accuracy was attempted by collecting fly ash in filter bags for ten minute test runs. Due to the ease with which the fly ash was fluidized, turndown of the feeder flow rates was very difficult without the back pressure of the fluidized bed. Hence, the very short duration coal tests were not satisfactory for the evaluation of the system.

Combustion tests were started with minimal control of the fly ash feed. During the first hot tests it was discovered that bridging in the thirty foot column of fly ash limited the total fly ash feed rate. A development effort to eliminate the bridging problem at high throughput rates was conducted using the closed loop test circuit shown in Figure 2.8. This led to the possibility of operating the system partially full of fly ash with a rotary feeder to seal the system. Combustion tests were restarted and development of the seal leg column continued during the test program.



Bin Vent Filter
FLOW DIAGRAM for TESTING FLYASH INJECTION SYSTEM
Flyash Silo
Rotary Valve
Blower
Bag Filter
Rotary Valve
Petrocarb Flyash Injector
Blower
Bag Filter
Weigh Hopper
Load Cells
Knife Valve
Blower

FIGURE 2.8

Combustion Heat Removal

The waterwalls of the FBM were covered with refractory to raise the operating bed temperature above the normal range of 1500-1500°F. Coverage was increased in vertical increments in an attempt to reach a heat balance of 2000°F bed temperature, 5 percent oxygen in the flue gas and 24 inch static bed depth. It was found that operation at these conditions could not be sustained in the FBM.

Heat balance tests firing coal with the refractory installed indicated that steam production was greater than predicted at the deeper bed depths. This was believed to be the result of increased splashing of bed material into the region of the steam drum as the bed depth increased. With maximum coverage of the waterwalls and the bottom of the steam drum insulated, nearly 30 percent of the heat released by combustion still produced steam in the boiler. To achieve bed temperatures of 2000°F and higher, the operating ranges for bed depth and superficial velocity had to be reduced to reduce the heat removal by steam production. It was interesting to note that in a transition from coal to fly ash feed, a step change in steam production would occur, indicating an increased percentage of combustion above the bed near the steam drum. This means that it was more difficult to raise bed temperature on fly ash than on coal firing.

Another problem which made high temperature operation difficult to sustain was changing operating parameters during a test run. The starting bed material for most tests consisted of limestone products and coal ash with a mean particle size of approximately 1000 microns. As a test progressed, the continuous feeding of fly ash fines and the low superficial velocities caused a build up of fines in the bed, resulting in a final bed mean particle size as low as 400 microns. Due to the higher heat transfer rate with decreased bed particle size, bed temperatures attained at the start of a test could not be reached at the end of a test period. Bed depth also increased heat transfer. This could be controlled, however, by draining bed material. Changes in composition of the fly ash feed required frequent readjustment of the fuel feed rate. When firing the lower carbon content fly ash, increase in the fuel rate would sometimes result in a bed temperature decrease due to heat loss to the inerts. Since the carbon content of the fly ash feed was not predictable, flue gas oxygen and bed temperature varied throughout the test runs.

TEST RESULTS

Carbon Burnup

Fourteen hot test runs were made over a period of about five months. Due to a general improvement in equipment performance and operating techniques during the test program, the later test results are believed to be more reliable. In two of the early test runs and in the last test run, the bed temperature exceeded the ash fusion point and the bed solidified. Data for selected times during the test runs is presented in Table 2.3 with the percent carbon burnup. It should be noted that for the fly ash burned, carbon burnup is, for practical purposes, the same as combustion efficiency due to the low hydrogen content of the fuel.

In Figure 2.9, the data in Table 2.3 is plotted on a graph from the one feeder test program reported in 1972. The equation for calculating the combustion efficiency in this figure was developed by statistical analysis of the one feeder test data. As indicated, the combustion efficiency observed in the multiple feed point test program is higher than that calculated by the previously developed equation. This was an unexpected result since the area served by each feeder in the multiple feed point tests was greater than the area fed by one feeder in the previous test program.

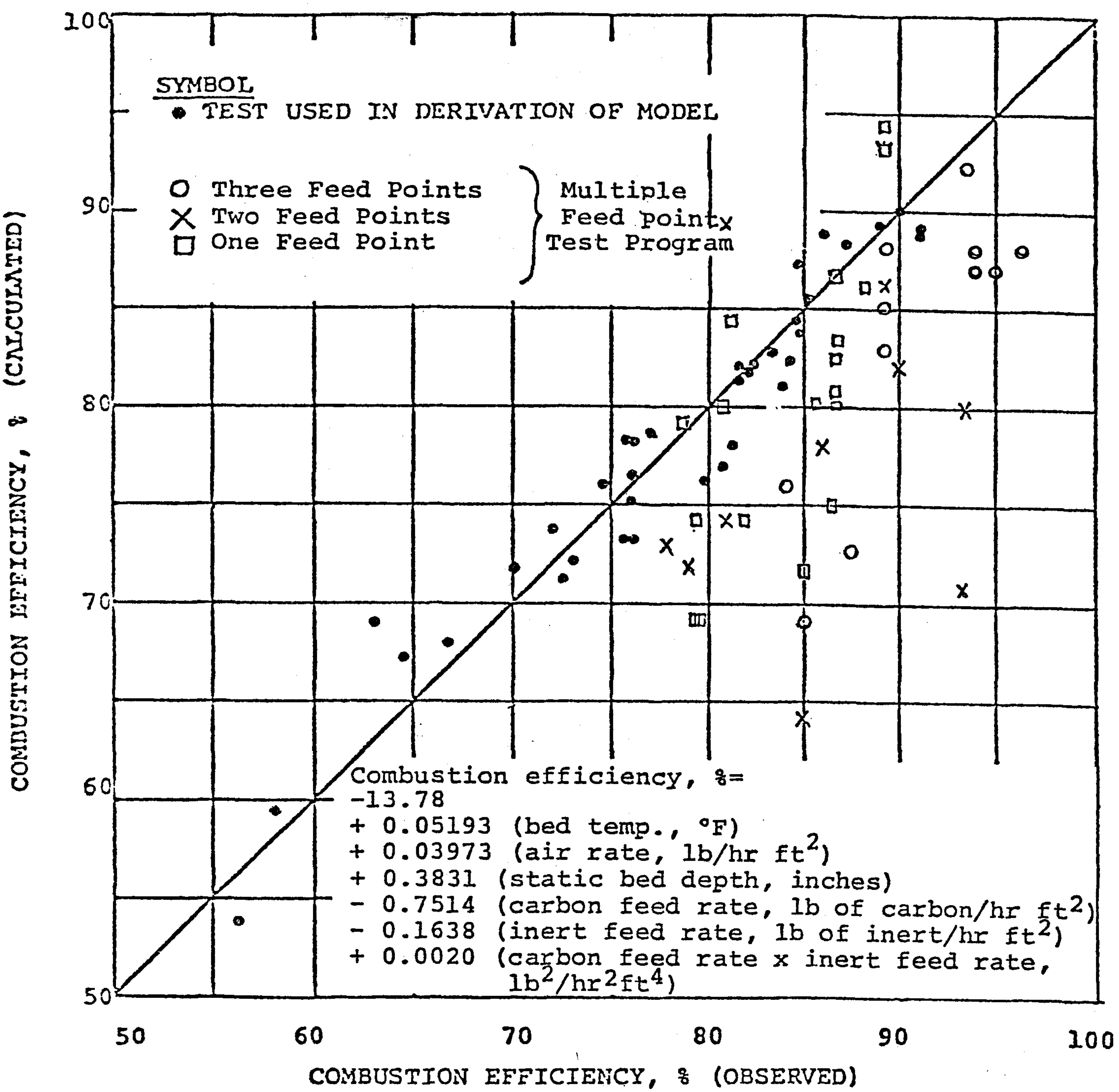
Number of Feed Points

An important result of the test program was the comparison of operation using one, two and three fly ash feed points. Based on the data evaluated to date and the operating experience, it was concluded that one feeder does not provide sufficient distribution in the FBM and that three feeders were only marginally better than two feeders.

Control of operation using one feeder was marginal due to the difficulty in transition from coal to fly ash feed and the tendency for temperature excursions to start slightly above 2000°F bed temperature. Control of two and three feeder operation was simpler by comparison. The carbon burnup for three and two feeder operation can best be compared in the data from Test 644-11 in which a switch from three to two feeders was accomplished at relatively steady conditions at the desired bed temperature and excess oxygen range.

Operating Conditions

Bed temperature was considered to be the most critical factor in obtaining high carbon burnup. As experienced in previous tests, the closer the bed temperature was to the ash fusion temperature (about 2150°F) the better the carbon burnup. This result is thought to be due to the effects of agglomeration



COMPARISON OF OBSERVED AND CALCULATED COMBUSTION EFFICIENCY USING EQUATION FROM PREVIOUS ONE FEEDER TESTS

FIGURE 2.9

TABLE 2.3

Summary of Test Results

Test No.	Time	No. of Feeders	%O_2 in Flue Gas(avg.)	Bed Temp °F	Bed Depth Static in.	Super. Veloc. ft/sec	Fly Ash Rate lb/hr	% Carbon Burnup
4	13:00	3	5.4	1895	18.6	7.3	960	87.5
6	13:30	3	3.7	1935	18.5	6.5	900	84.2
	15:00	3	2.3	1930	15.3	6.2	930	86.5
	15.40	3	1.0	1985	12.3	5.6	900	85.4
7	11:30	2	2.9	2010	21.5	8.7	862	90.1
	12:00	2	3.5	1960	22.6	7.2	1050	92.7
	13:00	2	1.7	1898	20.5	6.1	840	78.1
	13:30	2	2.5	1888	18.6	6.5	860	78.7
	14:00	2	2.6	1906	19.2	6.8	1050	84.6
	14:30	2	1.7	1916	17.7	6.9	840	80.1
8	10:00	1	3.0	1955	19.6	6.9	540	88.1
	11:00	1	2.7	1937	19.6	7.3	620	86.8
	12:00	1	2.2	1943	17.7	7.1	650	86.9
	12:30	1	3.2	1940	17.7	6.8	590	87.0
	13:30	1	2.3	1963	17.2	6.6	730	86.8
	14:00	1	2.5	1982	16.7	7.0	660	86.7
	14:30	1	2.8	1975	16.2	6.9	630	86.3
9	12:30	2	3.6	1985	21.6	8.7	850	92.9
	13:00	2	5.6	1865	22.6	7.5	540	89.3
	13:30	2	3.7	1920	21.6	7.9	540	81.8
	14:00	2	4.2	1900	22.1	7.9	780	86.0
10	11:00	1	2.0	1940	21.2	8.8	660	82.2
	11:30	1	2.3	1920	19.7	8.8	900	79.5
	12:00	1	1.7	1970	19.7	8.2	803	78.0
	12:30	1	1.4	1960	19.7	8.0	780	81.0
	13:00	1	1.8	1961	19.7	8.3	940	81.9
	13:30	1	1.8	1950	20.7	8.1	930	85.3
	14:00	1	2.3	1925	20.7	7.5	940	78.8
	15:00	1	2.3	1940	22.2	7.4	960	79.5
11	10:30	3	3.4	2000	16.0	7.1	480	89.3
	11:00	3	6.7	1960	15.5	8.3	390	93.2
	11:30	3	5.3	1940	16.0	7.8	615	95.0
	12:00	3	5.9	2010	16.0	9.6	615	96.1
	13:00	2	5.2	2030	16.0	9.7	615	94.0
	13:30	2	6.2	2000	15.4	9.6	615	94.6
	14:00	2	3.0	2025	15.5	9.6	615	89.2
12	11:30	1	1.7	1990	15.9	8.5	405	88.8
	13:00	1	1.5	1975	15.8	8.4	417	89.2
	13:30	1	3.3	1940	16.7	8.3	519	86.8

of ash to bed particles and improved fuel reactivity at the higher temperatures. The relative influence of these effects is unknown at present. Better carbon burnup was achieved at lower temperatures than expected. This was an encouraging result in terms of the anticipated difficulty of maintaining the bed temperature of a large carbon burnup cell near the ash fusion point. Based on the current test program it appears that satisfactory carbon burnup can be achieved within a reasonable operating range of bed temperatures.

Excess air was considered the next most important operating condition and was measured in terms of percent oxygen in the flue gas. The oxygen measurement was made at a point where the flue gas had cooled to about 1000°F. Reliability of the oxygen measurement was very poor due to the extremely high dust loading in the flue gas at low combustion efficiencies. In general, the test data indicates that at higher oxygen concentrations carbon burnup improved, as would be expected.

Bed depth and superificial velocity are important operating conditions since they affect the mixing and directly determine the residence time of the gas in the bed. Due to the difficulties of sustaining high bed temperatue with high excess air, superficial velocity was limited to the range of about 6 to 9 ft/sec and bed depth was lowered as required to maintain temperature. Increasing bed depth and lowering superficial velocity tends to increase the fraction of fuel burned in the bed, an important consideration in maintaining high bed temperature and high carbon burnup.

2.3.2 Automatic Control Tests

During the reporting period a series of tests was conducted to place the automatic combustion control system in operation. During this tuning period several changes were made both to the control equipment and the control logic. Prior to the tuning procedure a steam flow controller was installed so that the steam pressure could respond to load changes.

The first control loop to be placed into operation was loop one, the uptake controller. Tuning this loop was relatively simple since the controller was a standard P-I type controller. Initial adjustments were made with a cold furnace. Final tuning was completed during hot tests.

The second control loop placed on line was loop 3, the bed level and SO_2 controller. This control loop required the greatest amount of modifications to the control system. Originally, the control logic compared the plenum pressure and the FD damper position to infer bed level. This system was found to be unacceptable because the bed depth could change as much as 3 inches with no noticeable change in plenum pressure. To eliminate this problem, constant purge pressure taps were installed in the low bed and freeboard positions. The low bed-freeboard was thus a direct bed level indication eliminating the need to compare the plenum pressure and the FD damper position. Tuning of this loop was then relatively straightforward. The SO_2 proportional controller was adjusted by making a step change in the limestone feed rate, observing the response of the SO_2 and calculating the necessary gain.

The final control loop to be placed in operation was loop 2, the bed temperature controller. The first step of the tuning procedure involved setting the gains of the O_2 controller and the air flow computer. The initial gains for these two controllers were calculated from theoretical considerations. The O_2 controller gain was set by estimating the amount of extra coal feed necessary to decrease the percentage of O_2 by 1 percent. The air flow computer compensates for changes in air density due to changes in preheat.

The second step in the tuning procedure was determination of the proper fuel to air ratio (F/A). This was necessary because the control logic maintains bed temperature with the firing rate (both coal and air). A low F/A would result in a decrease in bed temperature with an increase in the firing rate. Thus, proper F/A is vital to temperature control. In order to determine the proper F/A, step changes corresponding to various ratios were made manually to the fuel and the air feed rates. Bed temperature response was then observed. For a given F/A, step changes were made to increase the firing rate and again to decrease the rate. In this manner the satisfactory F/A was found to be 3:1. It should be noted, that the fuel to air ratio corresponds to the pneumatic signals which control the coal feed rate and air damper position. Furthermore, boiler construction such as refractory on the waterwalls, will affect the fuel to air ratio. Therefore, this procedure must be followed to establish the proper ratio each time a significant change is made to the FBM.

Once the proper fuel/air ratio was established the P-I bed temperature controller was tuned. Finally, the gain of the steam pressure controllers were set by making step changes in the firing rate, observing the response, and calculating the necessary gain.

A total of seven automatic control tests were conducted, including two tests of approximately 80 hours duration. Since that time, various control loops have been utilized during other tests. Usually loops 1 and 3 operate in the automatic mode. Loop 2 is usually operated in manual to prevent changes in the air flow rate. Since most of the tests conducted require careful control of operating parameters, such as air flow and coal feed rate, loop 2 is usually turned off during these tests.

2.3.3 Coal Feeder Development Tests

Conversion of Petrocarb Injection System

A coal feeder evaluation program was started during the reporting period. The Petrocarb, Inc. fly ash injection system was modified for use as a coal injection system with various feeding arrangements. Due to the injectors's fuel size limitation, coal used was reprocessed to 1/8 inch or smaller. Figure 2.10 shows the coal flow diagram used in the evaluation program. This set up provided the greatest flexibility in switching between the Carman vibrating feeder table (needle feeders) and the Petrocarb, Inc. injector during test runs, which permitted ready comparison of the two systems.

Coal feeding arrangements consisted of under bed tee and mushroom feeders and over bed feeders as seen in Figure 2.11. The over bed feeders utilized two separate feed lines from the Petrocarb, Inc. injector to insure even feed distribution and were located in the center of the FBM as shown. They also have varying injection angles. The under bed feeder arrangement consisted of a welding tee or mushroom cap located in the FBM as shown in Figure 2.12. Testing procedure consisted of light-off using the existing coal feeding equipment. The FBM was run at steady state for sufficient data gathering and then the select coal feeder was turned on. Data was taken throughout the test for combustion efficiency comparisons. Limestone was fed throughout the test through the Carman feeder table only. Static bed depth of 15 inches and temperatures of 1500°F to 1600°F were maintained with four percent excess O_2 prior to switching to the Petrocarb injector. After switching to the injector, coal and air rates were held constant and changes in temperature and O_2 observed.

Summary of Test Results

A total of eight coal feeder tests were conducted during the reporting period. Table 2.4 is a summary of the steady state operating conditions obtained during the tests. Table 2.5 shows heat balances from all of the tests except Test No. 646-10.

Test Nos. 646-2 and 646-4 Of the first four tests, Nos. 646-1 through 4, short steady state conditions were achieved only in Test No. 646-2 using the overbed feeders, and Test No. 646-4 using a single underbed mushroom feeder shown in Figure 2.11. Plugging problems in the seal leg of the Petrocarb injector prevented further testing. To eliminate the seal leg problems, the surge vessel was moved and placed directly on top of the primary injector. In this configuration, the Petrocarb system was able to deliver coal without plugging problems.

COAL FLOW DIAGRAM- PETROCARB INJECTOR

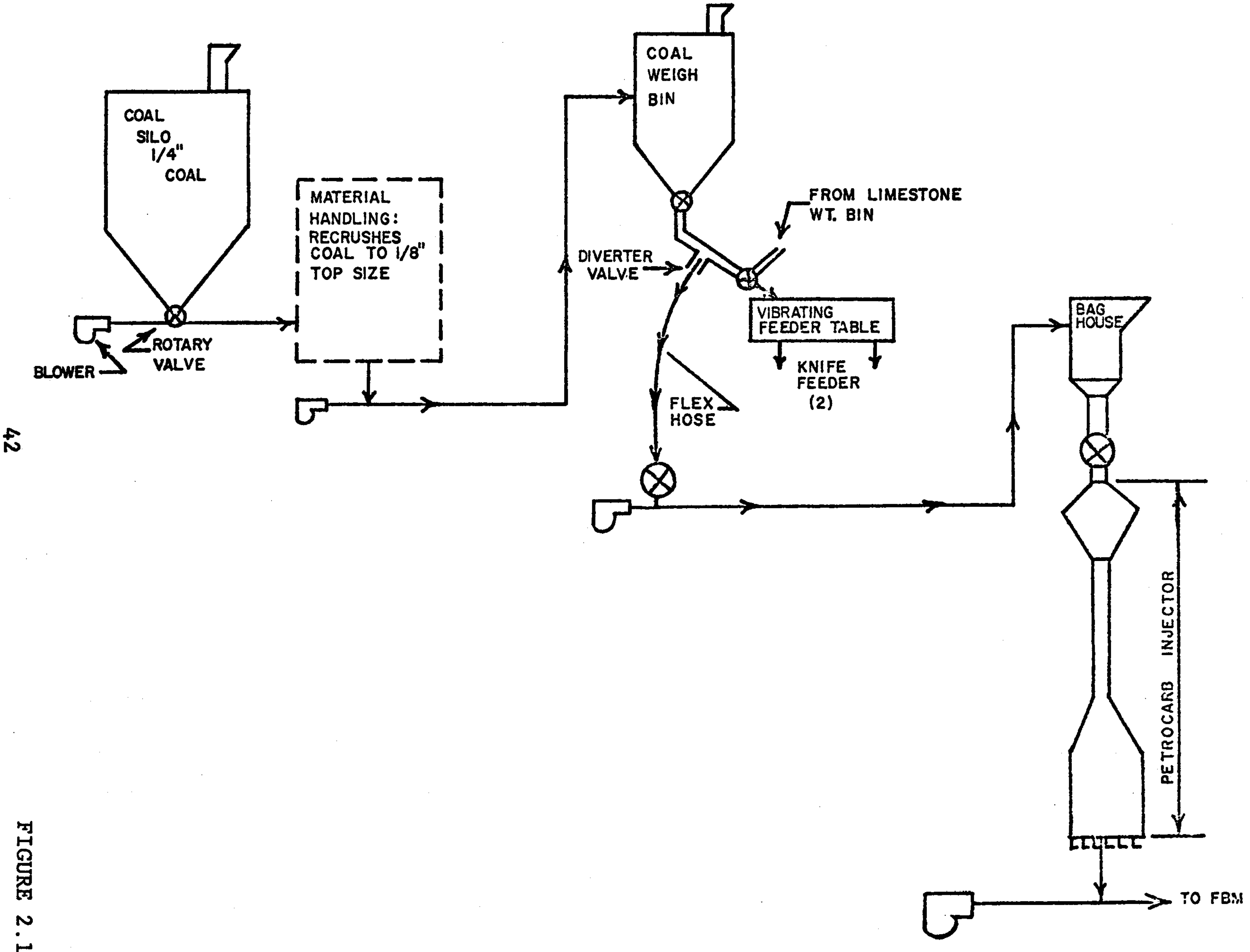
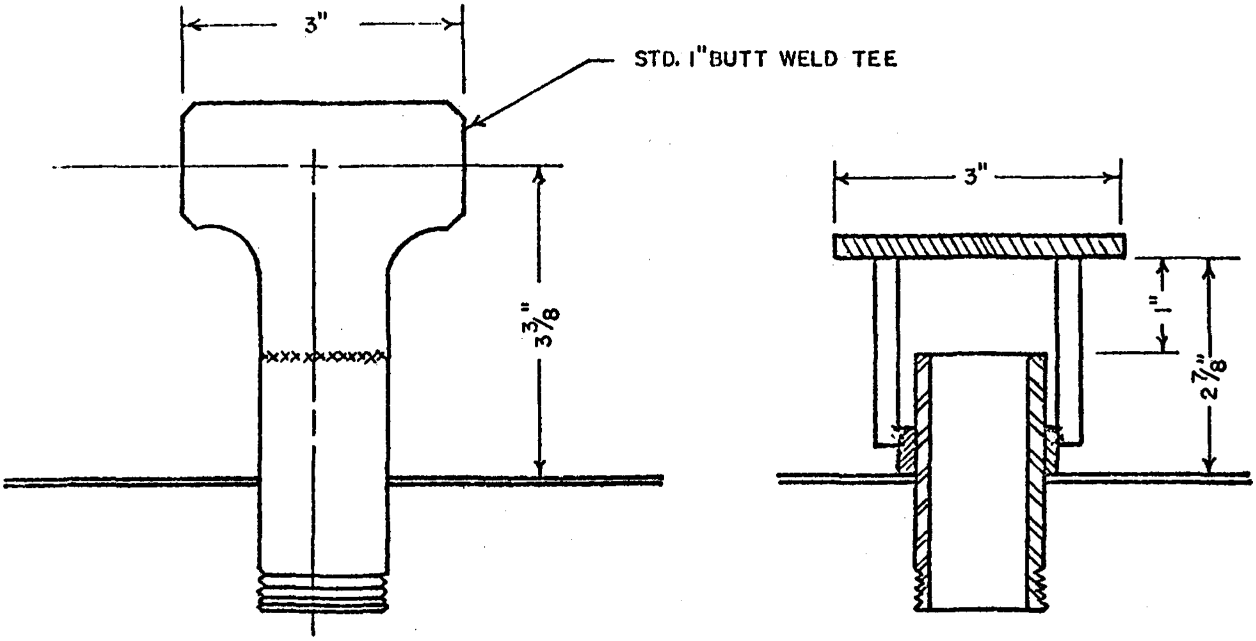


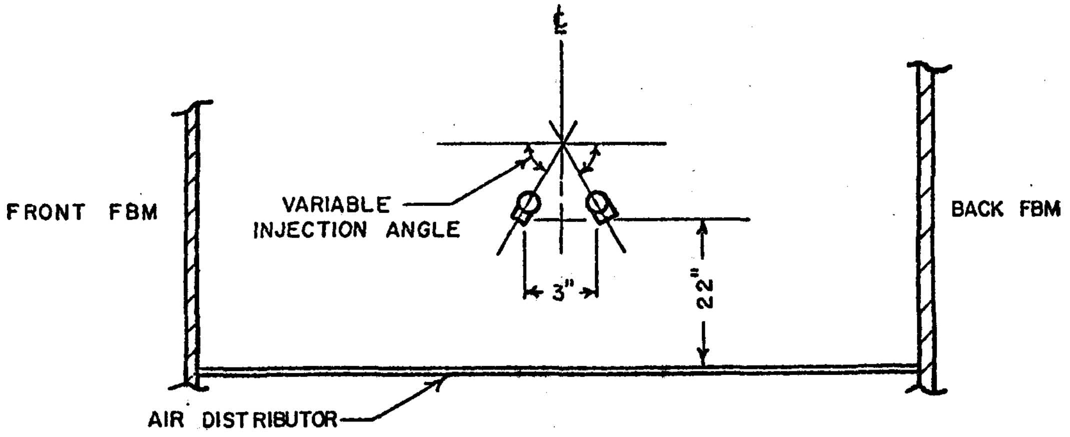
FIGURE 2.10



TEE

MUSHROOM

UNDER BED FEEDER ARRANGEMENTS



COAL FEEDER ARRANGEMENT

FIGURE 2.11

UNDER BED COAL FEEDER LOCATIONS (MUSHROOM or TEE)

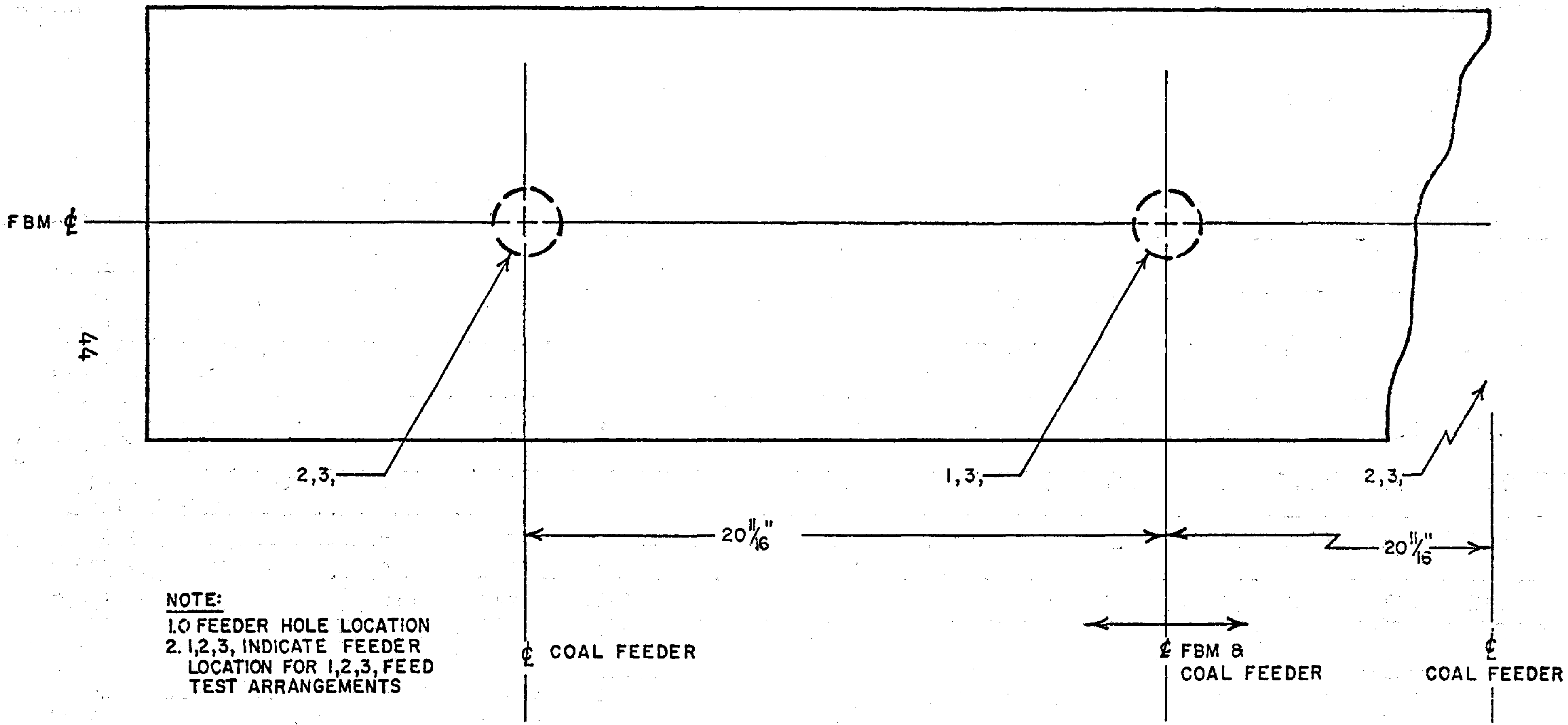


FIGURE 2.12

Sheet 1 of 3

TABLE 2.4

Summary Test Results Steady State Operating Conditions*

Test No.	646-2		646-4		646-5					646-6			
Time	10:00	12:30	10:00	11:30	14:00	10:30	13:00	13:30	14:45	11:00	13:00	14:00	14:30
Feeder Arrangement	Needle Feeders	Over Bed Feeder (∿60° from Horz.)	Needle Feeders	Under Bed Feeder 1 Mush., 1" Clear.	Needle Feeders	Needle Feeders	Under Bed Feeder 1 Mush., 1" Clearance			Needle Feeders	Under Bed Feeder 2 Mush., 1/2" Clearance		
Bed temp., °F	1570	1470	1640	1590	1660	1590	1625	1620	1615	1640	1565	1625	1585
O_2 in flue gas,%	4.5	3.0	4.5	4.8	4.2	3.0	2.0	3.0	3.0	3.1	6.2	4.7	6.2
SO_2 ppm	--	--	--	--	--	--	260	240	255	225	440	895	545
Coal feed rate, lb/hr	535	567	537	516	539	528	528	515	535	555	539	541	503
Fly ash collection rate, lb/hr	101	108	158	178	188	160	156	150	160	130	110	136	136
Stack dust rate, lb/hr	35.1	32.9	21.5	17.5	22.3	13.3	18.4	15.6	18.4	12.7	22.1	14.5	17.2
Total air flow rate, lb/hr	6296	6463	5886	5873	5868	6093	6023	6015	6072	6922	6615	6703	6703
Superficial velocity, ft/sec	10.6	10.4	10.3	10.0	10.4	10.4	10.5	10.4	10.5	12.1	11.1	11.6	11.4
Static bed depth, inches	16.9	16.8	15.7	15.9	15.9	16.3	16.4	16.4	16.4	15.4	15.5	15.5	15.5
EOBH (effective operating bed ht.), inches	--	--	18.3	18.8	20.0	21.0	21.0	21.0	20.5	22.0	20.5	21.0	21.0
Coal Injection velocity, ft/sec	--	94.3	--	79.2	--	--	83.1	83.6	74.8	--	75.4	73.4	74.0
Combustion efficiency,%	86.0	86.4	76.3	70.6	83.2	80.5	80.5	79.7	81.6	86.4	86.7	85.8	85.5

* Revised Calculations (6/7/1977)

Sheet 2 of 3

TABLE 2.4 Continued

Summary Test Results Steady State Operating Conditions

Test No.	646-7				646-8			646-9				
Time	10:00	13:00	14:00	14:45	11:30	13:30	14:00	09:30	12:00	13:00	13:30	14:45
Feeder Arrangement	Needle Feeders	Under Bed 1 Mush., 1/2" Clearance		Needle Feeders	Needle Feeders	Over Bed @ 60°	Needle Feeders	Needle Feeders	Over Bed Feeders @ 90°			Under Bed 3/4" Tee
Bed temp., °F	1570	1575	1600	1590	1560	1570	1570	1570	1555	1560	1540	1525
O_2 in flue gas	4.0	1.3	2.5	3.5	4.4	1.9	4.2	5.0	2.6	1.0	1.5	5.3
SO_2 ppm	790	1300	1365	695	410	587	565	655	719	713	587	470
Coal feed rate, lb/hr	558	548	582	568	540	568	556	466	540	530	510	532
Fly ash collection rate, lb/hr	138	152	161	145	126	150	150	130	140	150	135	137
Stack dust rate, lb/hr	32.1	15.3	14.1	18.0	15.7	14.9	17.6	18.4	11.6	15.1	13.5	19.4
Total air flow rate, lb/hr	6921	6872	6893	6944	6960	6864	6824	7035	6017	6641	6865	6319
Superficial velocity, ft/sec	11.7	11.6	11.8	11.8	11.7	11.6	11.5	11.9	10.1	11.2	11.4	10.4
Static bed depth, inches	15.8	15.3	15.4	15.9	15.7	15.6	15.6	16.8	16.3	16.4	16.3	16.4
EOBH (effective operating bed. ht.), inches	21.0	20.5	21.0	21.5	21.0	21.5	21.5	21.0	21.5	22.0	22.5	21.0
Coal injection velocity, ft/sec	--	74.9	74.9	--	--	77.1	--	--	78.7	79.1	85.9	67.6
Combustion efficiency, %	82.4	79.0	80.6	84.1	83.3	83.1	84.1	81.0	80.0	78.6	80.1	78.3

TABLE 2.4 Continued Sheet 3 of 3

Summary Test Results Steady State Operating Conditions

Test No.	646-10				
Time	10:00	11:30	12:30	13:30	14:30
Feeder Arrangement	Needle Feeders	Under Bed 3/4" Tee Feeder With Dead Zone			
Bed temp.,°F	1595	1590	1580	1570	1590
O_2 in flue gas	4.3	2.8	3.3	3.5	4.2
SO_2 ppm	651	1176	1158	1046	1313
Coal feed rate, lb/hr	548	530	555	548	555
Fly ash collection rate, lb/hr	140	150	150	150	150
Stack dust rate, lb/hr	15.3	13.2	16.6	12.6	13.0
Total air flow rate,lb/hr	6775	7060	6973	6967	6963
Superficial velocity,ft/sec	11.6	12.0	11.8	11.8	11.9
Static bed depth,inches	15.3	15.4	15.4	14.9	15.0
EOBH (effective operating bed. ht.), inches	22.0	21.0	21.5	21.5	20.5
Coal injection velocity, ft/sec	--	59.8	60.2	60.3	59.8
Combustion efficiency,%	--	--	--	--	--

Table 2.5

Heat Balance for Coal Feeder Tests*

Test No.	646-2		646-4			646-5				646-6			
Time	10:00	12:30	10:00	11:30	14:00	10:30	13:00	13:30	14:45	11:00	13:00	14:00	14:30
Feeder Arrangement	Needle Feeders	Over Bed Feeder (∿60° from Horz.)	Needle Feeders	Under Bed Feeder 1 Mush., 1"Clear.	Needle Feeders	Needle Feeders	Under Bed Feeder 1 Mush., 1" Clearance			Needle Feeders	Under Bed Feeders 2 Mush., ½" Clearance		
Fuel Input, Btu/hr	6,527,000	6,917,400	6,551,400	6,295,200	6,575,800	6,441,600	6,441,000	6,283,000	6,527,000	6,771,000	6,575,800	6,600,200	6,136,600
CO losses, Btu/hr	25,988	72,162	493,723	588,746	41,072	52,459	7,453	66,779	66,804	33,179	15,472	37,636	40,195
Hydrocarbon losses, Btu/hr	25,565	58,017	75,165	227,345	14,450	4,338	215,718	193,280	193,351	0	0	0	0
Fly Ash losses, Btu/hr	692,453	610,357	923,670	969,753	980,895	1,044,258	968,183	974,107	880,390	845,060	795,553	860,159	789,820
Dust losses, Btu/hr	170,008	202,852	35,440	37,550	37,355	47,860	42,507	36,275	43,959	36,308	52,964	34,092	46,982
Bed Drain losses, Btu/hr	0	0	21,845	28,257	33,391	104,787	21,909	6,316	16,681	9,109	12,718	7,394	11,900
Combustion Efficiency,%	86.0	86.4	76.3	70.6	83.2	80.5	80.5	79.7	81.6	86.4	86.7	85.8	85.8

Test No.	646-7				646-8			646-9				
Time	10:00	13:00	14:00	14:45	11:30	13:30	14:00	09:30	12:00	13:00	13:30	14:45
Feeder Arrangement	Needle Feeders	Under Bed Feeder 1 Mush., ½" Clearance		Needle Feeders	Needle Feeders	Over Bed at 60°	Needle Feeders	Needle Feeders	Over Bed Feeders at 90			Under Bed Feeder 3/4" Tee
Fuel Input, Btu/hr	6,807,600	6,685,600	7,100,400	6,929,600	6,588,000	6,929,600	6,856,400	5,685,200	6,588,000	6,466,000	6,222,000	6,490,400
CO losses, Btu/hr	104,775	139,430	127,120	26,440	47,701	60,524	33,593	66,200	139,013	161,931	151,887	112,167
Hydrocarbon losses, Btu/hr	0	205,269	172,353	0	26,703	8,555	13,936	30,970	182,788	223,689	141,941	356,961
Fly Ash losses, Btu/hr	973,416	999,247	1,023,429	1,028,422	961,380	1,037,259	975,189	917,973	894,045	918,602	846,760	840,987
Dust losses, Btu/hr	104,943	35,863	41,819	33,327	48,637	41,987	52,595	28,161	25,918	19,386	26,766	40,679
Bed Drain losses, Btu/hr	13,986	23,374	15,789	17,038	14,006	21,076	13,793	35,588	74,643	57,644	71,982	55,764
Combustion Efficiency.%	82.4	79.0	80.6	84.1	83.3	83.1	84.1	81.0	80.0	78.6	80.1	78.3

* Revised Calculations (6/7/1977)

Test No. 646-5 This test consisted of injecting coal through one under-bed mushroom feeder shown in Figure 2.11. Clearance between the nozzle and cap assembly was one-inch in this test. As seen from the results, the combustion efficiencies at 13:00 and 13:30 were 80.5 percent, over two percent lower than the combustion efficiency obtained with the two Carman feeders. At 14:00 hours, coal transport air velocity to the feeder was lowered from 70.6 ft/sec. to 64.6 ft/sec. to see what effect it would have. As seen from the results at 14:30, the combustion efficiency increased to 82.0 percent, very near the combustion efficiency obtained with the two Carman feeders during this test.

Test No. 646-6 This test consisted of operating two under bed mushroom feeders with a ½ inch clearance between the nozzle and cap assembly. A low transport air velocity was used for this test. Combustion efficiencies obtained in this test showed no difference between the two under bed mushroom feeders and the two Carman feeders.

Test No. 646-7 This test was a single mushroom feeder test similar to 646-5 with the exception of the feeder dimensions. In this test the gap between the nozzle and the cap assembly was reduced to ½ inch. The lower transport air velocity was used. The FBM was operated with the two Carman feeders, both at the beginning and at the end of the test to verify the combustion efficiency. As seen from Table 2.4, combustion efficiencies of 80.5 percent were obtained with the one mushroom feeder, while efficiencies of 84.1 percent were obtained with the two Carman feeders.

Test No. 646-8 This test consisted of feeding coal through the overbed feeders with an injection angle of 60° from horizontal and an injection velocity of about 75 feet per second. In comparing the combustion efficiencies of the needle feeders at 1400 hours with the overbed feeders as shown in Table 2.4, there was a one percent decrease in efficiency using the overbed feeders. The only measured difference between both steady state periods, was the O_2 in the flue gas. With the needle feeders, the excess oxygen was 4.2 percent and with the overhead feeders the excess oxygen was 1.9 percent.

Test No. 646-9 Overbed feeding with an injection angle of 90° from horizontal directly into the center of the bed, and feeding through an underbed 3/4 inch welding tee similar to the one shown in Figure 2.11, located in the center of the FBM, 2-½ inches above the grid plate, was accomplished during this test. As seen from the results in Table 2.4, the overbed feeders did not perform as well as the needle feeders. At 1200 and 1300 hours the overbed feed combustion efficiency was 80 percent and 78.6 percent compared to 81 percent for the needle feeders. When the injection velocity was increased as shown for the 13:30 hours period, the combustion efficiency was not greatly improved.

A more scattered temperature distribution was recorded throughout the bed when utilizing the overbed feeders indicative of poor mixing of the fuel in the bed. Bed temperatures varied as much as 100°F between different areas in the FBM. The underbed tee feeder also did not compare favorably with the needle feeders. This feeder showed a combustion efficiency of 78.3 percent at 14:45 hours. In both cases the CO losses and the hydrocarbon losses were much greater than with the needle feeders as can be seen in Table 2.5. This resulted in a lower combustion efficiency possibly due to poor coal distribution throughout the bed as suggested by a scattered temperature distribution.

Test No. 646-10 In this test, one-third of the grid plate area was blocked off in the center of the FBM to determine the effect on performance of the underbed tee feeder used previously in Test No. 646-9. The blocked off grid plate created a dead zone around the injection point. The purpose of the dead zone was to determine if better combustion efficiency could be achieved by feeding coal into a region of down flow in the FBM. Due to a laboratory hydrocarbon analyzer malfunction, insufficient data was taken to complete heat balance calculations and combustion efficiency comparisons. As seen from Table 2.4, there seemed to be no significant improvement over the needle feeders.

Conclusions

Results of the alternate coal feed methods test program suggest some general conclusions about coal feeding to a fluidized bed combustor. Two basic types of coal feeders were tested, underbed and overbed. Underbed coal feeders consisted of the conventional needle feeders (from the Carman feeder table), mushroom feeders, and a tee feeder. The type of underbed feeder appeared to have no significant effect upon the combustion efficiency. The number of feed points appears to be much more important. For the Alexandria FBM two feed points are necessary. From Table 2.5 it is seen that for all the single feed point tests, Nos 646-4, 5, 7, and No. 646-0 at 14:45 hours, the transition from two needle feeders to the single underbed feeder was marked by a noticeable increase in the amount of hydrocarbons in the stack. In contract, Test No. 646-6 showed essentially no difference in hydrocarbons between the two needle feeders and the two mushroom feeders. This agrees with the results of coal feed testing early in the hardware development stage when attempts to feed the FBM with one underbed feeder from the end of the unit resulted in excessive unburned hydrocarbons.[1] One possible explanation for the increase in

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[1] Final Report Vol. I, R & D Report No. 36, "Development of Coal Fired Fluidized-Bed Boilers", p. 46.

overbed coal feeding, appears to be possible. However, a strong downward injection velocity is necessary to overcome the upward force of the fluidized air. Furthermore, overbed coal feeding increases the potential for coal combustion above the bed, as discussed in Test No. 644-9. Coal burning in this region of low excess O_2 could favor CO formation and thereby reduce overall combustion efficiency.

2.3.4 Light-Off Procedures Study (Activity No. 1106)

During the reporting period, tests were run to develop alternative light-off procedures. Three alternate light-off procedures were attempted. The first procedure used No. 4 heating oil to heat the bed to ignition temperatures; the second involved a normal light-off procedure on a partially slumped bed and the third used a duct burner to preheat the air.

Light-off of a fluidized bed requires heating the bed to the ignition temperature of coal char and then introducing coal at a steady feed rate. Normally, heat is supplied to the bed by gas or oil burners. Some heat can be supplied to the bed by the volatiles in the coal if any inventory of coal (usually 5 to 10 percent of the bed by weight) is present in the bed prior to light-off. In this case the burning volatiles help raise the bed temperature to the ignition temperature of the char (usually between 800 and 1000°F). Normally, light-off is accomplished at or near minimum fluidization. This insures some mixing in the bed while minimizing the sensible heat loss from the bed. Thus, light-off time of a fluidized bed can be reduced by at least three approaches. The first involves addition of fuel to the bed prior to light-off. The second involves blocking portions of the grid plate, thereby reducing the total air flow through the cell and reducing the sensible heat loss. The third involves using a duct burner to preheat the air.

Oil Assisted Light-Off

The first alternate light-off procedure involved the addition of No. 4 heating oil to the bed prior to light-off. Six gallons of heating oil and 50 pounds of coal were added to a 1000 pound charge of spent bed material (about 12" static bed depth). The bed was then fluidized for 15 minutes to thoroughly mix the oil in the bed and allow the oil to be absorbed into the bed material. The air flow was then reduced to minimum fluidization and the light-off burner turned on. The light-off burner remained on until the bed temperature rose to 800°F. It was necessary to keep the light-off burner on during this time in order to keep the oil ignited. The burner was turned off when the temperature reached 800°F after which the oil continued to burn by itself. By the time the bed temperature reached 1000°F, the oil was completely consumed and the coal had begun to burn. The total time for light-off to a 1500°F bed temperature was under 8 minutes using this method. Three problems were encountered using this light-off procedure. The first was the fast temperature rise in the bed. The bed temperature rose to 1000°F in just under three minutes. This could be a problem in a large bed if too much fuel is present. The second problem involved the formation of several small ash clinkers in the bed if the bed was slumped before all of the oil was consumed. The third problem was cuased by burning oil being carried over to the dust collector and igniting fly ash in the collector.

Reduced Air Flow Light-Off

The second light-off procedure tested attempted to reduce the sensible heat loss from the bed during light-off. This was done by covering the back two-thirds of the grid plate with cardboard prior to adding bed material. This resulted in a small (1½ ft. x 2 ft.) fluidized region near the burner while the rest of the bed remained slumped. Thus, only the front third of the bed had to be heated to ignition temperature with the overbed light-off burner.

The burning bed then supplied heat to light-off the remaining bed as the cardboard burned back. The anticipated benefit of such a light-off scheme is that a large fluidized bed could be started up with less overbed burner capacity than required for the entire fluidized bed.

The procedure used for this light-off was as follows. The back two-thirds of the grid plate was covered with cardboard held in place by wood strips and wire. Bed material (11.5 inches, static) was then placed on top of the grid and mixed with 65 pounds of coal. The air flow was turned on and as the cardboard burned back, the air flow was increased to maintain minimum fluidization.

Figure 2.13 shows the temperatures normally encountered during light-off, two inches above the air distributor and 9, 45, and 63 inches back from the light-off burner. In a normal light-off procedure the front portion of the bed is heated to about 1000°F. The air velocity is then increased to mix the bed to a uniform temperature and the temperature all rise at the same time. Figure 2.14 shows the same temperatures for the light-off with no fluidization of the back two-thirds of the grid plate. Here the front third is ignited, and as the cardboard burned back (at a rate of about 3.5 inches per minute) the rest of the bed ignited.

Figures 2.13 and 2.14 indicate that the total time to light-off the bed is about the same for both techniques. This is because the total fuel input from the burner and the coal is the same in both cases. However, it is believed that with

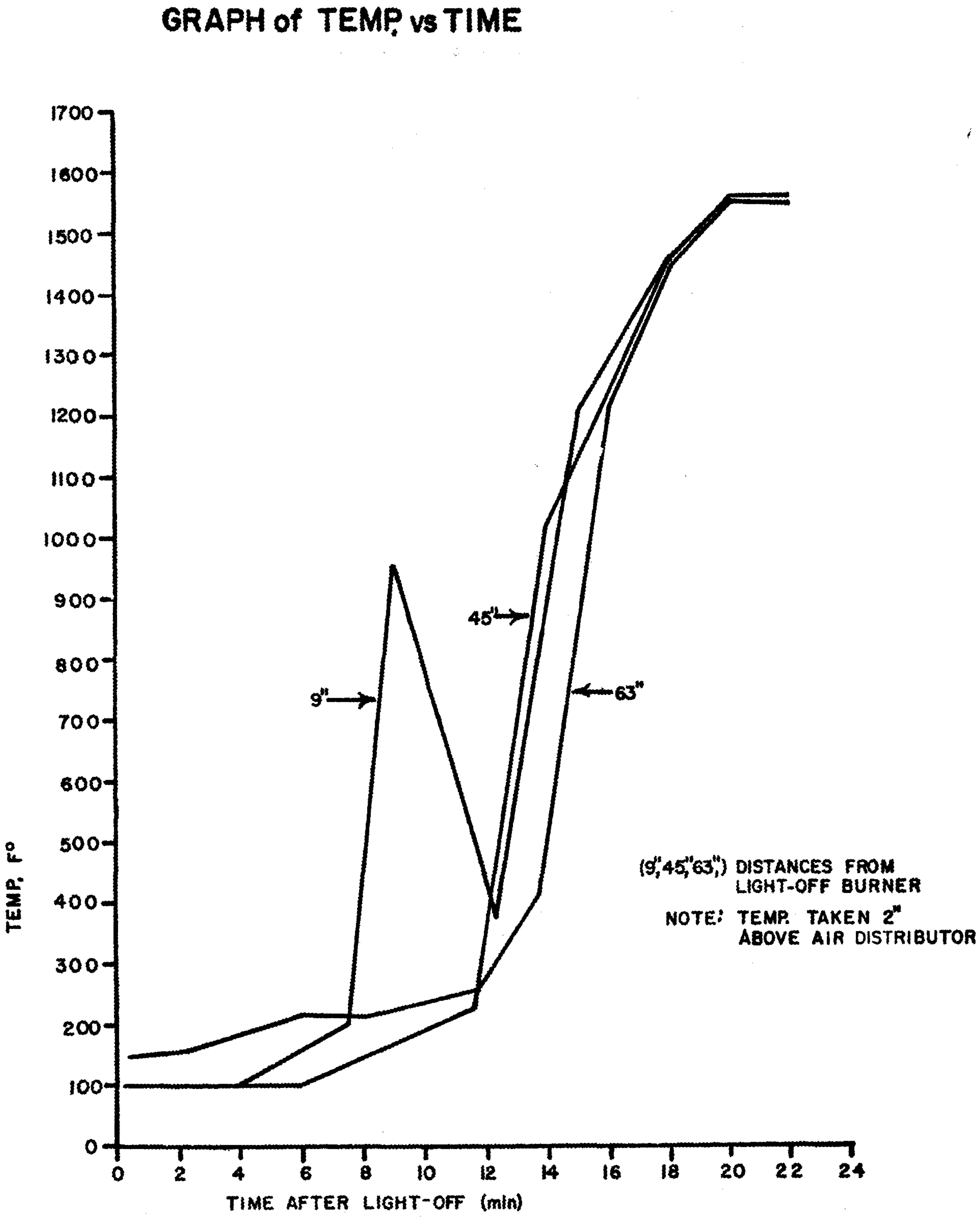

GRAPH of TEMP. vs TIME
TEMP. F°
1700
1600
1500
1400
1300
1200
1100
1000
900
800
700
600
500
400
300
200
100
0
0
2
4
6
8
10
12
14
16
18
20
22
24
TIME AFTER LIGHT-OFF (min)
9"
45"
63"
(9",45",63") DISTANCES FROM LIGHT-OFF BURNER
NOTE: TEMP. TAKEN 2" ABOVE AIR DISTRIBUTOR

FIGURE 2.13

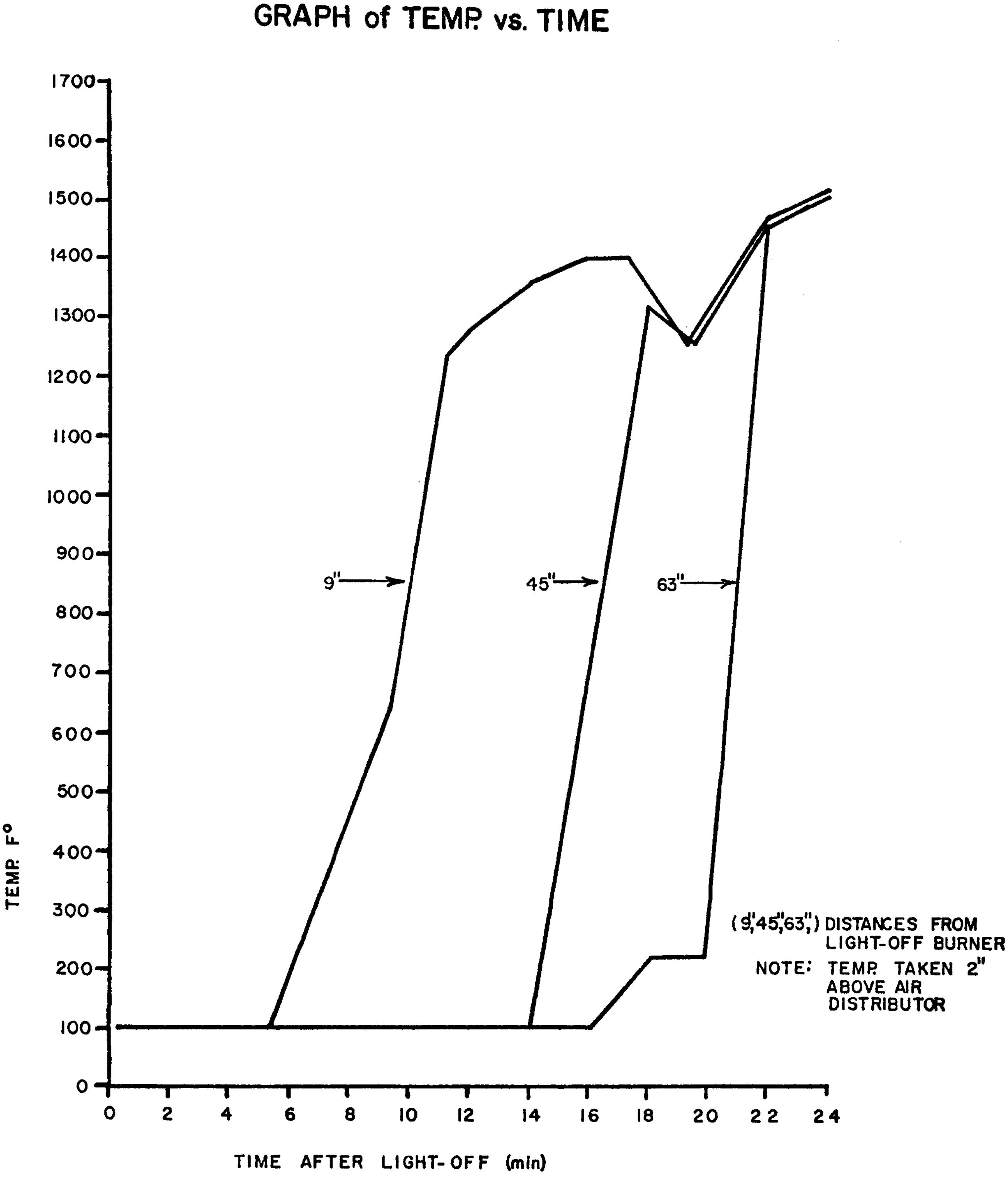

GRAPH of TEMP. vs. TIME
TEMP. F°
1700
1600
1500
1400
1300
1200
1100
1000
900
800
700
600
500
400
300
200
100
0
0
2
4
6
8
10
12
14
16
18
20
22
24
TIME AFTER LIGHT-OFF (min)
9"
45"
63"
(9",45",63") DISTANCES FROM LIGHT-OFF BURNER
NOTE: TEMP. TAKEN 2" ABOVE AIR DISTRIBUTOR

FIGURE 2.14

only one-third of the bed fluidized, light-off could have been accomplished with a smaller capacity light-off burner.

Duct Burner Assisted Light-Off

The third light-off procedure used the duct burner. During these tests the existing duct burner located between the F.D. damper and the plenum, was used to rapidly preheat the fluidizing air to the desired temperature. Bed temperatures were recorded during this time. When bed temperatures rose less than one degree per minute, coal feed was initiated. In all cases the overbed burner was turned on to ignite the volatiles. In some tests (those with bed temperatures near 800°F) the overbed burner was on only for one minute. In other tests, the overbed burner remained on until the bed temperature reached 900°F. The heat released by the burning volatiles was sufficient to heat the bed to the coal ignition temperature. During the testing it was observed, as expected, that prior to igniting the overbed burner, the bed temperature did not rise to the plenum air temperature due to the heat transfer from the bed to the waterwalls.

A heat balance for this system shows that the heat lost from the fluidizing air, $W\ C_p\ (T_{plenum}-T_{bed})$, is equal to the heat transferred to the water, $UA\ (T_{bed}-T_{water})$. Equating the two heats and solving for the overall heat transfer coefficient (U) gives

$$U = \frac{WC_p\ (T_{plenum}-T_{bed})}{A\ (T_{bed}-T_{water})} \qquad (1)$$

where

w = air flow rate lb/hr
C_p = heat capacity of air
A = bed-waterwall contact area

Since the heat transferred to the water is affected primarily by the bed temperature and bed depth, a higher bed temperature for a given plenum air temperature will be achieved by using high superficial velocities and shallow beds.

Table 2.6 shows the result of several tests using the duct burner for air preheat. The value of the overall heat transfer coefficient U calculated using equation (1) is also contained in this table. An average U of 44.8 Btu/hr/ft^2/°F was obtained. This number compares favorably with previously measured values.

TABLE 2.6

Steady State Light-Off Conditions

Bed Depth Inches Static	Superficial Velocity ft/sec	Plenum Temp. °F	Bed Temp. °F	Water Temp. °F	U Btu/hr/ft^2°F
16.5	5.5	541	427	114.4	42.3
16.5	5.5	613	479	150.3	47.3
10.5	5.4	680	570	155.5	43.4
10.5	6.2	850	710	232.1	47.9
11.5	6.0	880	720	181.0	42.9

The results of these tests can be used to estimate the bed temperatures that would be obtained by preheating the air for the MFB unit. Foster Wheeler Corporation estimates the highest safe plenum temperature allowable at Rivesville is 650°F. Using this plenum temperature, an overall heat transfer coefficient of 40 Btu/hr/ft^2/°F, a two foot static bed depth, 200°F water temperature and a superficial velocity of 6 ft/sec, the calculated Cell D equilibrium bed temperature would be 580°F. At this bed temperature, coal could begin to devolatilize. The volatiles can be ignited with an over-bed burner. Once this happens, the bed temperature should rise rapidly to the coal ignition temperature (above 800°F). Thus the initial light-off time for Cell D could be reduced.

If the same procedure was used to light-off one of the main cells after Cell D has been lit, the equilibrium bed temperatur could be calculated in the same manner. In this case the water temperature should be up to 475°F. Assuming the same bed depth and superficial velocity as before, gives an equilibrium bed temperature of 630°F for a plenum air temperature of 650°F.

2.3.5 Fly Ash Reinjection, Activity No. 407

A program of testing was begun during the reporting period to study the effectiveness of burning coal and fly ash simultaneously at 1600°F. The purpose of these tests was to burn coal and fly ash simultaneously at the low end of the expected CBC operating temperature range where sulfur capture would be optimum. Test results would also be applicable to industrial applications of FBC technology in which elimination of the Carbon Burn-Up Cell could have economic advantages depending on combustion efficiency.

The first series of tests was conducted to determine the once through percent carbon burnup under these conditions. Two tests, Nos. 650-2 and 650-4, were conducted in which coal and fly ash were burned simultaneously in the FBM at 1600°F. The fly ash used in these tests was obtained from previous FBM operations. In Test No. 650-4, the limestone feed rate and the bed removal rate were eliminated to increase the excess oxygen in the stack over that in Test No. 650-2, in which the limestone feed and bed removal rates were at normal rates. In both tests, data was collected for two steady-state periods when coal alone was fed and when fly ash and coal were fed simultaneously. The data obtained from these tests is summarized in Table 2.7.

TABLE 2.7

Test Data for Tests650-2 and 650-4

Test No.	650-2				650-4			
Time	11:30	12:30	13:30	14:00	11:30	13:30	14:15	14:45
Materials Feed Rate, lb/hr								
Coal Feed Rate	593	591	579	577	546	535	530	564
Limestone Feed Rate	300	293	293	293	0	0	0	0
High Carbon Fly Ash Feed Rate	0	0	427	427	0	318	332	0
Low Carbon Fly Ash Collection Rate	170	165	320	340	160	300	300	180
Bed Removal Rate	175	180	215	240	0	0	0	0
Dust Loading	19	20.9	31.1	30	16.93	26.32	18.19	16.84
Bed Temp., °F	1620	1610	1570	1570	1630	1625	1570	1610
Superficial Velocity, ft/sec.	11.0	10.3	10.0	9.9	9.5	9.0	8.7	8.8
Bed Depth (static) in.	19.9	19.7	19.7	19.5	19.6	20.1	20.2	20.7
% Excess O_2	1.6	2.2	2.1	2.0	3.2	1.0	1.3	1.2

The objective of these tests was to determine the percent carbon burnup of the fly ash being fed. Due to the delay in obtaining carbon analyses, the percent burnup of the fly ash based on LOI data (given in Table 2.8) was calculated. Table 2.9 shows this percent burnup of the fly ash based on LOI data in each test and the formula used to calculate this quantity. In making the calculations, it was assumed that the amount of fly ash produced from the coal is the same whether or not coal or coal plus fly ash is burned. Since the coal feed rates were slightly less when coal and fly ash were fed than when coal alone was fed, the fly ash produced from the coal, when coal and fly ash were fed together, was calcualted by multiplying the reference fly ash produced from burning coal only, by a coal feed ratio. This feed rate ratio is the coal feed rate when coal and fly ash are burned, divided by the reference coal feed rate when coal alone was burned. Each steady state value of the fly ash produced by burning coal alone was taken as the reference. Since two reference points and two combined coal and fly ash points were taken for each test, all four combinations of reference points and combined points were used to calculate the percent burnup.

The results show that in Test Nos. 650-2 and 650-4 the values of the average percent burnup based on LOI data were 59.0 percent and 61.1 percent respectively. If it is assumed that 60 percent of the carbon in the fly ash is burned on each pass through the boiler, then three passes are needed to give over 94 percent burnup of fly ash based on LOI data.

The next series of tests conducted, Nos. 650-6 through 650-9, were closed cycle tests where the fly ash produced was recycled to the FBM. In Test No. 650-7 some of the fly ash was bled off. Table 2.10 lists a summary of the operating conditions from these tests. Chemical analysis of the samples from these tests have not been completed at this writing.

During Test Nos. 650-6 and 650-8, one interesting observation was that the feed rate of recycled fly ash (as indicated by the pressure in the primary injection vessel of the Petrocarb injector) reached an equilibrium value. This indicates that in a total recycle situation, a steady state can be achieved where the amount of recycled fly ash burned equals the amount of fly ash produced by the burning coal.

In order to examine this observation more closely, it was first necessary to calibrate the Petrocarb injection pressure versus feed rate. In this way some measure of the recycle feed rate could be obtained. The next series of tests, Nos. 650-9 through 650-11, were run in the open loop so that the fly ash feed rate could be calibrated. Figure 2.15 shows the result of this calibration. The calibration is good above pressures of 1.4 psi. However, below 1.4 psi there was considerable scatter in the data (which is not shown).

TABLE 2.8

Loss on Ignition Data of Coal, Fly Ash and Dust for Tests 650-2 and 650-4

Test No.	650-2				650-4			
Time	11:30	12:30	13:30	14:00	11:30	13:30	14:15	14:45
Fly Ash Fed Into Boiler Lbs			36.62	38.62		47.03	40.85	
Fly Ash Which Leaves Boiler	50.82	48.94	44.23	42.38	58.75	46.24	48.96	50.12
Coal Lbs	85.12	85.12	85.12	85.12	80.48	80.48	80.48	80.48
Dust Lbs	20.38	19.48	20.29	24.80	22.79	23.27	18.99	25.79

TABLE 2.9

Calculated Percent Burnup of Fly Ash (Carbon in it) Based on Loss of Ignition

Test No.								
Coal Reference Time	11:30	11:30	12:30	12:30	14:45	14:45	11:30	11:30
Coal and Fly Ash Reference Time	14:00	13:30	14:00	13:30	13:30	14:15	14:15	13:30
Burnup Efficiency(%) Based on LOI	60.22	60.79	56.97	57.89	63.1	54.6	59.2	67.3

Average Burnup% based on LOI Test 650-2 - 59.01%
Average Burnup% based on LOI Test 650-4 - 61.1%

$$\text{Burnup of Fly Ash (\%)} = \frac{W_{FAin}C_{FAin} - (W_{FAo} - C_{FAo} - W_{FAcoal}C_{FAcoal}) - (W_{Do}C_{Do} - W_{Dcoal}C_{Dcoal}) - (W_{Bo}C_{Bo} - W_{Bcoal}C_{Bcoal})}{W_{FAin}C_{FAin}}$$

where:

W_{FAin}	is the fly ash feed rate.
C_{FAin}	is the LOI of fly ash feed.
W_{FAo}	is the fly ash collection rate from burning coal and fly ash.
C_{FAo}	is the LOI of fly ash collected from burning coal and fly ash.
W_{FAcoal}	is the fly ash produced from coal when burning coal and fly ash*.
C_{FAcoal}	is the LOI of fly ash collected by burning coal alone.
W_{Do}	is the dust collection rate from burning coal and fly ash.
C_{Do}	is the LOI of the dust when burning coal and fly ash.
W_{Dcoal}	is the dust collected from coal when burning coal and fly ash*.
C_{Dcoal}	is the LOI of the dust from burning coal alone.
W_{Bo}	is the bed removal rate when burning coal and fly ash.
C_{Bo}	is the LOI of the bed when burning coal and fly ash.
W_{Bcoal}	is the bed removal rate of coal when burning coal and fly ash*.
C_{Bcoal}	is the LOI of the bed when burning coal alone.

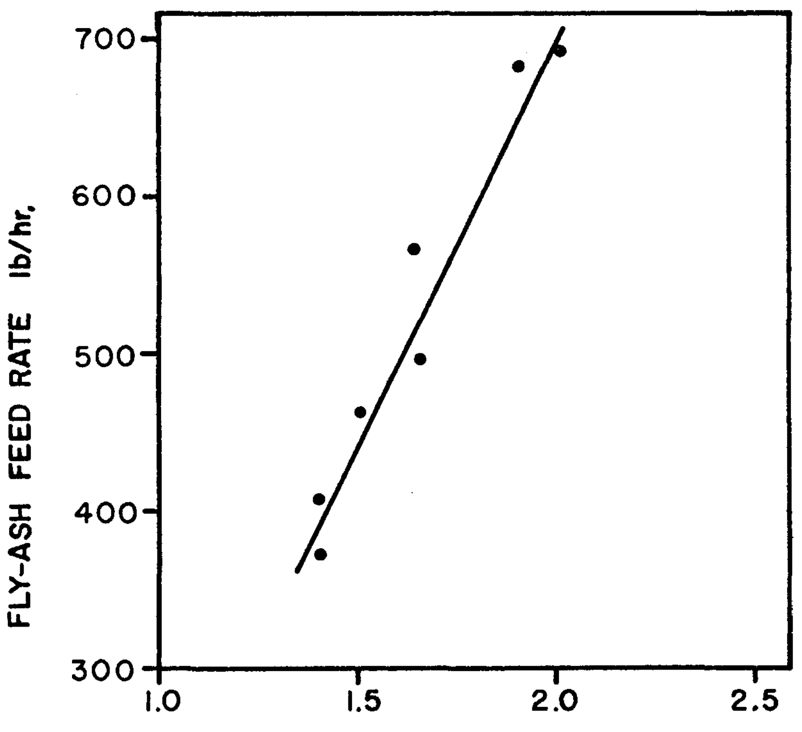
* The feed removal rate of a substance time the ratio of the coal feed rate when burning coal and fly ash to the coal feed rate method and results of burning of coal alone.

TABLE 2.10

Summary of Closed Cycle Fly Ash Reinjection Tests

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

CALIBRATION OF PETROCARB FLY-ASH INJECTOR



NOTE

Feeders 1 and 2 to FBM
Feeders 3 through 6 off

FIGURE 2.15

Testing of simultaneous coal and fly ash combustion will continue into the next reporting period. At present, it appears that high combustion efficiencies can be achieved with total fly ash recycle.

2.3.6 Corrosion/Erosion Studies

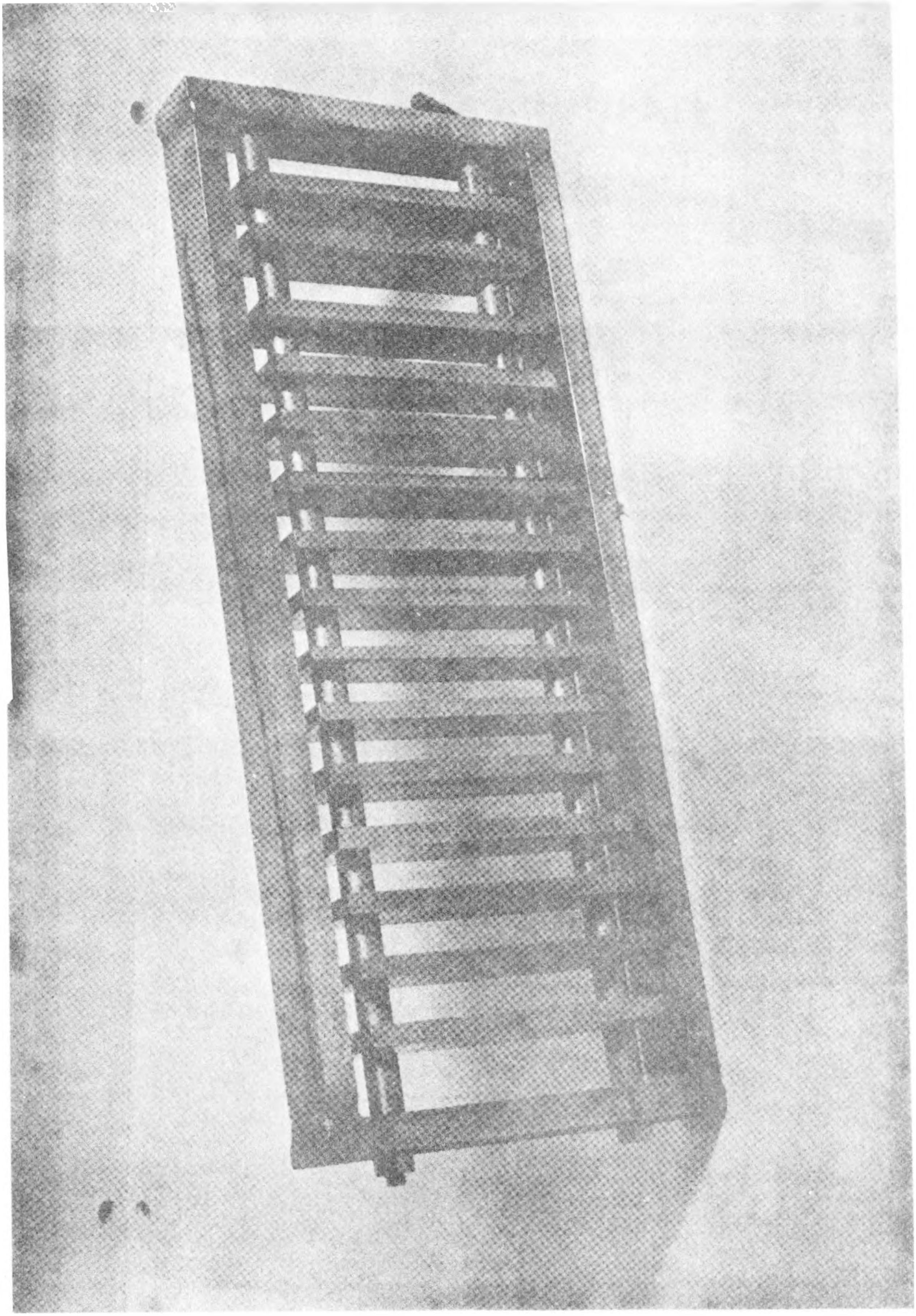
During the reporting period a program was initiated to evaluate a group of high temperature alloys for their corrosion and erosion resistance in a fluidized bed combustor. A corrosion test rack with a variety of metal samples was obtained from the International Nickel Company, Inc. for installation in the FBM. The test rack (Figure 2.16) was approximately 14" x 6" x 1" and the sample coupons were 4' x 1' x ¼". The rack contained one coupon each of the following materials:

1010 carbon steel
Alloy T-22
Type 304 stainless steel
Type 309 stainless steel
Type 310 stainless steel
Type 314 stainless steel
Hastelloy X
Incoloy 800
Incoloy 804
Inconel 600
Inconel 601
Inconel 690
In - 809
In - 811

In early January 1976, the test rack was installed below the horizontal bundle near the front of the bed. A photograph of the rack, after installation in the FBM, is shown in Figure 2.17. The rack was removed at the end of April 1976. During the period from January to April, the uncooled test rack was subjected to approximately 150 hours of hot operation at a variety of conditions, including bed agglomeration tests requiring many light-offs.

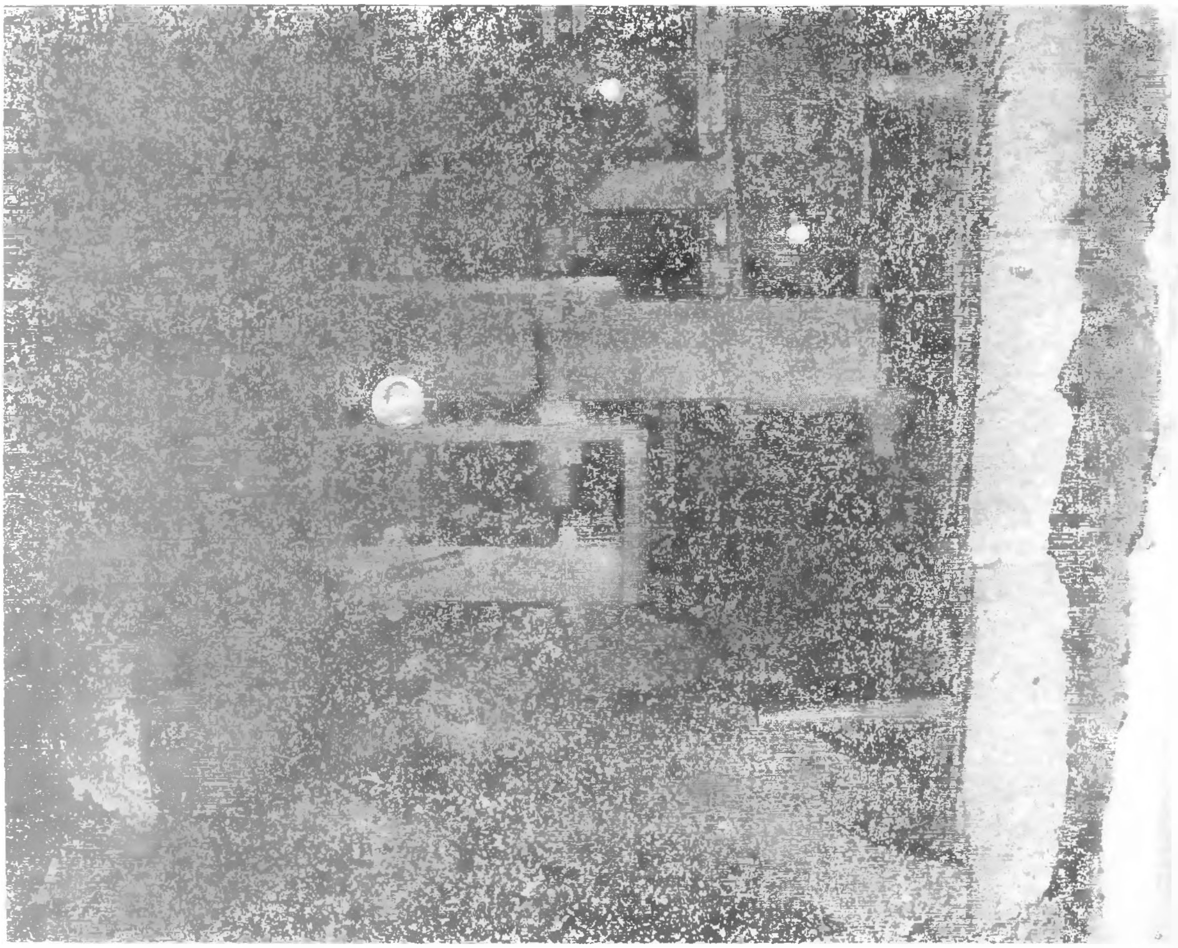
After removal from the FBM the test rack was returned to the International Nickel Company PDM Research Laboratory for metallographic analysis. The results of the metallographic analyses are shown in Table 2.11.

As seen in Table 2.11, a wide separation in alloy performance was achieved. Weight loss in mg./sq.cm is the dominant variable that should be examined. Using this parameter, it can be seen that Incoloy 800 was the most corrosion resistant alloy in the test followed by 314, 310, and 309 stainless steel. All of these alloys displayed excellent performance. Good performance was observed in Hastelloy X, IN-811 and 304 stainless steel. Fair performance was observed in Incoloy 804, Inconel 600, 1010 carbon steel and Inconel 690. The



CORROSION/EROSION TEST RACK PRIOR TO INSTALLATION IN THE FBM

FIGURE 2.16



CORROSION/EROSION TEST RACK AFTER INSTALLATION BELOW THE HORIZONTAL TUBE BUNDLE

FIGURE 2.17

TABLE 2.11

Corrosion/Erosion Test Results FBM Test Rack

Material	Weight Loss (mg/cm^2)	Metal Loss (One Half-in.)		Penetration (One Half - in.)		Comments
		Width (1")	Thickness (1/4")	Width (1")	Thickness (1/4")	
800	2.32	.0008	.0002	<.0001	<.0001	0
314	4.24	.0004	.0004	<.0001	<.0001	0,ss
310	4.29	<.0001	.0004	<.0001	<.0001	0
309	8.21	.0004	.0004	.0007	<.0001	0,ss
HAST-X	29.93	.0010	.0012	.0003	.0025	I0,s
811	31.18	.0005	.0001	.0001	.0001	0
304	36.05	.0016	.0011	.0017	.0026	I0,ss
671	92.86	.0133	.0004	<.0001	<.0001	0
804	147.85	.0140	.0153	<.0001	.0026	I0,ss
601	430.43	<.0001	.1080	<.0001	.0123	0,s
809	686.51	.0955	.0284	<.0001	<.0001	0,s
T-22	939.62	.0647	.0546	.0004	.0010	0,ss
600	1003.54	.1379	.0669	.0042	.0172	0,Is
1010	1135.14	.1018	.0799	.0004	.0004	0
690	1209.44	.3100	.0742	<.0001	.0141	0,s

major mode of corrosion for each alloy is indicated in the comments column. The key to the comments column is: O = oxidation, S = sulfidation, ss = slight sulfidation, IO = intergrannular oxidation, IS = intergrannular sulfidation.

The dominant mode of attack was oxidation. However, several alloys displayed extensive sulfidation attack very similar to that observed in gas turbines. These alloys include Incoloy 804, 809, and Inconel 600, 601 and 690.

Photomicrographs illustrating the mode of attach of the test rack samples are shown in Figures 2.18 through 2.25.

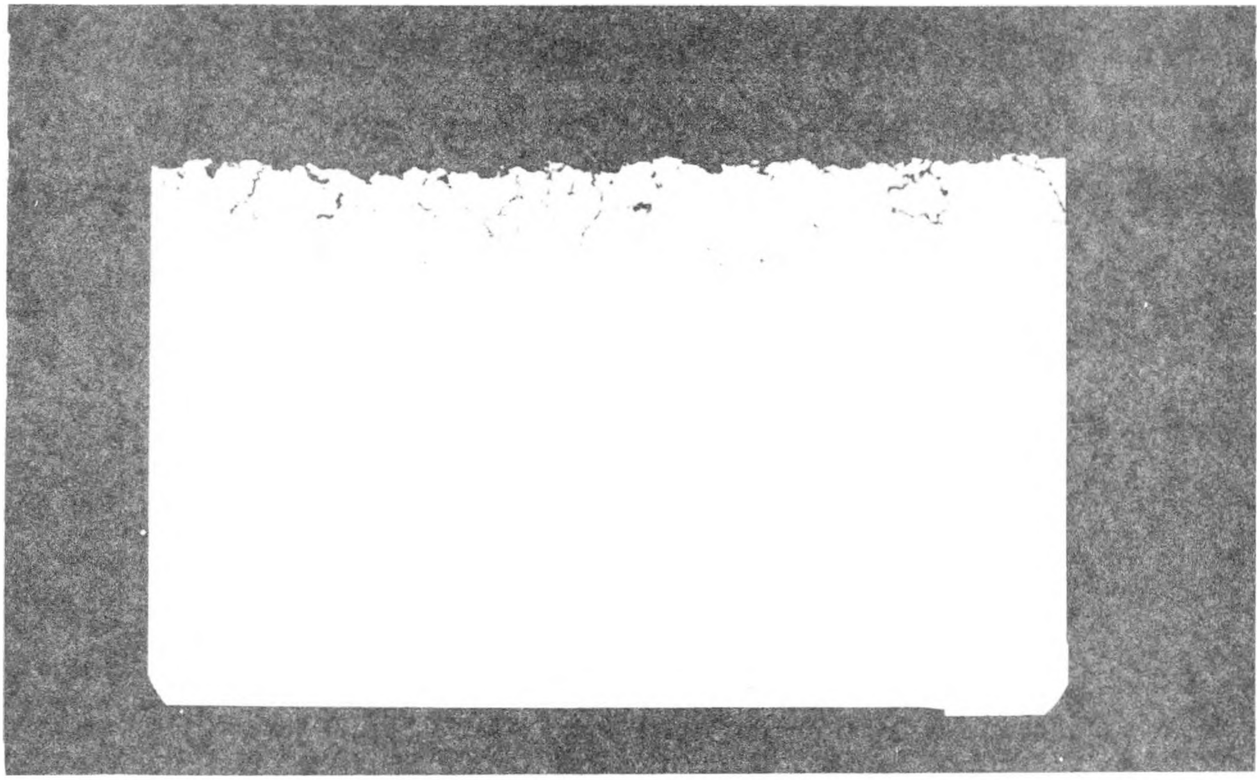
Sodium sulfate (Na_2SO_4) was fed during two runs in the corrosion/erosion test period. Each time the sodium sulfate was fed for approximately 30 minutes at a rate of 18 lbs/hr. This minor amount of sodium sulfate addition, in the view of the INCO metalurgists, caused irrepairable damage to the oxide scales that these alloys are intrinsically capable of developing. This would account for the poor performance of the higher nickel alloys. This type of poor performance is very common in gas turbine systems where high sulfur fuels and salt water ladened air are ingested. However, in view of the typical fluidized-bed environment, these test results must be considered biased by the sodium sulfate addition and therefore probably not typical of alloy performance in a fluidized bed furnace.

Due to the performance of the Inconel 601 test sample, pieces of the FBM horizontal tube bundle support ladder (also made of Inconel 601 but with no visible deterioration) were removed and are being examined metallurgically.

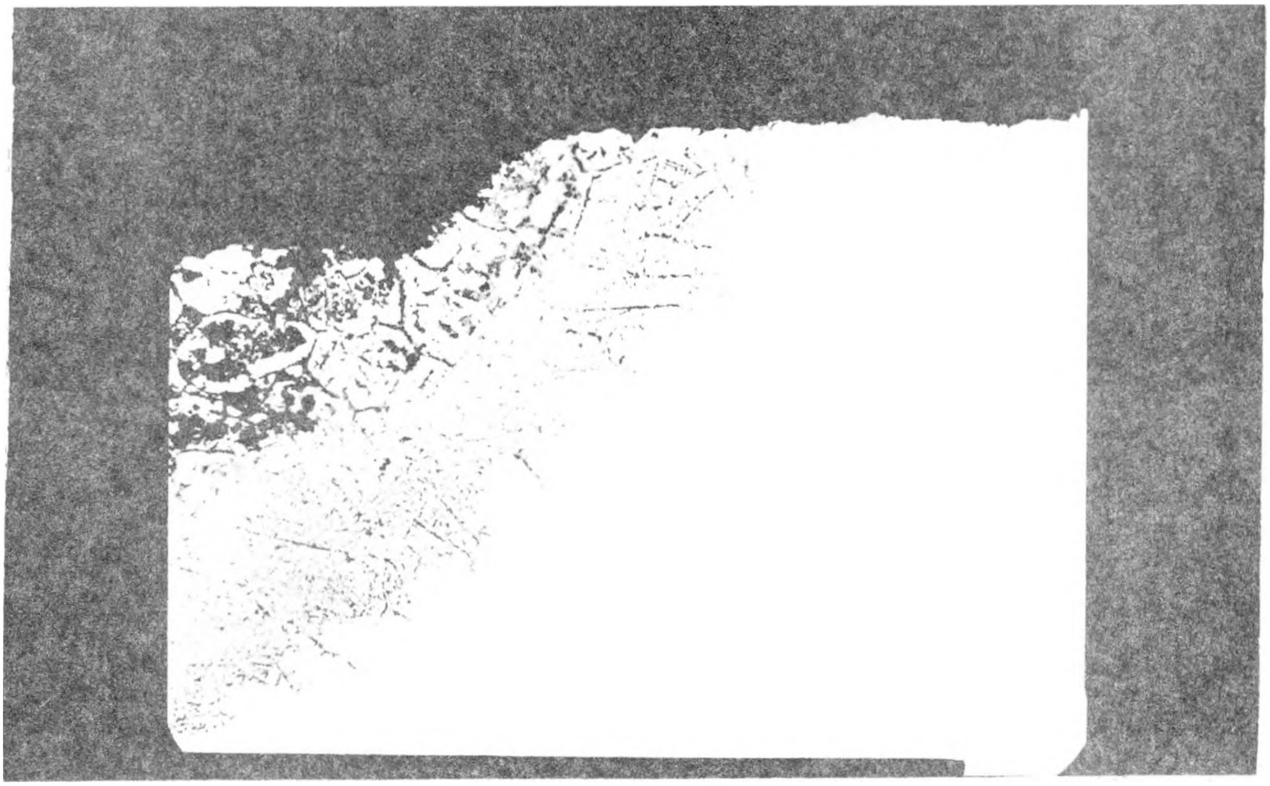
In order to examine the corrosion/erosion characteristics of the metal samples without the presence of sodium sulfate, two new test racks, identical to the previous one, were supplied by INCO. One rack was located in the bed 6½ inches above the grid plate while the other was located six inches below the steam drum. The test racks were removed after about 210 hours of operation and returned to INCO for analysis. The results of this second set of analyses is not complete yet.

2.4 Process Analysis Results

The results of tests obtained by measurement and data reduction are complemented by the results of process analysis. The general results obtained by process analysis are methods for checking test results, insight for understanding the process mechanisms and tools for the prediction of process behavior.

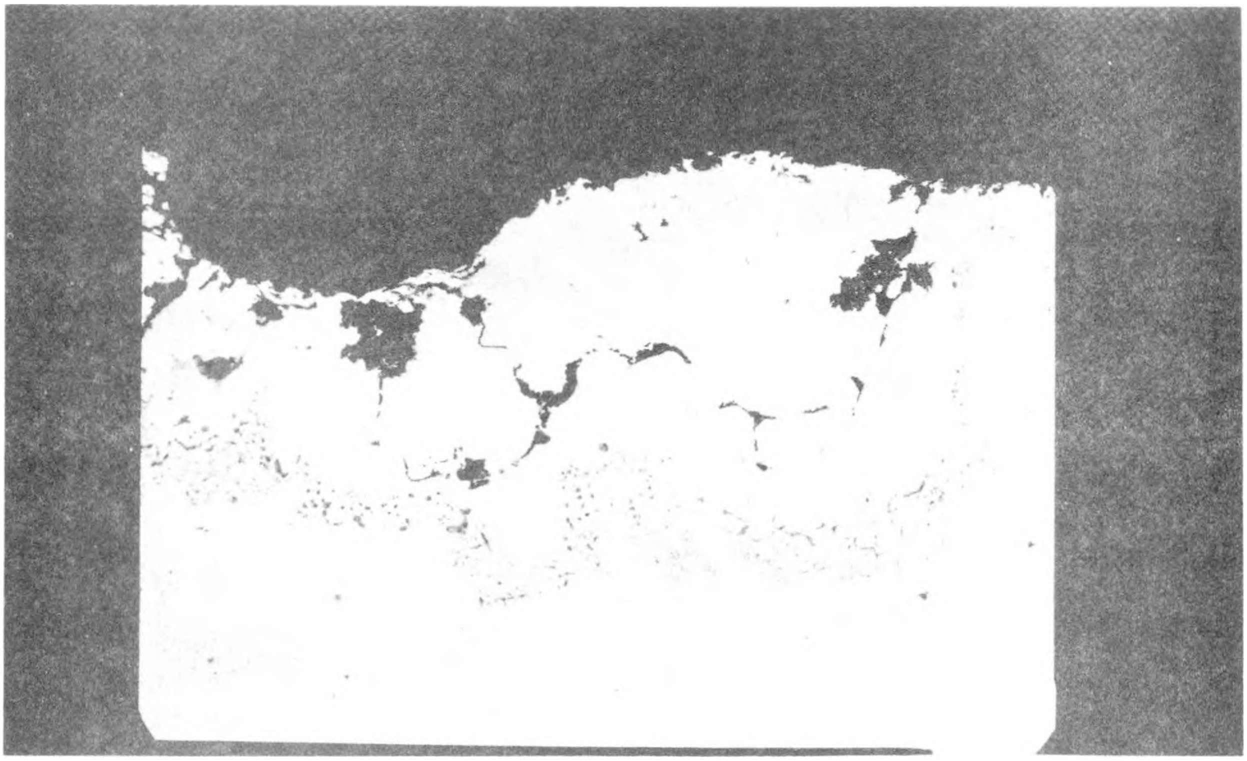


Type 304 Stainless 250X
Oxidation, Intergranular and Sulfidation

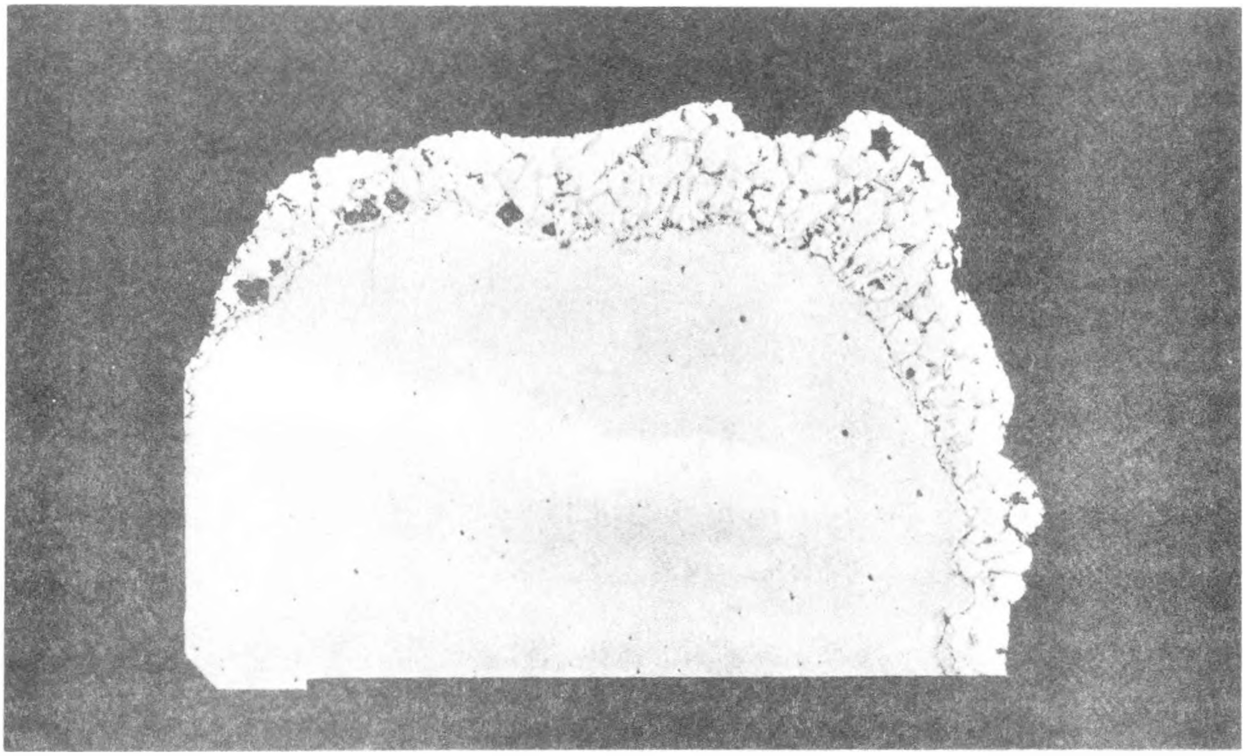


Incoloy 804 200X
Oxidation, Sulfidation and Intergranular

FIGURE 2.18

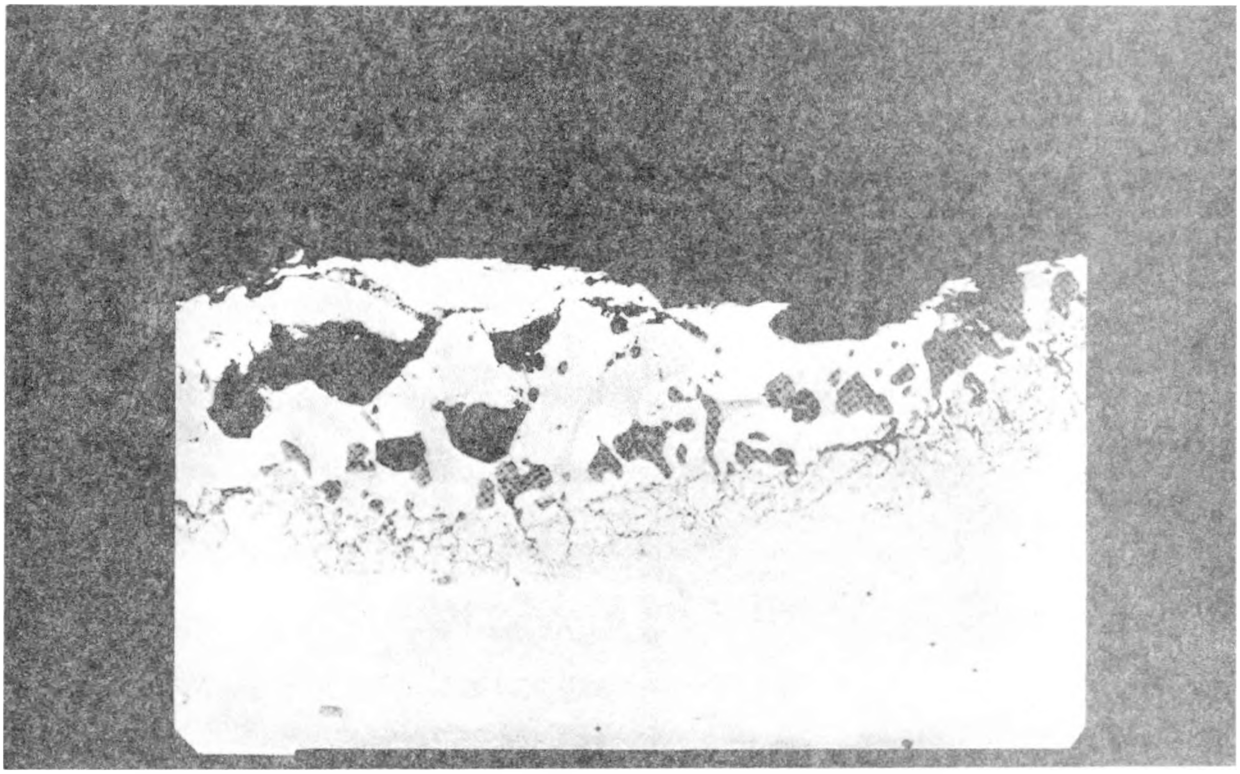


Inconel 600 250X Oxidation and Sulfidation

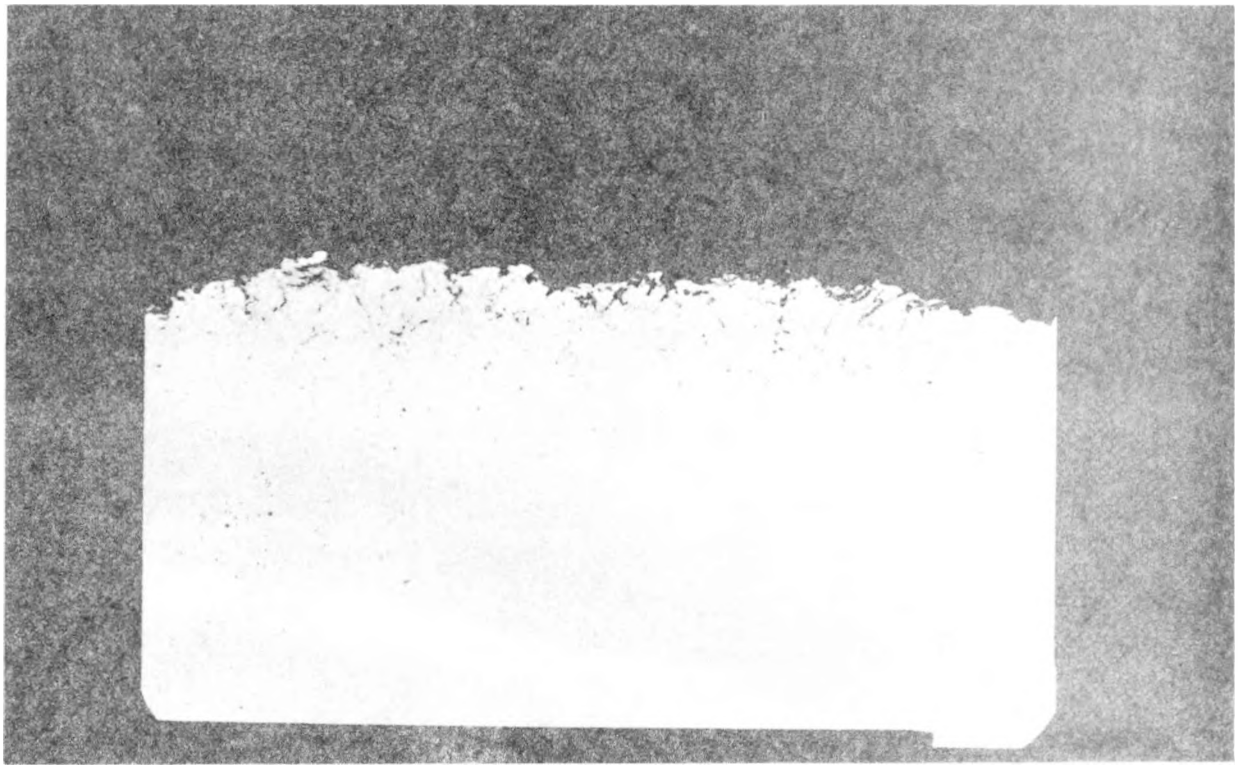


Inconel 601 50X Oxidation and Sulfidation

FIGURE 2.19

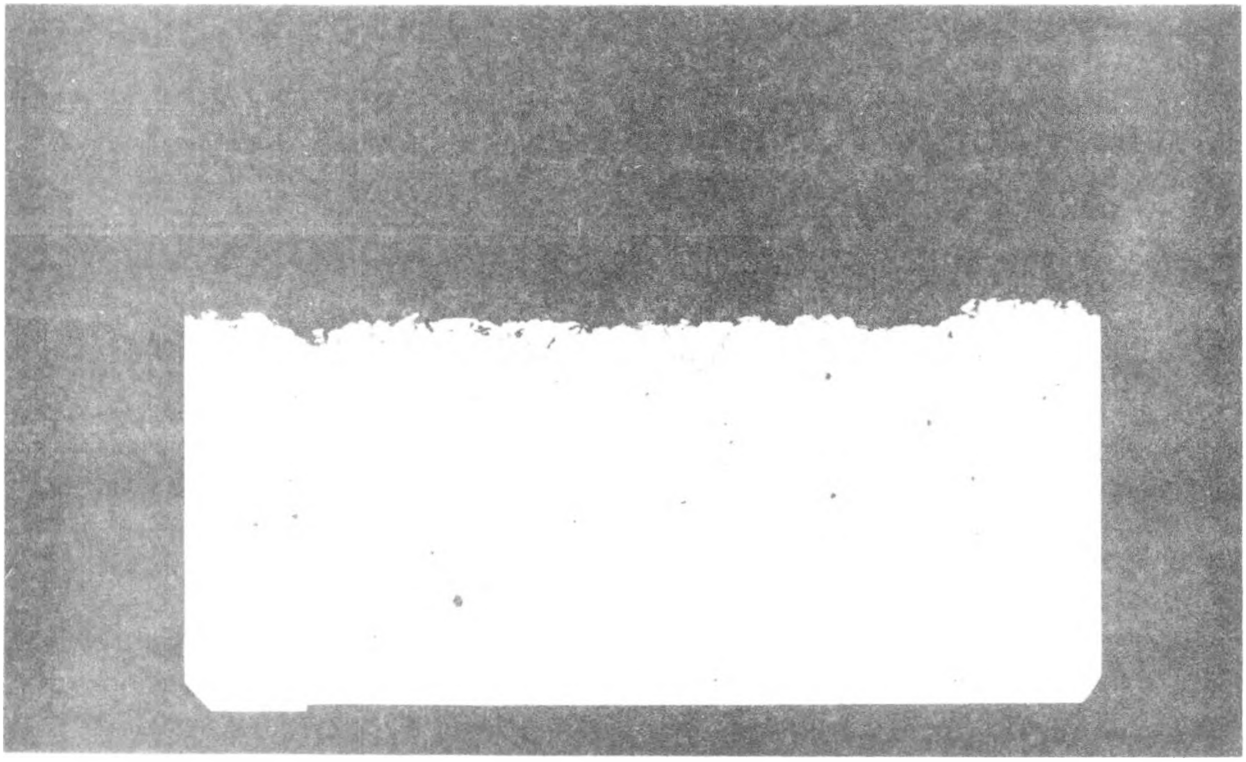


Inconel 690 250X Oxidation and Sulfidation

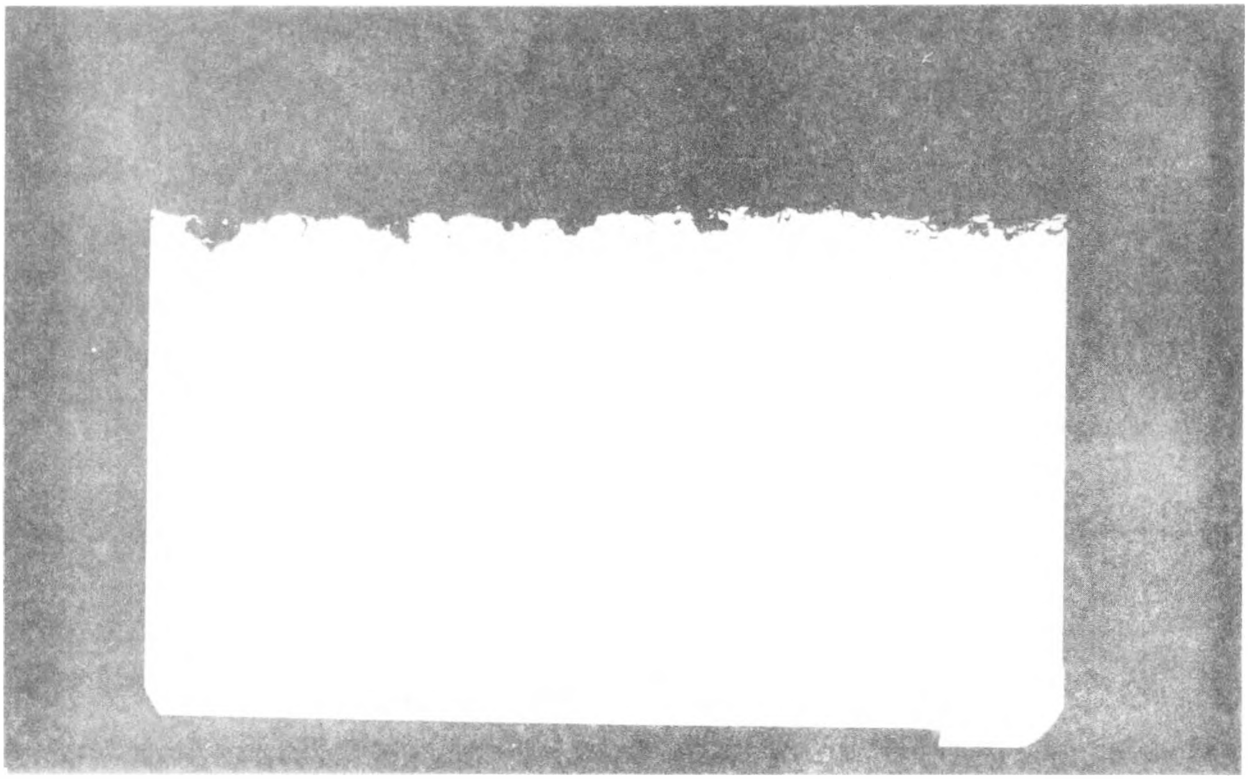


Alloy T-22 250X Oxidation and Some Sulfidation

FIGURE 2.20

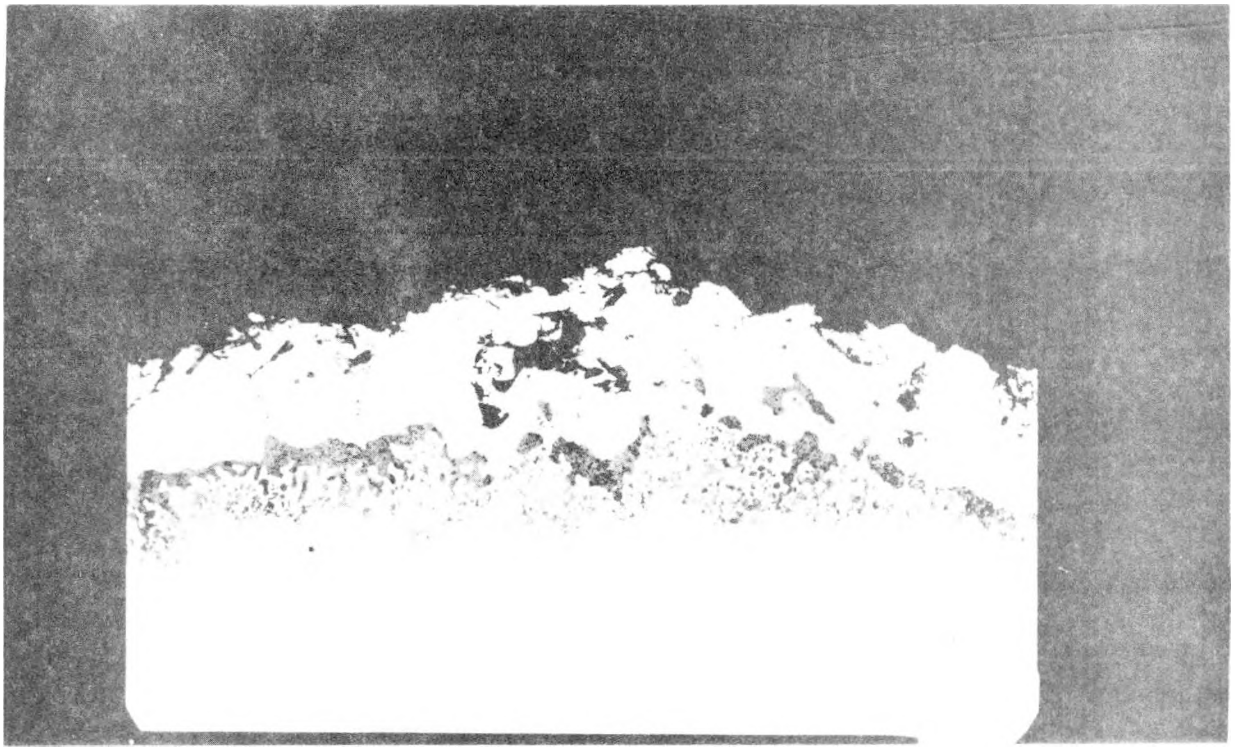


Type 309 Stainless 250X Oxidation and Sulfidation

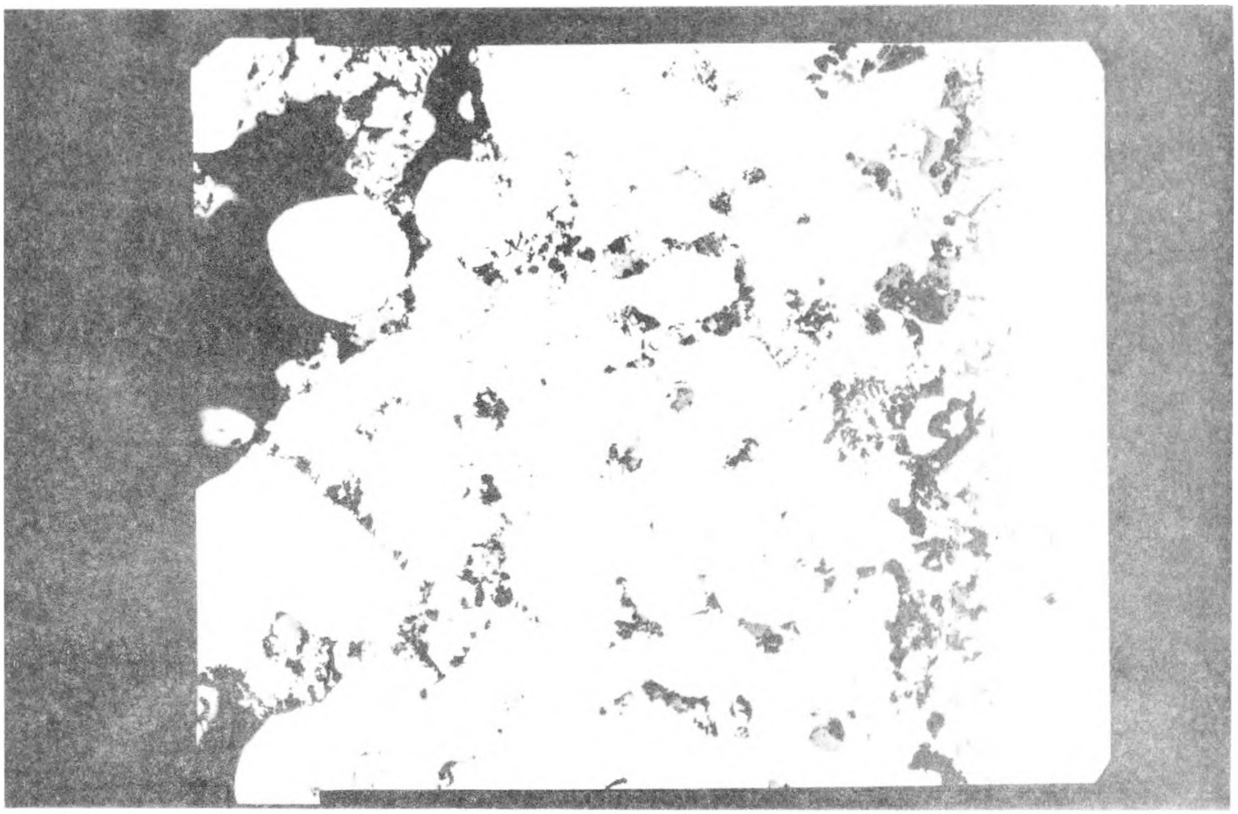


Type 314 Stainless 250X Oxidation and Sulfidation

FIGURE 2.21

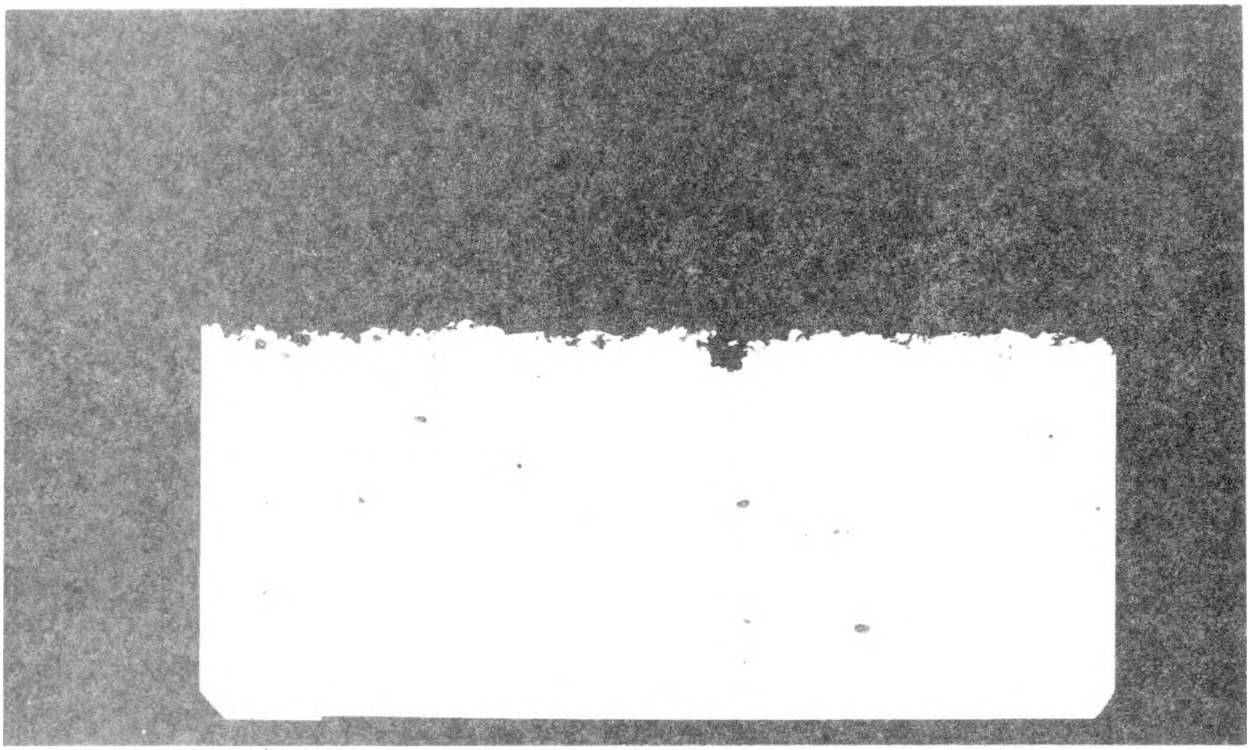


Incoloy 809 250X Sulfidation

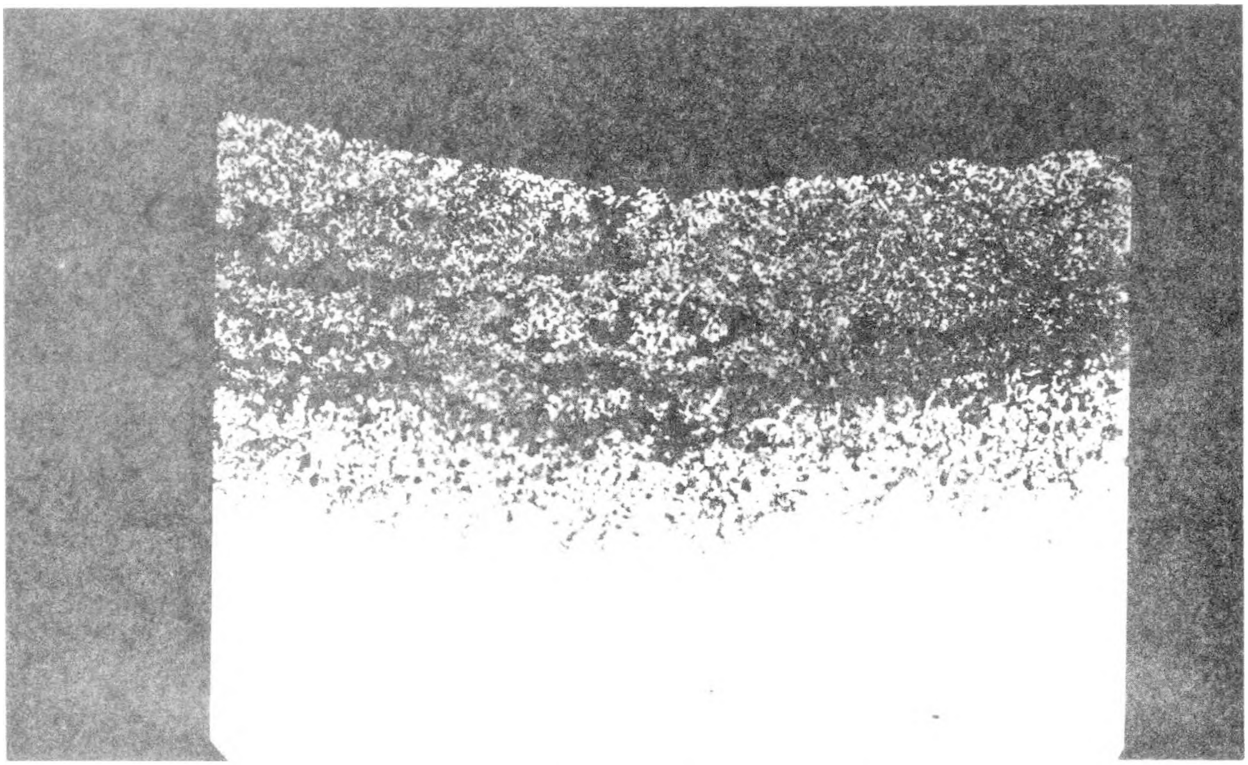


Incoloy 804 200X Sulfidation

FIGURE 2.22

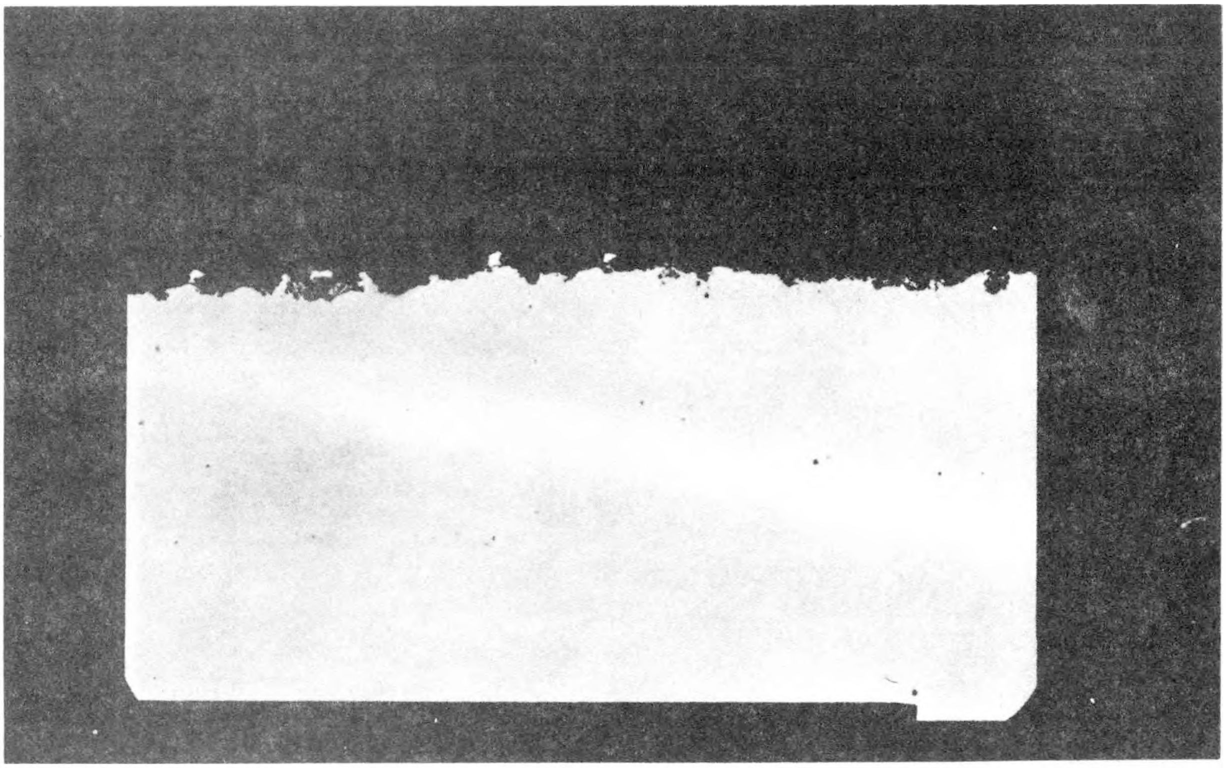


1010 Carbon Steel 250X Oxidation

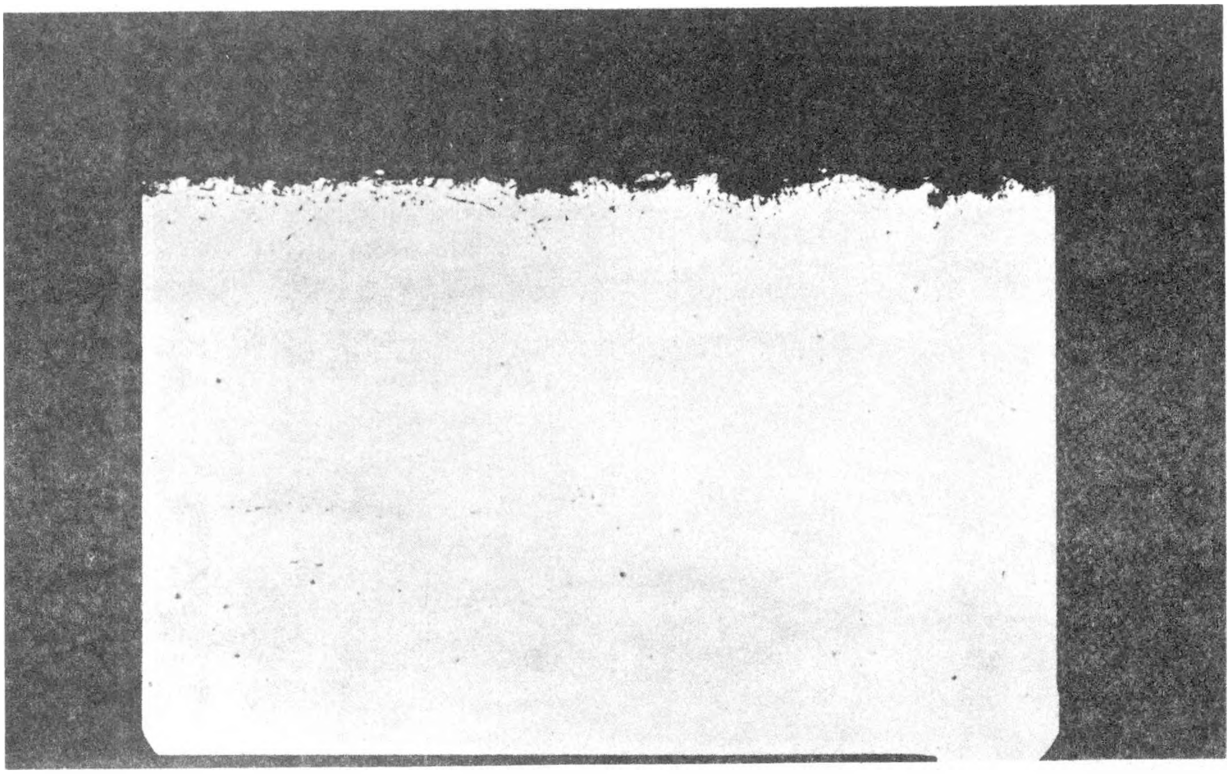


Inconel 671 250X Oxidation

FIGURE 2.23

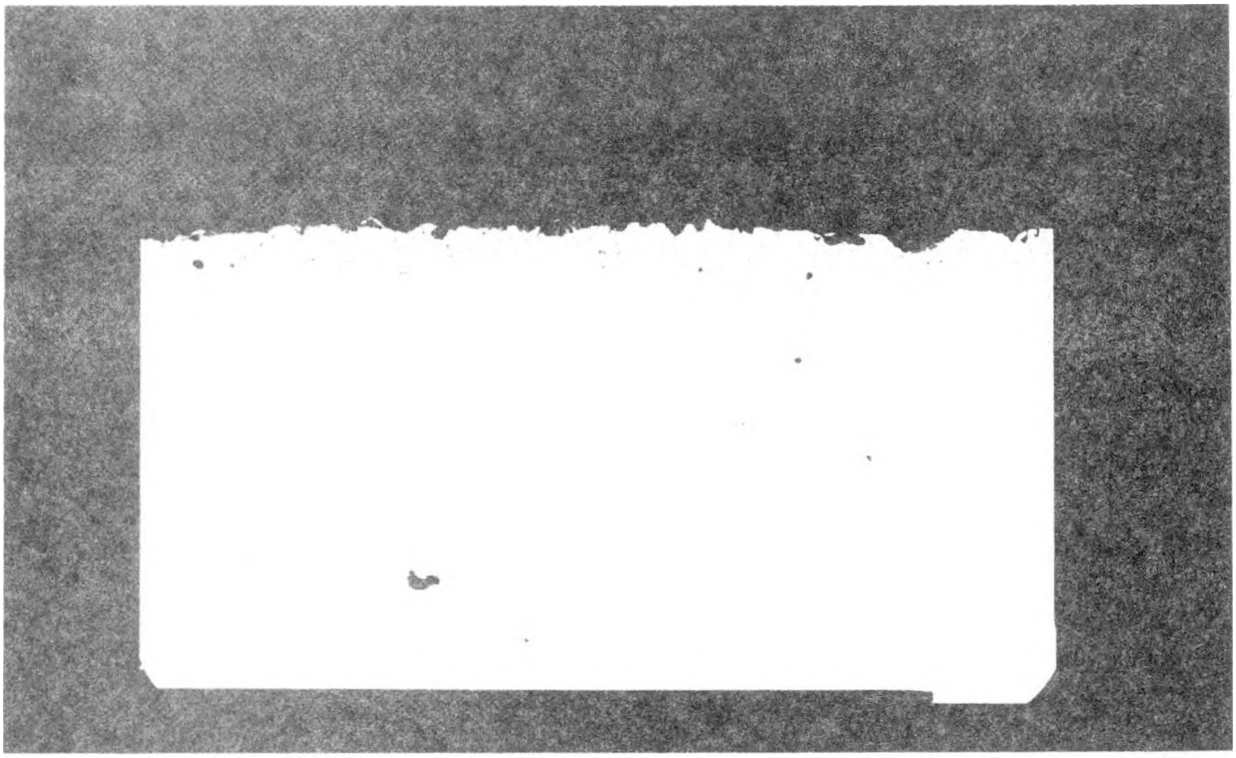


Incoloy 809 250X Oxidation

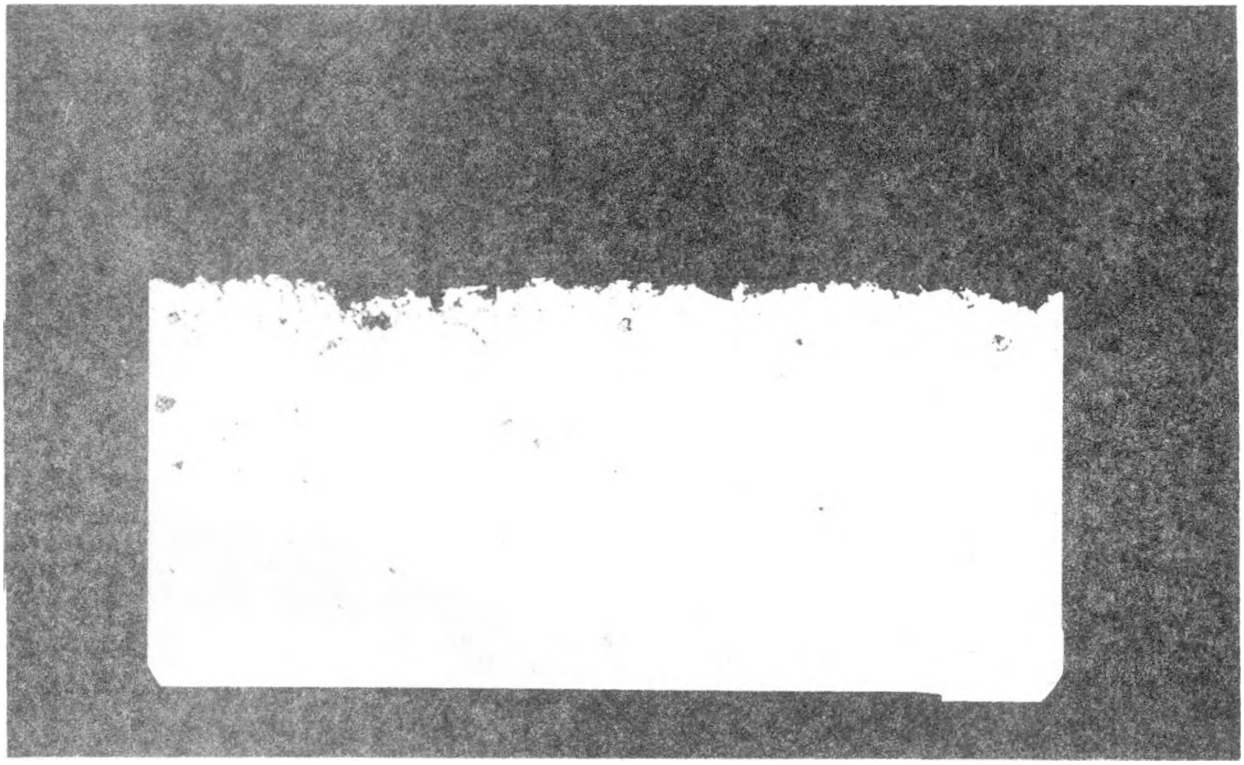


Inconel 811 200X Oxidation

FIGURE 2.24



Type 310 Stainless 250X Oxidation



Incoloy 800 200X Oxidation

FIGURE 2.25

In some cases, the analysis techniques are commonly used engineering methods adapted for the process and in other cases, the analysis method is developed specifically for the available test data. Examples of both are found in the following process analysis results for the reporting period.

2.4.1 Carbon-Burnup Cell Modeling

In order to estimate the relative importance of operating parameters in the CBC on carbon burnup, a simplified approach was attempted considering the burnout time of each particle to be given by:

$$t_b = Kd_o^2 \tag{1a}$$

where

K = rate constant, sec/ft^2

d_o = initial diameter, ft

The particle diameter at any time t can be found by differentiating and integrating Equation (2) to give

$$t_b - t = Kd^2 \tag{2}$$

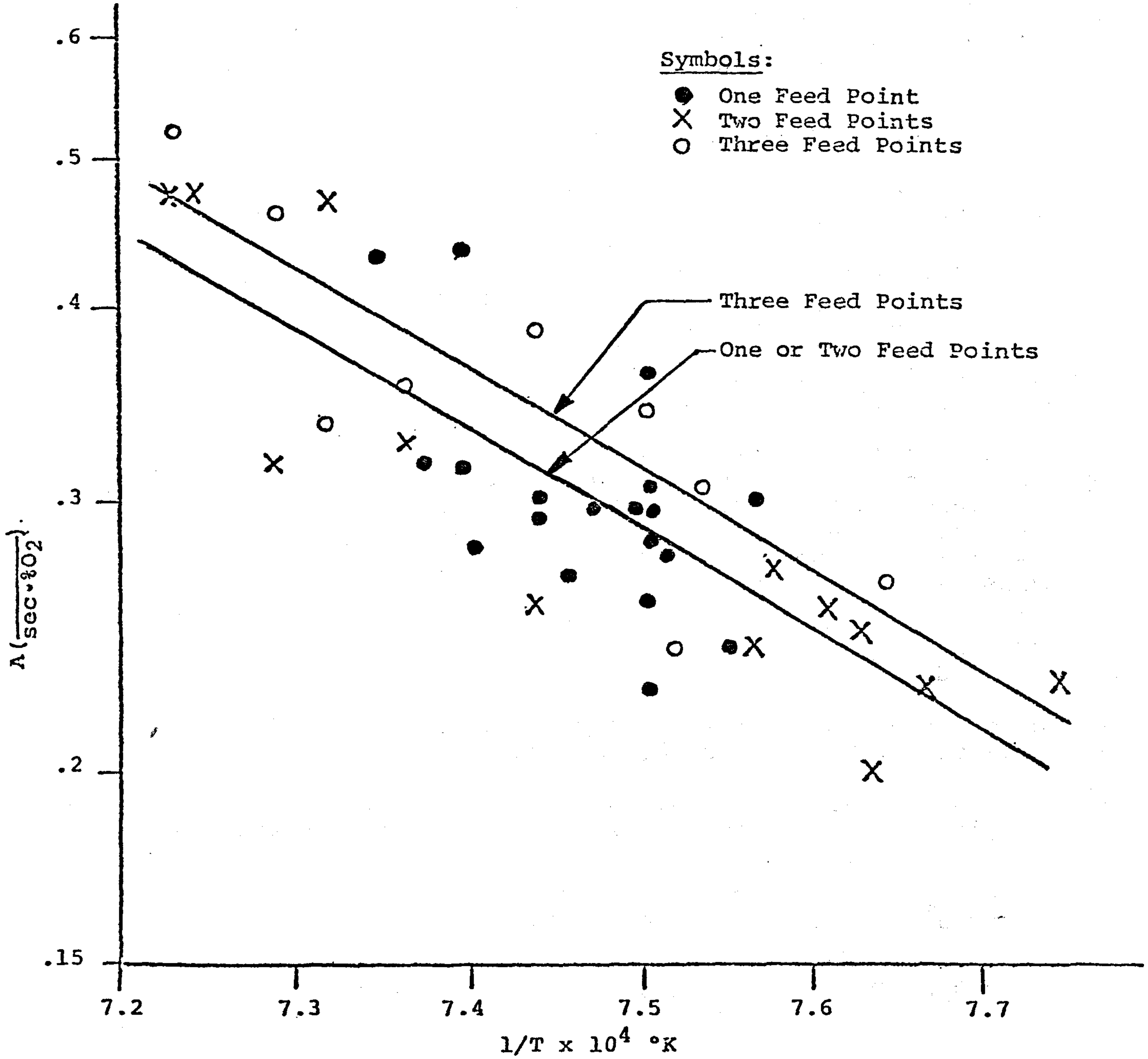
The fraction of carbon burnup is given by

$$C_B = 1 - \left(\frac{d}{d_o}\right)^3 = 1 - \left(1 - \frac{t}{Kd_o 2}\right)^{\frac{3}{2}} \tag{3}$$

The burnout time of a particle will be limited by the available oxygen. Thus, the constant K in Equation (3) can be assumed to be inversely proportional to the average oxygen concentration, c_o. The size distribution of the fly ash remained relatively constant throughout the test program so that the term d_o^2 can be assumed constant for this data. Furthermore, if we assume that the time for the reaction is proportional to the residence time in the bed (given by H/V Equation (3) can be rewritten as

$$C_B = 1 - \left(1 - \frac{AHCo}{V}\right)^{\frac{3}{2}} \tag{4}$$

A is constant which lumps all of the proportionally constants together. Figure 2.26 shows an Arrhenius plot of the constant A calculated from the test data. The straight lines in the figure are the least squares fits for the test data grouped by the number of feeders. Calculation of an activation energy from the average slope of the straight lines yields 29.8 KCal/g mole. The activation energy for combustion processes in this



ARRHENIUS PLOT OF MULTIPLE FEEDER TEST DATA

FIGURE 2.26

temperature range is about 25 KCal/g mole. The close agreement of these figures provides some confidence in this relatively simplistic model. It is interesting to note that the model showed the same predicted carbon burnup for one and two feed points and a slightly higher carbon burnup for three feed points.

Figures 2.27 and 2.28 show the calculated effects of bed temperature and excess oxygen for various valves of H/V. Due to the narrow range of the data, use of these figures should be limited to the range of the test data.

2.4.2 Light-Off By Hot Bed Transfer

Light-off of Cells A, B and C in the Rivesville MFB is accomplished by successive transfer of hot bed material from initial light-off of Cell D. In order to minimize the time required from initial light-off of Cell D to full load operation, an attempt was made to develop a correlation to predict transfer of bed material through a large opening with the resulting temperature-time response. This has resulted in a theoretical model which fits the bed transfer data obtained to date for light-off of Cell C.

The driving force for particle transfer between cells is the hydrostatic head difference of the fluidized materials in cells. This implies that transfer is complete when the hydrostatic pressure heads in the two connected cells are equal. At any time, the sum total mass of bed material in the two chambers is constant during the transfer operation. This mass is:

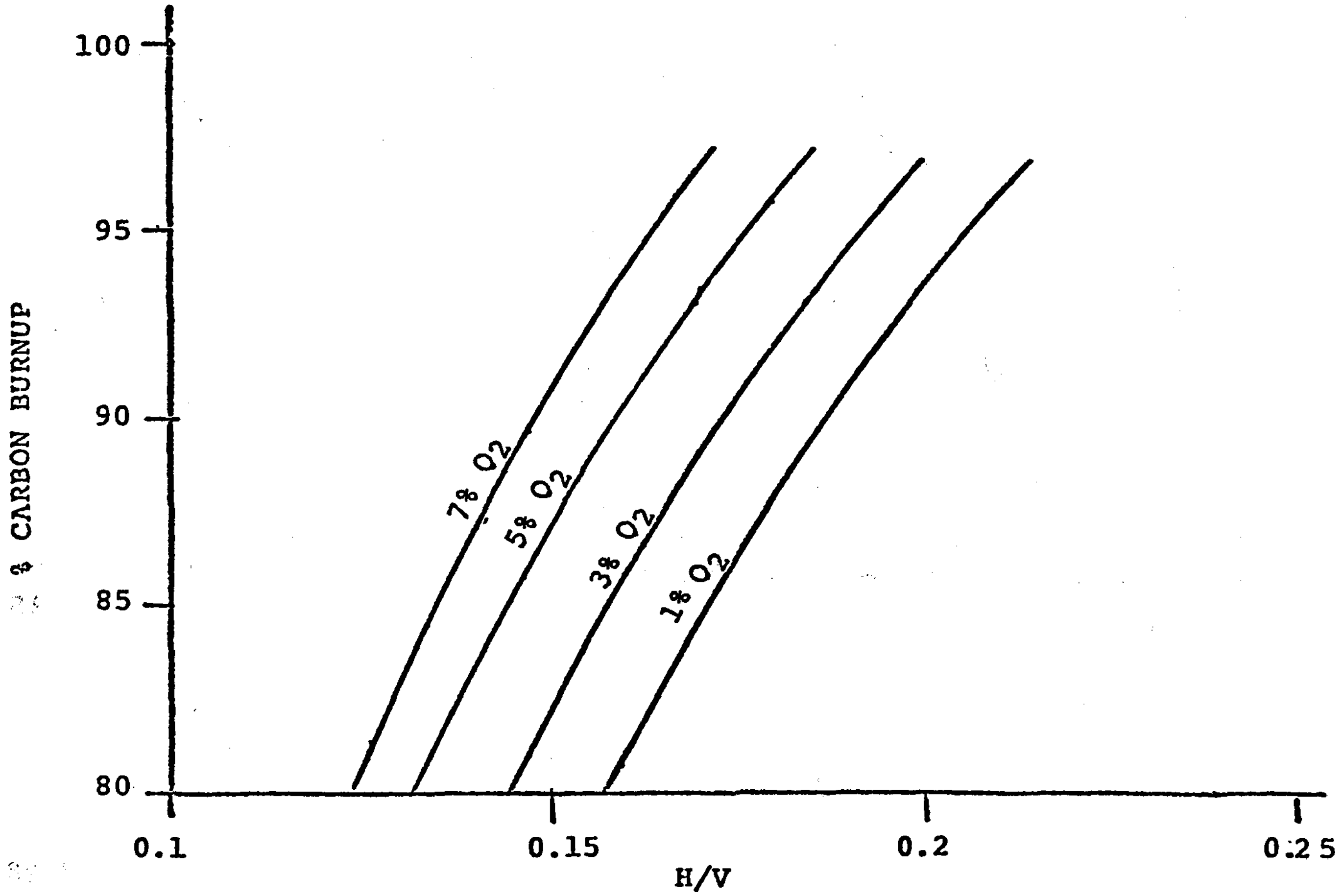
$$\rho_d h_d A_d + \rho_c h_c A_c \qquad (2)$$

where ρ is the bed density, h is the bed height, and A is the bed area in Cells D and C, respectively.

Setting this equal to the initial mass of material and solving for h_c yields the functional relation between the heights in the two beds as follows:

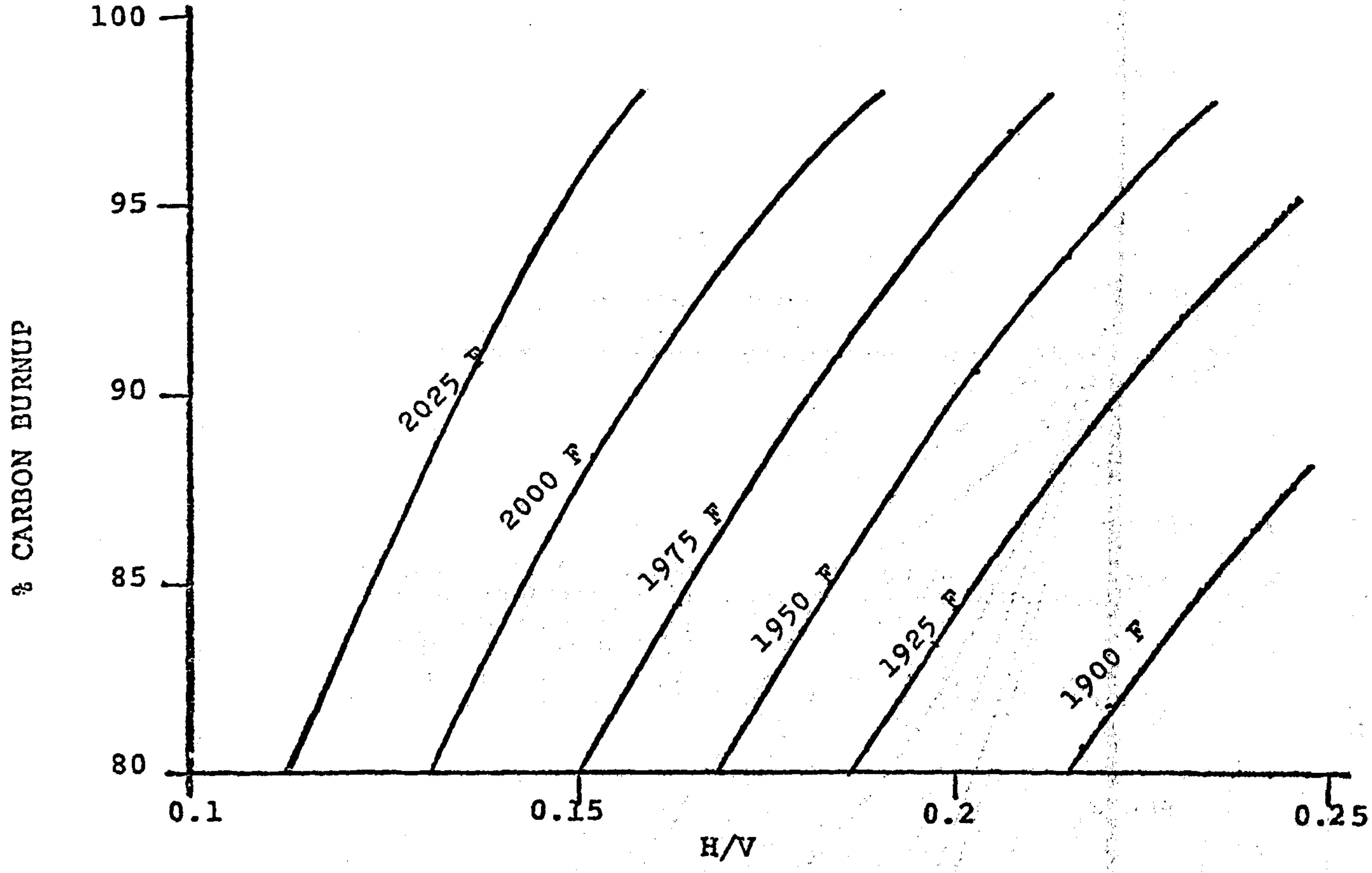
$$\rho_d h_d A_d + \rho_c h_c A_c = \rho_d h_{di} A_d + \rho_c h_{ci} A_c \qquad (3)$$

$$h_c = h_{ci} + \frac{\rho_d A_d}{\rho_c A_c} (h_{di} - h_d) \qquad (4)$$



CALCULATED EFFECT OF EXCESS OXYGEN ON CARBON BURNUP AT 2000°F WITH TWO FEED POINTS

FIGURE 2.27



CALCULATED EFFECT OF BED TEMPERATURE ON CARBON BURNUP AT 5% EXCESS O_2 WITH TWO FEED POINTS

FIGURE 2.28

heads in the two connected cells are equal. At any time, the sum total mass of bed material in the two chambers is constant during the transfer operation. This mass is:

$$\rho_d h_d A_d + \rho h_c A_c \tag{2}$$

where ρ is the bed density, h is the bed height, and A is the bed area in Cells D and C. respectively.

Setting this equal to the initial mass of material and solving for h_c yields the funtional relation between the heights in the two beds as follows:

$$\rho_d h_d A_d + \rho_c h_c A_c = \rho_d h_{di} A_d + \rho_c h_{ci} A_c \tag{3}$$

$$h_c = h_{ci} \frac{+\rho_d A_d}{\rho_c A_c} (h_{di} - h_d) \tag{4}$$

The parameter ρ_d/ρ_c, a dimensionless relative bed density, is found by measuring the final height in each cell and plugging it into Equation (4).

Applying the mechanical energy balance equation to the system yields the following equations:

$$\text{Kinetic Energy} = \text{Potential Energy} \tag{5}$$

$$\tfrac{1}{2} m V^2 = mgh \tag{6}$$

$$V = c\sqrt{2gh} \tag{7}$$

This is often used for discharge of fluids from an orifice where V is the discharge velocity through the orifice, h is the static pressure head across the orifice, and C is the orifice discharge coefficient. The orifice discharge coefficient is empirically determined and compensates for losses. For the case in point, Equation (7) can be modified to:

$$V = C\sqrt{2g(h_d - h_c)} = C\sqrt{2g\left[h_d - h_{ci} - \frac{\rho_d A_d}{\rho_c A_c}(h_{di} - h_d)\right]} \tag{8}$$

The change in volume of bed in Cell D per unit time is related to the discharge velocity times the open area of the orifice, A_o.

$$-A_d \frac{dh_d}{dt} = V A_o \tag{9}$$

Rearranging and integrating from t_o to t and from h_{di} to h_d:

$$\int_{h_{di}}^{h_d} \frac{dh_d}{V} = \frac{A_o}{A_d} \int_{t_o}^{t} dt \tag{10}$$

$$\int_{h_{di}}^{h_d} \left[\left(1 + \frac{\rho_d A_d}{\rho_c A_c}\right) h_d - h_{ci} - \left(\frac{\rho_d A_d}{\rho_c A_c}\right) h_{di}\right]^{-\frac{1}{2}} dh_d = -C(2g)^{\frac{1}{2}} \frac{A_o}{A_d} (t-t_o) \tag{11}$$

or

$$h_d = g \left(1 + \frac{d}{c} \quad \frac{A_o}{d}\right)\left(\frac{CA_o}{A_d}\right)^2 \left(t-t_o\right)^2$$

$$\sqrt{2g \quad (h_{di} - h_{ci})} \quad \left(\frac{CA_o}{A_d}\right)(t - t_o) + h_{di} \tag{12}$$

The orifice discharge coefficient C, and the time of initial transfer t_o, can be found by plotting observed time of transfer to a given height in Cell D against a calculated pseudo height assuming C = 1 and $t_o = 0$. The slope of the line is experimental C and the ordinate intercept is C t_o. Alternately, this can be fit by linear least squares regression. The transfer from cell to cell stops when the respective heights in the cells become equal. This "time of total transfer" can be calculated by setting the derivative of the height equal to zero. The result is :

$$\text{Time of Total Transfer} = t_o + \tag{13}$$

The temperature in Cell C as a function of time can be estimated as the bulk mixed temperature of the initial contents of Cells C and D. This assumes that the energy generated by combustion equals the energy lost to the surrounding container and gases in Cell C, that mixing is fast with respect to the transfer time and that the solids heat capacity is nearly constant. A simple enthalpy balance gives the temperature in Cell C as:

$$T_c = \frac{h_{ci}T_{ci}+XT_d(h_{di}-h_{ci})^{\frac{1}{2}}(t-t\ )-Y\ T_d(t-t_o)^2}{h_{ci}\quad +X\quad (h_{di}-h_{ci})^{\frac{1}{2}}(t-t_o)-Y\quad (t-t_o)\ 2} \qquad (14)$$

where:

$$X = \frac{\rho_d}{\rho_c}\,\frac{A_d}{A_c}\,\sqrt{2g}\left(\frac{A_o}{A_d}\right)C \qquad (15)$$

and

$$Y = -\frac{g}{2}\,\frac{\rho_d}{\rho_c}\,\frac{A_d}{c}\left(1+\frac{\rho_d}{\rho_c}\,\frac{A_d}{A_c}\right)\left(\frac{C\ A_o}{A_d}\right)^2 \qquad (16)$$

<u>Sample Calculations:</u> Recently at the Rivesville MFB, overall bed static pressure tap readings from the magnahelic pressure gage on Cell D were taken every ten seconds after the gate was visually observed to be open. The readings were as follows:

Time, Sec.	P, Inches of H_2O
0	33
10	28
20	22
30	18
40	17

50	17
60	16

The gate opening was taken to be 144 in^2. Cell D has a grid area of 72 ft^2. The superficial velocity in Cell D was 8.0 ft/sec at 1800°F. Cell C has a grid area of 120ft^2. The superficial velocity was 3.5 ft/sec. at its initial temperature of 400°F. The initial Cell C pressure drop was 7.5" H_2O and the final reading was 16" H_2O.

Since pressure drop readings rather than heights were used, the relative density ratio must be modified such that the overall bed pressure drop can be substituted for height in the above equations. Therefore, in the following discussion, overall bed pressure drop will be assumed equal to the "apparent" bed height. The minimum height in equation (12) was set equal to the final pressure drop in Cells C and D, 16" H_2O. The ratio $d/\rho c$ was found to be equal to 0.833. Then Equation (12) was fit to the above data. The best fit of this data yielded a discharge coefficient of 0.433 and a time of initial transfer of 3.62 seconds.

The bed height (overall pressure drop) in Cell D was calculated as a function of time as seen from Figures 2.29 and 2.30. The correlation was expanded by assuming the same constants and calculating the bed height as a function of time assuming various initial bed heights in Cell D conditions (figure 2.30). It can be seen that the curve fits the small amount of data available. Also as the relative height between cells increases, the rate of transfer increases as does the time of total transfer. The expected bulk mixing temperature versus time curves were calculated for Cell C for these two variables. Cell D initial height was varied at constant Cell C conditions (Figure 2.31) and Cell C initial height was varied at constant Cell D conditions (Figure 2.32). It can be observed that a criterion of 15 seconds to reach ignition temperature can be achieved if the relative bed height (overall pressure drop) ratio is 4.4 or greater. If the relative bed height was varied at constant Cell D conditions (Figure 2.32). It can be observed that a criterion of 15 seconds to reach ignition temperature can be achieved if the relative bed height (overall pressure drop) ratio is 4.4 or greater. If the relative bed height ratio is much less than three, a bulk mixing

TRANSIENT CELL D BED HEIGHT FOR VARIOUS INITIAL CELL D HEIGHTS

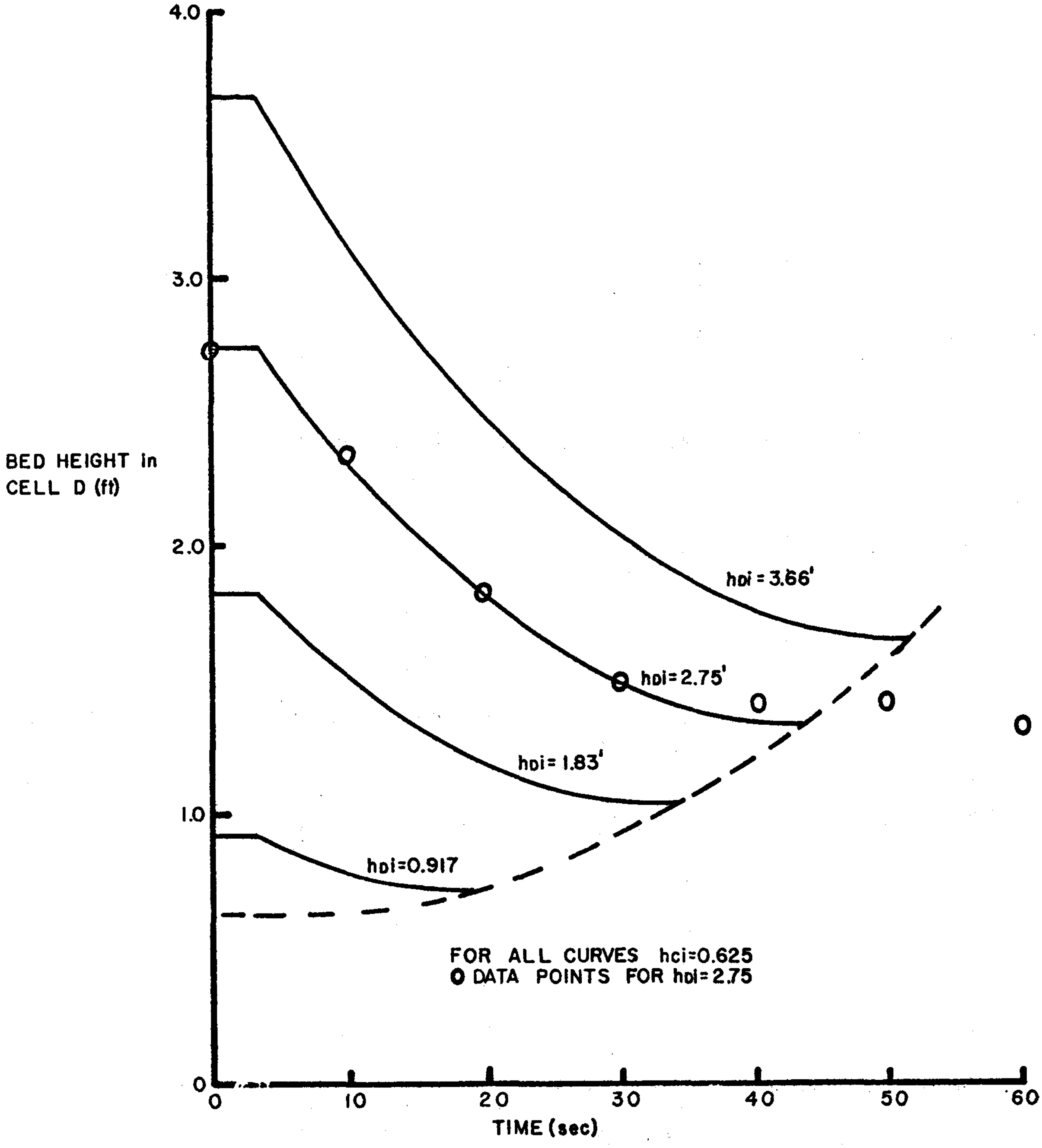


FIGURE 2.29

TRANSIENT CELL D BED HEIGHT FOR VARIOUS INITIAL CELL C HEIGHTS

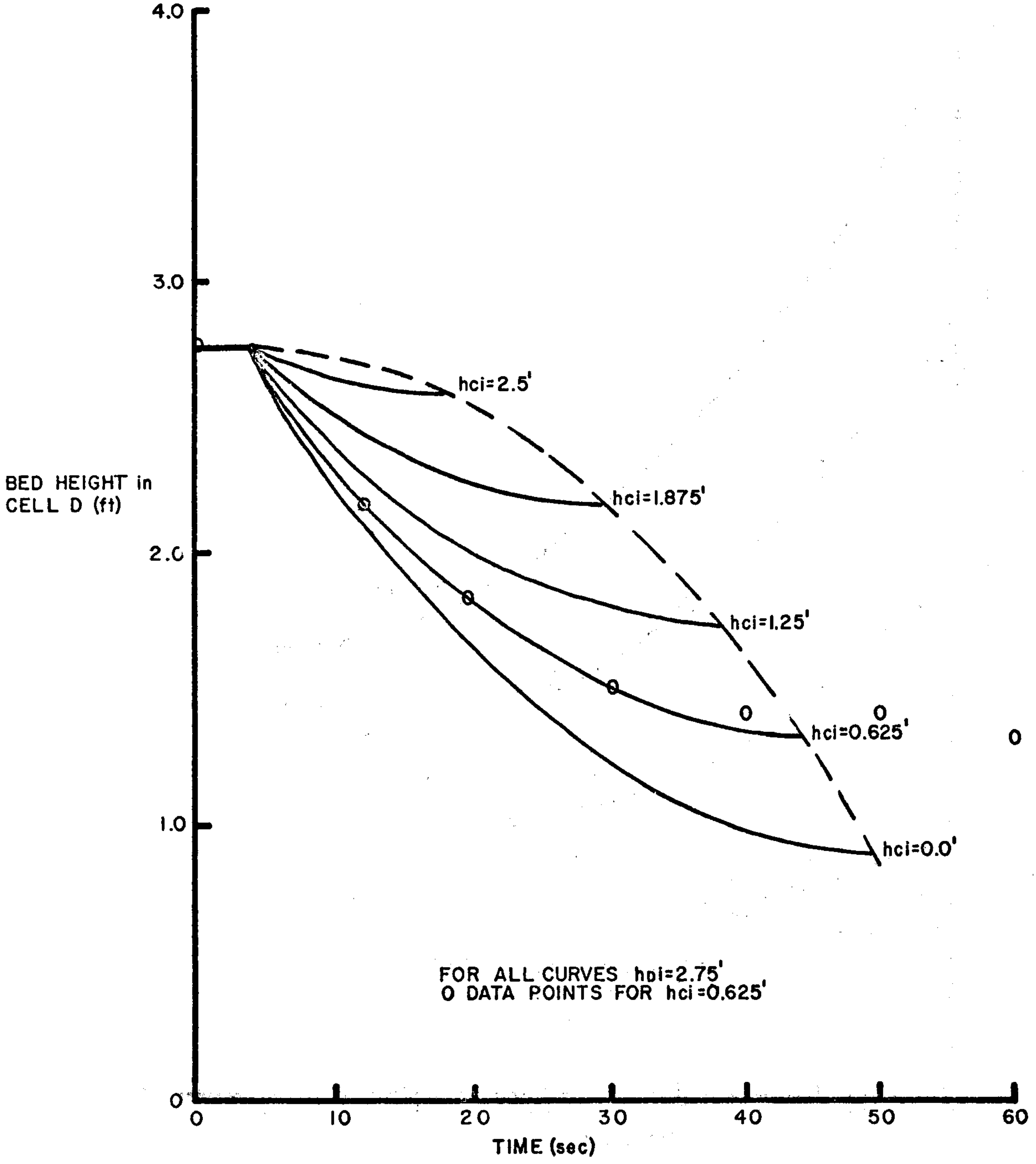
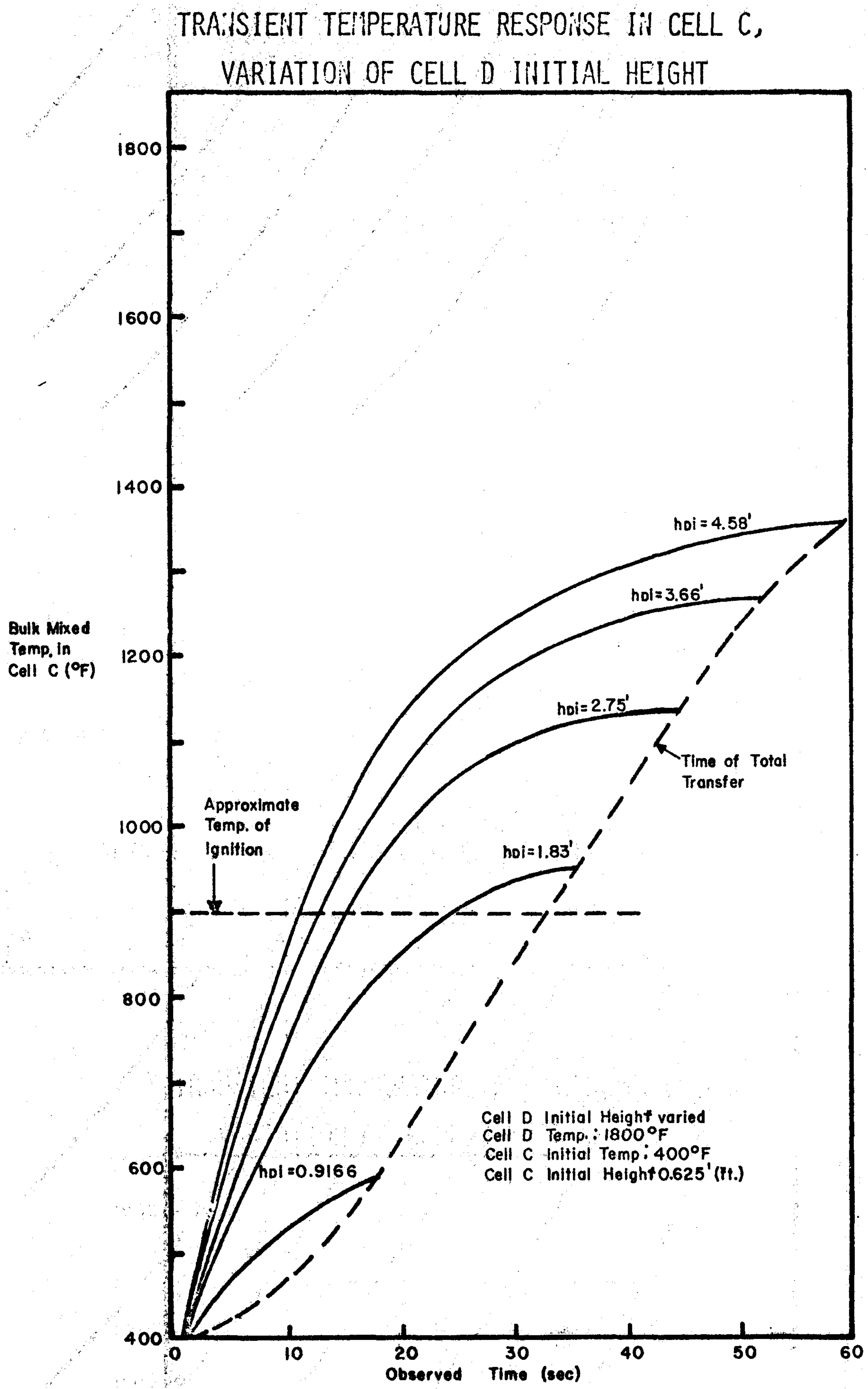


FIGURE 2.30



TRANSIENT TEMPERATURE RESPONSE IN CELL C, VARIATION OF CELL D INITIAL HEIGHT

FIGURE 2.31

TRANSIENT TEMPERATURE RESPONSE IN CELL C,

VARIATION OF CELL C INITIAL HEIGHT

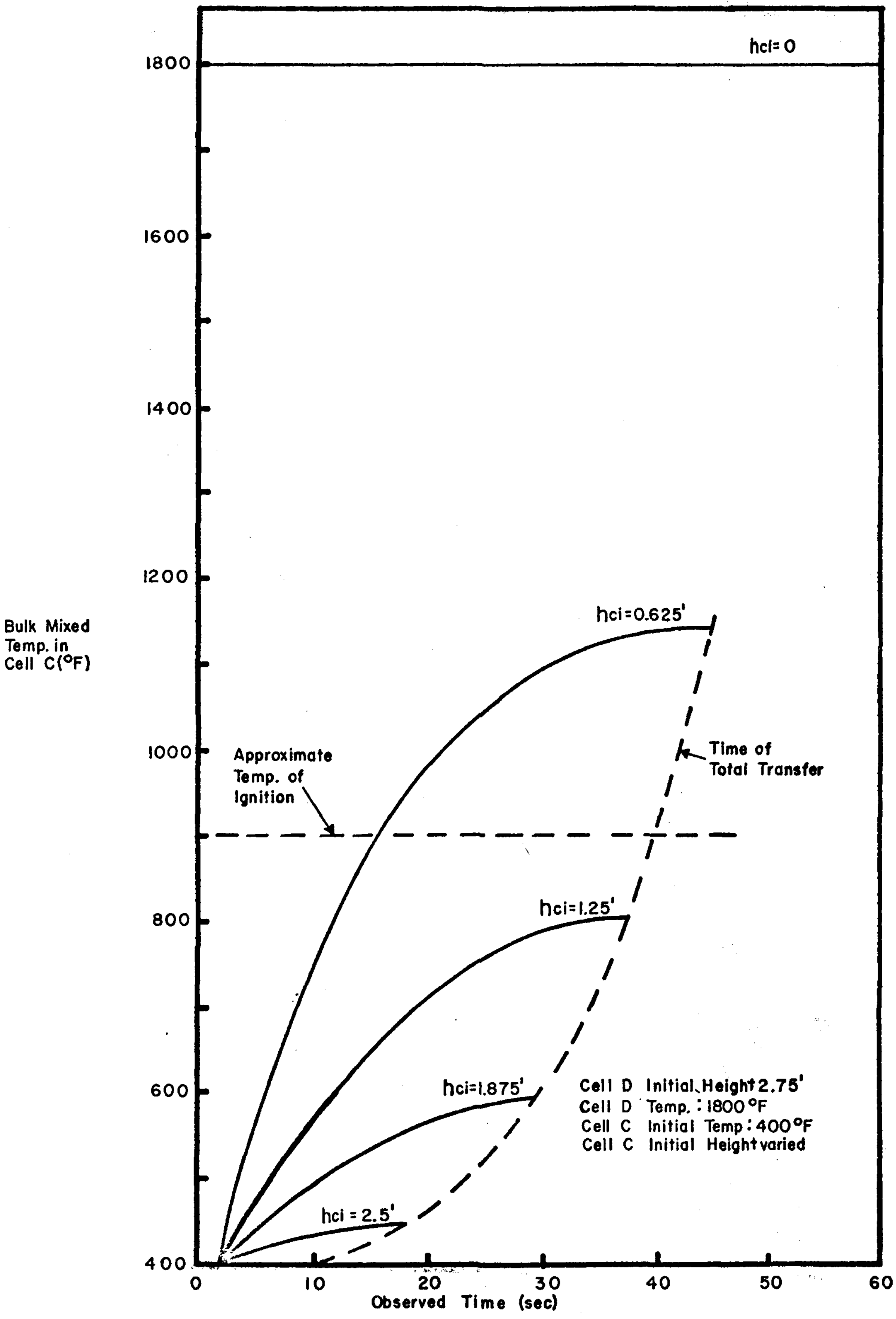


FIGURE 2.32

temperature of 900°F will never be achieved, unless ignition occurs before the hot and cold bed material mix. The ideal case is where no material exists initially in the receiving cell. In this case there is no cold diluent and the bed is immediately above ignition temperature.

Further data will be collected in order to verify this correlation model for temperature-time response.

2.4.3 Oversized Ash Concentrations

A solids material balance around the combustion zone of the fluidized bed can be used to determine the highest concentration level the oversized (+8 mesh) ash will attain in the bed. This saturation level is instructive in demonstrating the maximum benefit which can be derived from the use of the bed classifier. The analytical procedure consists of determining the concentration (and therefore mass) of oversized ash in the feed stream (coal + limestone) and the fraction of these materials which are removed from the solid phase during combustion/reaction. By knowing the mass of solids which remain after some specified combustion and reaction completion levels the concentration of oversized ash in the solid products can be determined.

The maximum oversized ash concentration can be determined by analyzing the process as a continuously stirred-tank reactor (CSTR). The system, as depicted in Figure 2.33 consists of the CSTR: solid inputs; gas phase outputs derived from solid inputs; and solid phase outputs derived from the solid inputs.

Assuming zero accumulation (steady-state operation) the material balance is written as:

input - gas phase output = solid phase output

Where the inputs are: coal, limestone and SO_3, captured as solid calcium sulfate, gas phase outputs are: coal ash and limestone fines, coal volatiles and combustibles, and limestone volatiles (CO_2 loss on calcination); and solid phase outputs are: unburned coal, coal ash, uncalcined limestone, calcined limestone and sulfated limestone.

Using these constituents the materials balance can be written:

$$F_C+F_L+.58F_L S_L X_{CaCO_3}\, e1 \quad - \quad (1-S_C)F_C+(1-S_L)F_L$$

$$X_A\ (1-S_A)F_C-X_A(1-S_C)F_C \quad + \quad X_{VC}F_C S_C e_2 \quad + \quad X_{VL}F_L S_L e_3 = F_B \qquad (1)$$

where

F_C = coal feed rate, lbs/hr

F_L = limestone feed rate, lbs/hr

F_b = solid products outlet rate, lbs/hr

S_C = fraction of coarse coal particles

S_L = fraction of coarse limestone particles

S_A = fraction of coarse coal ash particles

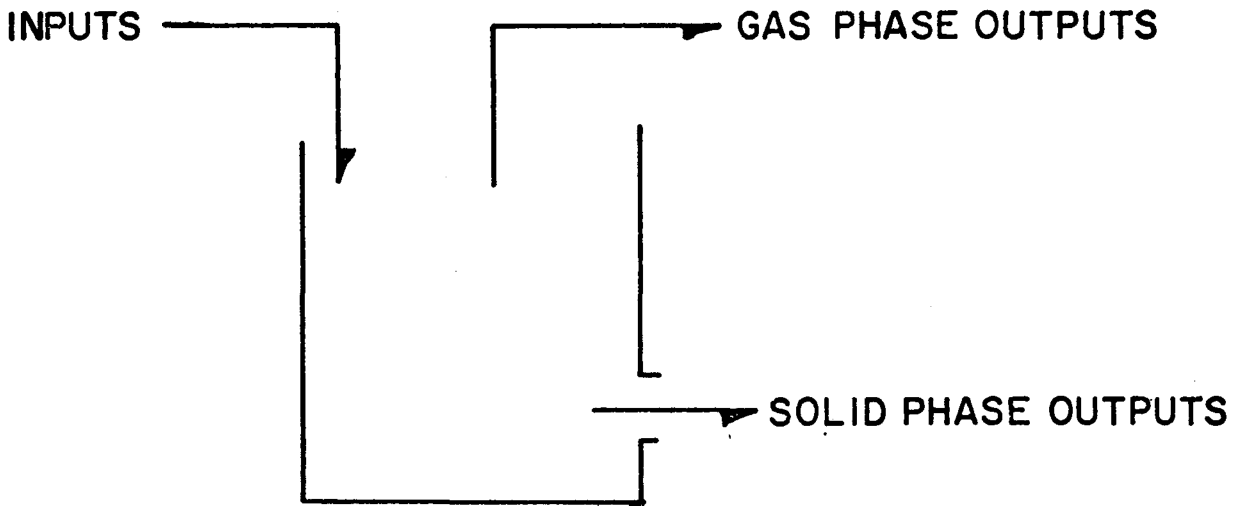
X_{CaCO_3} = fraction of $CaCO_3$ in limestone

X_{VC} = fraction of volatiles and combustibles in coal (loss on ignition

X_{VL} = fraction of volatiles in limeston (loss on calcination)

X_A = fraction of ash in coal

e_1 = extent of completion of $CaCO_3$ $CaSO_4$ reaction



CONTINUOUS STIRRED-TANK REACTOR MODEL FOR PERFORMING A SOLIDS MATERIALS BALANCE AROUND THE FLUIDIZED BED COMBUSTION ZONE.

FIGURE 2.33

e_2 = extent of completion of coal volatilization and combustion

e_3 = extent of completion of limestone volatilization

Coarse particles are those which are not transported out of the bed at operating superficial gas velocity. All fractions are on a weight basis.

The concentration of oversized ash (the fraction removed by a classifier) in the solids output stream, X_{OA}, can be defined as:

$$X_{OA} = \frac{X_A S_{OA} F_C}{F_B} \qquad (2)$$

where

S_{OA} = fraction of oversized (+8 mesh) coal ash

The limestone feed rate may be expressed in terms of the coal feed rate and a Ca/S molar ratio as follows:

$$F_L = \frac{40\ X_S\ (Ca/S)}{32\ X_{Ca}} F_C = \frac{1.25\ X_S\ (Ca/S)}{X_{Ca}} F_C \qquad (3)$$

where

X_S = fraction of sulfur in coal

X_{Ca} = fraction of calcium in limestone

Substituting equation (3) into equation (1) gives:

$$F_C \left\{ \left[1 + \frac{1.25\ (Ca/S) X_S}{X_{Ca}} + \frac{.725\ (Ca/S) X_S S_L X_{CaCO_3} e_1}{X_{Ca}} \right] - \left[(1-S_C) + \frac{1.25\ (1-S_L)\ (Ca/S) X_S}{X_{Ca}} + X_A (1-S_A) - X_A (1-S_C) + X_{VC} S_C e_2 + \frac{1.25\ X_{VL} S_L (Ca/S) X_S e_3}{X_{Ca}} \right] \right\} = F_B \tag{4}$$

Substituting equation (4) into equation (2) yields:

$$X_{OA} = \left\{ \frac{X_A S_{OA}}{\left[1 + \frac{1.25\ (Ca/S) X_S}{X_{Ca}} + \frac{.725\ (Ca/S) X_S S_L X_{CaCO_3} e_1}{X_{Ca}} \right] - \left[(1-S_C) + \frac{1.25\ (1-S_L)\ (Ca/S) X_S}{X_{Ca}} + \left(X_A (1-S_A) - X_A (L-S_C) \right) + X_{VC} S_C e_2 + \frac{1.25 X_{VL} S_L (Ca/S) X_S e_3}{X_{Ca}} \right]} \right\} \tag{5}$$

Equation (5) expresses the concentration of oversized ash in the solids output stream entirely as a function of assayed values of coal and limestone, except for the Ca/S molar feed ratio and the extent of reactions, e_1, e_2 and e_3. The Ca/S ratio is an operating condition and the extent of reactions are results of the operating conditions.

This analysis can be applied to the FBM system using Sewickley coal and Greer limestone. Chemical analyses and size analyses for these materials are shown in Tables 2.12 and 2.13 respectively.

Coarse particles have been taken as +40 mesh. Past bed and flyash size analyses have indicated this to be the approximate split point between elutriating and non-elutriating partiating particles at typical operating superficial gas velocities.

As stated previously, +8 mesh defines the size of the oversized particles.

From Table 2.12 and 2.13, the constants in equation (5) can be calculated as:

X_{VC} = .828

X_S = .044

X_{VL} = .35

X_{CaCO_3} = .75

$X_{Ca\frac{1}{2}}$ = .30

X_A = .172

S_C = .892

S_{OA} = .232

S_L = .98

S_A = .592

The Ca/S molar feed ratio and the extend of reactions (e_1, e_2, and e_3) can be determined from experimental results. Using the results of Test 621, a carefully monitored test of ten hours of steady-state operation (PER Monthly Progress Report No. 32), the Ca/S molar feed ratio was 2.5 and the extent of reaction completion for materials whose particle residence time greatly exceeded the gas residence time were as follows:

e_1 = calcium utilization ($CaCo_3 \rightarrow CaSo_4$) = 45%

TABLE 2.12

Greer Limestone Chemical Analysis

Constituent	Percent by Weight
$CaCO_3$	75.0
$MgCO_3$	4.0
Fe_2O_3	0.75
Al_2O_3	3.3
SiO_2	9.5
S	0.3
L.O.I.	37.4

Sewickley Coal Ultimate Analysis

As Fired Analysis	Percent by Weight
Moisture	1.40
Carbon	66.61
Hydrogen	4.73
Nitrogen	0.92
Chlorine	0.06
Sulfur	4.40
Ash	17.20
Oxygen (diff)	4.68
	100.00

TABLE 2.13

Size Distribution of Sewickley Coal, Coal Ash, and Greer Limestone

Sieve Size, U.S.S.	Sewickley Coal	Ash	Greer Limestone
- ½"+¼"	1.27	.72	
- ¼"+6"	20.94	13.65	.33
- 6+8	15.78	8.8	1.11
- 8+10	7.74	4.23	48.32
- 10+12	7.56	4.34	
- 12+14	6.66	3.10	42.66
- 14+16	6.17	3.86	
- 16+18	6.68	2.85	4.52
- 18+20	4.55	4.27	
- 20+25	3.70	3.45	.71
- 25+30	3.60	3.78	
- 30+40	4.54	6.10	.57
- 40+80	5.61	13.58	1.78
- 80	5.20	27.19	

e_2 = coal combustion = essentially 100%

e_3 = limestone calcination = essentially 100%

Applying these values to equation (5) the resulting oversized ash concentration is

X_{OA} = .08

This indicates that when Greer limestone and Sewickley coal are fed in a Ca/S molar ratio of 2.5, the oversized ash will build up to a maximum of 8 percent.

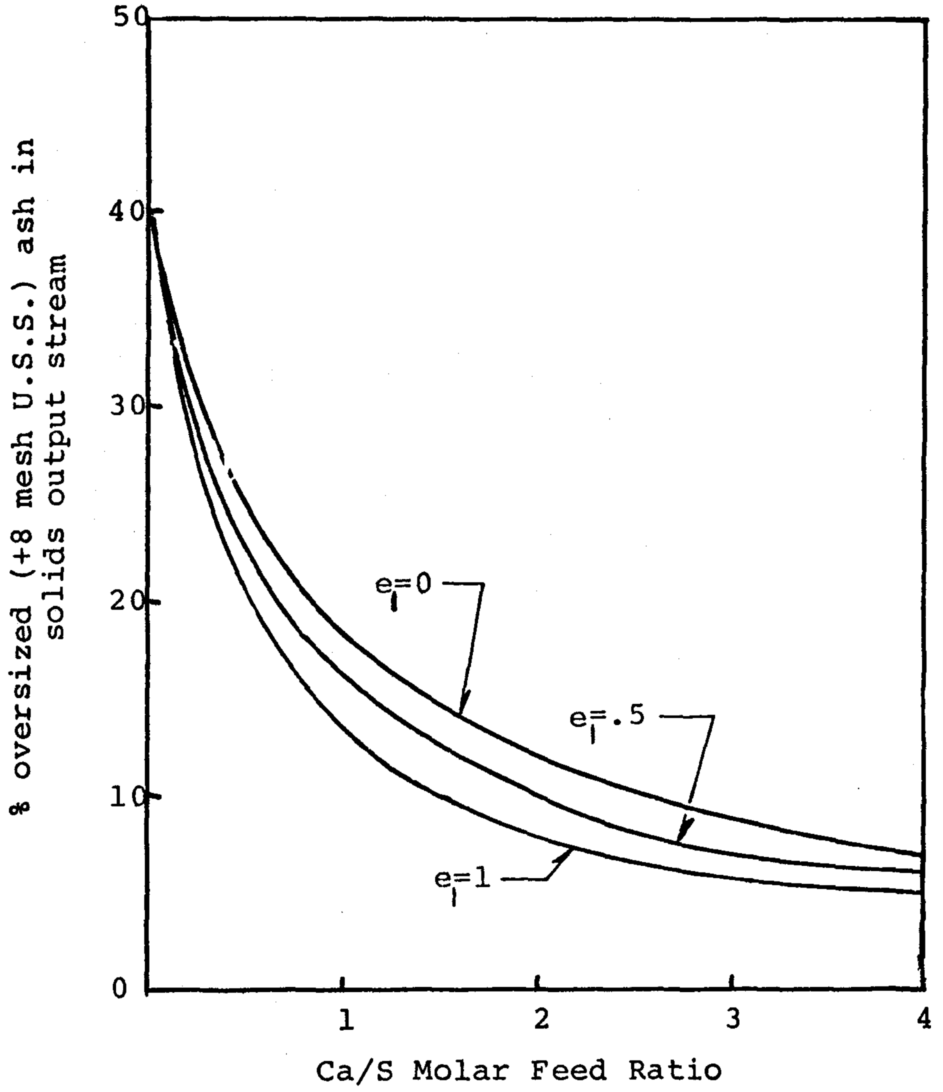
The oversized ash concentration can be calculated for a range of Ca/S molar feed ratios by determining the effect of Ca/S molar feed ratio on the extent of the various reactions.

For all reasonable operating conditions the rate of reaction of coal combustion and limestone calcination is very fast as evidenced by the consistently low (less than 1 percent) carbon content of bed samples. This has been found for a wide-range of Ca/S molar feed ratios. On the other hand, the extent of calcium utilization, e_1, is very dependent on the Ca/S molar feed ratio, as well as a large number of other parameters. (i.e., bed temperature, bed depth, excess O_2, type of limestone, catalyst addition, etc.)

Since the extent of calcium utilization is dependent on so many parameters, in order to establish an accurate value, it would have to be determined experimentally for each condition. However, it can be shown that the extent of calcium utilization does not significantly effect the oversized ash concentration and may therefore be estimated without invalidating the materials balance.

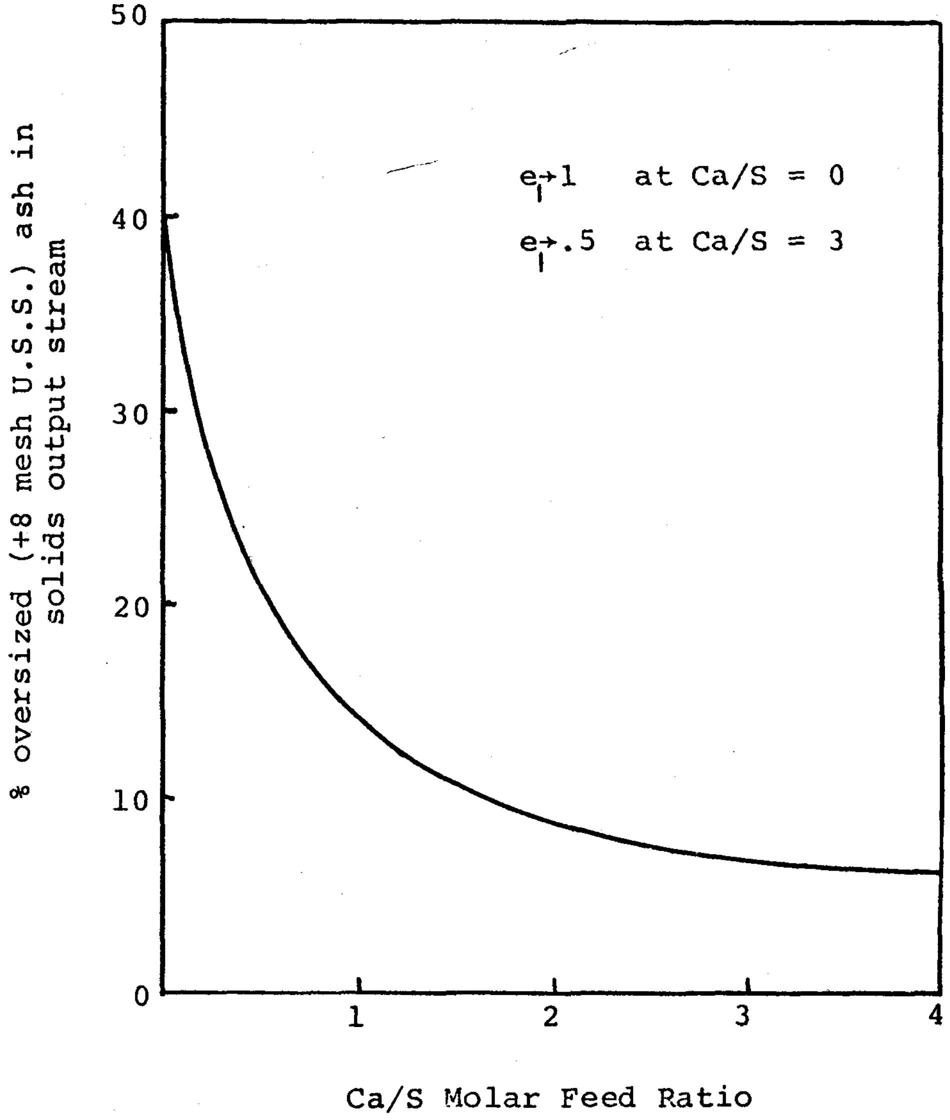
Using the previously determined values for Greer limestone and Sewickley coal, the oversized ash concentration was calculated for a range of Ca/S ratios and e_1 values ranging from 0 to 1 (calcium utilization of 0 percent to loo percent). The results are shown graphically in Figure 2.34. The upper and lower curves represent an envelope of possible calcium utilizations. It can be seen that for any Ca/S molar feed ratio the extent of calcium utilization does not significantly effect the oversized ash concentration. The three curves can be used to construct a typical oversized ash versus Ca/S molar feed ratio relationship.

As stated earlier, for a given set of operating conditions, the actual extent of calcium utilization will vary with the Ca/S molar feed ratio. At very low Ca/S feed ratios, the calcium utilization is very high (e_1 1) and at Ca/S ratios of around 3, the calcium utilization is typically around 50 percent (e_1 0.5). Therefore, the curves from Figure 2.34 can be used to construct a more realistic curve of oversized ash concentration vs. Ca/S molar feed ratio by following the curve of e_1 = 1 at very low CalS ratios and e_1 = 5 at Ca/S 3, Figure 2.35. It can be seen



THEORETICAL MAXIMUM OVERSIZED ASH CONCENTRATION IN THE BED AS A FUNCTION OF CA/S MOLAR FEED RATIO BASED ON SEWICKLEY COAL AND GREER LIMESTONE

FIGURE 2.34



THEORETICAL MAXIMUM OVERSIZED ASH CONCENTRATION IN THE BED AS A FUNCTION OF CA/S MOLAR FEED RATIO BASED ON SEWICKLEY COAL AND GREER LIMESTONE

FIGURE 2.35

that for Ca/S molar feed ratios of 2 or more, the fraction of oversized ash in the outlet stream is less than 10 percent.

It should be noted that the assumption of a CSTR type of reactor implies that the product stream has the same composition as the materials within the reactor. This is consistent with the fluidized bed vessel where the fluidized bed is well mixed and the outlet stream is of the same composition as the bed. Therefore, while this analysis was performed by determining an oversized ash concentration in the solids output stream, it is actually a calculation of the oversized ash concentration in the operating bed.

The preceding analysis was not based on a specific test rig and should therefore be applicable to any fluidized bed combustor. Different combustors may experience difference in calcium utilization due to variations in operating conditions, internals, etc.; however, the effect of calcium utilization e_1, on the fraction of oversized ash in the bed is small and the results would be similar regardless of the calcium utilization obtained. Since Sewickley coal and Greer limestone are expected to be used in the Rivesville MFB, Figure 2.35 can be assumed to be applicable to MFB operation if the assayed values in Tables 2.12 and 2.13 are valid. The chemical analyses should not change but the size analyses might, since a different type of crusher system will be used.

One factor which was not accounted for in the analysis is the possibility of particle attrition (reduction in particle size due to particle breakage in the fluidized bed).

Figure 2.35 indicated that at a typical Ca/S molar feed ratio of 3, the fraction of coal ash which can be removed from the bed by a classifier is approximately 8 percent. The effect of this nonsulfur sorbing fraction on the total bed sulfur sorption rate can be shown through the use of a mathematical model of the bed sulfur sorption process.

2.4.4 Application of Sulfur Sorption Process Model to FBC Design

The molar rate of sulfur capture per unit of bed, $dSo_2/A_n dT$, in units of lb moles/ft^2-hr, can be expressed in terms of the active sulfur sorbing area of limestone particles per unit area of bed A, and the average sulfur dioxide concentration in the bulk gas phase, c_b, by the following equation:

$$\frac{1}{A_b}\,\frac{dSO_2}{dt} = k^*AC_b \qquad (1)$$

where: k^* = the average effective reactivity of the sulfur sorbing area within the bed.

Units: $\dfrac{\text{moles } So_2 \text{ captured}}{ft^2 \text{ limestone/hr/} \quad \text{moles } SO_2 \,/\, ft^3 \text{ gas phase}}$

When discussing the effects of various coals or limestones on the sulfur capture, it is helpful to rearrange Equation 1 to separate the coal determined variables from the limestone determined variables. The required molar rate of sulfur capture is determined by the sulfur content and feed rate of the coal per square foot of bed, and the allowable emission limits for SO_2. The bulk gas phase concentration SO_2 is limited by the stoichiometry of combustion and the allowable emission limits for SO_2. Both of these terms are fixed for a specific coal and are, therefore, the coal derived variables. Rearranging Equation 1 to separate these variables gives:

$$\frac{1}{C_b A_b} \frac{dSO_2}{dt} = \text{constant} = k^*A \qquad (2)$$

Coal determined variables

Limestone determined variables

It can be seen from Equation 2, that it is necessary to provide active limestone surface area of adequate average reactivity in order to achieve the required sulfur capture performance of the system.

There are three means of enhancing the amount of active sulfur sorbing limestone particle surface area per square foot of grid area.

1. Increase the height of the bed at constant bed expansion. This increase of bed weight leads to a direct increase in the bed pressure drop which slightly reduces overall plant efficiency.

2. Decrease the particle radius at constant bed pressure drop. This increases the surface to mass ratio of the limestone at constant limestone mass; however, significant decreases in particle size lead to increased elutriation of the smaller particles.

3. Increase the fraction of limestone particles in the bed at constant pressure drop and constant bed expansion. This is accomplished by two types of bed cleansing operations: (1) elutriation of fine particles of coal derived materials and (2) physical removal of coal derived ash particles, in particular by a particle size classifier.

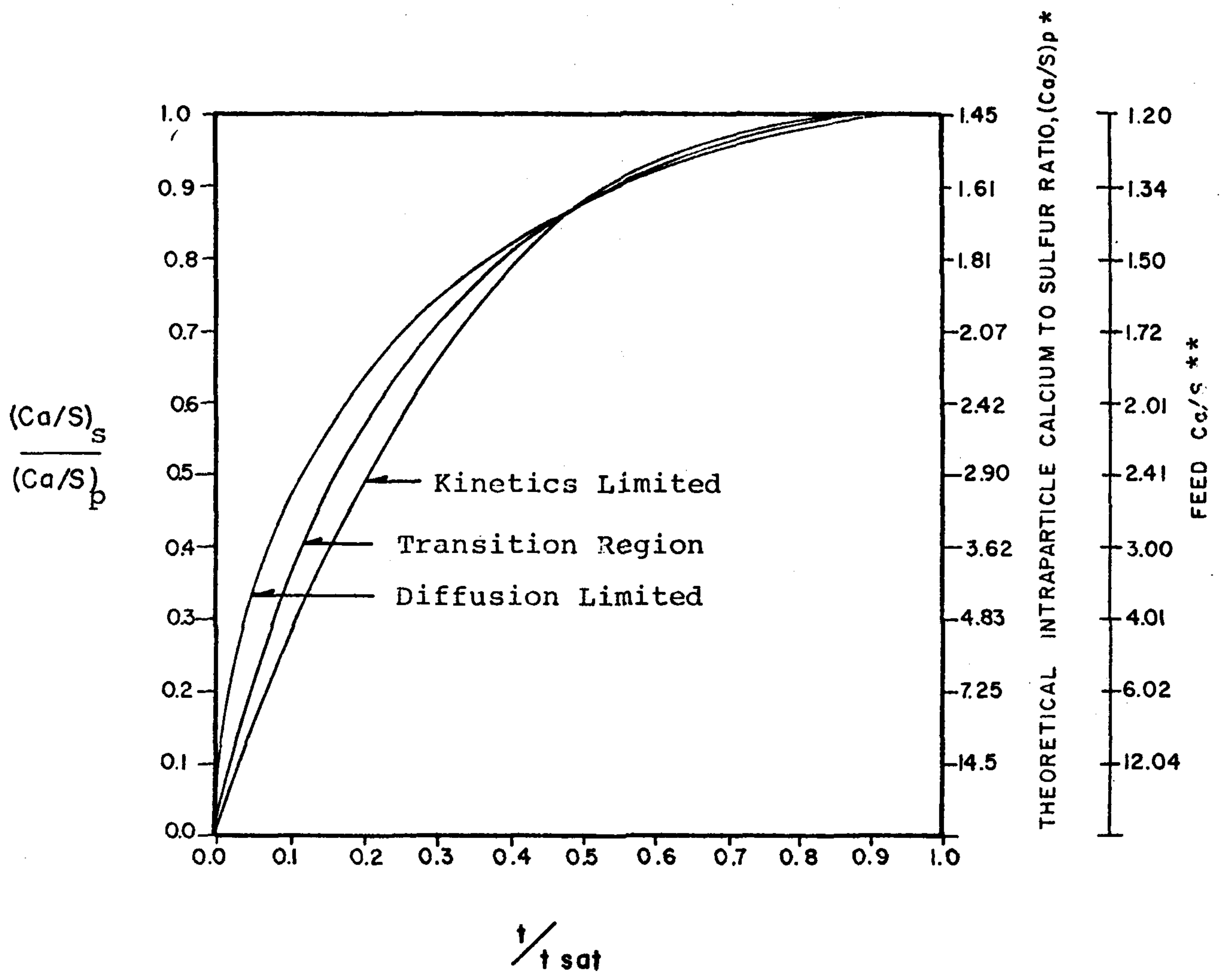
The Average Reactivity of Limestone Surface

In order to estimate the effect the average reactivity of the limestone surface has upon the rate of reaction, it is necessary to make several assumptions which define the system. In the present discussion, a first approximation assumption will be made that an "average" particle can be defined which has an "average" residence time in the bed, an "average" particle radius and an "average" reactivity of the sulfur sorbing surface. It has also been assumed that the shrinking core model of heterogeneous gas solid reactions applies, and that the external mass transfer resistance is very low compared to diffusion within the particles.

These assumptions allow the calculation of the extent of the reaction versus time. The shrinking core model predicts that given a sufficiently long reaction time t_s, the reaction for any particle will go to completion yielding a saturated internal particle calcium to sulfur ratio of $(Ca/S)_s$. An earlier analysis predicted that the calcium to sulfur ratio will be about 1.45 at saturation. The time to saturation of the particle varies proportionally to r_s/k for chemical reaction control and $r_s 2/d_e$ for diffusion through the particle shell control, where r_s is the particle radius, k is the first order kinetic rate constant, and D_e is the effective diffusivity of SO_2 through the sulfated shell. The dimensionless term $kr_s/6D_e$ defines the controlling resistance. A low calculated value implies the dominance of the chemical reaction step while a high value implies dominance of the intra-particle diffusion step.

Figure 2.36 shows the relationship between the internal particle molar calcium to sulfur ratio, $(Ca/S)_p$, and the time of reaction. The three lines plotted in Figure 2.36 are for the cases of chemical reaction kinetics limitation, diffusion through the product shell limitation, and for a transition region where both effects are important. It can be observed from Figure 2.36 that in all cases, the reaction is initially very rapid. In 10% of the saturation time, the reaction is between 27% and 47% complete. At 50% of the saturation time, the reaction is 87% complete.

The reactivity of the limestone surface area is proportional to the slope of the curves in Figure 2.36. It can be seen that with a Ca/S molar feed ratio of 3.0, at an 83% sulfur capture efficiency, the internal particle calcium to sulfur ratio will be 3.62 and the average residence time will be between 7% and 15% of the saturation time. The average slope of the line up to that point (proportional to the average rate of reaction) is between 5.7 for diffusion control (i.e., 04./.07) and 2.7 for kinetics control. If it were desired to feed limestone at a Ca/S feed ratio of 2.0 the average reactivity would be reduced



* 1.45 corresponds to theoretical saturation, $(Ca/S)_s$

** Intraparticle sulfur is related to feed sulfur assuming 83% sulfur capture.

EXTENT OF REACTION VERSUS TIME OF REACTION

FIGURE 2.36

3.1 and 2.4 respectively. This would require a threefold increase in the residence time of any particle in the bed for a 50% increase in sulfur captured per particle and a 33% decrease in limestone requirements.

A threefold increase in residence time could be accomplished by a number of methods; increasing the bed height at constant expansion by threefold; or increasing the bed area by a factor of three at constant bed height and Btu output; by a combination of bed cleansing and an increased bed height such that the total limestone surface area per square foot of grid is tripled.

It can be seen that the reactivity is closely related to the particle size, internal microstructure, and internal microchemistry. The inherent reactivity of the particle surface can be enhanced in three ways:

1. Reduction of particle size which reduces the diffusion resistance and increases the surface to volume ratio, but could lead to increased elutriation.
2. Use of an additive. NaCL additions in small amounts have been found to increase the inherent reactivity resulting in about a 1/3 savings in limestone requirements.
3. Preprocessing. There are some indications in the literature that calcining limestone in a 60% CO_2 atmosphere increases reactivity of the resulting sulfur sorbing particle.

Predicted Increased Calcium Utilization by Bed Cleansing

The classifier system, of interest in this analysis, is one which continuously screens a recycled side stream to remove oversized particles. These oversized particles have been found to be almost exclusively coal derived ash or debris. Since coal derived ash does not participate in sulfur capture, the selective removal of it should increase the sulfur capture efficiency of the bed by allowing higher bed concentrations of sulfur sorbing limestone. The following is an analysis of the possible expected enhancement of sulfur capture by the bed through the use of such a bed classifier system.

Size distribution information for Sewickley coal, coal derived ash, and Greer limestone is given in Table 2.13. From this table, it can be seen that while only 1.44% of the limestone is +8 mesh U.S.S. sieve size, 37.99% if tge ciakm as fed, and 23.17% of the resulting coal derived ash will be retained on an 8 mesh screen.

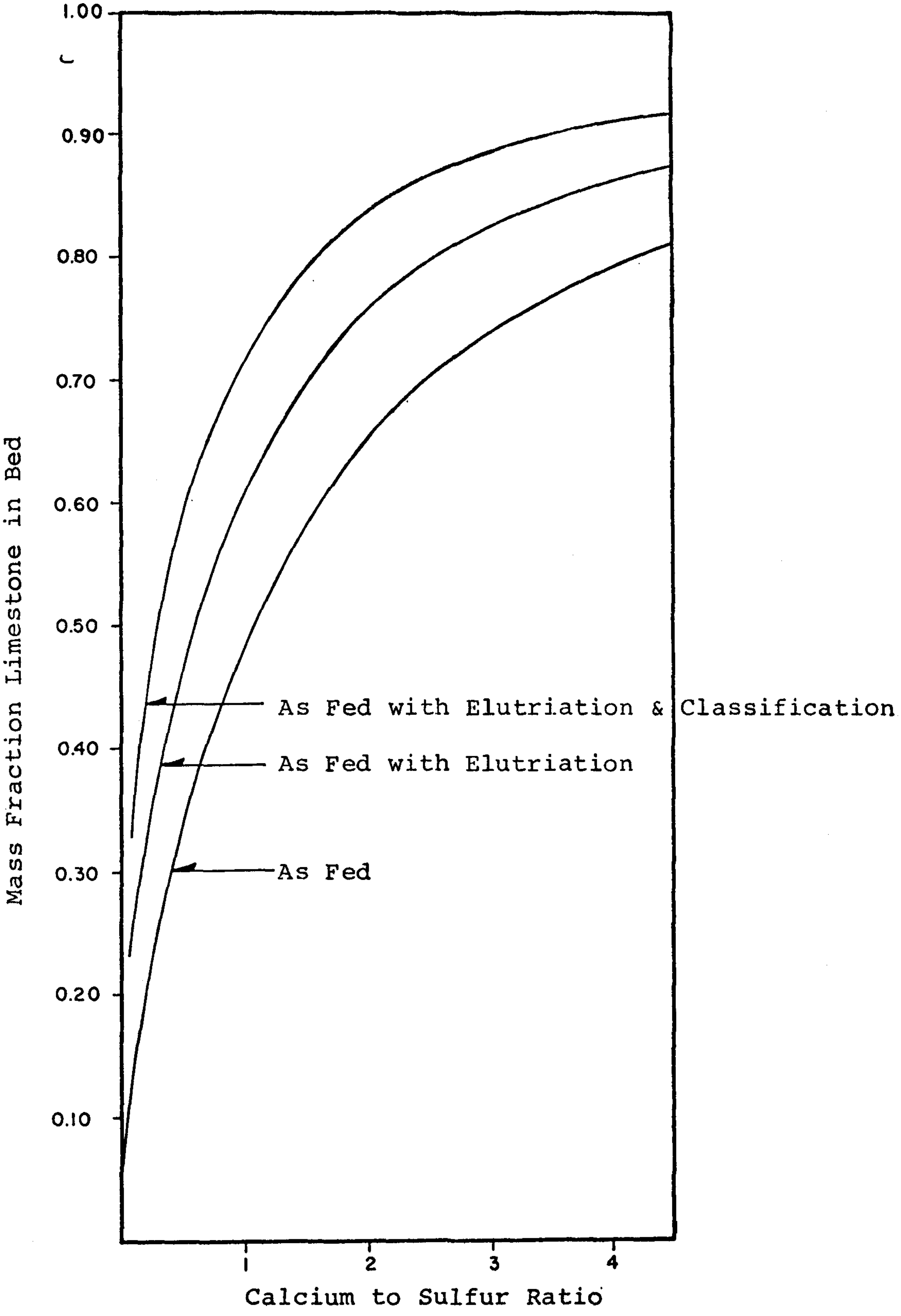
This would result in the limestone being slightly less reactive, due to the longer reaction time. However, the increased residence time will result in better calcium utilization. Figure 2.38 shows the effect of the increased residence time, Δt, on the feed Ca/S ratio. In this example, the classifier operation results in an 8.2% decrease in calcium requirements for the case of chemical kinetics limiting the reaction. For the case of diffusion limiting the reaction, the calcium requirements would be decreased by only 4.0%.

In order to duplicate the effect of 100% efficient classifier, the total amount of limestone sulfur sorbing area per square foot of grid area could be increased by a factor of 88.5/82.5 at a calcium to sulfur ratio of three. This could be accomplished by increasing the bed depth at constant bed expansion by the same factor (7.3%) and eliminating classifier operation. Thus, for a four foot expanded bed depth, the action of a classifier could theoretically be duplicated for Sewickley coal and Greer limestone by increasing the depth of the bed at constant expansion by about 3.5 inches. This would also result in incrementally longer residence times for the limestone which would then capture incrementally more sulfur resulting in about a 4.0% decrease in limestone feed rate if diffusion is limiting, or an 8.2% decrease in limestone requirements if chemical reaction controls the kinetics at a 3.0 Ca/S ratio.

In summary, the operation of a classifier has relatively minor effect on the limestone economy of the fluidized-bed system when operating at typical MFB design conditions. The effect of bed classifying on limestone economy for different design operating conditions and with other limestones must be evaluated for each specific case. Refer to Figure 2.37.

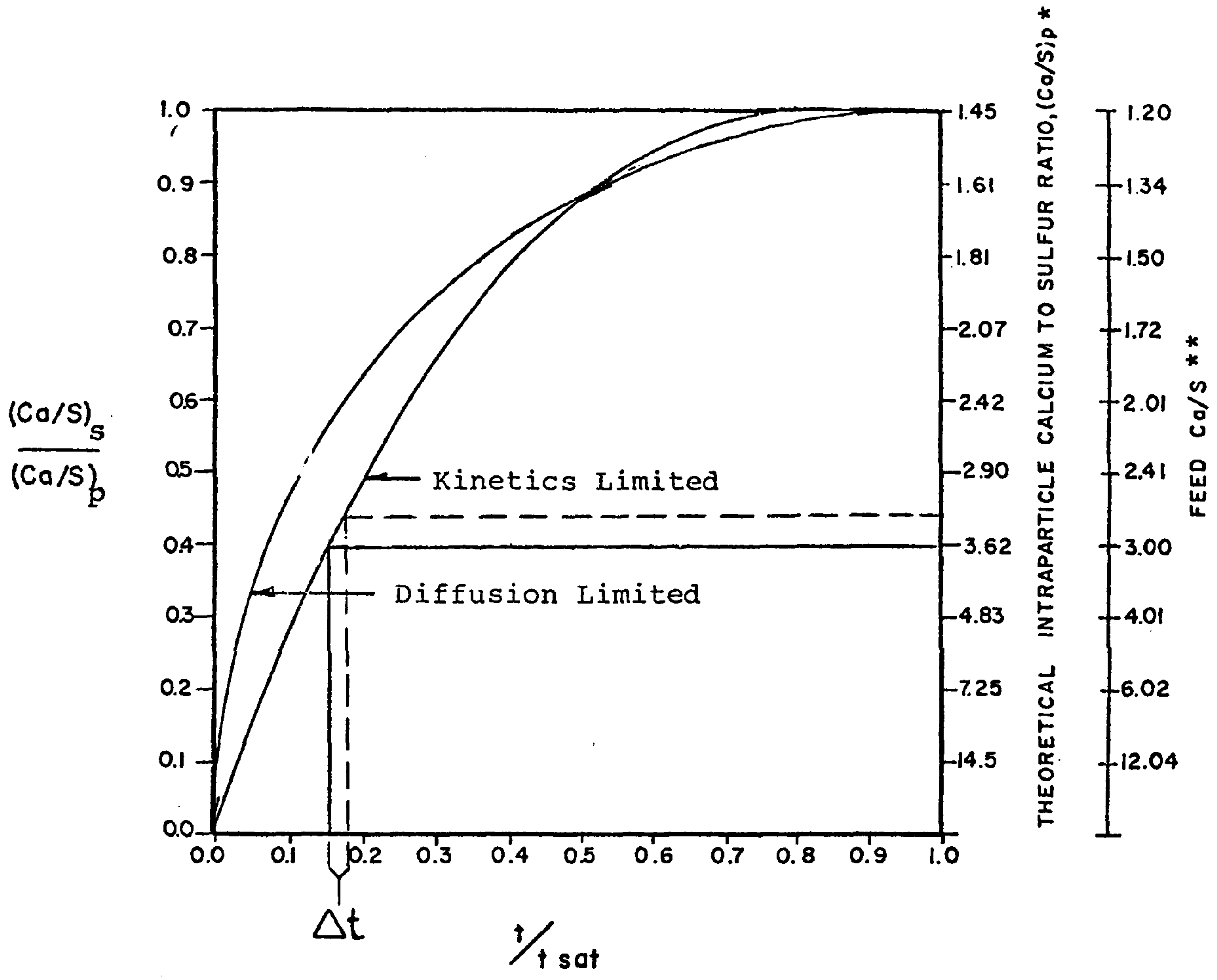
Predicted Effect of Increasing Bed Height

An estimate of the effect of increasing the height of the bed at constant bed expansion can be obtained from the kinetic rate curves of Figure 2.36. If the bed height is increased at constant Btu output and at constant bed expansion, the weight of the bed increases almost proportionally. Calculations from the preceding model indicate a significant decrease in limestone feed requirements to obtain equal sulfur capture efficiencies when the bed height is increased. This decrease in calcium to sulfur feed ratio results in a proportional increase in the reaction time for each particle which is partially offset by the increased dilution of the bed by coal derived ash.



MATERIAL BALANCES SHOWING CLASSIFIER PERFORMANCE

FIGURE 2.37



* 1.45 corresponds to theoretical saturation, $(Ca/S)_s$

** Intraparticle sulfur is related to feed sulfur assuming 83% sulfur capture.

EXTENT OF REACTION VERSUS TIME OF REACTION

FIGURE 2.38

Also, it can be noted that 1.78% of the limestone and 40.77% of the coal derived ash is -40 mesh.

In order convert these numbers into their respective contributions to the mass of the fluidized bed, changes in composition must be calculated. For this, it is assumed that the coal is 17.2% ash and 4.40% sulfur, whereas the limestone contains 30% calcium and 28% CO_2 which is lost upon calcicination being replaced by exactly half as many moles of SO_3.

With this information, a material balance which shows the effect of classifier operation can be derived. Figure 2.37 shows this material balance for three cases:

Case 1. As fed.

This curve shows the mass fraction of limestone for a range of calcium to sulfur feed ratios, that would exist in the bed if elutriation of the fines did not occur. This curve, while unattainable in practice, represents the minimum limestone mass fraction for a given calcium to sulfur feed ratio.

Case 2. As fed with elutriation.

For the calculation of this curve, it was arbitrarily assumed that the -40 mesh material elutriates carrying with it 40.77% of coal derived ash and 1.78% of the limestone derived material. The selective removal of the coal derived ash in this case has resulted in an increase in the mass fraction of limestone in the bed. This curve would more typically represent an operating fluidized bed with no classification.

Case 3. As fed with elutriation and classification.

This curve represents the mass fraction of limestone in the bed when the -40 mesh material elutriates and the +8 mesh fraction is totally removed by a continuous creening process.

It can be seen from Figure 2.37, that with a calcium to sulfur ratio of three, about 74% of the resulting bed mass would be limestone derived prior to elutriation. With elutriation of -40 mesh fines about 82.5% of the bed mass is limestone derived. This increases to about 88.5% by operation of a perfect classifier.

Figures 2.36 and 2.37 can be used to predict the effect of the classifier on calcium utilization in the following manner. At a Ca/S ratio of 3.0 we can see that the residence time of the limestone in the bed will increase by a factor of 88.5/82.5, or 7.3%.

This dilution of the bed can be obtained from the middle curve of Figure 2.37. Figure 2.39 shows the results of calculations combining these three effects for two cases. The cases presented are based upon a 4 foot bed depth with limestone feed rates of 3.0 and 4.0 molar calcium to sulfur ratios, respectively. Also shown, are the results of theoretical calculations, assuming diffusion limited kinetics and chemical reaction limited sulfur capture kinetics. It has not been determined which represents the actual kinetics, but diffusion limited kinetics are suspected to predominate. Also it is assumed that 83% of the feed sulfur is captured by the limestone in all cases.

It can be seen that considerable savings in limestone requirements can result from even small increases in bed height at constant expansion. Large increases in bed height may reduce the limestone requirements to near the theoretical minimum. This is done at an increase in overall pressure drop of the system which results in slightly lower overall electric power generation efficiency and perhaps, some additional capital costs. Nevertheless, bed height remains significant variable to be optimized in the design of any actual system.

Conclusions from Application of Model

1. Increasing the bed height at constant expansion can theoretically reduce limestone requirements significantly.

2. In some cases, this reduction in limestone requirements can be accomplished by decreasing the particle radius accompanied by corresponding increases in grid surface area.

3. The third effective means of significantly reducing limestone feed requirements consists of increasing the inherent reactivity of limestone. One way to do this is to add an additive such as NaCL.

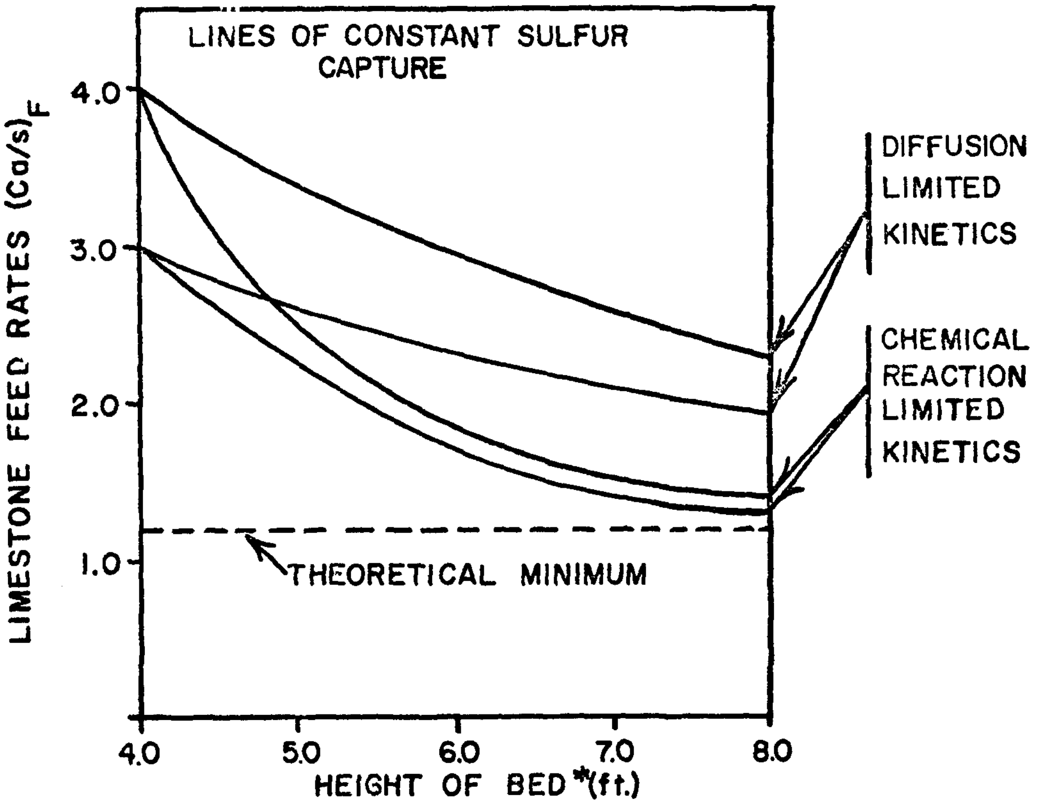
4. The operation of a bed classifier has a relatively minor effect on overall limestone economy at high calcium to sulfur feed ratios for the presently used size distributions. Therefore, in some cases it may be eliminated entirely.

2.5 Support Activities

2.5.1 MITRE Data Acquisition System

During the Reporting Period a computerized data acquisition system was installed at the Alexandria Laboratory. Equipment for the system was supplied by MITRE Corp.

THEORETICAL EFFECT OF BED HEIGHT ON LIMESTONE FEED REQUIREMENTS



*AT CONSTANT EXPANDED BED DENSITY AND CONSTANT STEAM OUTPUT

FIGURE 2.39

The Data Acquisition System shown in Figure 2.40 includes the following components:

NEFF System 620/Series 400 Multiplexer, including:
- 12-bit A/D converter
- Display control panel
- Nine (9) 16-channel differential multiplexer cards with ± 100v overload protection and screw terminal inputs
- Interface to PDP 11/04 computer

DEC PDP 11/04 computer with 8K core memory

DEC Real-time clock

Kennedy Model 9000 Tape Drive, with control unit for PDP 11/04

Electronic Development Corporation precision Voltage Source Model E100-E

Teletype Corporation ASR-33 Teletypewriter

Six (6) NEFF D/A converters, and computer interface card

Gould Model 2000 6-channel Strip Chart pen recorder

System Interface Panel (built and supplier by MITRE), including;
- phone jack input connections for 128 channels
- Lambda Electronics power supply Model LCS-D-24
- 110 ohm precision resitors

Figure 2.41 shows a schematic of the system.

A total of thirty pressure transmitters and differential pressure transmitters have been installed on the PDU and connected to the data acquisition system. The following measurements are recorded on magnetic tape by the system.

<u>Pressure Measurements</u>

1. Feedwater
2. Drum
3. Pressure drop across immersed bundle
4. Pressure drop from drum to mud drum
5. FD fan discharge
6. Inlet plenum
7. Lower bed
8. Freeboard
9. Pressure drop from plenum to feedboard
10. Pressure drop from lower bed to freeboard
11. Flue gas pressure below convection bundle
12. Flue gas pressure above convection bundle
13. ID fan inlet
14. Pressure drop across sample circulation blower
15. Dust collector inlet

MITRE COMPUTER EQUIPMENT

FIGURE 2.40

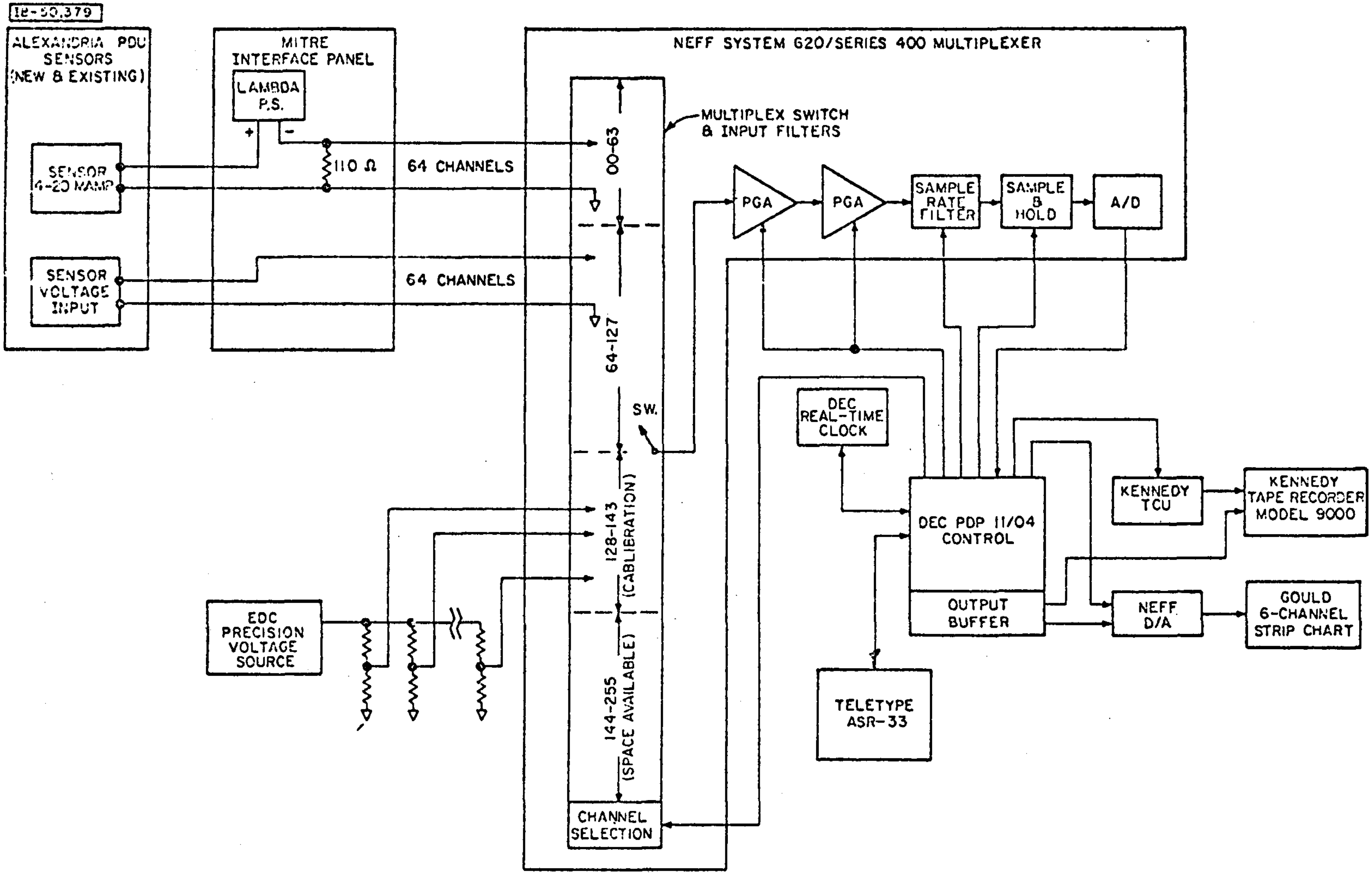


FIGURE 2.41

MULTI-CHANNEL DATA ACQUISITION SYSTEM INSTALLED AT ALEXANDRIA PDU

Flow Measurements

1. Water entering immersed bundle
2. Feedwater entering drum
3. Water leaving convection bundle
4. Water leaving door cooler
5. Water leaving support tubes
6. Stream leaving drum
7. Air entering plenum
8. Air flow to bag filter
9. Air to propane burner duct
10. Flue gas at ID fan inlet
11. Pneumatic transport air flow
12. Air flow to classifier
13. Air flow to separator (air preheater)

Temperature Measurements

1. Feedwater pump outlet
2. Auxiliary water supply
3. Immersed bundle outlet
4. Door cooler outlet
5. Support tube outlets (2)
6. Convection bundle outlet
7. Mud drum
8. Drum
9. Separator (preheater) outlet
10. FD fan inlet
11. Transport air
12. Air to propane duct burner
13. Plenum
14. Freeboard
15. Furnace outlets (2)
16. Flue gas entering convective heater
17. Flue gas leaving convective heater
18. Separator (preheater) inlet gas
19. Dust collector inlet
20. ID fan inlet
21. Classifier inlet
22. Classifier outlet

(In addition, there are several temperature measurement points in the immersed bundle, and in the fluidized bed.)

Other Measurements

1. Drum level
2. Classifier control signal
3. Coal feeder control signal
4. Limestone feeder control signal

Results of the test runs are available to the Alexandria staff in the form of printed raw data, printed reduced data, magnetic tape of raw data, and magnetic tape of reduced data.

2.5.2 By-Product Study

Shipments of FBM an CBC solid samples to other ERDA Contractors and outside facilities for use in applied research studies continued during the annual report period. Facilities to which samples were sent are as follows:

U.S.D.A. Beltsvile, Md.	11.7 tons sulfated bed material
U.S.D.A. Poultry Research Lab. Georgetown, Delaware	1.8 tons sulfated bed material
U.S.D.A. Valley Forge Lab. Valley Forge, Pa.	3 tons sulfated bed material
U.S.D.A. Agricultural Research Services West Virginia University Morgantown, W. Va.	5.2 ton sulfated bed material
Foster Wheeler Energy Corp. Livingston, New Jersey	10 lb. high carbon flyash
University of Denver Denver Research Institute Denver, Colorado	100 grams high carbon flyash
Dravo Lime Co. Research Center Neville Island Pittsburgh, Pennsylvania	50 lb. sulfated bed material
Dr. O.L. Bennett U.S.D.A. Agricultural Research Service Morgantown, W. Va.	15 lb. Greer Limestone
Ken Martira Lodge Cotterell Div. Dresser Industries Houston, Texas	15 lb. sulfated bed 15 lb. high carbon flyash 15 lb. low carbon flyash
Ralph Stone Co., Inc. Los Angeles, California	200 lb. sulfated bed material 250 lb. low carbon flyash

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

3.1 General

With completion of construction of the MFB Plant early in the reporting period, PER activities at Rivesville shifted from construction supervision and inspection to plant operation and maintenance. The MFB Plant Operations Group which was formally established in April 1976, was delegated the responsibility for initial start-up, operation, maintenance, repairs and modifications to the MFB unit and auxiliary systems and equipment.

3.2 Activities

During the reporting period, the Plant Operations Group engaged in the following major activities;

- Accepted, in behalf of ERDA, all MFB Plant equipment from Champion Construction & Engineering Co.,
- Tested MFB Auxiliary Systems in cold mode,
- Trained operations and maintenance personnel,
- Prepared MFB light-off procedure,
- Operated MFB Cells B, C and D for sustained periods of time,
- Debugged auxiliary equipment and performed maintenance and repair work necessitated by process problems or equipment failures.

3.3 Roster of Personnel

During the reporting period, the MFB Plant Operations Staff was increased to the level required for proper performance in the areas of designated responsibility. The Operations Staff comprised the following personnel:

Operations Manager	1
Assistant Operations Manager	1
Results Engineer	1
Maintenance Supervisor	1
Shift Supervisors	4
Control Room Operators	4
Equipment Operators	16
Laborers	3
Chemist	1
Plant Electrician	1
Field Accountant	1
Secretary	1

Maintenance Mechanics	3
Assistant Chemist	1
Assistant Results Engineer	1
Plant Instrument Man	1
Instrumentation Electrical Technician	1
Assistant Electrician	1
TOTAL	43

3.4 MFB Operation

During the reporting period, initial startup oil firing and subsequent coal light-off of Cell D was accomplished. Coal light-off of Cells C and B was also accomplished using the transfer gates between cells. The total MFB operating hours for this period were:

Oil Firing - 1024 hours
Coal Firing - 372 hours

3.4.1 Transfer of Material from Cell D to Cell C

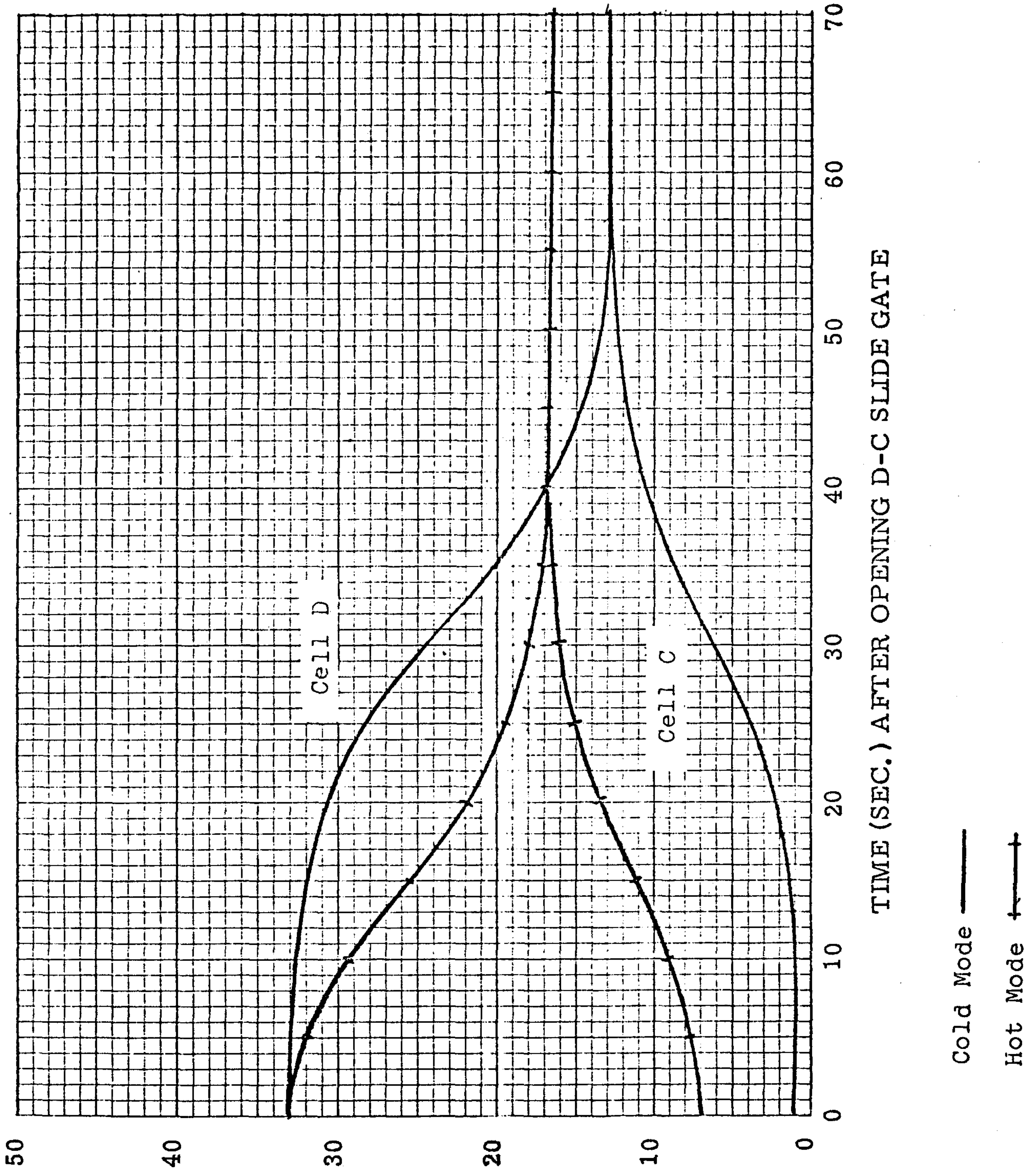
Test No. 1-77-2 was performed to determine the time required to transfer material from Cell D to Cell C. The test was run with a static bed height of 33" in Cell D and no material in Cell C. Cell D was fluidized, Cell C had appriximately 90,000 lbs/hr airflow and the slide gate was opened. The following data was recorded just prior to opening the slide gate and at ten (10) second intervals thereafter.

Low Bed Pressure - Cell D
Low Bed Pressure - Cell C
Pressure Differential - Annubar Cell D
Pressure Differential - Annubar Main Cells
Pressure Differential - Cell C
Pressure Differential _ Cell D

Test Results

As Figure 3.1 indicates, under a cold condition, the bed height in Cell D and cell C equalized approximately fifty (50) seconds after the slide gate was opened.

A similar hot mode test was run with a static bed height of 33" in Cell D and 7" in Cell C. As Figure 3.1 indicates, the bed height in Cell D and Cell C equalized approximately forty (40) seconds after the slide gate opened.



BED HEIGHT IN INCHES

EQUALIZATION OF CELLS D & C BED HEIGHT vs. TIME (SEC.)

FIGURE 3.1

3.4.2 MFB Light-off Procedure - Activity No. 1106

In order to standardize the light-off technique used for Cell D and to provide guidelines for the light-off of Cell C and subsequent light-off of Cell B and Cell A, a procedure was prepared for the complete light-off of the MFB unit.

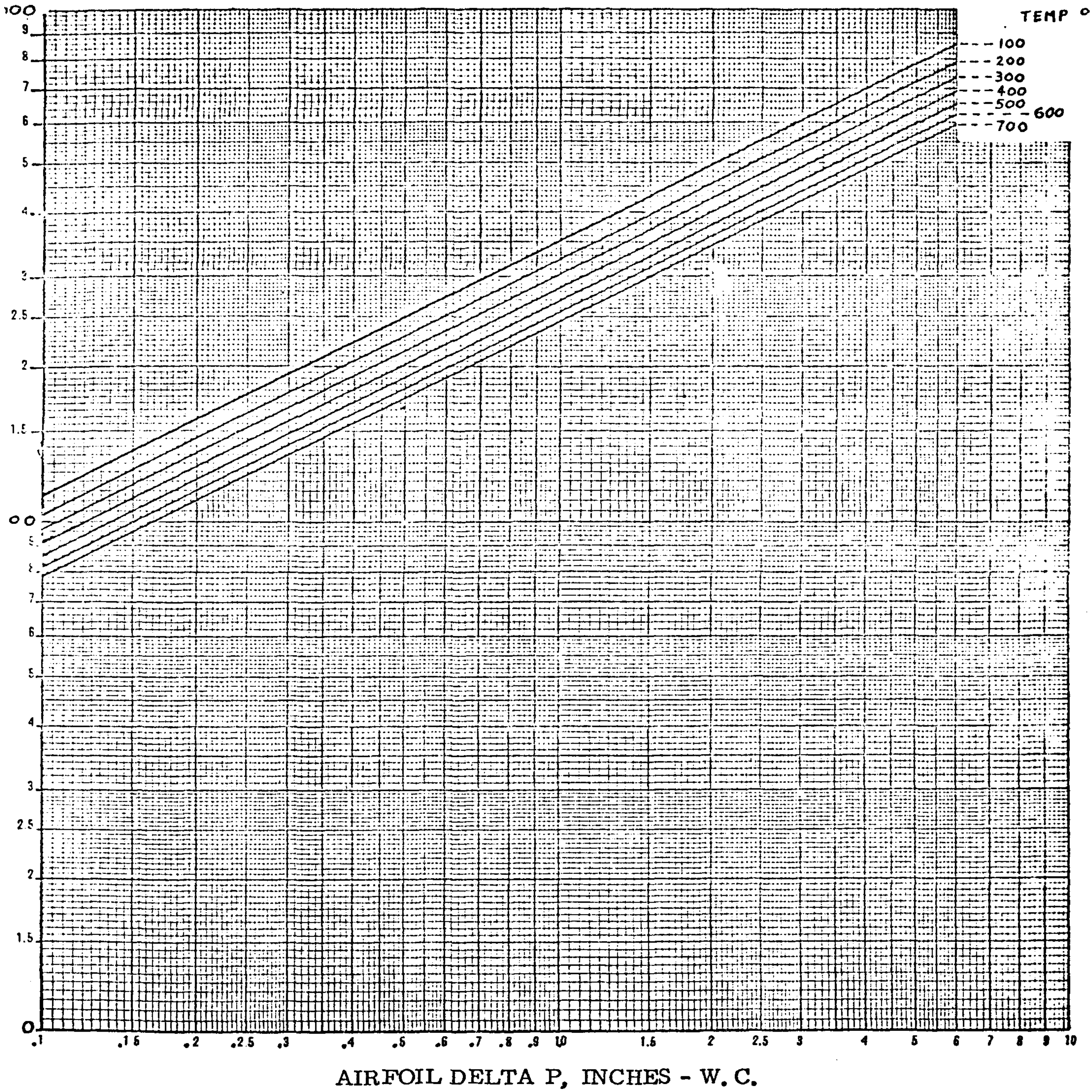
A series of tests were performed on the MFB unit and from the data obtained, a set of graphs were developed. On some tests, a pitot tube was used to check airfoil and annubar responses. Other tests gave relationships of airflow vs. grid plate differential pressure, material flow vs. feeder signal, air flow vs. damper position and water flow vs. valve position. These graphs and charts were included in the light-off procedure and are listed below:

Fig. 3.1-Equalization of Cells D & C Bed Height vs. Time
Fig. 3.2-Airflow vs. Airfoil Pressure Differential
Fig. 3.3-Main Cells Airflow vs. Main Annubar Delta P
Fig. 3.4-CBC Airflow vs. CBC Annubar Delta P
Fig. 3.5-Cell A Airflow vs. Gridplate Delta P
Fig. 3.6-Cell B Airflow vs. Gridplate Delta P
Fig 3.7-Cell C Airflow vs. Gridplate Delta P
Fig. 3.8-Cell D Airflow vs. Gridplate Delta P
Fig. 3.9-Cell A, B, C and D Coal Flow vs. Rotary Feeder Signal from Control Room
Fig 3.10-Cell A, B, C and D Limestone Flow vs. Rotary Feeder Signal from Control Room
Fig. 3.11-Static Bed Height vs. Fluidized Bed Pressure Drop
Fig. 3.12-Main Cell Airflow vs. Main Cell Airflow Scale
Fig. 3.13-CBC Airflow vs. CBC Airflow Scale
Fig. 3.14-Cell A Airflow vs. Damper Setting
Fig. 3.15-Cell B Airflow vs. Damper Setting
Fig. 3.16-Cell C Airflow vs. Damper Setting
Fig. 3.17-Cell D Airflow vs. Damper Setting
Fig. 3.18-Cell B Gridplate Delta P Required for Minimum Fluidization
Fig. 3.19-Cell C Gridplate Delta P Required for Minimum Fluidization
Fig. 3.20-Cell CBC Gridplate Delta P Required for Minimum Fluidization
Fig. 3.21-Equalized Bed Height for Cell C Light-off
Fig. 3.22-Equalized Bed Height for Cell B Light-off

3.4.3 MFB Light-off Summaries - Activity No. 1106

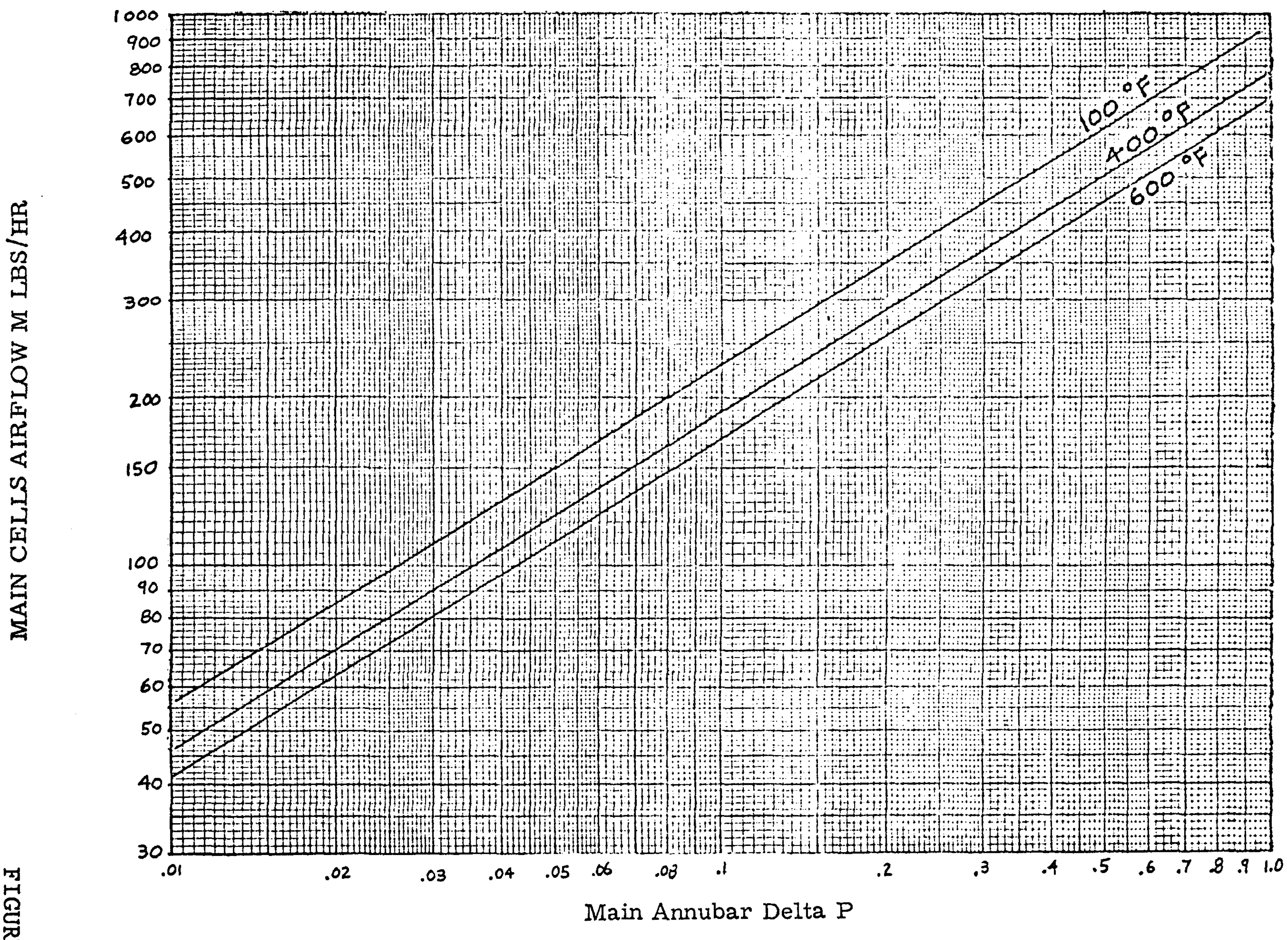
During this report period, final check-out, acceptance and commissioning of auxiliary equipment, systems, electrical interlocks and controls to permit initial light-off of the

5/6/77



AIRFOIL AIRFLOW MEASUREMENT

FIGURE 3.2



MAIN CELLS ANNUBAR DELTA P vs. AIR FLOW

FIGURE 3.3

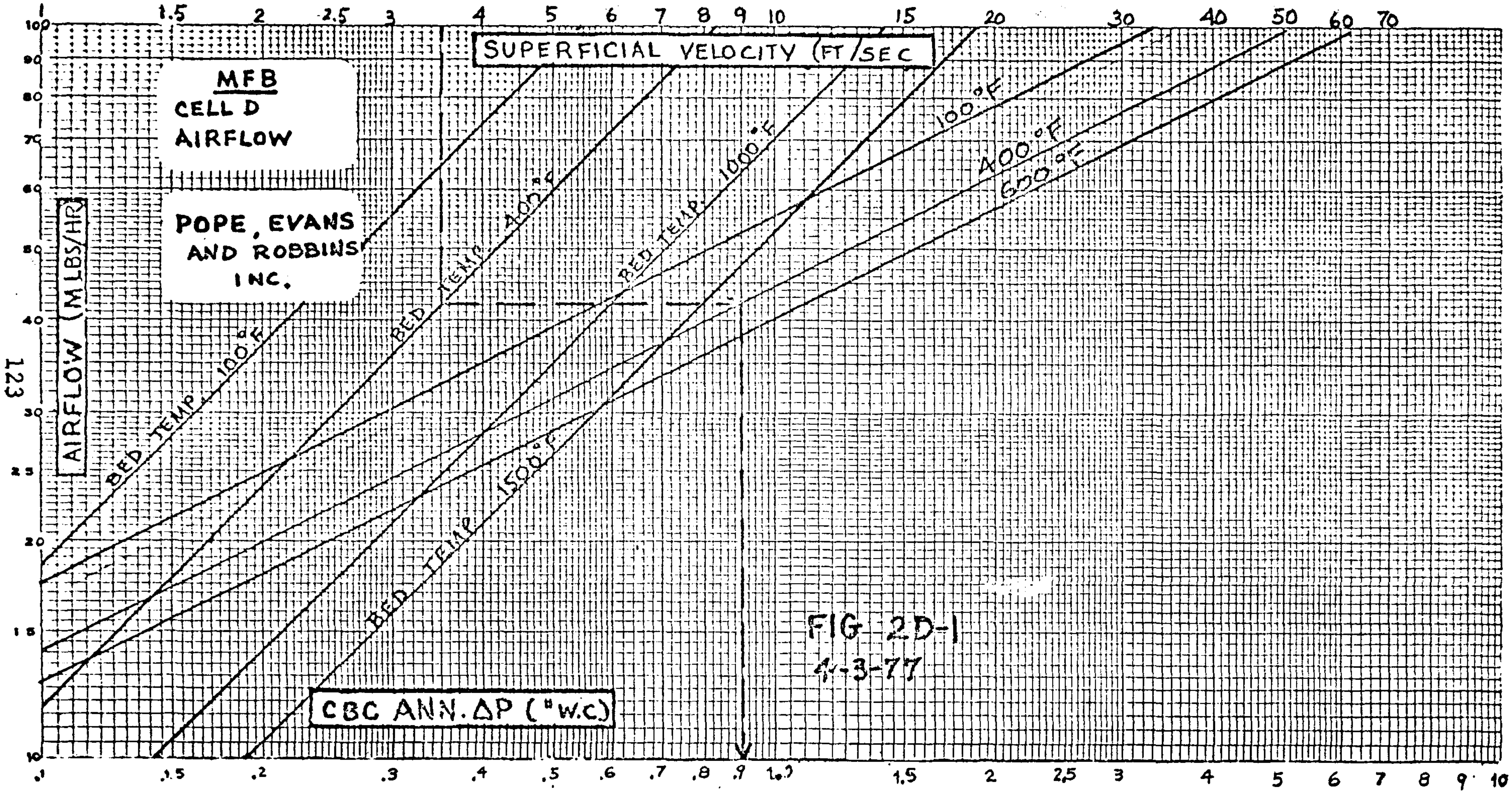
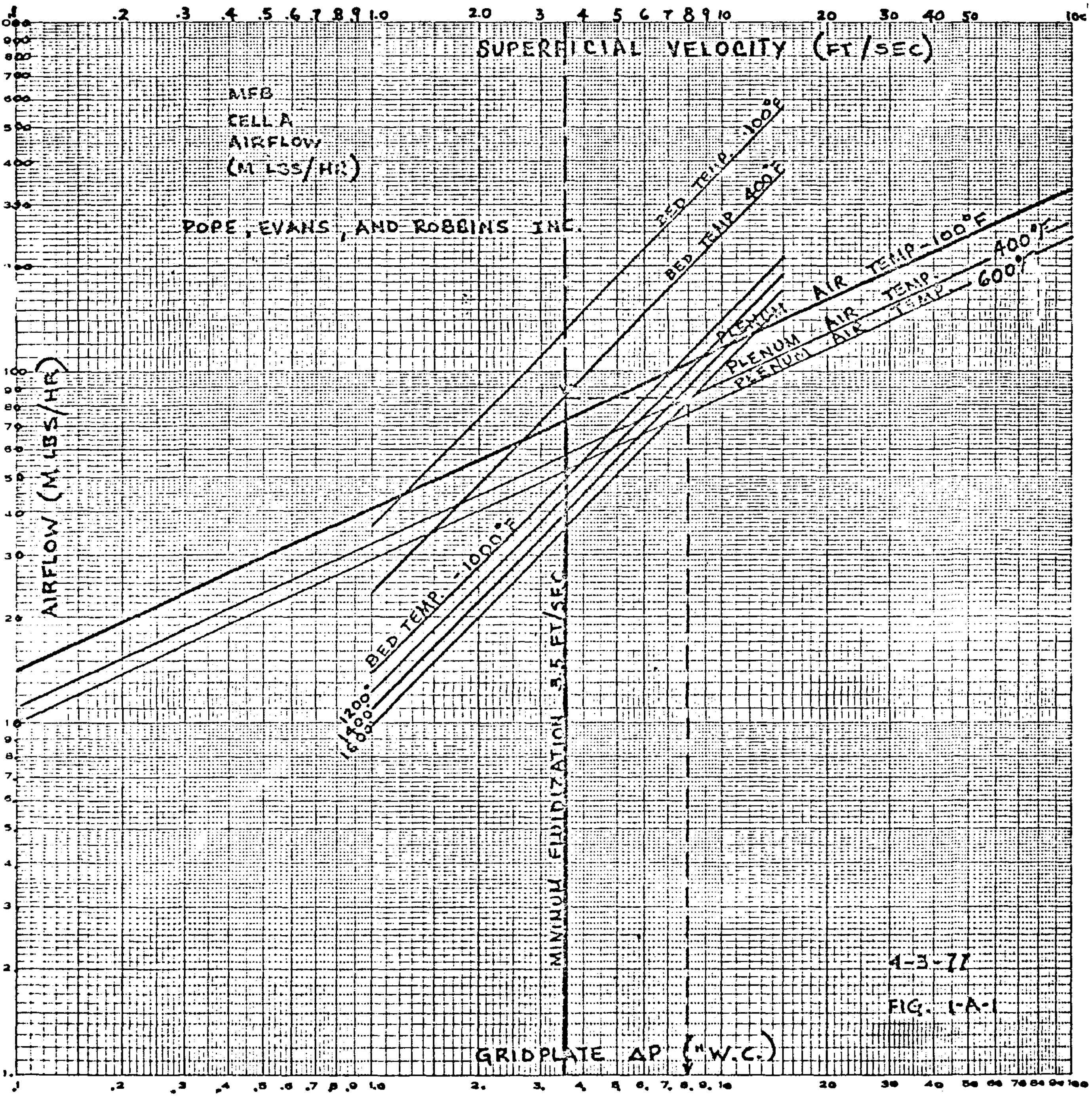
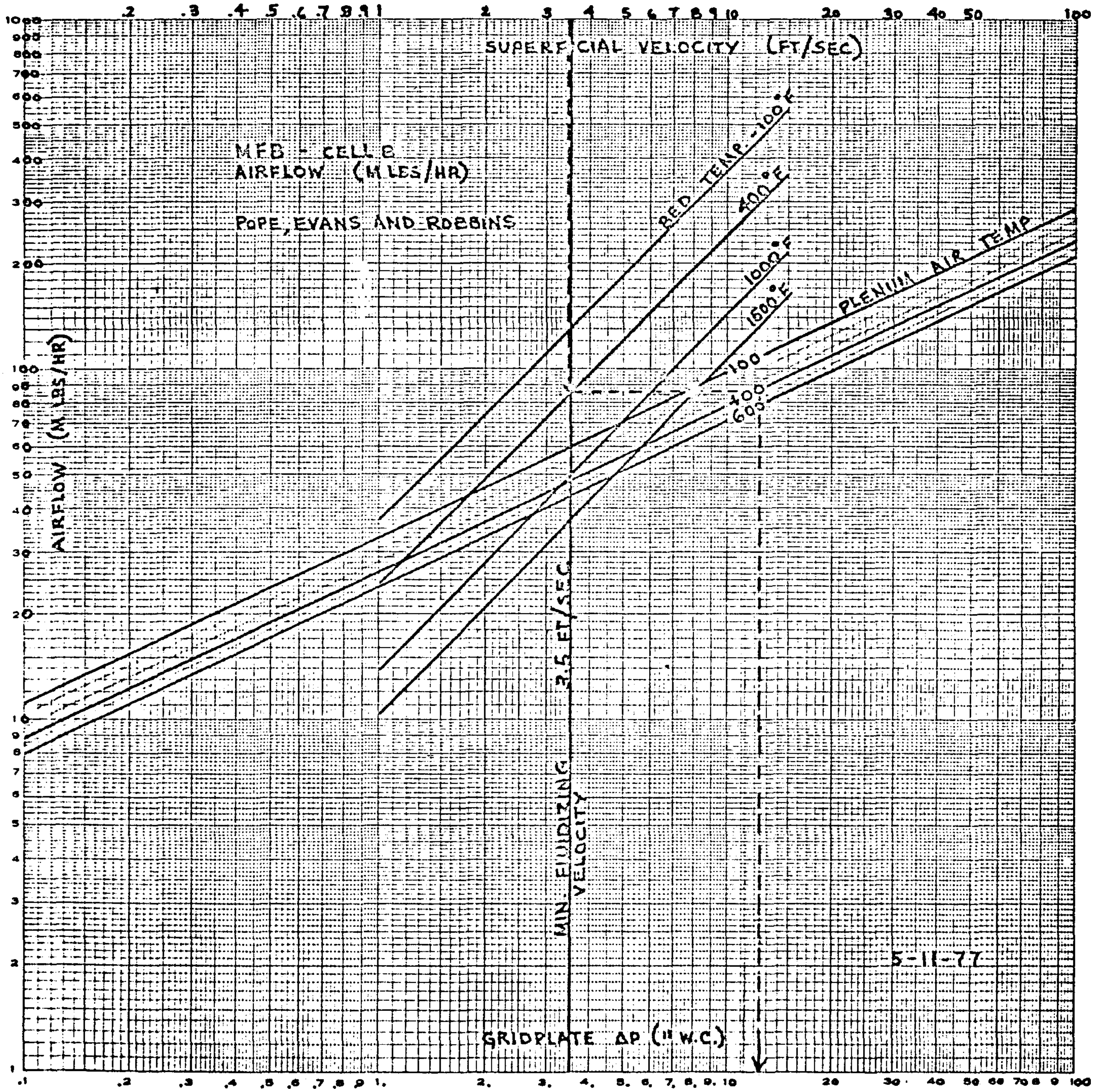

MFB
CELL D
AIRFLOW
POPE, EVANS AND ROBBINS INC.
SUPERFICIAL VELOCITY (FT/SEC
1
1.5
2
2.5
3
4
5
6
7
8
9
10
15
20
30
40
50
60
70
AIRFLOW (M LBS/HR)
100
90
80
70
60
50
40
30
25
20
15
10
BED TEMP. 100°F
BED TEMP. 400°F
BED TEMP. 1000°F
BED TEMP. 1500°F
100°F
400°F
600°F
FIG 2D-1
4-3-77
CBC ANN. ΔP ("W.C)
.1
.15
.2
.3
.4
.5
.6
.7
.8
.9
1.0
1.5
2
2.5
3
4
5
6
7
8
9
10

FIGURE 3.4

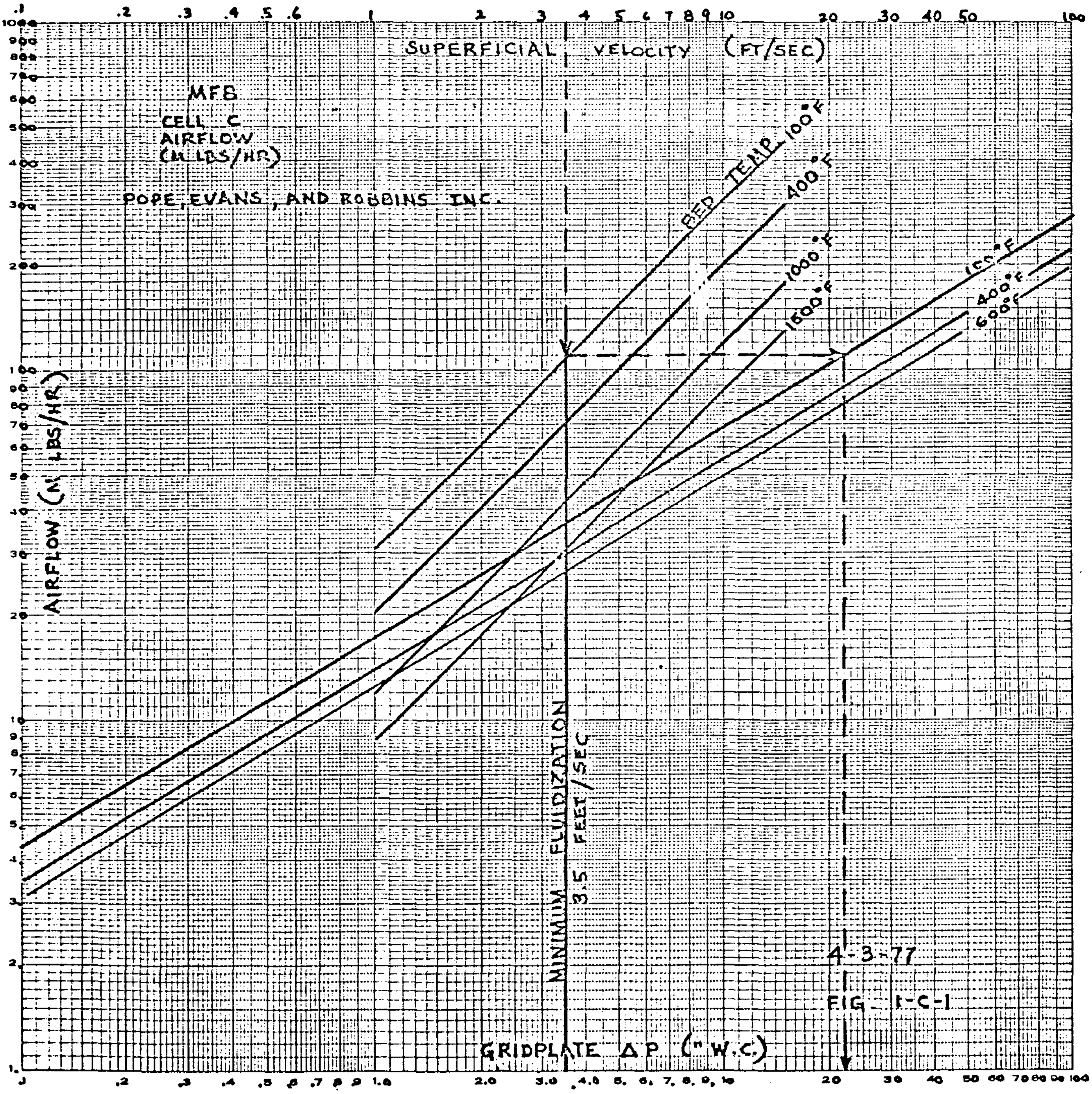


CELL A AIRFLOW vs. GRID PLATE DELTA P FIGURE 3.5



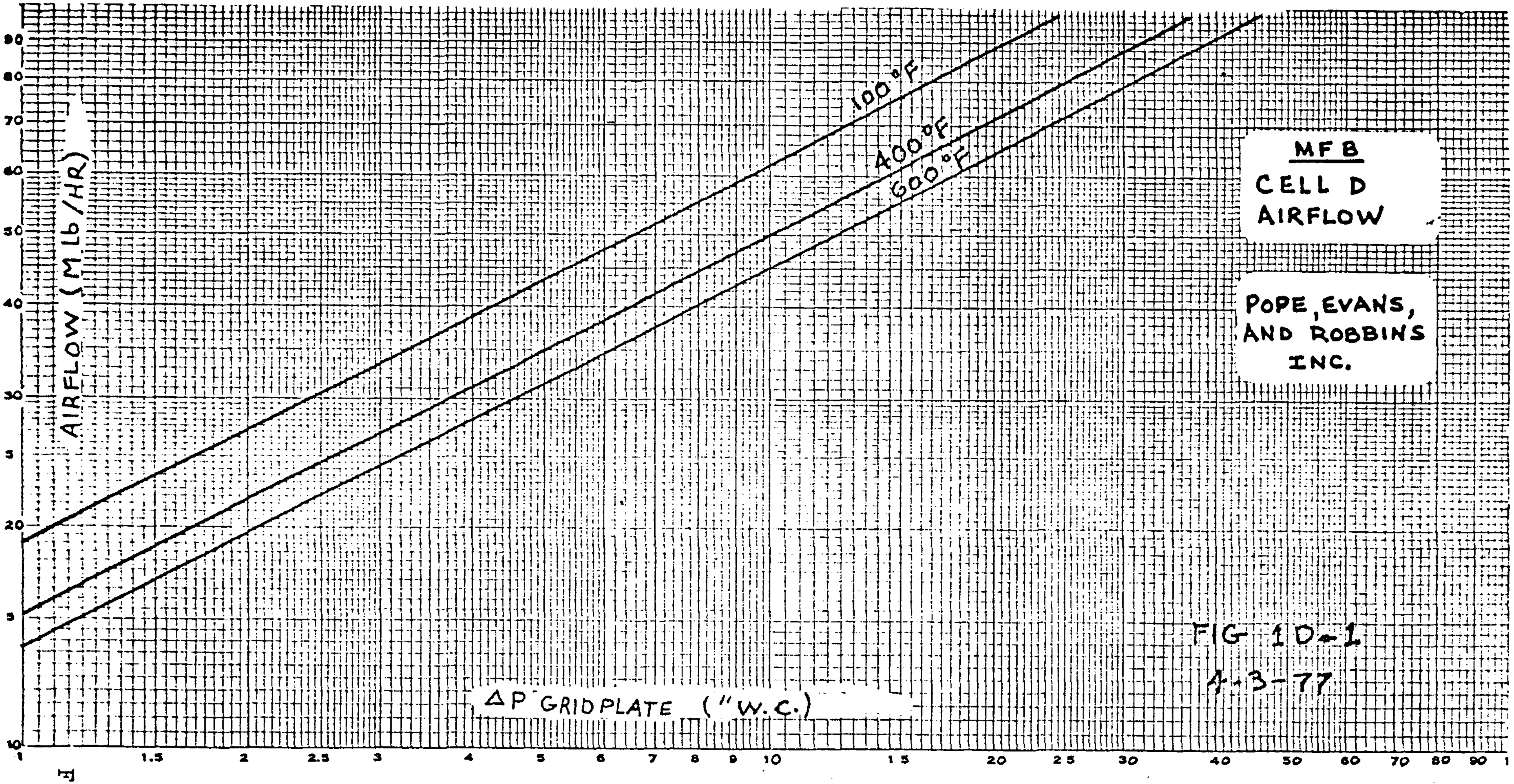
CELL B AIRFLOW vs. GRID PLATE DELTA P AND SUPERFICIAL VELOCITY

FIGURE 3.6



CELL C AIRFLOW vs. GRID PLATE DELTA P

FIGURE 3.7



MFB

CELL D
AIRFLOW

POPE, EVANS, AND ROBBINS INC.

FIG 1D-1
4-3-77

FIGURE 3.8

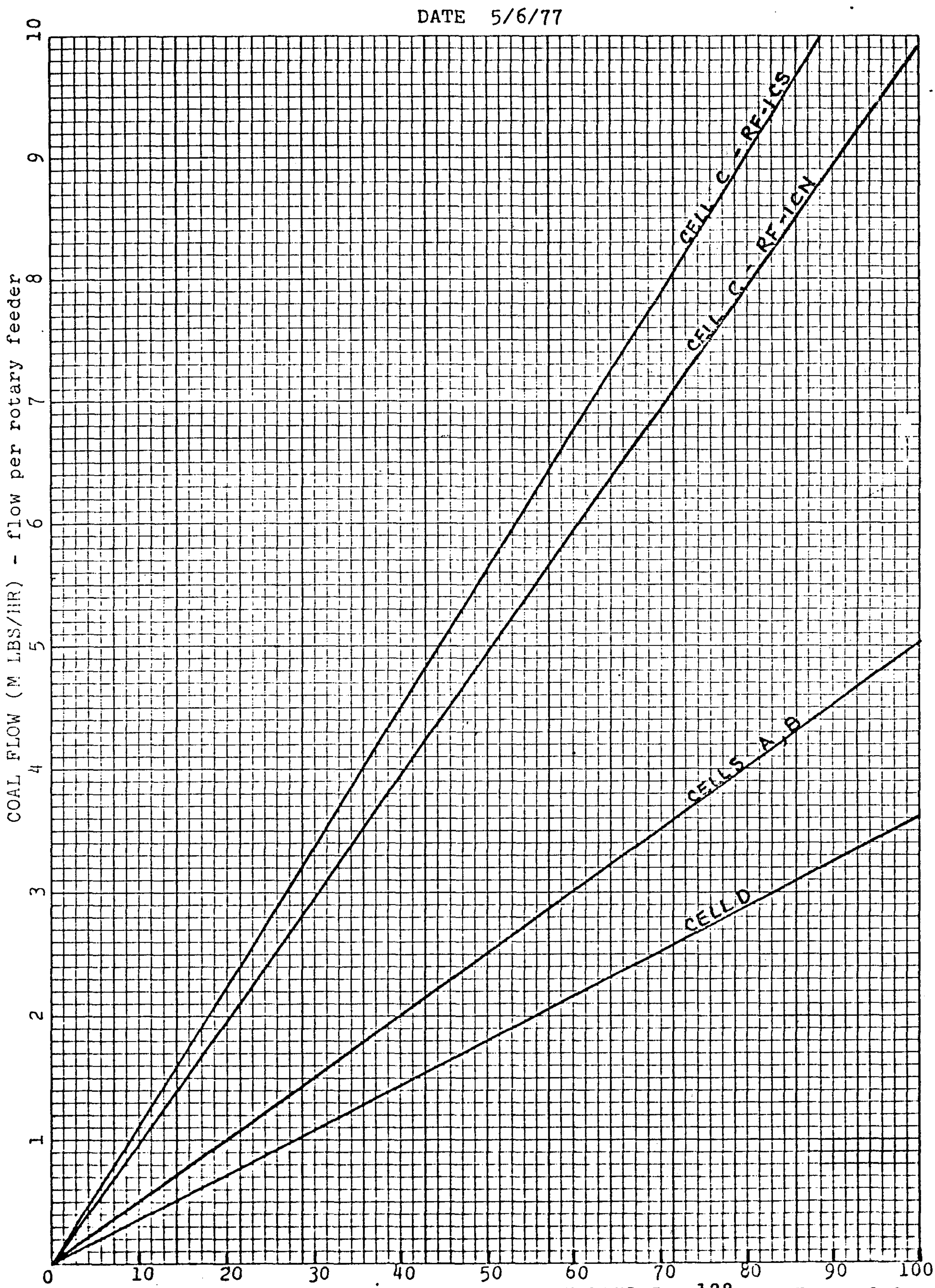
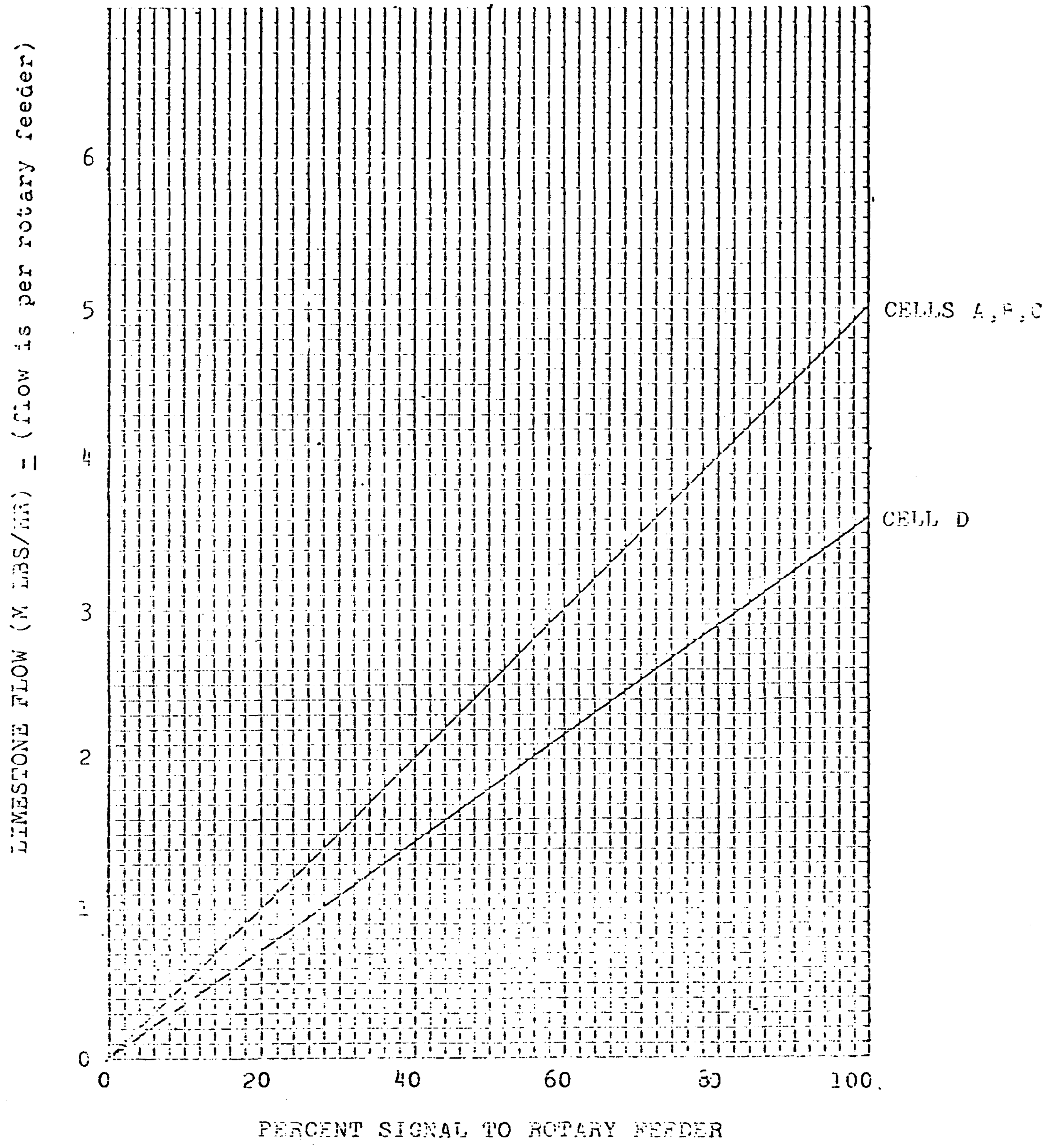

DATE 5/6/77
COAL FLOW (M LBS/HR) - flow per rotary feeder
10
9
8
7
6
5
4
3
2
1
0
10
20
30
40
50
60
70
80
90
100
PERCENT SIGNAL TO ROTARY FEEDER
CELL C - RF-1CS
CELL C - RF-1CN
CELLS A,B
CELL D

FIGURE 3.9

5/6/77



PERCENT SIGNAL TO ROTARY FEEDER VS. LIMESTONE FLOW(M LBS/HR)

FIGURE 3.10

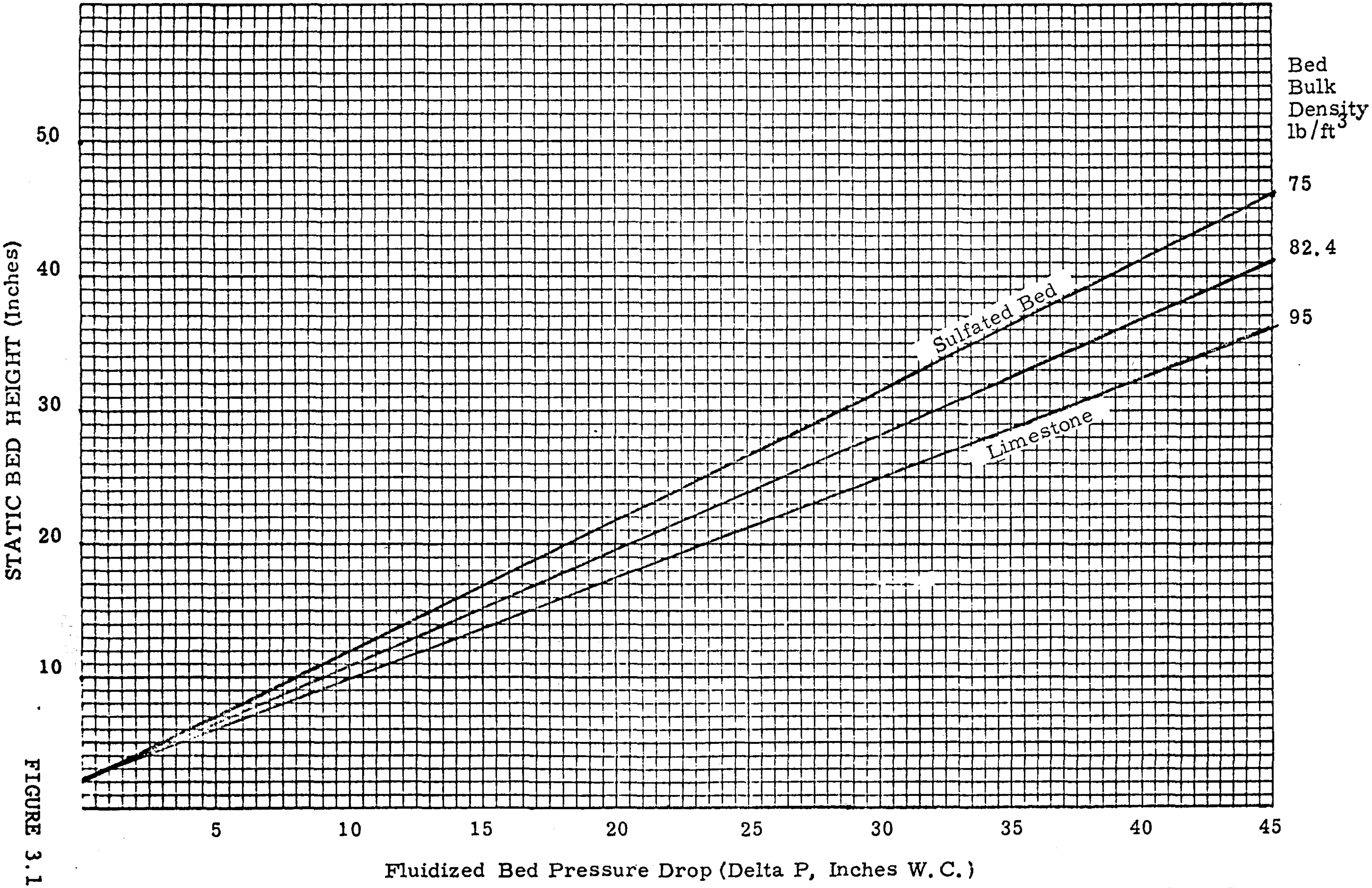
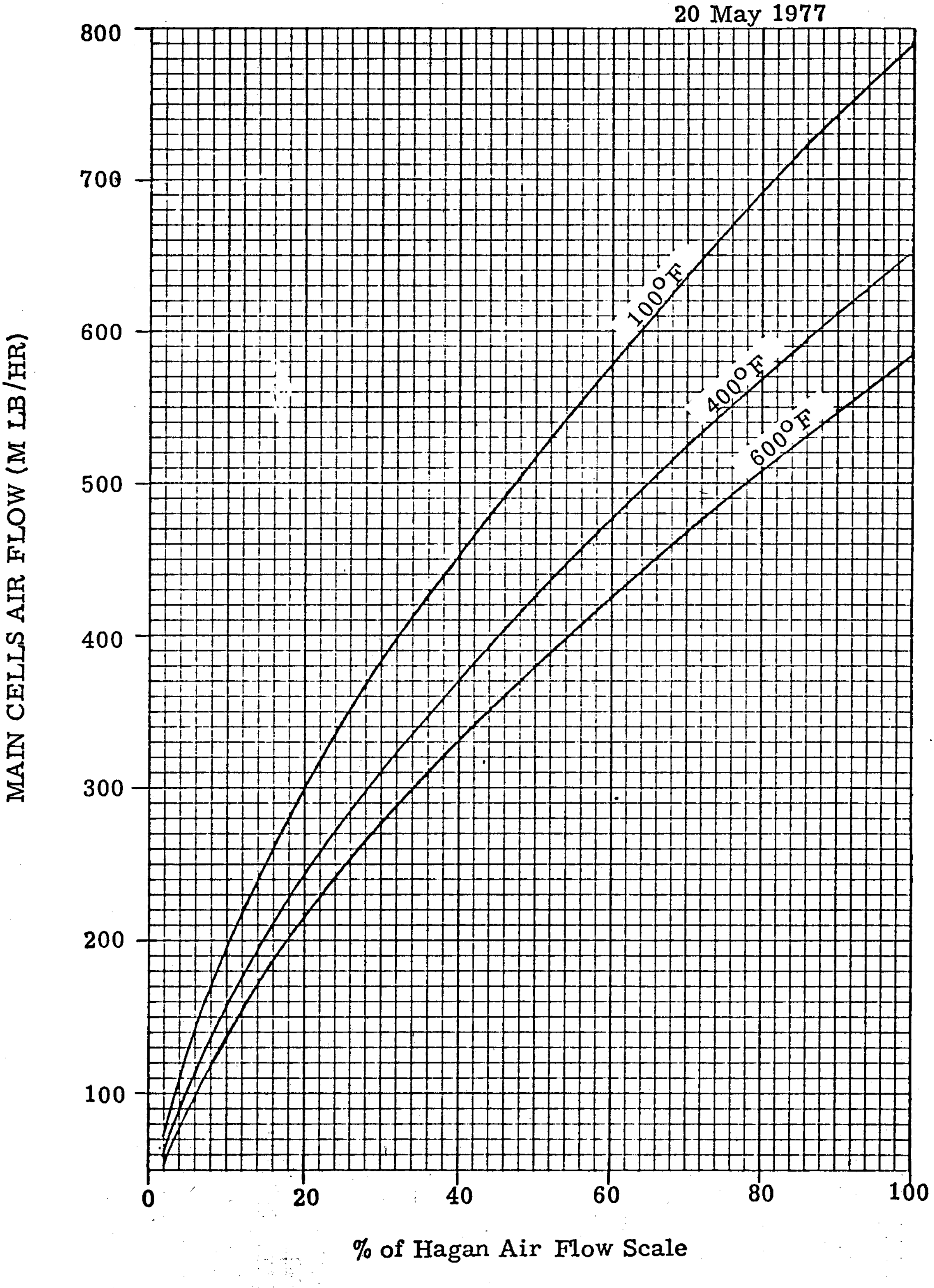


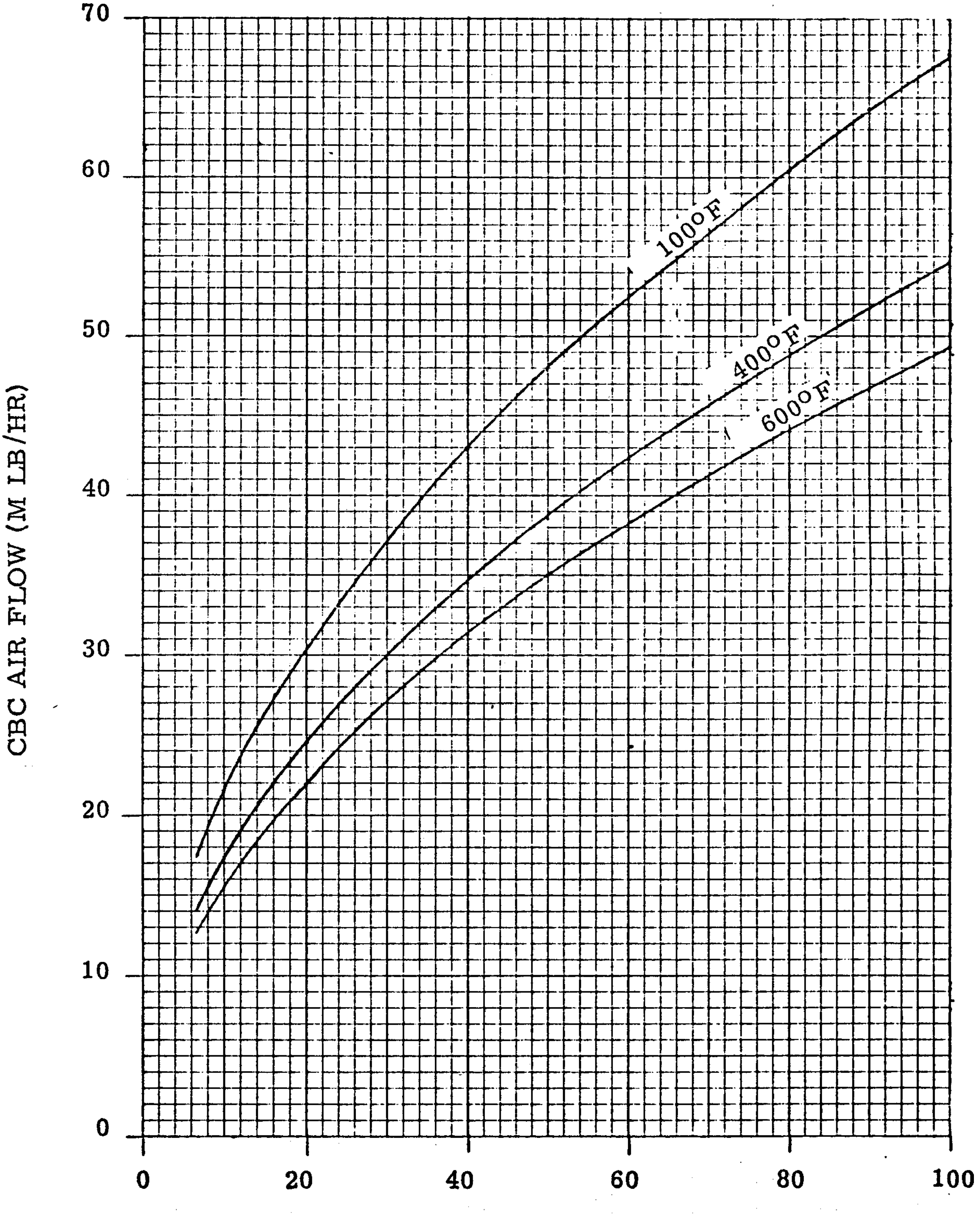
FIGURE 3.11
FLUIDIZED BED PRESSURE DROP vs. STATIC BED HEIGHT FOR VARYING BED BULK DENSITIES



% of MAIN CELLS AIR FLOW SCALE vs. AIR FLOW

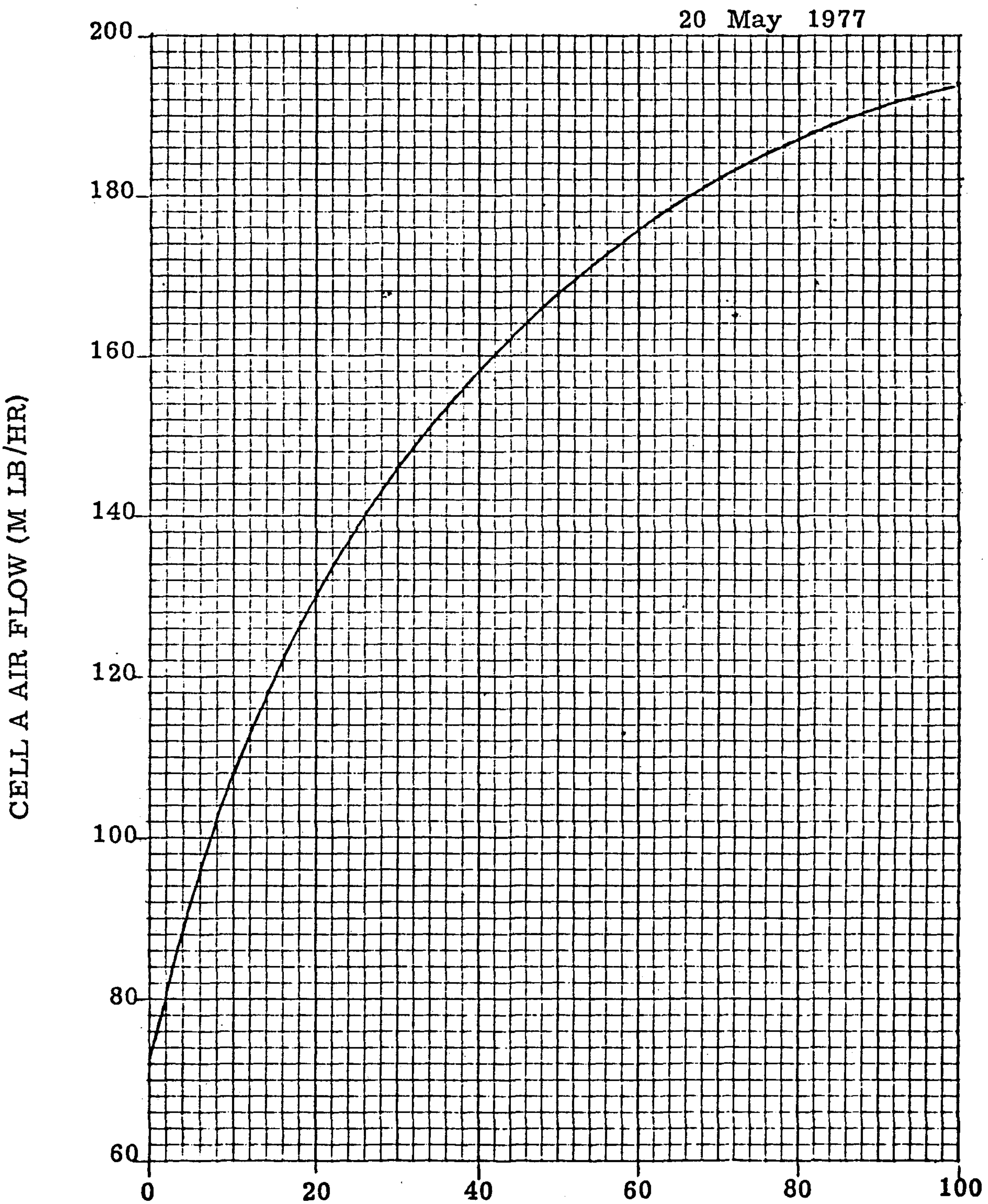
FIGURE 3.12

20 May, 1977



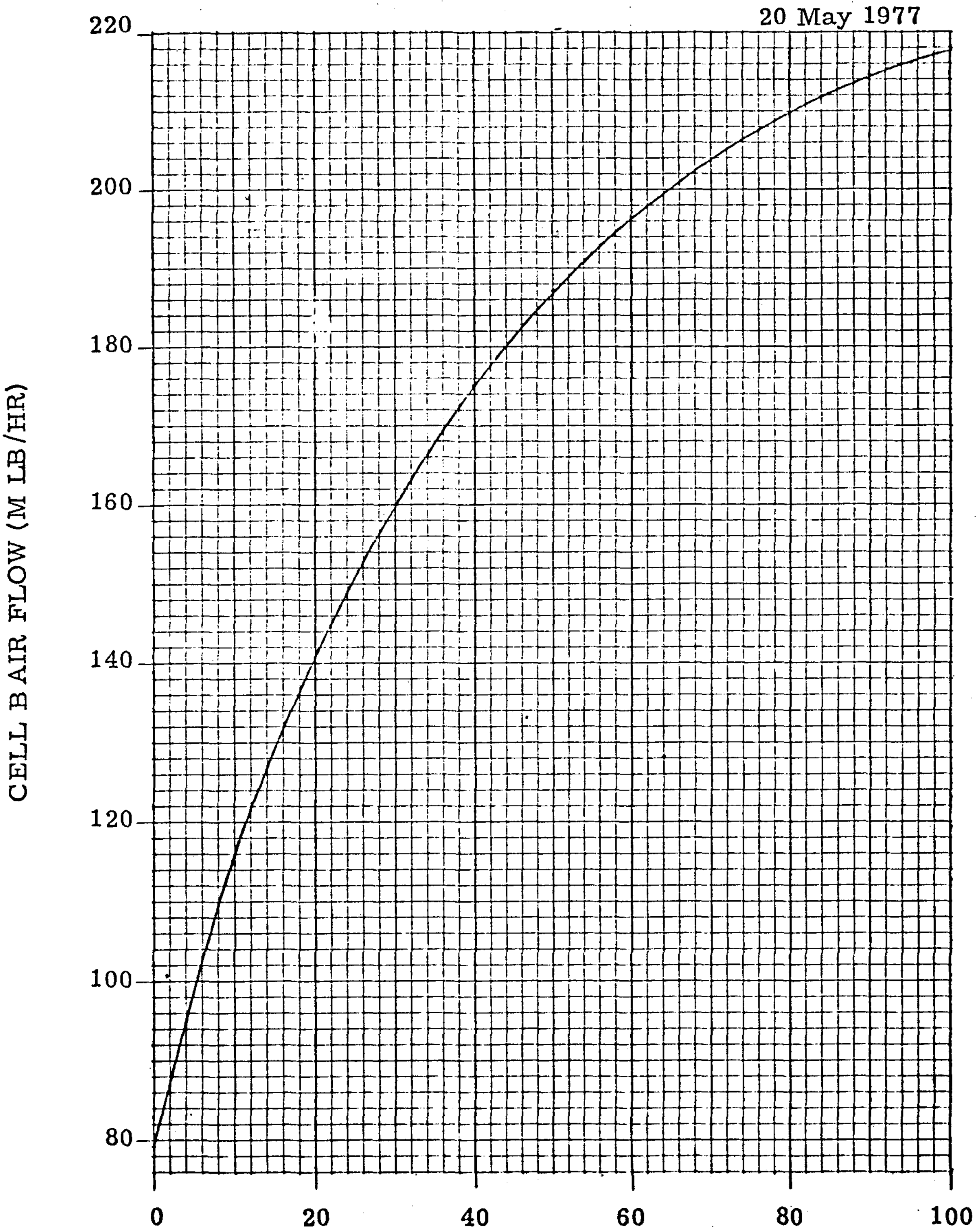
% OF CBC AIR FLOW SCALE vs. AIR FLOW

FIGURE 3.13



CELL A DAMPER SETTING vs. AIR FLOW

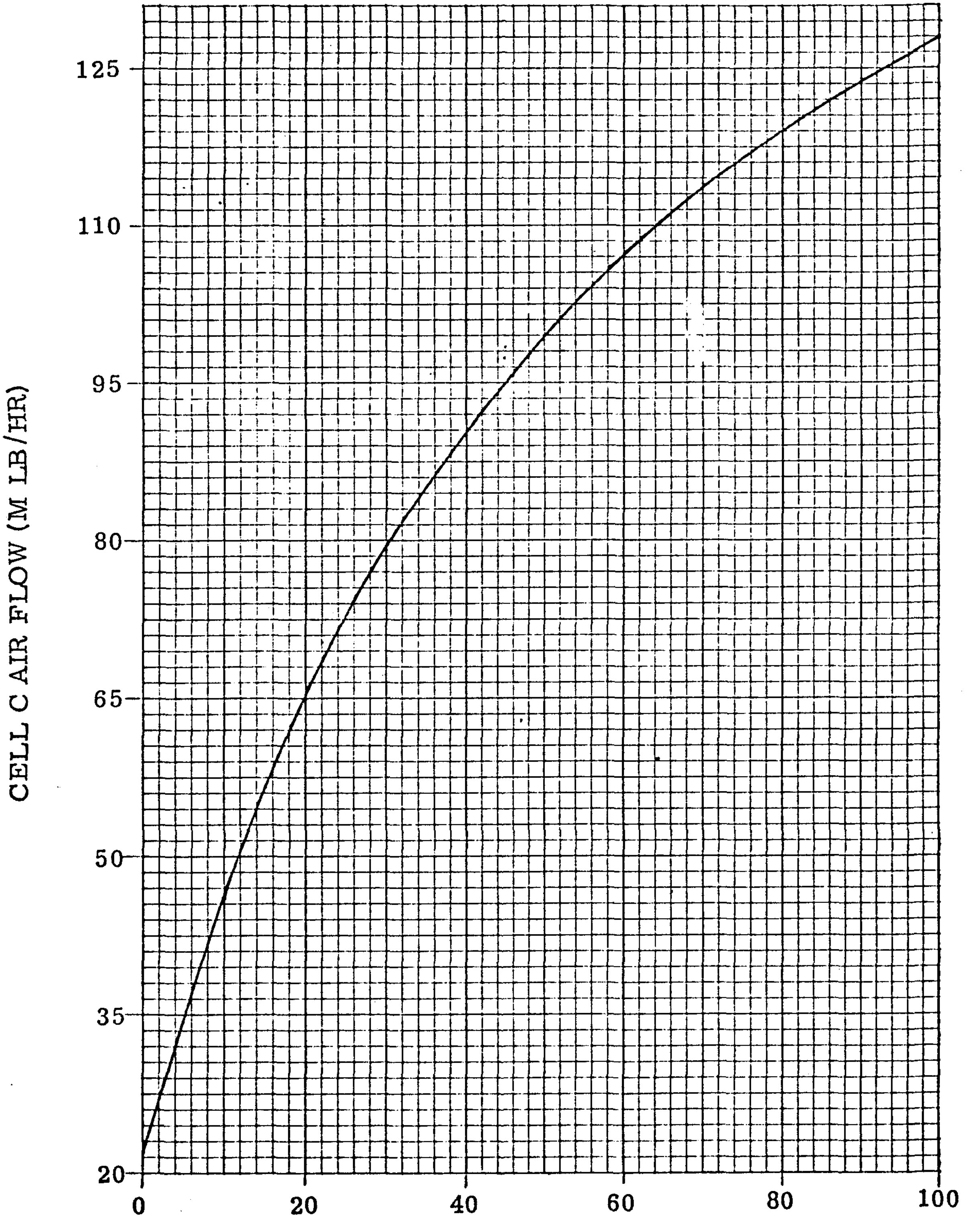
FIGURE 3.14



CELL B DAMPER SETTING vs. AIR FLOW

FIGURE 3.15

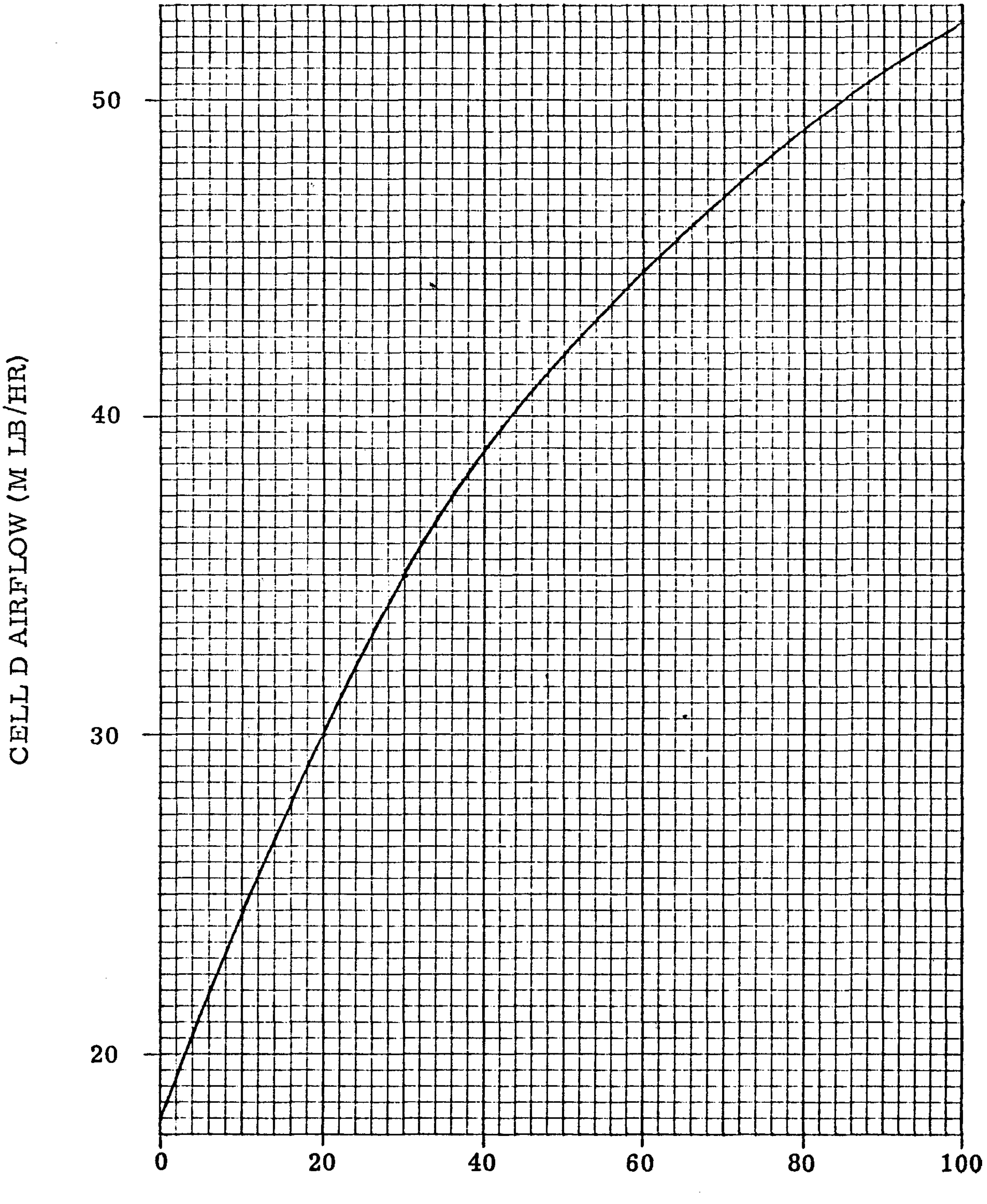
20 May 1977



CELL C DAMPER SETTING vs. AIR FLOW

FIGURE 3.16

20 May 1977



CELL D DAMPER SETTING vs. AIR FLOW

FIGURE 3.17

GRIDPLATE DELTA P REQUIRED FOR MINIMUM FLUIDIZATION

	CELL B	PLENUM TEMPERATURE					
		100°F	200°F	300°F	400°F	500°F	600°F
BED TEMPERATURE	100°F	19	--	--	--	--	--
	400°F	7.7	9.1	11.0	12.0	--	--
	1000°F	2.5	3.0	3.5	4.0	4.5	5.0
	1200°F	1.9	2.3	2.7	3.1	3.4	3.9
	1400°F	1.5	1.8	2.1	2.4	2.7	3.0
	1600°F	1.2	1.4	1.7	1.9	2.2	2.4
	1800°F	--	--	--	--	--	--
	2000°F	--	--	--	--	--	--

Now to use the table and find out if a bed is fluidized;

STEP 1 - Find out the cell plenum temperature and bed temperature.

STEP 2 - Find the square in the table where this bed temperature and this plenum temperature cross.

STEP 3 - Find out the gridplate delta P from the test stand.

STEP 4 - When the test stand gridplate delta P is higher than the delta P in the table, the bed is fluidized. When the test stand gridplate delta P is lower than the delta P in the table, the bed is static.

FIGURE 3.18

GRIDPLATE DELTA P REQUIRED FOR MINIMUM FLUIDIZATION

		PLENUM TEMPERATURE					
	CELL C	100°F	200°F	300°F	400°F	500°F	600°F
BED TEMPERATURE	100°F	22	--	--	--	--	--
	400°F	10.6	13	14	15.6	--	--
	1000°F	4.5	5.3	5.9	6.4	7.2	7.7
	1200°F	3.6	4.2	4.7	5.1	5.8	6.2
	1400°F	3.0	3.5	3.9	4.3	4.8	5.1
	1600°F	2.5	2.9	3.3	3.6	4.0	4.3
	1800°F	--	--	--	--	--	--
	2000°F	--	--	--	--	--	--

How to use the table and find out if a bed is fluidized:

STEP 1 - Find out the cell plenum temperature and bed temperature.

STEP 2 - Find the square in the table where this bed temperature and this plenum temperature corss.

STEP 3 - Find out the gridplate delta P from the test stand.

STEP 4 - When the test stand gridplate delta P is higher than the delta P in the table, the bed is fluidized. When the test stand gridplate delta P is lower than the delta P in the table, the bed is static.

FIGURE 3.19

CBC ANNUBAR DELTA P REQUIRED FOR MINIMUM FLUIDIZATION

		PLENUM TEMPERATURE					
	CELL CBC	100°F	200°F	300°F	400°F	500°F	600°F
BED TEMPERATURE	100°F	1.5	--	--	--	--	--
	400°F	.64	.77	.88	1.0	--	--
	1000°F	.22	.26	.31	.35	.39	.43
	1200°F	.17	.21	.24	.27	.30	.33
	1400°F	.14	.16	.19	.21	.24	.26
	1600°F	.12	.14	.15.	.17	.19	.21
	1800°F	.10	.11	.13	.145	.16	.18
	2000°F	.10	.10	.11	.12	.135	.15

How to use the table and find out if a bed is fluidized:

STEP 1 - Find out the cell plenum temperature and bed temperature.

STEP 2 - Find the square in the table where this bed temperature and this plenum temperature cross.

STEP 3 - Find out the CBC annubar delta P from the test stand

STEP 4 - When the test stand CBC annubar delta P is higher than the delta P in the table, the bed is fluidized. When the test stand CBC annubar Delta P is lower than the delta P in the table, the bed is static.

FIGURE 3.20

EQUALIZED BED HEIGHT FOR CELL C LIGHT-OFF

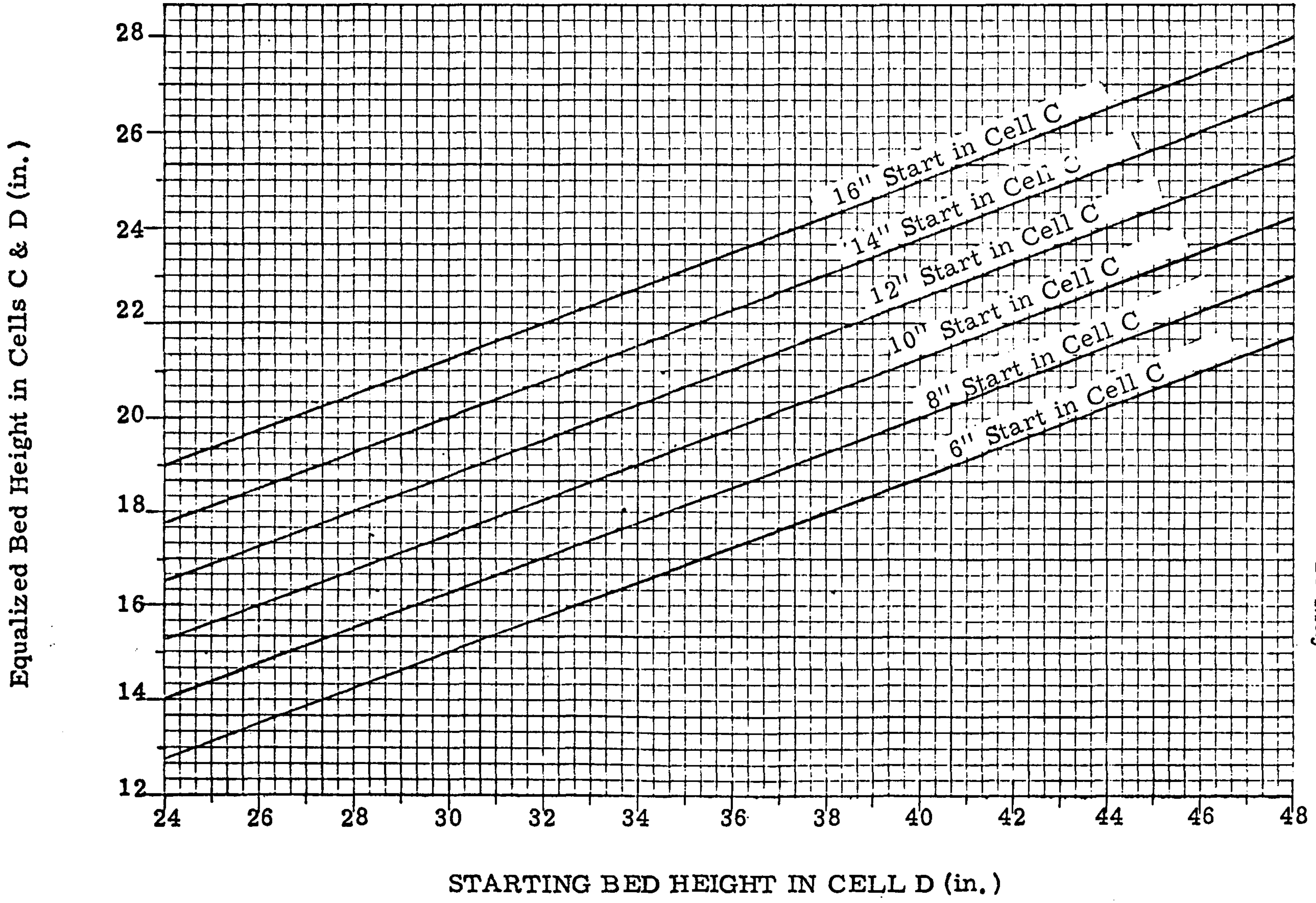
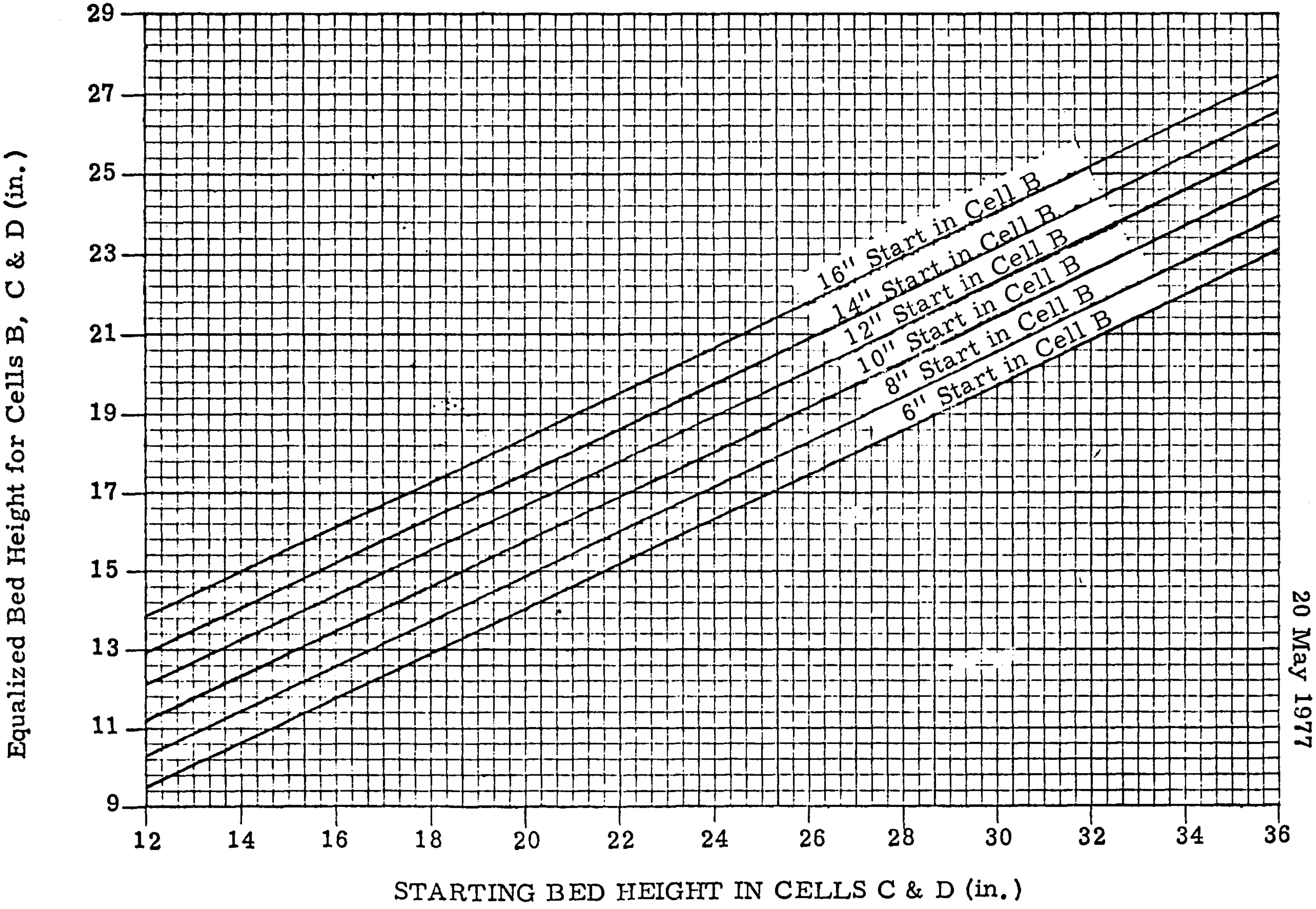


FIGURE 3.21

20 May 1977



EQUALIZED BED HEIGHT FOR CELL B LIGHT-OFF

FIGURE 3.22

20 May 1977

CBC cell and continuous oil firing to raise bed material temperatures to 800°F was initiated at 1900 hr on 29 November 1976.

Temperature increases were carefully controlled to permit evaluation of light-off procedures and continual check-out of the boiler combustion control panel instruments and controls for proper operation. In early December, an effective method for light-off of Cell D was developed and implemented.

Introduction of coal into Cell D was initiated at 1100 hr on 7 December 1976 and combustion sustained for a short period of time. Improvements were incorporated in successive light-offs, cutting down the time necessary to heat bed material and increasing the length of time coal combustion was sustained. A thirty-three (33) hour continuous Cell D coal burning (without fuel oil burner assistance) was accomplished on 21 December through 23 December 1976

Combustion of Coal Sustained, 7 December 1976

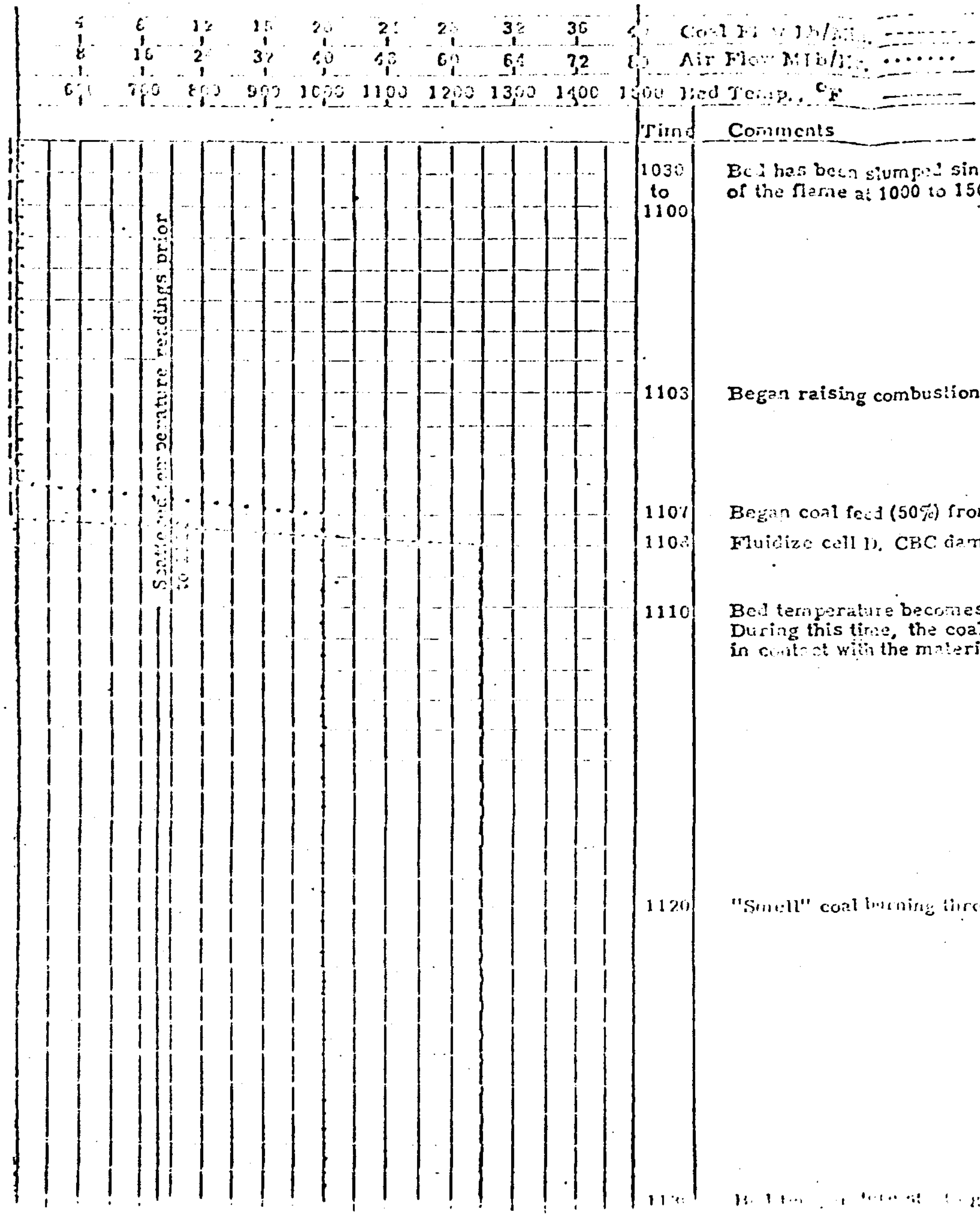
Table 3.1 shows the operating conditions at 1030 hr, one-half hour prior to start of coal feed. Table 3.2a through 3.2e is a plot of bed temperature, airflow and coal feed rate vs. time for the duration of coal firing. In the right-hand column are comments on the charges made during the coal injection period. The oxygen probe was not operational at that time, hence no attempt was made to keep the excess air at 3-4 percent oxygen while burning coal.

It was noted soon after coal feed began, that the four (4) injection needles which feed the west side of the bed were clogged. It was decided that temperature excursions above 1200 - 1300°F would risk damaging the clogged needles. Therefore, efforts were made to keep the bed temperature below 1300°F throughout the run in order to limit the risk of damage.

Cell C Initial Coal Light-off (Attempt No.4) 10 February 1977

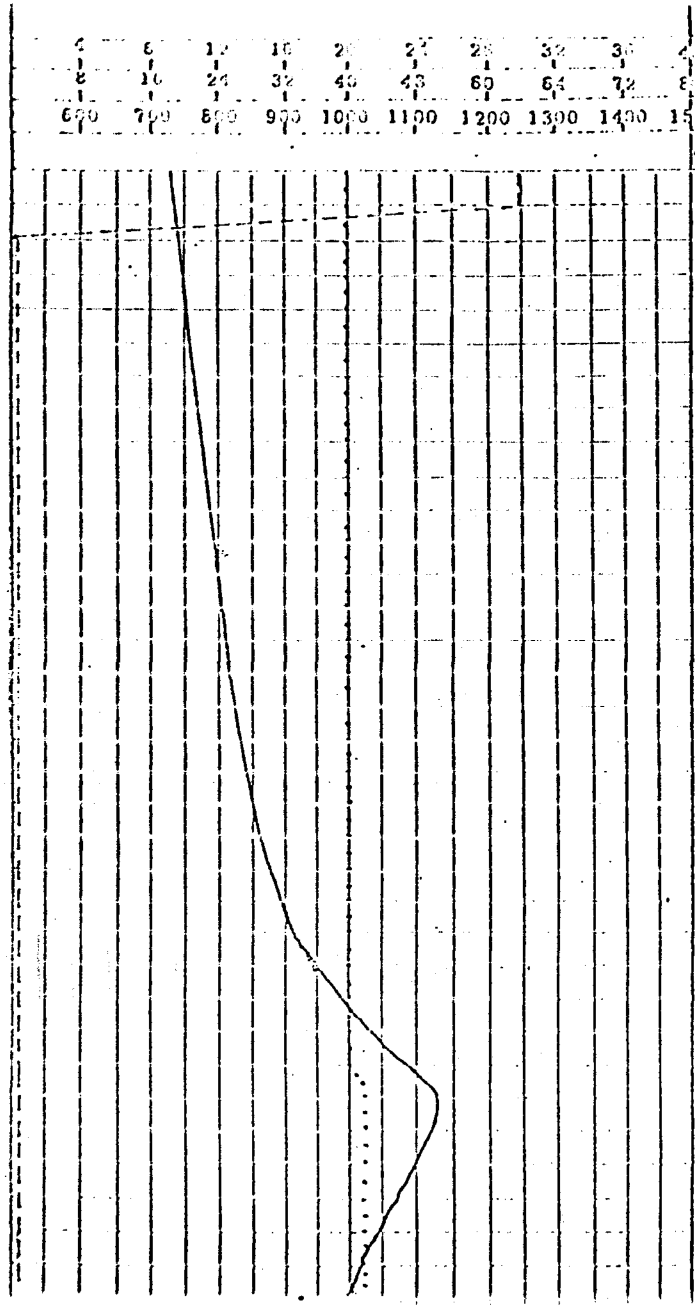
On 10 February 1977, coal fire was established in Cell D and at light-off attempt of Cell C, bed height was raised to 33" WG static, and bed temperature was 1800°F. Cell bed height was 7" - 8" static, with 90,000 lbs/hr airflow and combustion air temperature was 440°F.

Coal was introduced to Cell C two (2) minutes prior to the opening of D-C slide gate in order to provide a coal charge in Cell C. Approximately fifty (50) seconds after opening the



Time	Comments
1030 to 1100	Bed has been slumped since 0400 with local temperatures of the bed in the region of the flame at 1000 to 1500°F. The static bed height is approximately 25 inches.
1103	Began raising combustion air pressure to 57 inches in preparation to fluidize.
1107	Began coal feed (50%) from coal bin. Time to reach the MFB is approximately 45 sec.
1108	Fluidize cell D. CBC damper position 45%.
1110	Bed temperature becomes uniform at the six different thermocouple locations. During this time, the coal was migrating to the surface of the bed and coming in contact with the material from the hot zone
1120	"Smell" coal burning through pressure taps in the side of the unit
[illegible]	[illegible]

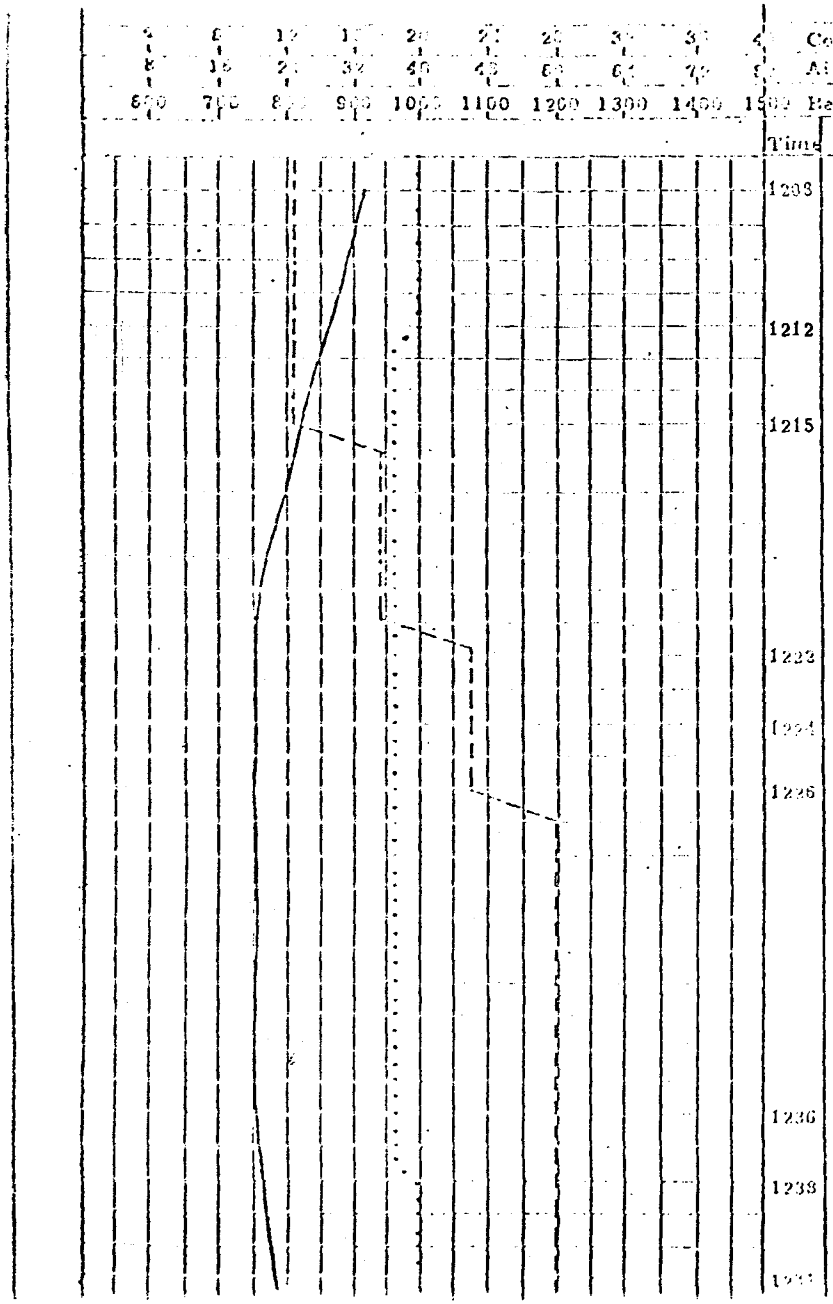
TABLE 3.1
Cell D Light-off
December 7, 1976



Coal Flow Lb/Min. - - - - -
Air Flow MLb/Hr. ....
Bed Temp., °F ———

Time	Comments
1132	Stopped coal feed (0%). Bin was almost empty - decided to refill the bin.
1137	Bed Temperature = 760°F.
1145	Bed Temperature = 820°F.
1150	
1155	
1158	Increased air flow (50%) in effort to slow rate of bed temperature increase.
1159	
1202	

TABLE 3.2a
Cell D light-off
December 7, 1976



Comments

1212 Reduced air flow to minimum possible (40%) without slumping.

1215 Coal Feed increased (30%)

1223 Increased coal feed (40%) in effort to reignite bed.

1224 Discovered that only 3 of 8 injection needles are feeding

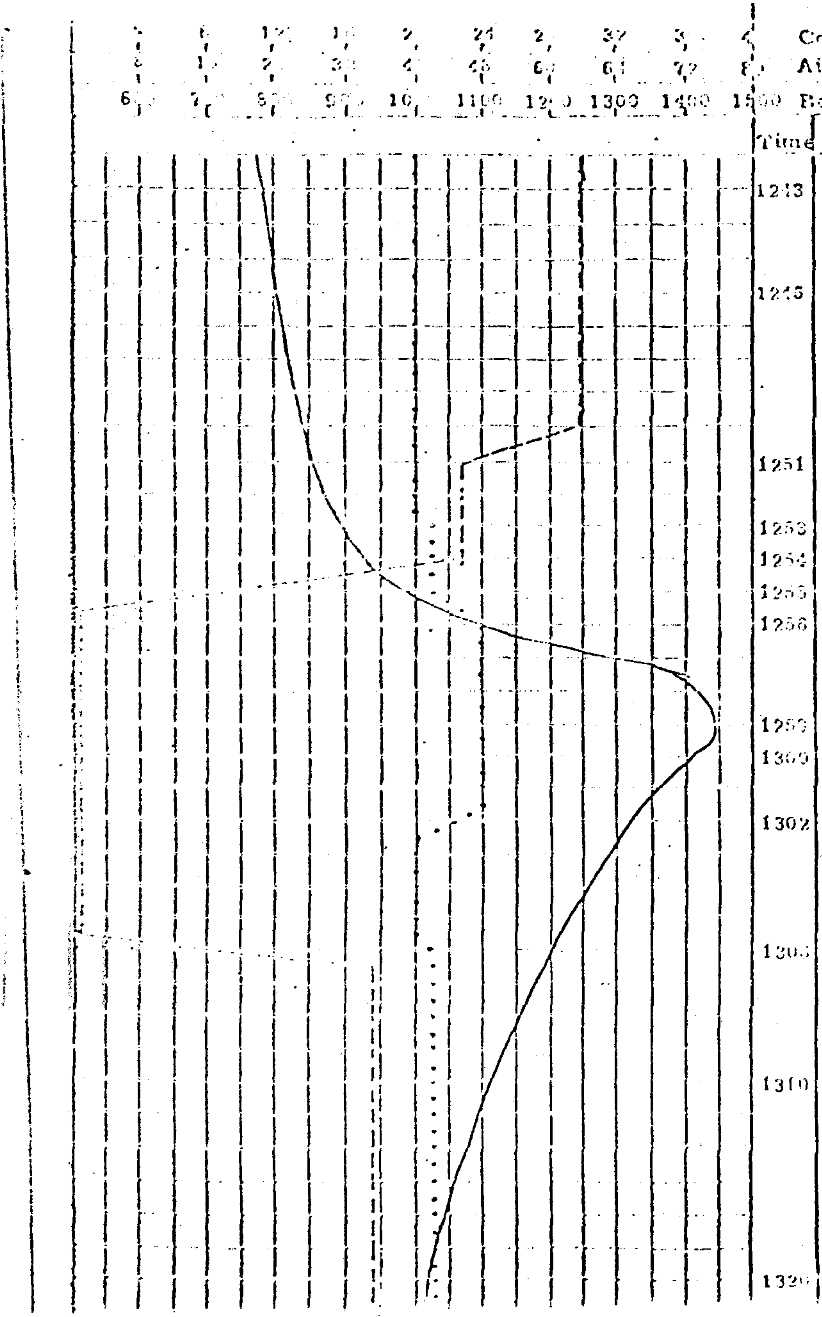
1226 Increased coal feed (50%)

1238 Increased air flow (45%) in effort to increase turbulence and start burners.

TABLE 3.2b

Cell D light-off

December 7, 1976



Comments

1251 Reduced coal feed (40%) in attempt to stabilize increase.

1253 Increased air flow (50%) to slow bed temperature increase.

1254 Stopped coal feed (0%) due to fear of temperature excursion.

1255 Increased air flow (60%)

1256 Increased air flow (70%) to drop bed temperature below margin of safety for possible clogged needle metal failure.

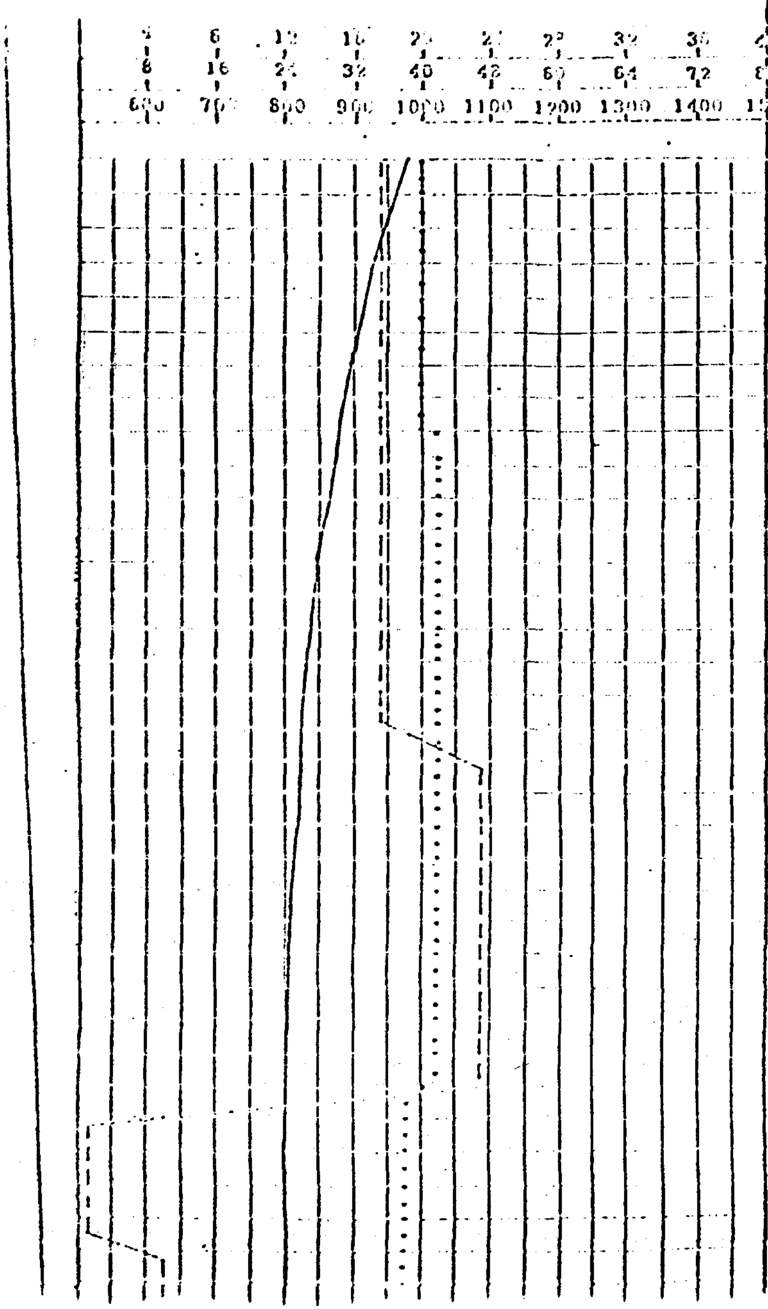
1302 Reduced air flow to normal (40%).

1305 Restarted coal feeder, increased air flow (50%).

TABLE 3.2c

Cell D light-off

December 7, 1976

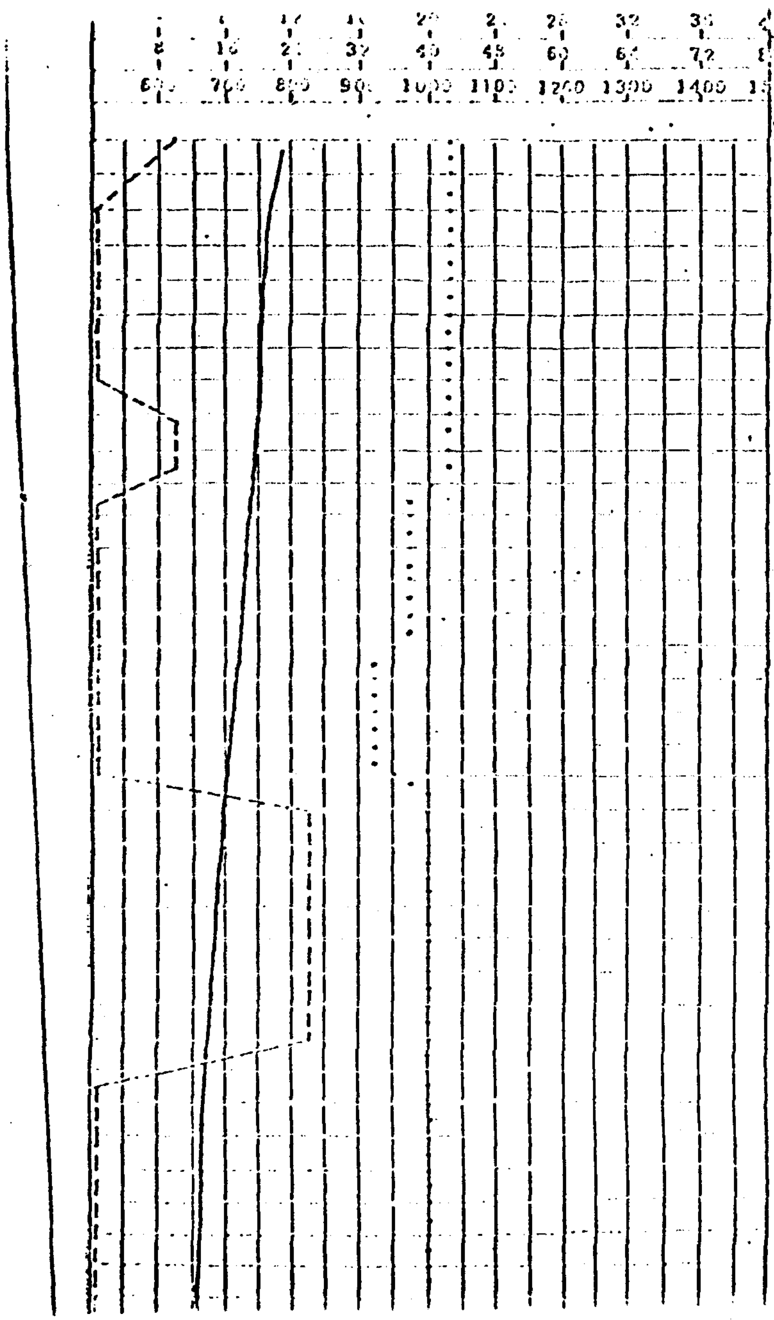


Time	Comments
1323	
1330	Air flow increased (50%).
1339	Increased coal feed rate (40%)
1340	
1350	Coal feed shut off, reduced air flow to below minimum fluidizing velocity (30%) in order to reignite coal in top portion of bed.
1355	Commenced coal feeding at low rate (10%).

TABLE 3.2d

Cell D light-off

December 7, 1976



Time	Comments
1400	
1400	
1402	*Coal reported to be not feeding at bin.*
1405	Coal feed resumed
1407	Coal feed ceased to prevent building up excess inventory of unburned coal.
1410	
1412	Reduced air flow to below minimum fluidizing velocity (38%) in effort to heat *and ignite top portion of bed.*
1415	
1416	Refluidized bed (45%) in effort to mix hot bed surface with rest of bed to reignite
1417	Coal Feed resumed.
1420	
1425	Cut coal feed to zero.
1428	Problems reported with MFB blowdown to MPC
1430	

TABLE 3.2e
Cell D light-off
December 7, 1976

slide gate, the bed material had equalized between the two (2) cells and one (1)minute after the cells had equalized, Cell C bed temperature started to rise sharply, leveling off at 1500°F to 1600°F. Cell C bed temperature then started dropping to 750°F - 800°F. Recording charts Figures 3.23 thru 3.26, indicate that bed temperature remained above 100°F for nine (9) minutes. (Also included: excess O$_2$, coal flow and airflow recorder charts). In an attempt to raise the bed temperature again, the coal feed was stopped and Cell C bed slumped for four (4) minutes. At this time, the bed temperature ranged from 450°F to 1400°F. The bed was then fluidized and coal feed started at which time the average bed temperature rose to 800°F. This same approach was attempted again, but was unsuccessful.

Summary of Events

Time (Hour)	Event
1740	RF-1CN & RF-1CS to 50% (5,000 lbs/hr coal)
1742	Slide Gate open
1744	RF-1CN & RF-1CS to 75% (7,600 lbs/hr).
1745	Bed temperature Cell C 1600°F Main Furnance uptake temperature 600°F.
	Cell C 8% excess O$_2$
	Drum Pressure rose from 750 psi to 1100 psi
	Bed height stablized at 14"
1755	RF1-CN and CS to 0%, slumped Cell C bed. Bed temperature rose to between 500°F and 1000°F
1800	Fluidized Cell C, RF1-CN and CS to 50% (5,000 lbs/hr coal flow)
1810	RF1-CN and CS to 0%
1812	RF1-CN and CS to 0%, slumped Cell C, bed temperature rose to between 500°F and 1580°F
1818	Fluidized cell C RF1-CN and CS to 100%
1822	RF1-CN and CS to 0%
1835	Closed slide gate

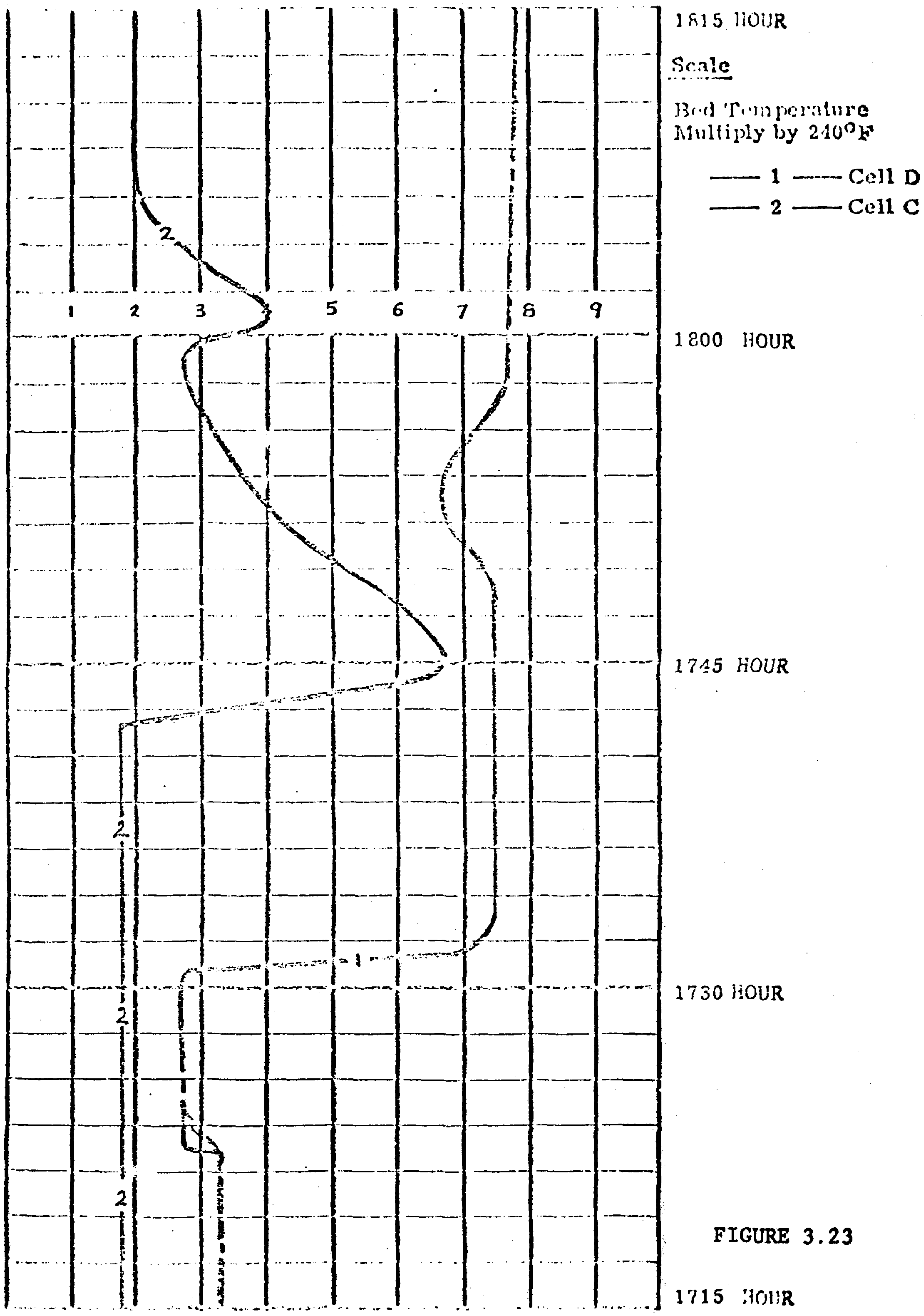


FIGURE 3.23

Cell C Light-Off - February 10, 1977

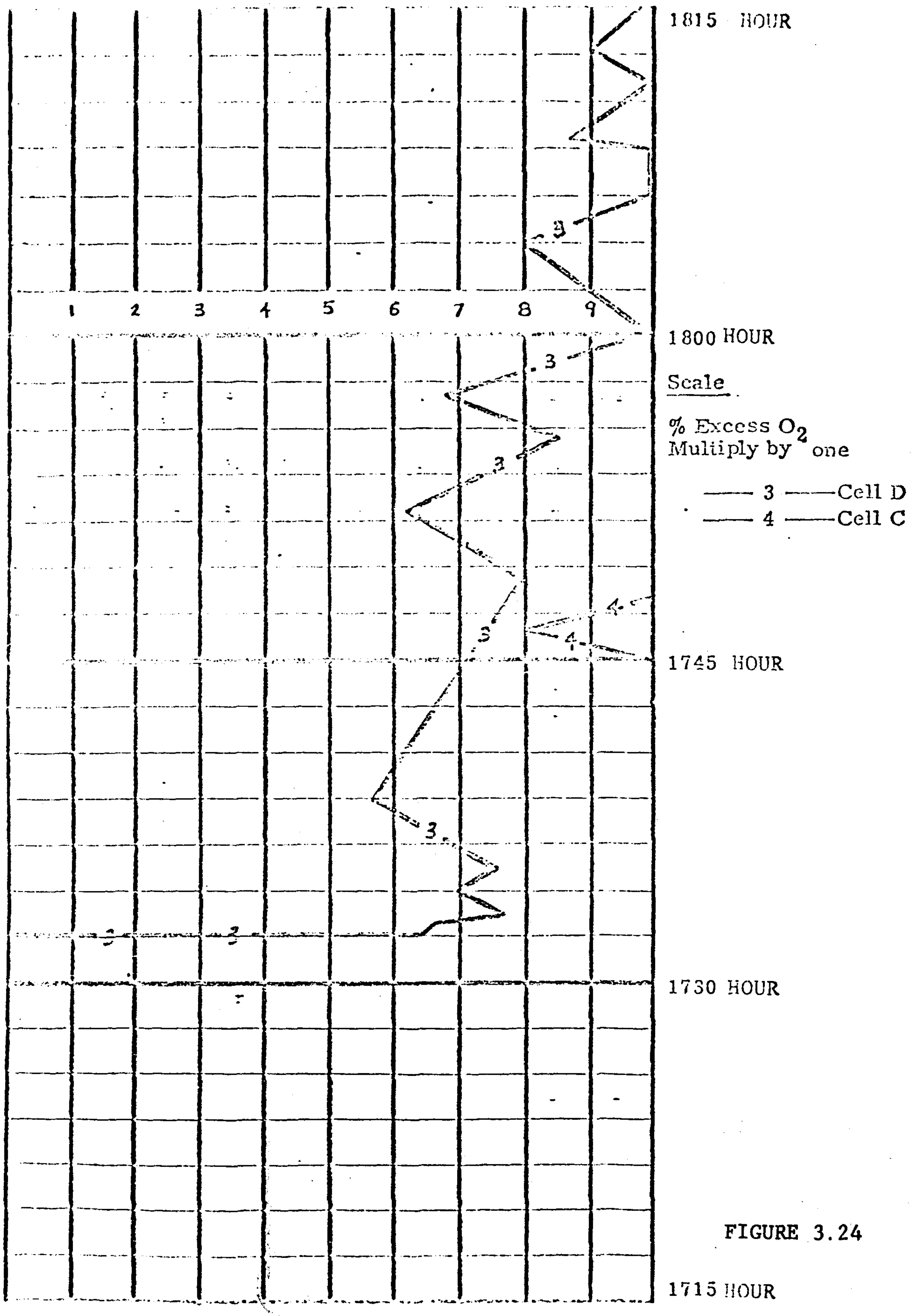
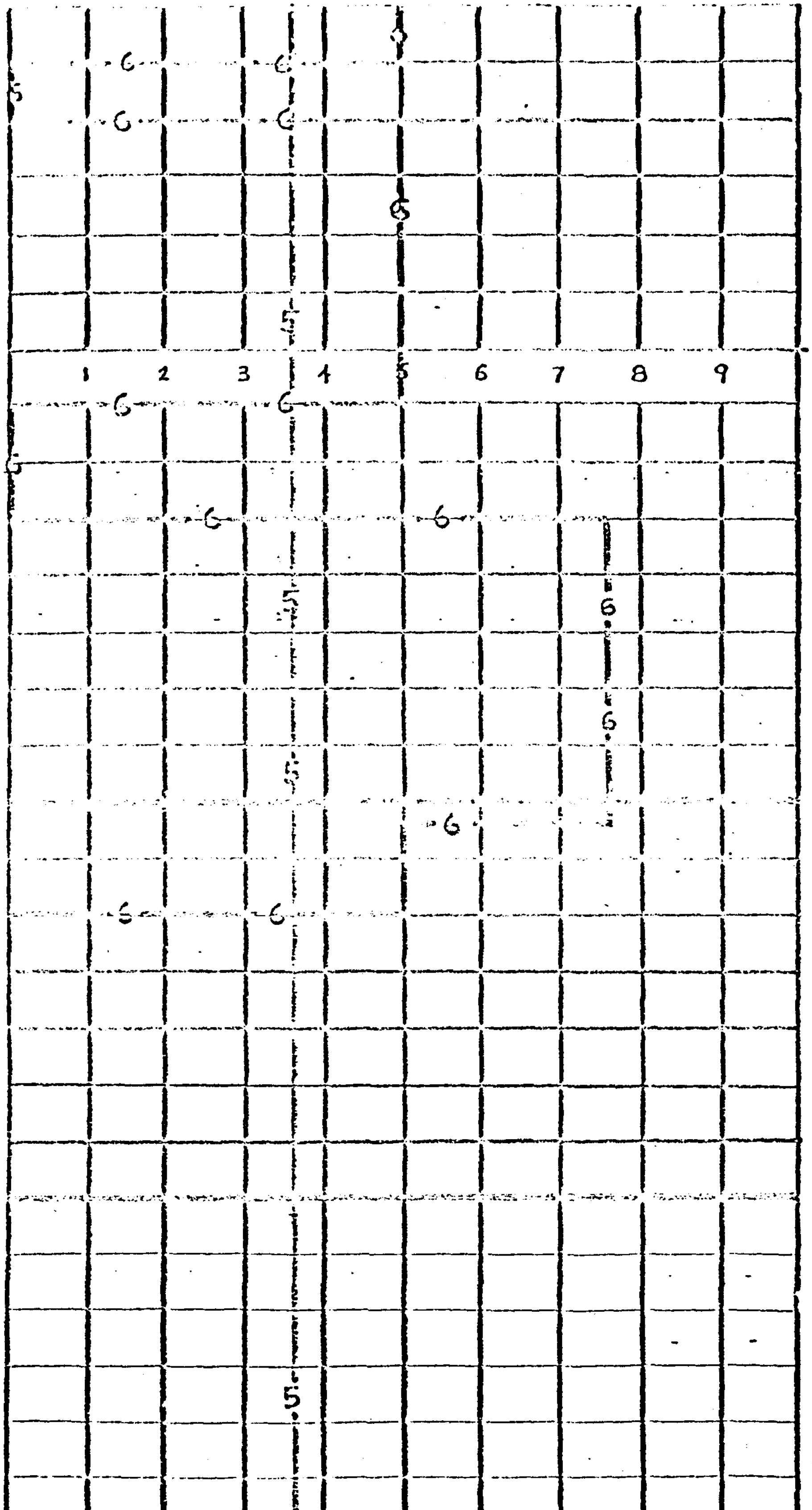


FIGURE 3.24

Cell C Light-Off, February 10, 1977



Scale

Coalflow
Multiply by 1000 lbs/hr

——— 5 ———Cell D
——— 6 — — -Cell C

FIGURE 3.25

Cell C Light-Off, February 10, 1977

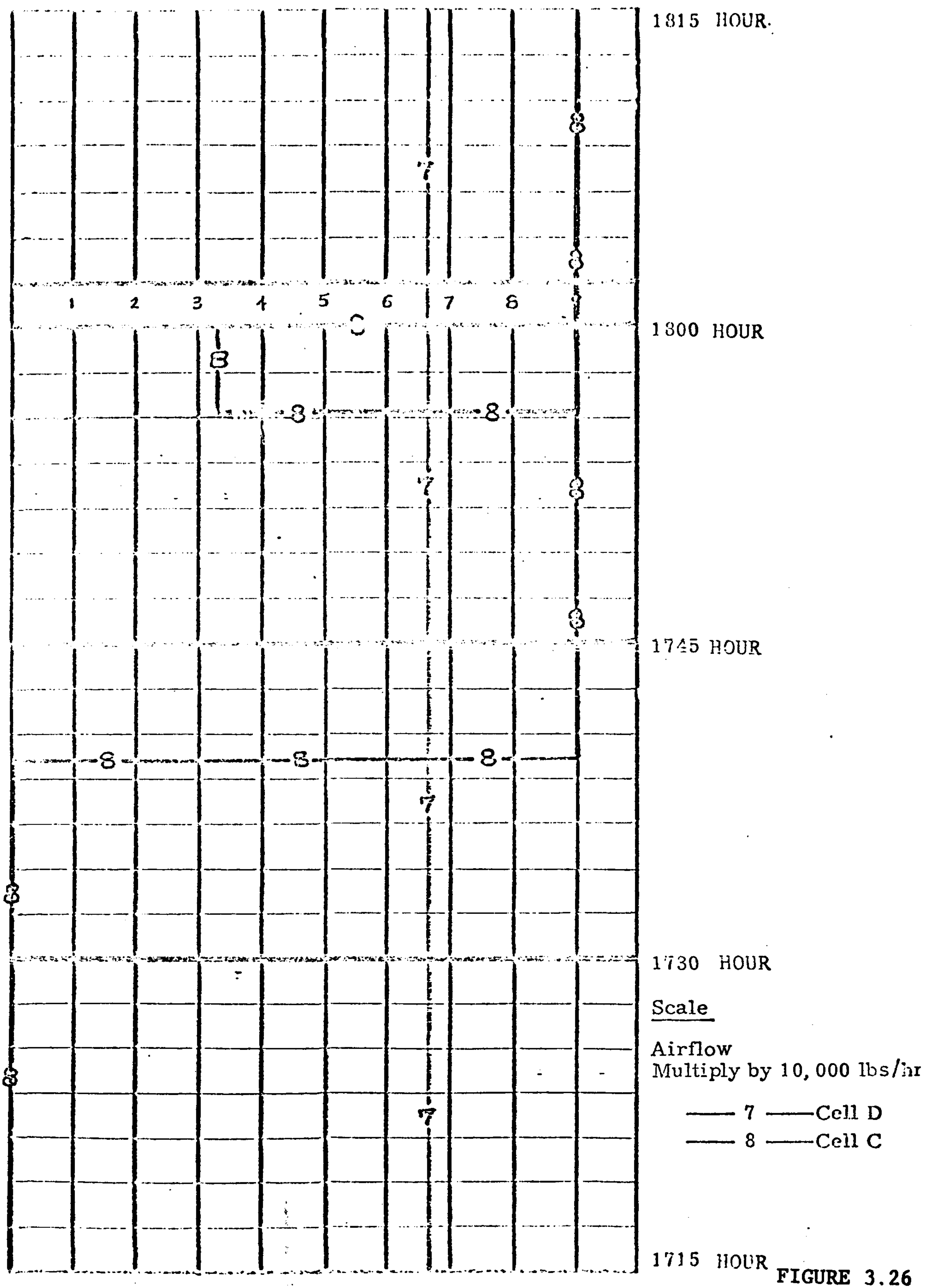


FIGURE 3.26

Cell C Light-Off, February 10, 1977

First prolonged C-Cell Coal Fire Operation, 5 April 1977

At 1105 on 5 April 1977, coal fire was established in Cell D in preparation for Cell C light-off. Cell D bed depth was increased to thirty-seven (37) inches and bed temperature increased to 1825°F.

At 1935, coal flow was started to Cell C and allowed to run two (2) minutes and twenty (20) seconds at 5,000 lbs/hr prior to opening the D-C transfer gate. Cell C temperature indication showed a sharp temperature rise within two (2) minutes. The bed temperature rose to 1300°F before coal feed was restarted and ignition took place. At 2014, D-C transfer gate was closed and Cell C bed temperature leveled off at 1760°F. During light-off, airflow was maintained at 3.5 ft/sec.

Cell C was operated independently of Cell D for two (2) hours and thirty-six minutes. Cell C was shut down at 2250 due to low water level in the MPC condensate tank.

Summary of Events, 5 April 1977

Time (hour)	Event
0850	Cell D check list complete
1058	Fluidized Cell D
1100	Started coal flow
1105	Coal fire in Cell D
1116	Burners OFF
1425	Dumping precipitator hoppers
1834	Cell C check list complete
1935	Started coal feed to Cell C at 5000 lbs/hr
1937	Opened D-C transfer gate
1940	Coal feed trouble
1059	Coal fire re-established
2014	Closed transfer gate
2119	Bed temperatures equal at 1760°F/16.5" bed height
2225	Trouble with coal feed

2245	Opened D-C transfer gate - Burners ON
2250	Stopped coal feed

Cell B Initial Coal Light-off, 19 April 1977

Initial coal light-off and prolonged coal fire operation was achieved during the same run of Cell B on 19 April 1977. A description of unit operation during this run is given.

At 2218 on 18 April 1977, Cell C coal fire was established and both Cells C and D were operated independently. Due to coal feeding problems from the south side Cell C furnace feed system and the resulting inability to feed linestone into Cell C continuously- the D-C transfer gate was re-opened in order to aid in maintaining Cell C during loss of coal feed and to build up bed height in Cell C and Cell D equally with the Cell D feed system.

The bed heights in Cells D and C were built up with the intention of lighting off Cell B. At 0100, the MPC Condensate Tank level dropped to 15 feet 6 inches from 18 feet 2 inches the hour before and the MPC demineralizer was being regenerated. An evaluation of the operating conditions, showed the feasibility of Cell B light-off. At 0122, Cell B was prepared for light-off. The following conditions were established.

MPC Condensate Tank water level - 14.2 feet
Cell D static bed height - 20 inches
Cell C statis bed height - 20 inches
Cell B static bed height - 7 1/2 inches
Cell D bed temperature - 1800°F
Cell C bed temperature - 1575°F
Cell B bed temperature - 475°F
Feedwater flow (all blowdowns closed) prior to lighting Cell B - 125,000 lbs/hr
Feedwater flow (all blowdowns closed) after Cell B light-off - 140,000 - 160,000 lbs/hr

The Cell B check-off list was completed and coal feed was established to Cell B at 0153 hr. A coal inventory of approximately 225 lbs. was built up in Cell B and C-B transfer gate opened at 0155. Coal fire was established in Cell B at approximately 1600°F and the transfer gate was closed. Cells B, C and D were then operated independently for one (1) hour and twenty-five (25) minutes, at which time the MFB unit was forced to shut down because of low condensate storage tank water level. Up to June 30, 1977, the Rivesville, W. Va. 30 MW_e MFB unit had accumulated 324.5 hrs of coal burning operation, as shown on Figure 3.27.

RIVESVILLE MFB UNIT OPERATIONAL HOURS

MONTH	CBC CELL (72 SF SIZE) OIL	CBC CELL (72 SF SIZE) COAL	CELL C (120 SF) COAL	CELL B (132 SF) COAL	CELL A (132 SF) COAL
Nov. 1976	23.0 (*)	--	--	--	--
Dec. 1976	285.5 (*)	29.0 (*)	--	--	--
Jan. 1977	203.0 (*)	42.0 (*)	--	--	--
Feb. 1977	135.0 (*)	79.5 (*)	--	--	--
March 1977	--	--	--	--	--
April 1977 (**)	172.5	131.5	12.5	1.5	--
May 1977	155.5	42.5	10.5	2.5	--
TOTAL:	974.5	324.5	23.0	4.0	--

(*) Pre-construction completion operation (general construction completion: March 1977)
(**) Initial plant start-up operation started

FIGURE 3.27

Summary of Events, 19 April 1977

Time (hour)	Event
000	Coal fires maintained from 18 April 1977 Cell D - 11 hours 6 minutes Cell C - 1 hour 40 minutes
0041	Started reinjection of bed material to Cell D and Cell C (2 rpm on each feeder)
0045	Started limestone feed to Cell C
0122	Started preparation for Cell B light-off. Completed Cell B check-off list.
0125	Stopped bed material reinjection system
0153	Established coal flow into Cell B at 5,000 lbs/hr
0155	Opened C-B transfer gate
0205	Closed C-B transfer gate: coal fire established in Cell B
0213	Closed D-C transfer gate
0257	Initiated MFB unit shutdown after being advised by the MPC Control Room that the condensate tank water level was close to the reserve level.
0302	Suspended shutdown activities upon being advised by MPC that condensate water would be made available for MFB operation until 0330. Re-established coal feed to Cell B after flow had been stopped for five (5) minutes.
0303	Opened C-B transfer gate: Coal fire re-established in Cell B.
0311	Closed C-B transfer gate
0330	Initiated unit shutdown; Shut-off coal flow to Cell B.

3.5 Coal Handling System Modifications

Due to difficulties in continuously feeding high fines content, high moisture content coal into the MFB, concentrated efforts have been made to locate sources of double screened, (-1/2 + 1/4") dry coal.

Unavailability of this coal had necessitated changes to the coal handling system to provide double screening capability within the plant. The following changes have been completed or are underway:

- Installation of by-pass chute around MPC Coal Crusher,
- Installation of by-pass chute around MFB Coal Crusher,
- Installation of additional screen in Rotex Vibrating Screen for double screening,
- Installation of chute from discharge of oversized coal Redler into coal bunker No. 5.

3.6 Equipment Failures

Routine inspections of the MFB Plant were conducted after each operational run. Various equipment failures were noted during the course of these inspections and are described below.

3.6.1 Induced Draft Fan Failure

a. Nature of Failure: Thirty-three bolts which attach inlet cones to I.D. Fan housing sheared.

One blade on I.D. Fan rotor had large crack.

Housing of I.D. Fan cracked where directional vane on outlet of I.D. Fan is attached to housing

b. Repairs: Welded housing on either side of directional vane.

Rectangular section of cracked blade cut out and new piece welded in.

Sheared bolts drilled out, replaced with through bolts, double nutted and tack welded.

All remaining inlet cone bolts were double nutted and tack welded.

3.6.2 Grid Plate Distortion/Failure

a. Nature of Failure: The gridplate was noted to have failed in the following manner under three (3) feed nozzles:

-Distortion
-Local Overheating
-Failure (cracked)

b. Repairs: The three (3) failure spots were cut out and new sections of grid plate welded in.

3.6.3 Burner Throat Refractory Failure

a. Nature of Failure: An inspection of the CBC revealed that the refractory in the north and south oil burner throats had eroded away after only 337 hours of oil firing operation. The break-up and erosion of the refractory was attributed to overheating and breakdown of the throat liner insulation causing thermal stress in the refractory material.

b. Repairs: The throat liner insulation was replaced and a thicker coal of refractory was applied to the burner throats to prevent future overheating of the liner insulation.

3.6.4 Electrostatic Precipitator Structural Failure

a. Nature of Failure:	During an internal inspection of the precipitator it was discovered that several of the structural angles that support walkbeams across the width of the precipitator had failed in buckling.
b. Repairs:	The damaged structural members were replaced and kicker plates for the walkbeam supports which had been mistakenly omitted were installed.

3.7 Operational Limitations

In addition to the above equipment problems, coal and demineralized water supply problems have limited operation of the MFB unit.

Continued operation of the MFB furnace feed system has indicated that some components are sensitive to the presence of moisture in the coal and a high percentage of coal fines. It is necessary to use sized and dried coal in order to obtain stable operations. The availability of this type of coal was extremely limited and could not be obtained on a regular basis. At present, coal handling systems modifications have been completed in order to transport low moisture content coal into the MFB Plant and double screen the coal just prior to entering the main storage bunker.

The MFB unit, being tied into MPC Boiler No. 7 feedwater system, can only be operated when Boiler No. 7 is operated. In addition to this problem which has necessitated shutdown of the MFB several times, the MPC condensate water tank has enough storage to make only 100,000 gallons of water available to the MFB. At present, two (2) portable mobile demineralizer trucks have been brought on site to supplement the MPC supply and provide water for sustained three (3) cell operation.

IV. DESIGN AND ENGINEERING

4.1 General

Since start-up of the MFB Plant in December 1976 and operation of the unit since that time, it became apparent that due to the uniqueness of the plant and the first of a kind application of systems and equipment, it was necessary to modify certain features of the installation to:

- improve performance
- increase reliability

Accordingly, with ERDA authorization, various tasks to accomplish these purposes were undertaken by the PER New York Office design and engineering staff.

During this reporting period, work on several such tasks was initiated and was completed or continued. A brief description of each follows.

4.2 CO_2 Fire Inerting System

Long-term storage of high sulfur coals in confined areas such as bins and bunkers, tends to encourage accelerated oxidation. This action in turn creates local regions of elevated temperature which can eventually lead to fire. Investigation of the MFB coal handling and storage system indicated that Bunkers 2, 5 and 6 are candidate locations for such occurances as the result of intermittent operation of the MFB unit during the testing and evaluation program. To prevent such occurances, a fire detection and protection system was provided for Bunkers 2, 5 and 6.

Available systems were investigated and evaluated and it was determined that a temperature monitoring and alarm system would be provided for the bunkers in conjunction with a CO_2 fire inerting system. The monitoring system utilizes thermocouples and monitors 40 points in Bunker No. 2. Temperature read-out is provided at a control and alarm panel for each bunker and actuates visual and audible alarms when the temperature at any thermocouple exceeds 160°F. At that time an operator actuates the CO_2 system and manually controls the flow of CO_2 to the location where high temperature is indicated.

Design drawings and specification were prepared for this system and competitive bids for construction/installation were solicited. Bids were evaluated and forwarded to ERDA with recommendations for award of a fixed price lump sum contract. With ERDA's concurrence and authorization, a contract was awarded to Automatic Sprinkler Corp. who were oficially notified to proceed on 29 June 1977. Automatic Sprinkler was directed to coordinate their activities and schedule their work in such a manner as to avoid interfering with or delaying in any way, operation of the MFB unit.

4.3 Limestone Delivery

Design of the limestone delivery system for the MFB plant was based on delivery by a pneumatic transport trailer equipped with a blower. The trailer would inferface with the MFB delivery connection and the blower would pneumatically transport the sized limestone to the storage bunker approximately 93 ft. above grade.

The pneumatic transport trailer used for deliveries at Rivesville, has a capacity of 20-22 tons and utilizing the truck mounted blower, it takes 4-5 hr. to complete delivery. This was objectionable for two reasons: (1) the delivery truck utilized a high traffic area and (2) a demurrage charge was incurred for delivery time over one hour. Investigations indicated that the truck mounted blower had a capacity of 200-250 cfm while 550-600 cfm was required to produce transport velocities which would permit unloading within an hour.

A performance specification was prepared for a trailer mounted gasoline engine driven blower of the required capacity. The unit would be positioned at the delivery connection, mate with a connectoon on the truck provided for this purpose and be used in place of the truck blower. Competitive quotations were solicited and a purchase order issued to Schwitzer Industrial Products. To insure specification compliance and satisfactory operation a pre-shipping inspection was performed at the plant and the unit released for shipment to Rivesville. Delivery is scheduled for early July.

4. 4 In-Plant Office

The test instrumentation program which will provide for continuous MFB data acquisition under operating conditions, utilizes fairly large gas sampling and analysis equipment. The optimum location for this equipment to minimize tubing runs, is the westerly office cubicle in the Control Room - Laboratory area. Results Engineering personnel had been accommodated in that area and it became necessary to provide an alternate area. An evaluation of potential available areas indicated that an area on the roof of the Control Room - Laboratory area was most suitable for this purpose. To minimize costs, it was decided to utilize an in-plant, prefabricated structure which would be delivered to Rivesville in knock-down form and erected in-place by MFB maintenance personnel.

Specifications and a sketch were prepared and competitive quotations solicited. A purchase order was issued to Global Equipment Co. for delivery in mid July.

To provide access to the office area, it was necessary to modify existing steel access platforms and install a new access stair. Sketches were prepared, local steel fabricators contacted, and competitive quotations solicited. A purchase order for fabrication and erection is scheduled to be issued in early July.

4.5 Waste Sample Storage Silos

Under normal full load operating conditions, the MFB unit produces approximately 7 tons of waste material per hour consisting of spent bed material and fly ash. Several programs are underway to explore useful applications of this material. To facilitate obtaining samples, a waste sample storage system was designed.

Anticipating that future MFB units would be set up in a different manner than Rivesville and the contributing elements of the waste products could have independent applications; provision was made for isolated sampling. Three silos were provided, one each for spent bed material, normal fly ash (from the electrostatic precipitator and Cyclone No. 2) and carbonous fly ash (from Cyclone No. 1). The existing pneumatic transport system to the MPC ash silo was tapped into pneumatically operated knife gate isolating valves provided and transport piping runs extended to the individual silos. The silos were mounted on a steel structure to permit truck access below for unloading. Appropriate instrumentation and controls were provided.

At that time, ERDA was in the process of reviewing the overall program sampling requirements and was advised by EPA of their requirement for 4000 tons of MFB waste products as discussed in detail in the following item. The system as designed would not practically provide for such quantities. A re-evaluation of the sampling program requirements was made and it was determined that a single 80 Ton silo into which the waste products would be programmed on an as required basis would be satisfactory. Design of the modified system was initiated and at the close of the reporting period was 80% complete.

Upon completion and review by ERDA, EDA and MPC, competitive construction bids will be solicited.

4.6 EPA Test Program

The Federal Environmental Protection Administration in order to establish performance and pollution standards for fluidized bed combustion units, will undertake a test program at the Rivesville MFB with ERDA authorization. To assist EPA, subcontract services of Battelle and Monsanto will be utilized.

In conjunction with this program, EPA also desires to establish performance and pollution standards for electrostatic precipitators and has solicited a proposal from the Buell Emission Control Division of Envirotech Corp. In addition to these programs, PER/FWEC will be involved in testing the performance and operating characteristics of the MFB. As a result of these activities, three test programs involving fire subcontractors, will be running concurrently at the MFB plant.

During this reporting period, PER coordinated the requirements of all participants regarding the type and method of sampling and location and size of ports for sampling and making traverses. In conjunction with MPC, a location for the EPA sampling/analysis trailers was established, drawings prepared for sampling duct runs and electric service. Designs were also prepared for a stack access platform, ladder and sampler traverse track required by EPA/Battelle.

As part of the EPA test program, Battelle will investigate groundwater contamination resulting from the leaching action of rainwater on MFB waste products deposited in a sanitary landfill. To accomplish this, Battelle requires approximately 4000 tons of MFB waste products. Investigations were made to acquire this material utilizing the three waste material sample silos described in the previous item. The evaluations indicate that because of the lengthy time period required to collect such quantities, two serious disadvantages would result:

- interference with normal MPC and MFB traffic, resulting from limestone delivery, MPC ash removal, MFB ash and spent bed material removal, MPC and MFB coal deliveries MPC refuse removal,
- interference with MFB operation resulting from need to shut down pneumatic transport systems each time shift of material to sample silos is made.

Further investigations indicated that a single large silo of approximately 80 tons capacity would replace the three small silos without incurring the disadvantages noted above.

4.7 Coal Dryer

The Bureau of Mines at the request of ERDA, visited the MFB plant to evaluate the safety aspects of the coal dryer installation. As the result of their evaluation, PER engineering investigated four alternate configurations and developed a matrix evaluating the safety features of each. The matrix and skethces of each of the alternates were presented to ERDA at a meeting in early May and it was decided that the coal dryer system would not be modified or used and only dry, double screened coal would be purchased for use in the MFB.

4.8 MFB Light-Off

Operation of the MFB unit since December 1976, indicated that approximately 10 hr. was required from start of ignition of the auxiliary fuel oil burners to establishing a coal fire in Cell D. This time period was considered overly long and resulted in consumption of large quantities of fuel oil and excessive use of feedwater for blowdown during this period. Additionally, it was extremely desirable to reduce this time to a more reasonable period for availability of the MFB for power generating purposes.

To reduce this time period, PER Engineering and FWEC evaluated the following alternate light-off schemes:

- Installation of oil fired duct burner(s),
- Installation of duct air heater(s),
- Installation of additional oil fired burner in Cell D,
- Installation of individual, over the bed, oil fired burners in Cell A thru C,
- Separation of the air plenum chambers for local fluidiation of individual cell sections,
- Injection of externally heated hot bed material to the individual cells,
- Individual small burners in each cell to heat the locally fluidized bed sections,
- Injection and ignition of liquid fuel in the individual cells,
- Combination of the above schemes.

The results of the evaluation were presented to ERDA at a meeting on 5 May 1977. Following discussion and evaluation of alternates it was agreed that the most practical immediate solution consisted of modifying the existing light-off oil burners to increase their capacity from 4 x 10^6 BTU/hr to 7 x 10^6 BTU/hr each and repositioning the burners so that the flame patterns are closer to the fluidized bed. FWEC, manufacturer of the boiler, was authorized to proceed with the necessary modifications during the next MFB outage.

4.9 Coal Feed System

Continued operation of the MFB unit highlighted operational difficulties with the coal feed system. In-house investigations for both immediate and long-term solutions were initiated which resulted in the temporary use of only double screened, dry coal. Feed problems were discussed with FWEC and their recommendations for improvement solicited. FWEC investigated and recommended installation of a gravimetric weight belt feeder for coal to replace the rotary feeder. The belt feeder eliminates blow back of air, bridging and ratholing of coal in the bins, and erratic coal feed rates resulting from build-up of material in the rotor assemblies. This recommendation was presented to ERDA at a meeting on May 5th and, based on ERDA's concurrence, FWEC was authorized to proceed with procurement and installation.

On May 26th, a workshop session was held at PER's offices, attended by all knowledgeable parties, at which a complete airing of all coal feed related problems and solutions were analyzed and evaluated. As a result of that session, the following actions were initiated:

- air supply to the vibrating feeder tables revised to provide for automatic control to eliminate operator error which in the past resulted in needle pluggage on several occasions.
- feed pipes from the vibrating feeder table to the cells modified by installation of an eductor at each pipe bend to provide transport air supply at 12-15 psig rather than depending on transport at 1.5 - 2 psig air supplied from the auxiliary air fan through the vibrating feeder table.
- investigate the feasibility and cost of increasing the diameter of the feed pipes from 1½ inch I.D. to 2 inch size to discourage pluggage.
- replace the rotary feed air lock valves at the vibrating feeder table(s) with new air locks having polished stainless steel rotor assemblies, to discourage pluggage of the rotors.

In addition, modify the Rotex screen to provide double screening capability and modify the coal transport system to return both fines and oversize coal to MPC. Additional screens were ordered and sketches prepared for the necessary modifications.

Coal Feed System

Operational problems with the MFB coal feed system and planned remedial actions were described in previous monthly progress reports. During this reporting period, the following activities occurred:

The following follow-up action was initiated as the result of the workshop session:

- FWEC was authorized to provide automatic control of the air supply to the vibrating feeder tables and to install weigh belt feeders to replace the coal rotary feeders. FWEC has placed orders for the required equipment/materials and is scheduling installation to start in mid-August to coincide with shutdown of the MFB for installation of test instrumentation.

- Replacement rotary air locks with stainless steel rotor assemblies were ordered from Neuman Industries Inc. Seven units were ordered to replace the existing Allen-Sherman-Hoff air locks. In order to expedite delivery, the Neuman units were ordered with "as is" rotor assemblies. On arrival at Rivesville, scheduled for early July, the rotor assemblies will be inspected for surface smoothness and, if necessary, they will be disassembled for additional polishing of the vanes to insure free motion of coal. Also, depending on the success achieved with consistent transport of fuel resulting from use of the educators, the rotary air locks may be completely eliminated.

4.10 Additional Activities

During this reporting period, New York engineering was also engaged in the following activities:

- Feedwater to the MFB is supplied by MPC, utilizing the deaerator and boiler feed pump for their No. 7 boiler. The MFB is solely dependent upon this source and can only be operated at the time that No. 7 unit is on the line and an adequate supply of feedwater is available. The decision to rely upon MPC feedwater supply was based on a cost benefit analysis prepared at the time of the MFB design in 1973-74. To reduce reliance on the MPC supply which has recently imposed severe limitations on operation of the MFB, investigations were initiated to find alternate sources of water to permit operation of the MFB during the period that steam is not exported to MPC and condensate is not recovered. Seven alternate schemes were evaluated, including means of recovering and condensing the waste stream. On an interim basis, arrangements were made to rent two 100 GPM portable demineralizer units, supplied by potable water from the Rivesville-Fairmont domestic supply.

- EPA test program requirements were coordinated with PER test program and location of test trailers established.

- Air supply for the fuel feed educators was investigated with the most feasible source on an interim basis being the auxiliary limestone blower. Permanent supply will be furnished by a long lead delivery electric motor driven, high capacity blower.

- A study was initiated to evaluate modifications to existing MPC boiler feedwater pump No. 4 to permit operation of the MFB at the time when no MPC units (Nos. 7 or 8) are on the line. The results of the study indicate that it is possible to modify the pump to reduce the capacity for compatibility with the MFB. However, cold start-up of Turbogenerator No. 5 with the MFB unit must be further investigated, before recommendations for modification of BFW Pump No. 4 can be made.

- Investigations were conducted into methods of cooling the instrument air supply which at the present time is above the recommended working range of the refrigerator-dryer.

- Sketches were prepared, additional screens ordered and visits made to Rivesville to assist the field forces in modifying the discharge chutes from the Rotex coal screen to provide double screening capability and return both fines and oversize coal to Bunker No. 5. By-pass chutes were also developed for the MFB crusher and MPC crusher to provide direct passage of delivered sized coal to Bunker No. 5. At the close of the reporting period, installation of chutes at the Rotex screen and by-pass around the MFB crusher were completed and material ordered for the by-pass around the MPC crusher.

MFB STEAM GENERATOR DESIGN, SYSTEMS DEVELOPMENT, MODEL TESTING, OPERATING INSTRUCTIONS AND SUPPLEMENTAL CONSTRUCTION

5.1 Background and Work Scope

This Interim Report describes the work completed by Foster Wheeler Energy Corporation from July 1, 1976 through June 30, 1977 under Subcontract No. 2-58-2126 to Pope, Evans and Robbins Incorporated. Work under this contract covers the design of a 300,000 lb/hr capacity, 1250 psig, 925°F superheated steam generator burning coal in a fluidized bed of limestone, the development and design of a calcium sulfate regeneration system, component testing for development support of the system design, preparation of operating instructions and supplemental field construction at the Rivesville Plant.

The work scope for this contract is divided into ten phases, Phase I through Phase V were completed prior to this reporting period and the results presented in previous Interim Reports. This report describes the scope of work and activities for the remaining phases which were worked on from July 1, 1976 to June 30, 1977. The work scope for those phases are as follows:

5.2 Phase VI - Operating Manual/Steam Generator Test Program

Prepare an operating instruction manual for the Rivesville steam generator and auxiliary equipment supplied by FWEC and prepare a steam generator test program for testing all systems installed at Rivesville.

5.2.1 Task 1

The operating and instruction manual will include a detailed description of the steam generator, its components and auxiliary equipment supplied by FWEC. A summary of performance information including temperatures, pressures and flow of steam, gas, air and fuel at full and minimum load points on the unit will be included. Safety precautions will be defined and procedures for hydrostatic testing, refractory drying, boiling out and chemical cleaning will be described. Instructions for boiler start-up, normal operation and shut-down, plus emergency situations will also be included. As an appendix to this manual the operating and maintenance instruction for all auxiliary equipment supplied by FWEC will be included.

5.2.2 Steam Generator Test Program

A test program will be prepared defining all tests, instrumentation and testing procedures required for data acquisition at the Rivesville unit. The test program will provide the ability to determine unit efficiency, combustion efficiency within each of the cells, heat transfer coefficients in the waterwalls and horizontal tube coils, flue gas analysis, fly ash and bed material analysis and effects of erosion and corrosion.

5.3 Phase VII - Field Construction and Supervision

This phase involves performing the following field construction items which are supplemental to the boiler construction performed under a separate subcontract and providing construction supervision.

5.3.1

Perform initial boil out and chemical cleaning of the fluidized bed steam generator after systems have been checked out and operational.

5.3.2

Supply and install sway snubbers on the boiler circulation piping to dampen vibrations imposed by the recirculation pumps. Supply of the sway snubbers includes snubber selection, procurement and installation on the boiler recirculation pump piping.

5.3.3

Install perforated sheet, air distribution grid assemblies in Cells A, B, C and D of the fluidized bed steam generator.

5.3.4

Install safety screens around the ignition burners on the fluidized bed steam generator.

5.3.5

Modify the air flow air dampers in Cells A, B, C and D of the fluidized bed steam generator to provide better air flow control characteristics.

5.3.6

Modify four coal injection feed sleeves to extend beyond the boiler circuit feeders on the fluidized bed steam generator to eliminate interference with the flange on the coal feed pipes. This modification will require revision to the length of the coal feed pipes to be inserted within these four sleeves.

5.3.7

Install the finishing superheater outlet header support framing required to accept additional load imposed by the steam piping.

5.3.8

Modify the plenum seal plates to permit removal of grid plate assemblies.

5.3.9

Install air flow control damper drive support frames and mount the damper drives (supplied by others) on the support frames attaching the damper drives to the damper linkage.

5.3.10

Install seal and aspirating piping required for the observation doors and coal feed pipe sleeves.

5.3.11

Install thermocouple assemblies to measure temperature of the D Cell air distribution grid.

5.3.12

Install thermocouples on the primary and finishing superheater outlet tubes and on the steam drum.

5.3.13

Install superheater header drains.

5.3.14

Install an air foil in the air duct between the air preheater and the steam generator for measurement of air flow.

5.4 Phase VIII - Preparation of Operating Instruction Manual

Work under this phase is divided into two tasks, one covering the preparation of preliminary operating instruction manual and the other covering the preparation of a detailed operating instruction manual.

5.4.1 Task I - Preliminary Operating Instruction Manual

This manual will contain procedures for start-up and operation of the MFB system, as well as safety precautions, maintenance, and spare part requirements. The manual will describe systems and equipment installed at the Rivesville site.

This preliminary manual will contain an overall system description, individual component descriptions and appropriate sketches showing system component interfaces. A generalized start-up and shut-down procedure will be included.

5.4.2 Task II - Detailed Operating Instruction Manual

A detailed manual for the fluidized bed system will be prepared which will expand upon the preliminary manual to provide specific plant operating procedures.

5.5 Phase IX - System Evaluation and Steam Generator Design Manual

Work under this phase is divided into two tasks, one covering the systems evaluation and the other covering preparation of a design manual for the Rivesville steam generator.

5.5.1 Task I - Systems Evaluation

The systems evaluation will consist of a study to develop alternative light-off techniques in addition to the existing ignition burners. Evaluation of the coal handling system, including recommendations for modifications to improve the system operation will be performed. Other evaluations of systems performance, modifications to the Rivesville facility or operating procedures will be performed as requested.

5.5.2 Task II - Steam Generator Design Manual

This manual will describe the design of the multicell fluidized bed steam generator which was designed and constructed by FWEC at the Rivesville Plant. The design manual will include general arrangement drawings of the steam generator and specifications for the FWEC supplied auxiliary equipment. The design manual will not include any proprietary information and will be specific to the Rivesville multicell fluidized bed steam generator.

5.6 Phase X - Test Instrumentation

Work under this phase includes the design, fabrication and installation of test instrumentation at the Rivesville facility in accordance with the test plan prepared under Phase VI (revised June 1977). In addition to the test instrumentation, a second task under this phase includes the design, procurement and installation of miscellaneous equipment to modify the Rivesville facility.

5.6.1 Task I - Test Instrumentation

Work under this task consists of the following items:

a. Detailed instrumentation design - Determination of the exact type and location of instrumentation required, including test connections, to be compatible with the "as built" configuration of the plant. Instruments required for readout of data will be identified, including range and accuracy required. Spare parts and necessary calibration equipment will also be identified.

b. Material procurement - Preparation of material recommendations and purchase orders for test instruments defined in Item (a) including material required for installation and interconnections.

c. Instrumentation installation - Installation of the test instrumentation defined in Item (a) and furnishing of required manpower and equipment.

d. Instrumentation checkout and calibration - Supervision for checkout and calibration of the test instrumentation defined in Item (a).

5.6.2 Task II - Miscellaneous Work

In addition to the test instrumentation described under Task I, the following miscellaneous work is included under this phase:

a. Observation doors - Design, procurement and installation of two additional observation doors and ports on the steam generator to provide for additional viewing capability of the carbon burn-up cell.

b. Fuel feed air supply control - Design, procurement and installation of an automatic control system for control of the auxiliary air valves to regulate air flow to the vibrating distributor tables.

c. Coal belt feeders - Procurement and installation of seven gravimetric belt feeders to replace the existing rotary coal feeders.

d. Modification of ignition burners - Procurement and installation of components and materials required to increase the capacity of each existing igntion burner from 4 million Btu/hr to 7 million Btu/hr.

5.7 Work Previously Completed

Work on Phase VI included preparation of a Boiler Operating Instruction Manual and a Test Instrumentation Plan for the Rivesville Unit. Work on the Operating Instruction Manual for the Steam Generator was completed in December 1975 and work on the Test Instrumentation Plan was completed in January 1976. However, following an ERDA review in May 1976, it was decided to change the original approach to the Test Instrumentation and an additional review in this area was carried out by FWEC during this report period. Work on Phase VII was partially completed during this report period. Work on Phase VII was partially completed during the time period covered by the 1976 Interim Report and continued into the time period covered by this Interim Report and is reported herein. Work on Phase VIII had been initiated as of the previous Interim Report and the remaining work is reported herein.

5.8 Work Completed - Phase VI

Development of a revised Test Instrumentation Plan, based on ERDA's review and estimates of costs was completed and in October 1976, FWEC was authorized to begin engineering work for the specification and design of the test instrumentation and plant modification. Subsequent work in this area is reported under the Phase X activities.

5.9 Work Completed - Phase VII

Work on this phase began in April 1976 and has been completed as follows:

5.9.1

Steam generator chemical cleaning - completed July 1976.

5.9.2

Supply and installation of sway snubbers - completed July 1976.

5.9.3

Installation of boiler air distributors - completed July 1976.

5.9.4

Ignition burners safety screens - completed September 1976.

5.9.5

Air flow control damper modifications - completed April 1976.

5.9.6

Coal injection feed sleeve modifications - completed September 1976.

5.9.7

Superheater outlet header support - completed September 1976.

5.9.8

Plenum seal plate modifications - completed August 1976.

5.9.9

Seal and aspirating air systems - completed August 1976.

5.9.10

Thermocouple assemblies - completed September 1976.

5.9.11

Superheater header drains - completed August 1976.

5.9.12

Air flow measurement air foil - completed August 1976.

5.10 Work Completed - Phase VIII

Verification of equipment locations and designations at Rivesville was carried out by a jobsite visit by FWEC personnel on July 7 - 9, 1976. As a result of this visit, comments received from operating personnel were incorporated into the operating instruction manual and revised manual sections were issued for review. Based on the review on August 30, 1976, two FWEC representatives began work at the Rivesville jobsite to expand and clarify sections of the manual. During this jobsite effort, system modifications that had not previously been incorporated in the operating instruction were identified. In October 1976, all of the sections of the operating instruction manual were submitted except for the sections covering the communications system and the fire protection systems.

In November 1976, approval of the manual was received and the manual was printed, assembled and bound and twenty-five copies delivered in January 1977.

Since completion of the manual, only minor efforts have been carried out involving revisions of the various sections based on operating information received from the Rivesville Plant.

5.11 Work Completed - Phase IX

During March 1977, several alternative light-off techniques were evaluated in order to reduce the time required to start coal ignition in Cell D. These included: (1) Evaluation of the plenum temperature differential, (2) direct fired air preheater, (3) increased Cell B ignitor capacity, (4) segmental fluidization, (5) oil injection into Cell D, and (6) separate ignition burners for Cells A, B and C. In addition to these alternate light-off techniques, FWEC recommended that the present rotary coal feeders be removed and replaced with gravimetric belt feeders. The following is a description of the above evaluations:

5.11.1 Temperature Evaluation

Evaluations for use of direct fired heaters required a review of the plenum design to establish the temperature differential it could withstand. The plenum chamber, air duct and dampers are constructed of carbon steel and can withstand an operating temperature of 800°F. The plenum chamber is welded directly to the boiler pressure parts and temperature differentials between the boiler pressure parts and the plenum chamber results in differential expansion. The evaluation indicated that no problems could be expected if a temperature differential of 300°F or less is maintained between the plenum walls and the boiler water temperature. At temperature differentials in excess of this level, some stress cracking could occur. However, such stress cracks could be repaired by welding over cracked areas with expansion boot covers. The maximum temperature that could be withstood by the present plenum design is 800°F, however, this would necessitate replacement of the air control damper gaskets which have a temperature limitation of 750°F.

5.11.2 Direct Fired Air Preheater

Three alternate methods of direct fired air preheater installations were evaluated. The first alternate involved a direct fired air preheater supply 20,000 lb/hr of preheated air at 800°F to the D Cell plenum chamber. This alternate had an estimated cost of equipment plus installation of $85,000.00 with a lead time requirement of approximately eighteen weeks. The second alternate involved preheating combustion air to 800°F in the main combustion air duct supplying all the boiler cells. Two 10,000,000 Btu/hr burners were considered, mounted in the combustion air duct at the main operating floor elevation. With the burners at this location, 800°F air could be directed to any one or all four cells. To protect the sidewalls of the duct in the vicinity of the burners, stainless steel radiation shields mounted off the duct sidewalls would be required. Combustion air for the burners would be supplied by the auxiliary foreced draft fan. The cost for this alternate was estimated at $38,000.00 with an estimated lead time of twenty weeks. The third alternate evaluated the use of an electric element for heating combustion air for Cell D. It was estimated that 1½ MW_e of electric power would be required to heat 20,000 lb/hr of air to 1050°F for ignition of Cell D. The heating elements would be located directly below the air distributor which is designed for temperatures in excess of 1050°F, thereby permitting the air to heated to coal ignition temperature without use of the existing ignition burners. Problems for implementing this alternate at Rivesville could be expected due to limited power availability. The cost of this system was estimated at $120 $120,000.00 plus installation cost.

5.11.3 Increased Cell D Ignition Burner Capacity

With the present 4 million Btu/hr capacity ignition burners, it takes approximately 8 to 10 hours from the lighting of the ignition burners to the time of coal ignition.

To reduce the light-off time, two alternates were evaluated. The first alternate considered increasing the velocity and temperature of the burner gases at the impact point with the bed, by increasing the velocity of the burner gases leaving the burner exhaust. This could be accomplished by moving the burner closer to the steam generator wall by modifying the burner mounting boxes to permit reducing the burner exhaust port from 11 inches to approximately 7 inch diameter. The second alternate involved replacement of the existing 4 million Btu/hr burners with larger burners capable of 7.5 million Btu/hr each. The purchase of the new 7.5 million Btu/hr burners was estimated at approximately $35,000.00. It was recommended that installation of the larger capacity burners be implemented.

5.11.4 Segmental Fluidization

Evaluation of segmetal fluidization was carried out for the purpose of improving light-off procedures and turn down ability. Segmental fluidization could be accomplished in Cells A, B or C by dividing the air plenum in half with a baffle. An arrangement for implementing this concept while maintaining use of the existing air control dampers was considered and the estimated cost to supply and install the baffle plates, damper operators and controls was approximately $75,000.00.

5.11.5 Oil Injection Into Cell D

Installation of a retractable, auxiliary oil gun/nozzle for direct oil injection into Cell D was considered. Installation of two oil gun/nozzles was evaluated with each nozzle adjacent to each ignition burner and angled so that the oil spray would pass through the flame of the oil ignition burners. Each oil gun/nozzle would be sized for 10 million Btu/hr heat input, firing No. 2 fuel oil. This approach would provide for rapid heat up of Cell D prior to coal ignition and was estimated to cost $15,000.00.

5.11.6 Separate Ignition Burners for Cells A, B, and C

The flat flame oil ignition burners previously evaluated in May 1976 were re-evaluated and due to uncertainties of the ability for these ignitors to heat the bed material and concern about direct flame impingement on heat transfer surfaces, further consideration of the flat flame oil ignitors was not pursued.

An alternate technique for direct injection of high temperature gases into Cells A, B and evaluated. This involved installation of a manifold for distribution of hot gases (1800°F) to multiple points along the sidewall of Cells A, B and C. Three alternatives were evaluated and the results indicated that the costs would be high with attendant installation problems. The costs for the three alternates evaluated ranged from $245,000.00 to $360,000.00. Due to these high costs, injection of hot gases through the boiler sidewall was not recommended.

5.11.7 Coal Belt Feeders

Maintaining continuous coal flow to the boiler cells has been difficult to achieve with the existing rotary feeders. Coal feed problems associated with these feeders have caused several unit shutdowns and considerable instability in operation during coal firing. Problems associated with the rotary feeders have included plugged feeder flights, feeder "jogging", feeder jamming, inability to maintain coal flow at the feeder inlet and inability to control the coal flow rate.

Based on utility plant experience with belt feed systems, a modification to remove the existing rotary feeders and replace them with gravimetric belt feeders was recommended.

Due to limited space at the Rivesville Plant, special feeders are required. A total of seven units are needed to replace the seven rotary feeders. The total cost for providing the belt feeders was initially estimated at $125,000.00.

At a meeting held at ERDA in April 1977, during which the various alternate light-off improvement techniques were reviewed, FWEC was released to procure and install the larger capacity ignition burners for Cell D.

A workshop session was convened at PER offices to analyze coal handling and coal feed problems. In addition to the techniques previously mentioned, an alternate involving the use of a vertical head of coal for the pressure seal between the vibrating feeder table and the bunkers was discussed. This alternative requires the gravimetric belt feeders to be located at a lower elevation thereby providing the vertical height needed for the pressure seal. This alternative was considered based on information concerning problems associated with the rotary air locks. The problems included the inability of the air locks to continuously pass coal to the higher pressure feeder tables without coal adhering to the rotor assemblies of the air lock valves. Subsequent to the session, it was decided to replace the rotary air locks and FWEC was authorized to proceed with procurement and installation of the belt feeder system originally recommended to ERDA in the April 1977 meeting.

FWEC was requested to expedite the approved modifications so that installation could occur during the outage scheduled to begin on August 1, 1977 for installation of the test instrumentation. In addition to the coal feed system modifications and increased burner capacities, FWEC was authorized to proceed with two additional observation doors on Cell D and automation of the air flow to the coal feeder tables.

5.12 Work Completed - Phase X

In October 1976, FWEC was released to begin engineering work for the specification and design of test instrumentation and plant modification for performance testing at Rivesville based on the test plan previously prepared under Phase VI.

Based on the initial schedule for this phase, approval for purchase of equipment was required by January 15, 1977 to avoid delays and potential cost increases. Release for purchase of equipment was finally received in March 1977 and efforts to expedite procurement were initiated. The initial schedule for this phase called for a unit outage in June 1977 for installation. However, due to the delay in release for procurement, the outage date slipped to August 1, 1977.

All items for the test instrumentation installation have been purchased and are expected at the jobsite in time for the August 1977 outage.

During June 1977, the original test plan issued under Phase VI was revised to incorporate changes associated with eliminating the automatic data acquisition system originally requested by ERDA. The revised test instrumentation plan will be issued to PER in July 1977.

VI. AUXILIARY PLANT SYSTEMS CONSTRUCTION

6.1 General

Work on Construction Subcontract No. 2 for procurement and installation of auxiliary plant equipment, interconnecting piping and electrical work, continued and was completed by Champion Construction & Engineering Co., Inc., and their subcontractors. The work included all the base contract items, Amendments and Change Orders. All work was completed by mid-March 1977 at which time Champion and their subcontractors demobilized and moved off the site.

6.2 Performance of Work

Early in the reporting period the Contractor completed installation of all major items of equipment. Inspections of completed installations were conducted by PER Operations Group and punch lists prepared. When corrective action was completed by the contractor, the equipment/systems were operated through their specified performance range and final acceptance from the contractor documented. By the end of November 1976, all major systems and equipment had been accepted and run-in to permit the first successful firing of the MFB unit in early December 1976.

The contractor continued installation/construction and correction of punch list items till mid-March. At that time all outstanding construction items were accepted as complete and maintenance and responsibility transferred to the PER Rivesville Operations Group.

Included among the contractors final activities was procurement and stocking of some designated spare parts.

APPENDIX A

FBM OPERATING CONDITIONS LOG

Test No. 644-1 Date October 25, 1976

	LIGHT-OFF		TRANSITION TO DEEP BED	
Time Duration (Hours)		2		2
Materials Rates				
bed depth (inches)		10.5		22.2
coal feed rate (lbs/hr)		319		340
limestone feed rate (lbs/hr)		0		0
salt feed rate (lbs/hr)		0		0
fly ash removal rate (lbs/hr)		22.2		48
bed removal rate (lbs/hr)				
classifier removal rate (lbs/hr)		0		0
dust loading (lbs/hr)		12.75		22.6
Air and Gas Rates				
main air flow rate (lbs/hr)		2730		2730
light-off burner air rate (lbs/hr)		0		0
light-off burner gas rate (CFH)		0		0
duct burner air rate (lbs/hr)		1000		900
duct burner gas rate (CFH)		205		190
overfire air rate (lbs/hr)		378		378
coal-feed air rate (lbs/hr)		440		430
classifier air rate (lbs/hr)		--		--
fly ash removal air rate (lbs/hr)		450		450
stack gas flow rate (lbs/hr)		7600		6800
Water Rates				
horizontal bundle water rate (lb/hr)				
steam rate (lbs/hr)		1550		1750
make-up (steam drum) rate (GPM)		--		--
economizer rate (GPM)		25.2		25.6
water door rate (GPM)		3.2		2.5
support tube rate (GPM)				
duct burner rate (GPM		--		--

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Test No. 644-1 Date October 25, 1976

Temperatures (°F)

	LIGHT-OFF		TRANSITION TO DEEP BED	
air @ FD fan inlet		70		70
air @ FD fan discharge		100		100
air after preheater		340		320
air @ plenum		570		510
gas below steam drum		1690		1620
gas before economizer		1180		1190
gas after economizer		970		940
gas after preheater		660		630
gas before ID fan		530		500
bed (average)		1950		1690
inlet to classifier		--		--
return from classifier		--		--
horizontal bundle inlet				
horizontal bundle outlet				
economizer inlet		62		62
economizer outlet		120		130
water door inlet		62		62
water door outlet		150		154
support tube inlet				
support tube outlet				
air distributor (average)		--		--
O_2 in flue gas (vol %)		4.5		4.5
Mean Particle Size (Microns)				

FBM OPERATING CONDITIONS LOG

Test No. 644-6 Date Nov. 23, 1976

Time Duration (Hours)	3.5	2.25	2.75	
Materials Rates				
bed depth (inches)	18-27	25	15	
coal feed rate (lbs/hr)	274	320 → 0	0	
limestone feed rate (lbs/hr)	436	0	0	
salt feed rate (lbs/hr)	-	-	-	
fly ash removal rate (lbs/hr)	83.8	303(Avg.)	313	
bed removal rate (lbs/hr)	0	-	-	
classifier removal rate (lbs/hr)	-	-	-	
dust loading (lbs/ft^3)	-	-	97	
high carbon fly ash rate (lb/hr)	-	740(Avg)	875	
Air and Gas Rates				
main air flow rate (lbs/hr)	3700	3600	2600	
light-off burner air rate (lbs/hr)	-	-	-	
light-off burner gas rate (CFH)	-	-	-	
duct burner air rate (lbs/hr)	-	-	-	
duct burner gas rate (CFH)	-	-	-	
fly ash injection air rate (lb/hr)	378	366	366	
coal-feed air rate (lbs/hr)	385	430	211	
classifier air rate (lbs/hr)	-	-	-	
fly ash removal air rate (lbs/hr)	-	-	-	
stack gas flow rate (lbs/hr)	8400	7900	6900	
Water Rates				
horizontal bundle water rate (lb/hr)	-	-	-	
steam rate (lbs/hr)	750	1650	1580	
make-up (steam drum) rate (GPM)	-	-	-	
economizer rate (GPM)	25.2	25.0	24.8	
water door rate (GPM)	2.3	1.7	2.5	
support tube rate (GPM)	-	-	-	
duct burner rate (GPM	-	-	-	

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Test No. 644-6 Date Nov. 23, 1976

Temperatures (°F)				
air @ FD fan inlet	42	45	41	
air @ FD fan discharge	79	80	81	
air after preheater				
air @ plenum	230	270	250	
gas below steam drum	1550	1880	1950	
gas before economizer	1130	1350	1300	
gas after economizer	920	1040	--	
gas after preheater	650	750	700	
gas before ID fan	400	400	390	
bed (average)	1560	1900	1970	
inlet to classifier	--	--	--	
return from classifier	--	--	--	
horizontal bundle inlet	--	--	--	
horizontal bundle outlet	--	--	--	
economizer inlet	50	51	50	
economizer outlet	98	130	116	
water door inlet	50	50	50	
water door outlet	90	132	130	
support tube inlet	--	--	--	
support tube outlet	--	--	--	
air distributor (average)				
O_2 in flue gas (vol %)	4.2	4.0	1.5	
Mean Particle Size (Microns)				

FBM OPERATING CONDITIONS LOG

st No. 644-7 Two feeder test Date Nov. 30, 1976

Time Duration (Hours)	2-1/2	1	3-1/2	
Materials Rates	LIGHT-OFF	TRANSITION FROM COAL TO FLY ASH		SHUT DOWN
bed depth (inches)			20.4	
coal feed rate (lbs/hr)			0.0	
limestone feed rate (lbs/hr)			0.0	
high carbon fly ash feed rate (lb/hr)			862.5	
low carbon fly ash collection rate (lb/hr)			190	
bed removal rate (lbs/hr)			--	
classifier removal rate (lbs/hr)			--	
dust loading (lb/hr)			74	
Air and Gas Rates				
main air flow rate (lbs/hr)			3150	
light-off burner air rate (lbs/hr)			0.0	
light-off burner gas rate (CFH)			0.0	
duct burner air rate (lbs/hr)			0.0	
duct burner gas rate (CFH)			0.0	
fly ash injection air rate (lb/hr)			585	
coal-feed air rate (lbs/hr)			97.7	
classifier air rate (lbs/hr)			--	
fly ash removal air rate (lbs/hr)			420	
stack gas flow rate (lbs/hr)			7300	
Water Rates				
horizontal bundle water rate (lb/hr)				
steam rate (lbs/hr)			1475	
make-up (steam drum) rate (GPM)			--	
economizer rate (GPM)			13.75	
water door rate (GPM)			2.2	
support tube rate (GPM)			--	
duct burner rate (GPM			0.0	

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Test No. 644-7 Two feeder test Date Nov. 30, 1976

mperatures (°F)

air @ FD fan inlet	LIGHT OFF	TRANSITION FROM COAL TO FLY ASH	35	SHUT DOWN
air @ FD fan discharge			78	
air after preheater			--	
air @ plenum			220	
gas below steam drum			1834	
gas before economizer			1255	
gas after economizer			850	
gas after preheater			600	
gas before ID fan			400	
bed (average)			1835	
inlet to classifier			--	
return from classifier			--	
horizontal bundle inlet				
horizontal bundle outlet				
economizer inlet			50	
economizer outlet			162	
water door inlet			50	
water door outlet			104	
support tube inlet				
support tube outlet				
air distributor (average)			--	
O_2 in flue gas (vol %)			1.1	
Mean Particle Size (Microns)				

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FBM OPERATING CONDITIONS LOG

Test No. 644-8 One feeder test　　Date Dec. 2, 1976

Time Duration (Hours)	2-1/2	2-1/2	3	
Materials Rates	LIGHT-OFF	TRANSITION TO FLY ASH FROM COAL		
bed depth (inches)			17.3	
coal feed rate (lbs/hr)			0	
limestone feed rate (lbs/hr)			0	
high carbon fly ash feed rate (lb/hr)			678	
low carbon fly ash collection rate (lb/hr)			227	
bed removal rate (lbs/hr)			--	
classifier removal rate (lbs/hr)			--	
dust loading (lb/hr)			60	
Air and Gas Rates				
main air flow rate (lbs/hr)			2760	
light-off burner air rate (lbs/hr)			--	
light-off burner gas rate (CFH)			--	
duct burner air rate (lbs/hr)			--	
duct burner gas rate (CFH)			--	
fly ash injection air rate (lb/hr)			820	
coal-feed air rate (lbs/hr)			158	
classifier air rate (lbs/hr)			--	
fly ash removal air rate (lbs/hr)			--	
stack gas flow rate (lbs/hr)			6550	
Water Rates				
horizontal bundle water rate (lb/hr)				
steam rate (lbs/hr)			1400	
make-up (steam drum) rate (GPM)			--	
economizer rate (GPM)			10.2	
water door rate (GPM)			1.6	
support tube rate (GPM)				
duct burner rate (GPM			--	

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Test No. 644-8 One feeder test　　Date Dec. 2, 1976

Temperatures (°F)

air @ FD fan inlet			51.4	
air @ FD fan discharge	LIGHT-OFF	TRANSITION TO FLY ASH	90.6	
air after preheater			--	
air @ plenum			190	
gas below steam drum			1895	
gas before economizer			1210	
gas after economizer				
gas after preheater			600	
gas before ID fan			370	
bed (average)			1980	
inlet to classifier			--	
return from classifier			--	
horizontal bundle inlet				
horizontal bundle outlet				
economizer inlet			53.3	
economizer outlet			196.8	
water door inlet			53.3	
water door outlet			176.5	
support tube inlet				
support tube outlet				
air distributor (average)			--	
O_2 in flue gas (vol %)			2.0	
Mean Particle Size (Microns)			--	

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FBM OPERATING CONDITIONS LOG

st No. 644-9　　　　Date Dec. 15, 1976

Time Duration (Hours)	2-1/2	1	2-1/4	
Materials Rates	LIGHT-OFF AND TRANSITION TO FLY ASH			
bed depth (inches)		20.8	21	
coal feed rate (lbs/hr)		0	0	
limestone feed rate (lbs/hr)		0	0	
high carbon fly ash feed rate (lb/hr)		580	665	
low carbon fly ash collection rate (lb/hr)		1200	1870	
bed removal rate (lbs/hr)		--	53.3	
classifier removal rate (lbs/hr)		0	0	
dust loading (lb/hr)				
Air and Gas Rates				
main air flow rate (lbs/hr)		3700	3700	
light-off burner air rate (lbs/hr)		0	0	
light-off burner gas rate (CFH)		0	0	
duct burner air rate (lbs/hr)		0	0	
duct burner gas rate (CFH)		0	0	
fly ash injection air rate (lb/hr)		345	345	
coal-feed air rate (lbs/hr)		260	260	
classifier air rate (lbs/hr)				
fly ash removal air rate (lbs/hr)				
stack gas flow rate (lbs/hr)		6900	6900	
Water Rates				
horizontal bundle water rate (lb/hr)				
steam rate (lbs/hr)		1450	1500	
make-up (steam drum) rate (GPM)		--	--	
economizer rate (GPM)		23.6	23.6	
water door rate (GPM)		3.8	3.8	
support tube rate (GPM)				
duct burner rate (GPM		0	0	

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Test No. 644-9　　　　Date Dec. 15, 1976

mperatures (°F)

air @ FD fan inlet	LIGHT-OFF AND TRANSITION TO FLY ASH	42	43	
air @ FD fan discharge		81	82	
air after preheater		--	--	
air @ plenum		230	230	
gas below steam drum				
gas before economizer		1250	1260	
gas after economizer		920	920	
gas after preheater		620	610	
gas before ID fan		385	385	
bed (average)		1990	1940	
inlet to classifier		--	--	
return from classifier		--	--	
horizontal bundle inlet				
horizontal bundle outlet				
economizer inlet		44	45	
economizer outlet		106	112	
water door inlet		44	45	
water door outlet		100		
support tube inlet				
support tube outlet				
air distributor (average)		--	--	
O_2 in flue gas (vol %)		3	4	
Mean Particle Size (Microns)		--	--	

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FBM OPERATING CONDITIONS LOG

Test No. 644-10 Date Dec. 16, 1976

Time Duration (Hours)	3-1/2	1	2-1/2	
Materials Rates	↑ LIGHT-OFF AND TRANSITION TO FLY ASH ↓			
bed depth (inches)				
coal feed rate (lbs/hr)		0	0	
limestone feed rate (lbs/hr)		0	0	
high carbon fly ash feed rate (lb/hr)		803	940	
low carbon fly ash collection rate (lb/hr)		382	332	
bed removal rate (lbs/hr)		--	--	
classifier removal rate (lbs/hr)		--	--	
dust loading (lb/hr)				
Air and Gas Rates				
main air flow rate (lbs/hr)		3310	3010	
light-off burner air rate (lbs/hr)		--	--	
light-off burner gas rate (CFH)		--	--	
duct burner air rate (lbs/hr)		--	--	
duct burner gas rate (CFH)		--	--	
fly ash injection air rate (lb/hr)				
coal-feed air rate (lbs/hr)		243	260	
classifier air rate (lbs/hr)		--	--	
fly ash removal air rate (lbs/hr)		--	--	
stack gas flow rate (lbs/hr)		7800	8000	
Water Rates				
horizontal bundle water rate (lb/hr)		--	--	
steam rate (lbs/hr)		1770	1800	
make-up (steam drum) rate (GPM)		--	--	
economizer rate (GPM)		23.8	24.4	
water door rate (GPM)		3.9	3.9	
support tube rate (GPM)		--	--	
duct burner rate (GPM		--	--	

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Test No. 644-10 Date Dec. 16, 1976

Temperatures (°F)				
air @ FD fan inlet		50.2	50.2	
air @ FD fan discharge		87.4	86.9	
air after preheater		293	290	
air @ plenum		345	340	
gas below steam drum		2000	1974	
gas before economizer		1373	1350	
gas after economizer		1026	1010	
gas after preheater		744	732	
gas before ID fan		488	430	
bed (average)		1976	1946	
inlet to classifier		--	--	
return from classifier		--	--	
horizontal bundle inlet		--	--	
horizontal bundle outlet		--	--	
economizer inlet		44	44	
economizer outlet		132	131	
water door inlet		44	44	
water door outlet		124	120	
support tube inlet		--	--	
support tube outlet		--	--	
air distributor (average)		--	--	
O_2 in flue gas (vol %)		2.0	2.5	
Mean Particle Size (Microns)		--	--	

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FBM OPERATING CONDITIONS LOG

Test No. 645-1

Date Jan. 12, 1977

Time of Day (Hours)					
Time Duration (Hours)	2	2	0.33	1.5	
MATERIALS RATES (lb/hr)		Automatic Control Test		Petrocarb Performance Test	
bed depth (inches)	LIGHT-OFF		TRANSITION TO FLY ASH	15	
coal feed rate				0	
limestone feed rate				0	
high carbon fly ash feed rate					
low carbon fly ash collection rate					
bed removal rate					
classifier removal rate				0	
dust loading					
AIR AND GAS RATES (lb/hr)					
main air flow rate				3450	
light-off burner air rate				0	
light-off burner gas rate				0	
duct burner air rate				0	
duct burner gas rate (CFH)				0	
fly ash injection air rate				384	
coal-feed air rate				0	
classifier air rate				-	
fly ash removal air rate				-	
stack gas flow rate				2000	
WATER RATES (GPM)					
steam rate				1150	
make-up (steam drum) rate				-	
economizer rate				-	
water door rate				2.6	
duct burner rate				-	

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FBM OPERATING CONDITIONS LOG

Test No. 645-1

Date Jan. 12, 1977

Time of Day (Hours)					
Time Duration (Hours)	2	2	0.33	1.5	
TEMPERATURES (°F)		Automatic Control Test			
air @ FD fan inlet	LIGHT-OFF		TRANSITION TO FLY ASH	34	
air @ FD fan discharge				75	
air after preheater				-	
air @ plenum				220	
gas below steam drum				1953	
gas before economizer				1215	
gas after economizer				965	
gas after preheater				700	
gas before ID fan				440	
bed (average)				1970	
inlet to classifier				-	
return from classifier				-	
economizer inlet				41	
economizer outlet				103	
water door inlet				41	
water door outlet				104	
air distributor (average)				-	
O_2 in flue gas (vol %)				6.8	
Mean Particle Size (Microns)				-	

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FBM OPERATING CONDITIONS LOG

Test No. 645-2

Date Jan. 18, 1977

Time of Day (Hours)	10:35	11:00	11:30	12:00	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	25.5	25.1	21.6	21.2	
coal feed rate	346	312	164	312	
limestone feed rate	348	164	132	216	
high carbon fly ash feed rate	-	-	-	-	
low carbon fly ash collection rate	-	-	-	-	
bed removal rate	-	-	-	-	
classifier removal rate	-	-	-	-	
dust loading	-	-	-	-	
AIR AND GAS RATES (lb/hr)					
main air flow rate	3820	3820	2400	4650	
light-off burner air rate	0	0	0	0	
light-off burner gas rate	0	0	0	0	
duct burner air rate	0	0	0	0	
duct burner gas rate (CFH)	0	0	0	0	
fly ash injection air rate	424	424	424	424	
coal-feed air rate	440	460	450	440	
classifier air rate	-	-	-	-	
fly ash removal air rate	-	-	-	-	
stack gas flow rate	6800	6900	5500	8200	
WATER RATES (GPM)					
steam rate	-	-	-	-	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	-	-	-	-	
duct burner rate	0	0	0	0	

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FBM OPERATING CONDITIONS LOG

Test No. 645-2

Date Jan. 18, 1977

Time of Day (Hours)	12:30	13:00	13:25	14:00	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	21.3	21.3	22.6	24.3	
coal feed rate	218	216	182	214	
limestone feed rate	432	564	432	372	
high carbon fly ash feed rate	-	-	-	-	
low carbon fly ash collection rate	-	-	-	-	
bed removal rate	-	-	-	-	
classifier removal rate	-	-	-	-	
dust loading	-	-	-	-	
AIR AND GAS RATES (lb/hr)					
main air flow rate	4650	4650	2000	4850	
light-off burner air rate	0	0	0	0	
light-off burner gas rate	0	0	0	0	
duct burner air rate	0	0	0	0	
duct burner gas rate (CFH)	0	0	0	0	
fly ash injection air rate	424	424	424	424	
coal-feed air rate	435	445	435	430	
classifier air rate	-	-	-	-	
fly ash removal air rate	-	-	-	-	
stack gas flow rate	8400	8300	5600	7000	
WATER RATES (GPM)					
steam rate	-	-	-	-	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	-	-	-	-	
duct burner rate	0	0	0	0	

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FBM OPERATING CONDITIONS LOG

Test No. 645-2
Date Jan. 18, 1977

Time of Day (Hours)	14:30	14:45			
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	24.3	24.3			
coal feed rate	288	390			
limestone feed rate	468	240			
high carbon fly ash feed rate	-	-			
low carbon fly ash collection rate	-	-			
bed removal rate	-	-			
classifier removal rate	-	-			
dust loading	-	-			
AIR AND GAS RATES (lb/hr)					
main air flow rate	4500	5700			
light-off burner air rate	0	0			
light-off burner gas rate	0	0			
duct burner air rate	0	0			
duct burner gas rate (CFH)	0	0			
fly ash injection air rate	424	424			
coal-feed air rate	430	430			
classifier air rate	-	-			
fly ash removal air rate	-	-			
stack gas flow rate	5700	9000			
WATER RATES (GPM)					
steam rate					
make-up (steam drum) rate	-	-			
economizer rate	-	-			
water door rate					
duct burner rate	0	0			

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FBM OPERATING CONDITIONS LOG

Test No. 645-2
Date Jan. 18, 1977

Time of Day (Hours)	10:35	11:00	11:30	12:00	
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	24.3	27.1	30.6	30.2	
air @ FD fan discharge	64.9	66.5	73.4	67.5	
air after preheater	-	-	-	-	
air @ plenum	190	190	200	200	
gas below steam drum	172	180	183	174	
gas before economizer	1150	1190	1115	1200	
gas after economizer	900	930	790	980	
gas after preheater	590	620	430	710	
gas before ID fan	380	400	315	455	
bed (average)	1700	1780	1860	1720	
inlet to classifier					
return from classifier					
economizer inlet					
economizer outlet					
water door inlet					
water door outlet					
air distributor (average)					
O_2 in flue gas (vol %)					
Mean Particle Size (Microns)					

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FBM OPERATING CONDITIONS LOG

Test No. 645-2

Date Jan. 18, 1977

Time of Day (Hours)	14:30	14:45			
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	32.8	33.5			
air @ FD fan discharge	71.5	70.1			
air after preheater	-	-			
air @ plenum	1490	1540			
gas below steam drum	162	168			
gas before economizer	1550	1500			
gas after economizer	930	1010			
gas after preheater	680	755			
gas before ID fan	445	495			
bed (average)	1560	1570			
inlet to classifier					
return from classifier					
economizer inlet					
economizer outlet					
water door inlet					
water door outlet					
air distributor (average)					
O_2 in flue gas (vol %)					
Mean Particle Size (Microns)					

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FBM OPERATING CONDITIONS LOG

Test No. 645-2

Date Jan. 18, 1977

Time of Day (Hours)	12:30	13:00	13:25	14:00	
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	30.4	32.1	32.0	31.8	
air @ FD fan discharge	68.1	70.0	74.5	72.4	
air after preheater	-	-	-	-	
air @ plenum	200	1505	200	1550	
gas below steam drum	168	167	168	169	
gas before economizer	1170	1520	1560	1470	
gas after economizer	970	960	790	910	
gas after preheater	710	705	510	640	
gas before ID fan	470	465	360	420	
bed (average)	1660	1630	1750	1570	
inlet to classifier					
return from classifier					
economizer inlet					
economizer outlet					
water door inlet					
water door outlet					
air distributor (average)					
O_2 in flue gas (vol %)					
Mean Particle Size (Microns)					

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FBM OPERATING CONDITIONS LOG

Test No. 644-11

Date Jan. 27, 1977

Time of Day (Hours)	11:30	12:00	12:30	13:00	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	16.0	16.0	15.5	16.0	
coal feed rate	-	-	-	-	
limestone feed rate	-	-	-	-	
high carbon fly ash feed rate	615	615	615	615	
low carbon fly ash collection rate	180	180	170	170	
bed removal rate	98	149	79	60	
classifier removal rate	-	-	-	-	
dust loading	38.0	52.7	-	-	
AIR AND GAS RATES (lb/hr)					
main air flow rate	3400	4100	4100	4100	
light-off burner air rate	-	-	-	-	
light-off burner gas rate	-	-	-	-	
duct burner air rate	-	-	-	-	
duct burner gas rate (CFH)	-	-	-	-	
fly ash injection air rate	379	382	378	378	
coal-feed air rate	400	580	580	580	
classifier air rate	-	-	-	-	
fly ash removal air rate	-	-	-	-	
stack gas flow rate	3300	3300	3300	3300	
WATER RATES (GPM)					
steam rate	-	-	-	-	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	4.1	4.05	4.0	4.05	
duct burner rate	-	-	-	-	

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FBM OPERATING CONDITIONS LOG

Test No. 644-11

Date Jan. 27, 1977

Time of Day (Hours)	13:30	14:00	14:30	15:10	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	15.4	15.5	15.5	15.6	
coal feed rate	-	-	-	-	
limestone feed rate	-	-	-	-	
high carbon fly ash feed rate	615	615	615	615	
low carbon fly ash collection rate	170	186	450	-	
bed removal rate	98	98	98	-	
classifier removal rate	-	-	-	-	
dust loading	40.8	20.9	-	49.8	
AIR AND GAS RATES (lb/hr)					
main air flow rate	3800	4100	4100	4100	
light-off burner air rate	-	-	-	-	
light-off burner gas rate	-	-	-	-	
duct burner air rate	-	-	-	-	
duct burner gas rate (CFH)	-	-	-	-	
fly ash injection air rate	376	376	392	405	
coal-feed air rate	183	620	580	595	
classifier air rate	-	-	-	-	
fly ash removal air rate	-	-	-	-	
stack gas flow rate	3300	3300	3300	3300	
WATER RATES (GPM)					
steam rate	-	-	-	-	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	4.05	4.0	4.0	4.0	
duct burner rate	-	-	-	-	

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FBM OPERATING CONDITIONS LOG

Test No. 644-11

Date Jan. 27, 1977

Time of Day (Hours)	11:30	12:00	12:30	13:00	
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	40.8	41.0	41.8	43.2	
air @ FD fan discharge	80.1	78.2	79.7	80.3	
air after preheater	-	-	-	-	
air @ plenum	275	290	300	290	
gas below steam drum	1906	2017	2038	2041	
gas before economizer	1215	1310	1320	1305	
gas after economizer	975	1070	1085	1080	
gas after preheater	725	790	795	785	
gas before ID fan	470	520	530	525	
bed (average)	1970	2030	2050	2070	
inlet to classifier	-	-	-	-	
return from classifier	-	-	-	-	
economizer inlet	41	41	41	41	
economizer outlet	113	125	125	125	
water door inlet	41	41	41	41	
water door outlet	114	127	127	131	
air distributor (average)	-	-	-	-	
O_2 in flue gas (vol %)	5.0	5.5	5.3	6.0	
Mean Particle Size (Microns)	-	-	-	-	

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FBM OPERATING CONDITIONS LOG

Test No. 644-11

Date Jan. 27, 1977

Time of Day (Hours)	13:30	14:00	14:30	15:10	
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	41.8	41.3	45.7	45.0	
air @ FD fan discharge	80.7	82.9	84.2	82.8	
air after preheater	-	-	-	-	
air @ plenum	300	295	270	275	
gas below steam drum	2003	2055	1524	829	
gas before economizer	1300	1300	1185	1160	
gas after economizer	1070	1075	990	985	
gas after preheater	780	785	735	760	
gas before ID fan	520	530	505	515	
bed (average)	2010	2070	1500	1490	
inlet to classifier	-	-	-	-	
return from classifier	-	-	-	-	
economizer inlet	41	41	41	41	
economizer outlet	122	126	114	104	
water door inlet	41	41	41	41	
water door outlet	125	131	103	71	
air distributor (average)	-	-	-	-	
O_2 in flue gas (vol %)	6.4	5.5	8.0	-	
Mean Particle Size (Microns)	-	-	-	-	

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FBM OPERATING CONDITIONS LOG

Test No. 644-12

Date Feb. 3, 1977

ime of Day (Hours)	10:00	10:30	11:00	11:30	12:00
ime Duration (Hours)					
ATERIALS RATES (lb/hr)					
bed depth (inches)	18.6	17.3	15.8	15.9	15.6
coal feed rate	-	-	-	-	-
limestone feed rate	-	-	-	-	-
high carbon fly ash feed rate	60	60	405	405	563
low carbon fly ash collection rate	14	58	86	86	212
bed removal rate	-	-	-	-	-
classifier removal rate	-	-	-	-	-
dust loading	-	-	-	-	-
IR AND GAS RATES (lb/hr)					
main air flow rate	2730	2730	2850	2730	2650
light-off burner air rate	-	-	-	-	-
light-off burner gas rate	-	-	-	-	-
duct burner air rate	-	-		660	660
duct burner gas rate (CFH)	-	-	-	-	-
fly ash injection air rate	286	283	287	282	
coal-feed air rate	445	445	190	190	190
classifier air rate	-	-	-	-	-
fly ash removal air rate	-	-	-	-	-
stack gas flow rate	6200	5300	6300	6150	6200
TER RATES (GPM)					
steam rate	1570	1570	1500	1500	1500
make-up (steam drum) rate	-	-	-	-	-
economizer rate	-	-	-	-	-
water door rate	3.8	2.3	2.3	2.3	-
duct burner rate	-	-	-	-	-

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FBM OPERATING CONDITIONS LOG

Test No. 644-12

Date Feb. 3, 1977

Time of Day (Hours)	12:40	13:00	13:30	14:00	14:15
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	15.5	15.8	16.7	16.2	16.2
coal feed rate	-	-	-	-	-
limestone feed rate	-	-	-	-	-
high carbon fly ash feed rate	417	417	519	828	828
low carbon fly ash collection rate	100	100	117		
bed removal rate			390	620	620
classifier removal rate	-	-	-	-	-
dust loading	-	11.4	23.5	-	68.5
AIR AND GAS RATES (lb/hr)					
main air flow rate	2700	2700	2700	3550	3600
light-off burner air rate	-	-	-	-	-
light-off burner gas rate	-	-	-	-	-
duct burner air rate		650	610		
duct burner gas rate (CFH)	-	-	-	-	-
fly ash injection air rate	292	289	289	281	283
coal-feed air rate	190	190	180		165
classifier air rate	-	-	-	-	-
fly ash removal air rate	-	-	-	-	-
stack gas flow rate	6200	6100	6000	5050	5100
WATER RATES (GPM)					
steam rate	1500	1500	1500	1920	1920
make-up (steam drum) rate	-	-	-	-	-
economizer rate	-	-	-	-	-
water door rate	2.3	2.3	2.3	2.3	2.3
duct burner rate	-	-	-	-	-

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FBM OPERATING CONDITIONS LOG

Test No. 644-12

Date Feb. 3, 1977

Time of Day (Hours)	10:00	10:30	11:00	11:30	12:00
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	37	40	41	42	44
air @ FD fan discharge	76	78	80	81	83
air after preheater	-	-	-	-	-
air @ plenum	278	290	296	297	296
gas below steam drum	1823	1799	1716	1899	1843
gas before economizer	1145	1177	1129	1190	1231
gas after economizer	838	876	825	860	899
gas after preheater	604	660	619	632	670
gas before ID fan	508	504	483	476	468
bed (average)		1890	1815	1990	1895
inlet to classifier	-	-	-	-	-
return from classifier	-	-	-	-	-
economizer inlet	41	41	41	41	41
economizer outlet	94	91	91	99	107
water door inlet	41	41	41	41	41
water door outlet	104	141	138	145	145
air distributor (average)	-	-	-	-	-
O_2 in flue gas (vol %)	7.0	5.7	9.5	6	-
Mean Particle Size (Microns)	-	-	-	-	-

Taylor O_2				2.4	
Bailey O_2				0.8	0.4
Wsths O_2					
Thermox O_2				6.6	

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FBM OPERATING CONDITIONS LOG

Test No. 644-12

Date Feb. 3, 1977

Time of Day (Hours)	12:30	13:00	13:30	14:00	14:17
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	45	46	49	52	53
air @ FD fan discharge	84	86	89	89	91
air after preheater	-	-	-	-	-
air @ plenum	305	305	304	308	312
gas below steam drum	1928	1922	1907	1950	1916
gas before economizer	1201	1185	1199	1316	1290
gas after economizer	856	834	845	995	980
gas after preheater	642	627	616	724	727
gas before ID fan	477	473	442	545	560
bed (average)	1990	1975	1940		
inlet to classifier	-	-	-	-	-
return from classifier	-	-	-	-	-
economizer inlet					
economizer outlet	41	41	41	41	41
water door inlet	100	98	100	120	116
water door outlet	155	152	149	162	159
air distributor (average)	-	-	-	-	-
O_2 in flue gas (vol %)	-	-	-	-	-
Mean Particle Size (Microns)	-	-	-	-	-

Taylor O_2		4.8	2.5		
Bailey O_2		1.5	1.0		
Wsths O_2					
Thermox O_2		7	2.2		

FBM OPERATING CONDITIONS LOG

Test No. 644-13

Date Feb. 8, 1977

Time of Day (Hours)	12:00	12:30	13:00	13:30	14:00
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	20.4	19.7	20.0	19.7	19.5
coal feed rate	-	-	-	-	-
limestone feed rate	-	-	-	-	-
high carbon fly ash feed rate	796	860	720	954	1037
low carbon fly ash collection rate	←	279	→	275	→
bed removal rate	-	-	-	-	-
classifier removal rate	-	-	-	-	-
dust loading	-	98.8	85.0	-	-
AIR AND GAS RATES (lb/hr)					
main air flow rate	2720	2720	2720	2720	2280
light-off burner air rate	0	0	0	0	0
light-off burner gas rate	0	0	0	0	0
duct burner air rate	650	650	610	690	670
duct burner gas rate (CFH)	200	200	200	200	200
fly ash injection air rate	327	327	284	284	284
coal-feed air rate	458	130	130	130	130
classifier air rate	-	-	-	-	-
fly ash removal air rate	-	-	-	-	-
stack gas flow rate	3700	3700	3850	3850	4400
WATER RATES (GPM)					
steam rate					
make-up (steam drum) rate					
economizer rate					
water door rate					
duct burner rate					

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FBM OPERATING CONDITIONS LOG

Test No. 644-14
Date Feb. 10, 1977

Time of Day (Hours)	13:00	13:30	14:00		
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
[illegible]	20	19.5	21		
bed depth (inches)	28.5	28.7	30.2		
coal feed rate					
limestone feed rate					
high carbon fly ash feed rate	588	650	806		
low carbon fly ash collection rate	223				
bed removal rate					
classifier removal rate					
dust loading					
AIR AND GAS RATES (lb/hr)					
main air flow rate	2750	2750	2750		
light-off burner air rate					
light-off burner gas rate					
duct burner air rate	670	685	685		
duct burner gas rate (CFH)	227	215	227		
fly ash injection air rate	239	239	240		
coal-feed air rate	215	195	215		
classifier air rate					
fly ash removal air rate					
stack gas flow rate	3500	3460	3750		
WATER RATES (GPM)					
steam rate					
make-up (steam drum) rate					
economizer rate					
water door rate					
duct burner rate					

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FBM OPERATING CONDITIONS LOG

Test No. 644-14
Date Feb. 10, 1977

Time of Day (Hours)	13:00	13:30	14:00		
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	72	72	63		
air @ FD fan discharge	104	106	101		
air after preheater	303	304	299		
air @ plenum	700	700	692		
gas below steam drum	1910	1903	1869		
gas before economizer	1260	1280	1261		
gas after economizer	970	986	973		
gas after preheater	719	715	711		
gas before ID fan	556	577	561		
bed (average)	1950	1921	1838		
inlet to classifier					
return from classifier					
economizer inlet	42	42	42		
economizer outlet	110	113	113		
water door inlet	115	112	115		
water door outlet					
air distributor (average)					
O_2 in flue gas (vol %)					
Mean Particle Size (Microns)					

O_2 Taylor	8.5	8.1	8.1
O_2 Bailey	2.9	2.4	4.0
O_2 Waths	5.75	5.3	5.7

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FBM OPERATING CONDITIONS LOG

Test No. 646-2

Date March 18, 1977

Time of Day (Hours)	10:00	12:30			
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	16.9	16.8			
coal feed rate	535	567			
limestone feed rate	355	354			
high carbon fly ash feed rate	-	-			
low carbon fly ash collection rate	101	108			
bed removal rate	-	-			
classifier removal rate	-	-			
dust loading	35.1	32.9			
AIR AND GAS RATES (lb/hr)					
main air flow rate	5700	5750			
light-off burner air rate	-	-			
light-off burner gas rate	-	-			
duct burner air rate	-	-			
duct burner gas rate (CFH)	-	-			
fly ash injection air rate	310	335			
coal-feed air rate	415	420			
classifier air rate	-	-			
fly ash removal air rate	-	-			
stack gas flow rate	3800	2700			
WATER RATES (GPM)					
steam rate	-	-			
make-up (steam drum) rate	-	-			
economizer rate	-	-			
water door rate	5.4	4.5			
duct burner rate	0	0			

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FBM OPERATING CONDITIONS LOG

Test No. 646-2

Date March 18, 1977

Time of Day (Hours)	10:00	12:30			
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	129	125			
air @ FD fan discharge	36	58			
air after preheater	-	-			
air @ plenum	-	-			
gas below steam drum	1482	1352			
gas before economizer	1230	1130			
gas after economizer	1100	1135			
gas after preheater	810	820			
gas before ID fan	570	570			
bed (average)	1570	1470			
inlet to classifier	-	-			
return from classifier	-	-			
economizer inlet	-	-			
economizer outlet	-	-			
water door inlet	-	-			
water door outlet	90	91			
air distributor (average)	-	-			
O_2 in flue gas (vol %)	4.5	3.0			
Mean Particle Size (Microns)	-	-			

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FBM OPERATING CONDITIONS LOG

Test No. 646-4

Date March 30, 1977

Time of Day (Hours)	10:00	11:30	14:03		
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	15.7	15.9	15.9		
coal feed rate	537	516	-		
limestone feed rate	-	452	-		
high carbon fly ash feed rate	-	-	-		
low carbon fly ash collection rate	166	178	-		
bed removal rate	228	192	258		
classifier removal rate	0	0	0		
dust loading	21.3	17.3	-		
AIR AND GAS RATES (lb/hr)					
main air flow rate	5200	5200	-		
light-off burner air rate	0	0	0		
light-off burner gas rate	0	0	0		
duct burner air rate	0	0	0		
duct burner gas rate (CFH)	0	0	0		
fly ash injection air rate	315	298	0		
coal-feed air rate	410	420	-		
classifier air rate	0	0	0		
fly ash removal air rate	-	-	-		
stack gas flow rate	790	790	-		
WATER RATES (GPM)					
steam rate	-	-	-		
make-up (steam drum) rate	-	-	-		
economizer rate	-	-	-		
water door rate	5.3	5.3	5.1		
duct burner rate	0	0	0		

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FBM OPERATING CONDITIONS LOG

Test No. 646-4

Date March 30, 1977

Time of Day (Hours)	10:00	11:30	14:03		
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	78.3	85.6			
air @ FD fan discharge	110.2	116.7			
air after preheater	-	-			
air @ plenum	-	-			
gas below steam drum	1418	1330			
gas before economizer	1240	1170			
gas after economizer	1160	1160			
gas after preheater	740	710			
gas before ID fan	560	560			
bed (average)	1640	1590			
inlet to classifier	0	0			
return from classifier	0	0			
economizer inlet	0	0			
economizer outlet	0	0			
water door inlet	0	0			
water door outlet	135	127			
air distributor (average)	-	-			
O_2 in flue gas (vol %)	2.0	2.5			
Mean Particle Size (Microns)	-	-			

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FBM OPERATING CONDITIONS LOG

Test No. 646-5

Date April 14, 1977

Time of Day (Hours)	11:30	12:00	13:00	13:30	14:45
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	16.4	16.4	16.4	16.4	16.4
coal feed rate	476	476	517	517	550
limestone feed rate	300	300	300	300	300
high carbon fly ash feed rate	-	-	-	-	-
low carbon fly ash collection rate	142	142	156	156	133
bed removal rate	760	1068	300	480	3280
classifier removal rate	-	-	-	-	-
dust loading	13.0	12.3	12.2	12.6	15.0
AIR AND GAS RATES (lb/hr)					
main air flow rate	5400	5400	5300	5300	5400
light-off burner air rate	0	0	0	0	0
light-off burner gas rate	0	0	0	0	0
duct burner air rate	0	0	0	0	0
duct burner gas rate (CFH)	0	0	0	0	0
fly ash injection air rate	292	292	292	292	279
coal-feed air rate	415	415	420	410	420
classifier air rate	0	0	0	0	0
fly ash removal air rate	0	0	0	0	0
stack gas flow rate	7200	7200	7150	7050	7050
WATER RATES (GPM)					
steam rate	-	-	-	-	-
make-up (steam drum) rate	-	-	-	-	-
economizer rate	-	-	-	-	-
water door rate	5.3	5.4	5.4	5.4	5.4
duct burner rate	0	0	0	0	0
ECBH	21	22.5	21	21	20.5

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FBM OPERATING CONDITIONS LOG

Test No. 646-5

Date April 14, 1977

Time of Day (Hours)	11:30	12:00	13:00	13:30	14:45
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	83	85	83	85	85
air @ FD fan discharge	113	116	116	117	116
air after preheater	-	-	-	-	-
air @ plenum	350	355	370	370	370
gas below steam drum	1302	1362	1360	1296	1312
gas before economizer	1140	1180	1185	1180	1175
gas after economizer	1135	1160	1170	1165	1165
gas after preheater	710	720	735	735	740
gas before ID fan	490	500	515	515	515
bed (average)	1560	1520	1625	1620	1615
inlet to classifier	0	0	0	0	0
return from classifier	0	0	0	0	0
economizer inlet	0	0	0	0	0
economizer outlet	0	0	0	0	0
water door inlet	0	0	0	0	0
water door outlet	130	134	138	129	132
air distributor (average)	0	0	0	0	0
O_2 in flue gas (vol %) Westinghouse	3.0	1.5	2.0	3.0	3.0
O_2 in flue gas (vol %) Bailey	3.0	2.5	1.5	3.0	2.5
Mean Particle Size (Microns)	-	-	-	-	-

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FBM OPERATING CONDITIONS LOG

Test No. 646-6

Date April 18, 1977

Time of Day (Hours)		11:00	13:00	14:00	14:30
Time Duration (Hours)	↑				
MATERIALS RATES (lb/hr)					
bed depth (inches)		15.4	15.5	15.5	15.5
coal feed rate		555	595	529	529
limestone feed rate		294	294	294	294
high carbon fly ash feed rate		-	-	-	-
low carbon fly ash collection rate		264	135	110	110
bed removal rate		19	192	192	192
classifier removal rate		0	0	0	0
dust loading		12.7	22.1	14.5	17.2
AIR AND GAS RATES (lb/hr)					
main air flow rate	LIGHT-OFF	6100	5900	6000	5950
light-off burner air rate		0	0	0	0
light-off burner gas rate		0	0	0	0
duct burner air rate		0	0	0	0
duct burner gas rate (CFH)		0	0	0	0
fly ash injection air rate		316	324	317	316
coal-feed air rate		410	415	410	410
classifier air rate		0	0	0	0
fly ash removal air rate		0	0	0	0
stack gas flow rate		7100	6900	7100	7100
WATER RATES (GPM)					
steam rate		-	-	-	-
make-up (steam drum) rate		-	-	-	-
economizer rate		-	-	-	-
water door rate		5.5	-	5.3	5.4
duct burner rate	↓	0	0	0	0
EOBH		22	20.5	21	21

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Test No. 646-7

Date April 20, 1977

Time of Day (Hours)	10:00	11:30	13:00	14:00	14:45
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	15.8	15.8	15.3	15.4	15.9
coal feed rate	558	476	548	582	568
limestone feed rate	301	301	301	305	305
high carbon fly ash feed rate	0	0	0	0	0
low carbon fly ash collection rate	138	140	152	161	145
bed removal rate	161	206	190	190	190
classifier removal rate	0	0	0	0	0
dust loading	32.1	15.7	12.6	14.1	15.3
AIR AND GAS RATES (lb/hr)					
main air flow rate	6100	6100	6100	6200	6200
light-off burner air rate	0	0	0	0	0
light-off burner gas rate	0	0	0	0	0
duct burner air rate	0	0	0	0	0
duct burner gas rate (CFH)	0	0	0	0	0
fly ash injection air rate	315	317	317	317	317
coal-feed air rate	415	420	420	420	415
classifier air rate	0	0	0	0	0
fly ash removal air rate	0	0	0	0	0
stack gas flow rate	7050	7100	7100	7100	7100
WATER RATES (GPM)					
steam rate	-	-	-	-	-
make-up (steam drum) rate	-	-	-	-	-
economizer rate	-	-	-	-	-
water door rate	5.5	5.5	5.5	5.5	5.5
duct burner rate	0	0	0	0	0
EOBH	21	21	21.5	21	21.5

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Test No. 646-7

Date April 20, 1977

Time of Day (Hours)	10:00	11:30	13:00	14:00	14:45
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	78	82	82	79	79
air @ FD fan discharge	104	108	110	109	108
air after preheater	-	-	-	-	-
air @ plenum	350	355	375	385	360
gas below steam drum	1410	1365	1367	1376	1420
gas before economizer	1235	1190	1245	1245	1260
gas after economizer	1150	1190	1230	1240	1175
gas after preheater	720	750	775	785	735
gas before ID fan	520	535	555	565	535
bed (average)	1570	1490	1560	1600	1590
inlet to classifier	0	0	0	0	0
return from classifier	0	0	0	0	0
economizer inlet	0	0	0	0	0
economizer outlet	0	0	0	0	0
water door inlet	0	0	0	0	0
water door outlet	138	135	134	134	140
air distributor (average)	-	-	-	-	-
O_2 in flue gas (vol %) Westinghouse Bailey	4.0 6.0	1.9 4.5	1.3 4.0	2.5 3.5	3.5 5.5
Mean Particle Size (Microns)	-	-	-	-	-

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Test No. 646-8

Date April 25, 1977

Time of Day (Hours)	11:30	13:30	14:00		
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	15.7	15.6	15.6		
coal feed rate	540	568	562		
limestone feed rate	340	315	315		
high carbon fly ash feed rate	-	-	-		
low carbon fly ash collection rate	126	150	150		
bed removal rate	193	185	165		
classifier removal rate	0	0	0		
dust loading	15.7	14.9	17.6		
AIR AND GAS RATES (lb/hr)					
main air flow rate	6216	6104	6082		
light-off burner air rate	0	0	0		
light-off burner gas rate	0	0	0		
duct burner air rate	0	0	0		
duct burner gas rate (CFH)	0	0	0		
fly ash injection air rate	328	341	328		
coal-feed air rate	416	419	414		
classifier air rate	0	0	0		
fly ash removal air rate	0	0	0		
stack gas flow rate					
WATER RATES (GPM)					
steam rate	-	-	-		
make-up (steam drum) rate	-	-	-		
economizer rate	-	-	-		
water door rate	5.5	5.5	5.5		
duct burner rate	0	0	0		
ECBH	21	21.5	21.5		

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FBM OPERATING CONDITIONS LOG

Test No. 646-8

Date April 25, 1977

Time of Day (Hours)	11:30	13:30	14:00		
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	77	76	73		
air @ FD fan discharge	102	105	103		
air after preheater	-	-	-		
air @ plenum	310	340	335		
gas below steam drum	1371	1358	1387		
gas before economizer	1225	1235	1240		
gas after economizer	1140	1230	1150		
gas after preheater	720	760	730		
gas before ID fan	520	550	525		
bed (average)	1560	1570	1570		
inlet to classifier	0	0	0		
return from classifier	0	0	0		
economizer inlet	0	0	0		
economizer outlet	0	0	0		
water door inlet	0	0	0		
water door outlet	140	138	144		
air distributor (average)	-	-	-		
O_2 in flue gas (vol %) Westinghouse Bailey	3.4 5.3	1.8 2.0	3.3 5.0		
Mean Particle Size (Microns)	-	-	-		

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Test No. 646-9

Date April 28, 1977

Time of Day (Hours)	9:30	12:00	13:00	13:30	14:45
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	16.8	16.3	16.4	16.3	16.4
coal feed rate	466	540	530	510	532
limestone feed rate	320	315	315	315	270
high carbon fly ash feed rate	-	-	-	-	-
low carbon fly ash collection rate	130	140	150	135	137
bed removal rate	170	183	183	183	150
classifier removal rate	0	0	0	0	0
dust loading	18.5	11.7	15.2	13.5	19.4
AIR AND GAS RATES (lb/hr)					
main air flow rate					
light-off burner air rate	0	0	0	0	0
light-off burner gas rate	0	0	0	0	0
duct burner air rate	0	0	0	0	0
duct burner gas rate (CFH)	0	0	0	0	0
fly ash injection air rate					
coal-feed air rate					
classifier air rate	0	0	0	0	0
fly ash removal air rate	0	0	0	0	0
stack gas flow rate					
WATER RATES (GPM)					
steam rate	-	-	-	-	-
make-up (steam drum) rate	-	-	-	-	-
economizer rate	-	-	-	-	-
water door rate	5.5	5.5	5.5	5.5	5.5
duct burner rate	0	0	0	0	0
EOBH	21	21.5	22	22.5	21

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FBM OPERATING CONDITIONS LOG

Test No. 646-9

Date April 28, 1977

Time of Day (Hours)	9:30	12:00	13:00	13:30	14:45
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	69	89	87	88	86
air @ FD fan discharge	101	117	118	119	117
air after preheater	-	-	-	-	-
air @ plenum	306	460	380	370	410
gas below steam drum	1394	1328	1347	1325	1310
gas before economizer	1230	1240	1255	1250	1175
gas after economizer	1125	1230	1250	1250	1190
gas after preheater	720	775	790	790	735
gas before ID fan	500	555	560	560	540
bed (average)	1570	1555	1560	1540	1525
inlet to classifier	0	0	0	0	0
return from classifier	0	0	0	0	0
economizer inlet	0	0	0	0	0
economizer outlet	0	0	0	0	0
water door inlet	0	0	0	0	0
water door outlet	128	121	123	122	121
air distributor (average)	-	-	-	-	-
Westinghouse Bailey O_2 in flue gas (vol %)	3.2 5.0	2.6	1.0	1.5	5.3
Mean Particle Size (Microns)	-	-	-	-	-

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Test No. 646-10

Date May 16, 1977

Time of Day (Hours)	10:00	11:30	12:30	13:30	14:30
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	15.3	15.4	15.4	14.9	15.0
coal feed rate	548	530	555	548	555
limestone feed rate	315	308	278	280	280
high carbon fly ash feed rate	-	-	-	-	-
low carbon fly ash collection rate	140	150	150	150	150
bed removal rate	200	160	175	195	205
classifier removal rate	-	-	-	-	-
dust loading	15.3	13.2	16.6	12.6	13.0
AIR AND GAS RATES (lb/hr)					
main air flow rate	6015	6311	6223	6223	6223
light-off burner air rate	-	-	-	-	-
light-off burner gas rate	-	-	-	-	-
duct burner air rate	-	-	-	-	-
duct burner gas rate (CFH)	-	-	-	-	-
fly ash injection air rate	350	335	337	331	330
coal-feed air rate	411	414	414	414	411
classifier air rate	-	-	-	-	-
fly ash removal air rate	-	-	-	-	-
stack gas flow rate	-	-	-	-	-
WATER RATES (GPM)					
steam rate	-	-	-	-	-
make-up (steam drum) rate	-	-	-	-	-
economizer rate	-	-	-	-	-
water door rate	5.4	5.4	5.4	5.4	5.4
duct burner rate	0	0	0	0	0

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FBM OPERATING CONDITIONS LOG

Test No. 646-10

Date May 16, 1977

Time of Day (Hours)	10:00	11:30	12:30	13:30	14:30
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	80	79	83	85	85
air @ FD fan discharge	106	109	112	113	115
air after preheater	-	-	-	-	-
air @ plenum	320	340	350	350	350
gas below steam drum	1399	1341	1345	1323	1335
gas before economizer	1230	1190	1190	1180	1190
gas after economizer	1105	1175	1175	1175	1165
gas after preheater	730	770	770	770	765
gas before ID fan	580	625	625	625	630
bed (average)	1595	1590	1580	1570	1590
inlet to classifier	0	0	0	0	0
return from classifier	0	0	0	0	0
economizer inlet	0	0	0	0	0
economizer outlet	0	0	0	0	0
water door inlet	0	0	0	0	0
water door outlet	116	109	107	108	109
air distributor (average)	-	-	-	-	-
O_2 in flue gas (vol %) Taylor	6.2	5.6	5.4	5.7	5.5
Bailey	5.3	2.3	2.4	2.4	2.3
W.H.	3.2	3.3	4.2	4.6	6.0
Mean Particle Size (Microns)	-	-	-	-	-

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FBM OPERATING CONDITIONS LOG

Test No. 650-1

Date May 19, 1977

Time of Day (Hours)	11:30	13:00			
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	18.3	17.3			
coal feed rate	600	586			
limestone feed rate	290	300			
high carbon fly ash feed rate	-	369			
low carbon fly ash collection rate	150	330			
bed removal rate	185	263			
classifier removal rate	-	-			
dust loading	15.6	-			
AIR AND GAS RATES (lb/hr)					
main air flow rate	5771	5946			
light-off burner air rate	-	-			
light-off burner gas rate	-	-			
duct burner air rate	-	-			
duct burner gas rate (CFH)	-	-			
fly ash injection air rate	326	319			
coal-feed air rate	402	397			
classifier air rate	-	-			
fly ash removal air rate	-	-			
stack gas flow rate	-	-			
WATER RATES (GPM)					
steam rate	-	-			
make-up (steam drum) rate	-	-			
economizer rate	-	-			
water door rate	5.4	5.4			
duct burner rate	-	-			

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FBM OPERATING CONDITIONS LOG

Test No. 650-1

Date May 19, 1977

Time of Day (Hours)	11:30	13:00			
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	93	95			
air @ FD fan discharge	121	124			
air after preheater	-	-			
air @ plenum	340	375			
gas below steam drum	1452	1489			
gas before economizer	1240	1295			
gas after economizer	1135	1160			
gas after preheater	740	785			
gas before ID fan	585	630			
bed (average)	1615	1630			
inlet to classifier	0	0			
return from classifier	0	0			
economizer inlet	0	0			
economizer outlet	0	0			
water door inlet	0	0			
water door outlet	119	121			
air distributor (average)	-	-			
O_2 in flue gas (vol %) Taylor Bailey W.H.	4.8 2.9 2.0	4.3 2.3 2.2			
Mean Particle Size (Microns)	-	-			

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FBM OPERATING CONDITIONS LOG

Test No. 650-2

Date May 23, 1977

Time of Day (Hours)	11:30	12:30	13:30	14:00	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	19.9	19.7	19.7	19.5	
coal feed rate	593	591	577	577	
limestone feed rate	300	293	293	293	
high carbon fly ash feed rate	0	0	427	427	
low carbon fly ash collection rate	170	165	320	340	
bed removal rate	175	180	215	240	
classifier removal rate	-	-	-	-	
dust loading	19	20.9	31.1	30	
AIR AND GAS RATES (lb/hr)					
main air flow rate	5602	5229	5190	5140	
light-off burner air rate	-	-	-	-	
light-off burner gas rate	-	-	-	-	
duct burner air rate	-	-	-	-	
duct burner gas rate (CFH)	-	-	-	-	
fly ash injection air rate	332	333	327	329	
coal-feed air rate	411	405	405	405	
classifier air rate	-	-	-	-	
fly ash removal air rate	-	-	-	-	
stack gas flow rate	-	-	-	-	
WATER RATES (GPM)					
steam rate	-	-	-	-	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	5.4	5.4	5.4	5.4	
duct burner rate	0	0	0	0	

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FBM OPERATING CONDITIONS LOG

Test No. 650-2

Date May 23, 1977

Time of Day (Hours)	11:30	12:30	13:30	14:00	
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	85	89	85	86	
air @ FD fan discharge	114	117	116	116	
air after preheater	-	-	-	-	
air @ plenum	335	440	440	445	
gas below steam drum	1482	1468	1459	1466	
gas before economizer	1230	1260	1260	1270	
gas after economizer	1120	1140	1130	1135	
gas after preheater	615	760	760	770	
gas before ID fan	570	590	590	605	
bed (average)	1620	1610	1570	1570	
inlet to classifier	0	0	0	0	
return from classifier	0	0	0	0	
economizer inlet	0	0	0	0	
economizer outlet	0	0	0	0	
water door inlet	0	0	0	0	
water door outlet	121	122	121	122	
air distributor (average)	-	-	-	-	
O_2 in flue gas (vol %) Taylor	4.1	4.6	4.6	6.5	
Bailey	1.6	2.2	2.1	2.0	
W.H.	2.1	2.9	3.2	3.0	
Mean Particle Size (Microns)	-	-	-	-	

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Test No. 650-4

Date May 26, 1977

Time of Day (Hours)	10:30	11:30	13:30	14:15	14:45
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	20.1	19.6	20.1	20.2	20.7
coal feed rate	550	546	535	530	564
limestone feed rate	0	0	0	0	0
high carbon fly ash feed rate	0	0	318	332	0
low carbon fly ash collection rate	128	160	300	300	180
bed removal rate	0	0	0	0	0
classifier removal rate	0	0	0	0	0
dust loading	17.42	16.93	26.32	18.19	16.84
AIR AND GAS RATES (lb/hr)					
main air flow rate	4750	4732	4439	4456	4456
light-off burner air rate	0	0	0	0	0
light-off burner gas rate	0	0	0	0	0
duct burner air rate	0	0	0	0	0
duct burner gas rate (CFH)	0	0	0	0	0
fly ash injection air rate	328	324	326	326	247
coal-feed air rate	405	402	400	394	400
classifier air rate	0	0	0	0	0
fly ash removal air rate	0	0	0	0	0
stack gas flow rate	-	-	-	-	-
WATER RATES (GPM)					
steam rate	-	-	-	-	-
make-up (steam drum) rate	-	-	-	-	-
economizer rate	-	-	-	-	-
water door rate	5.4	5.4	5.4	5.4	5.4
duct burner rate	0	0	0	0	0

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FBM OPERATING CONDITIONS LOG

Test No. 650-4

Date May 26, 1977

Time of Day (Hours)		10:30	11:30	13:30	14:15	14:45
Time Duration (Hours)						
TEMPERATURES (°F)						
air @ FD fan inlet		85	87	91	91	89
air @ FD fan discharge		115	117	121	122	123
air after preheater		-	-	-	-	-
air @ plenum		400	406	412	408	408
gas below steam drum		1400	1397	1415	1390	1393
gas before economizer		1200	1210	1200	1200	1190
gas after economizer		1085	1100	1065	1050	1070
gas after preheater		670	685	685	675	670
gas before ID fan		545	555	545	535	530
bed (average)		1615	1630	1625	1570	1610
inlet to classifier		0	0	0	0	0
return from classifier		0	0	0	0	0
economizer inlet		0	0	0	0	0
economizer outlet		0	0	0	0	0
water door inlet		0	0	0	0	0
water door outlet		115	113	116	113	114
air distributor (average)		-	-	-	-	-
O_2 in flue gas (vol %)	Bailey	3.4	3.2	1.0	1.3	1.2
	W.H.	4.8	4.85	3.5	4	4.3
Mean Particle Size (Microns)		-	-	-	-	-

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FBM OPERATING CONDITIONS LOG

Test No. 650-5

Date June 2, 1977

Time of Day (Hours)	10:45	11:15	12:15	12:30	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	18.5	18.5	19.1	19.1	
coal feed rate					
limestone feed rate	0	0	0	0	
high carbon fly ash feed rate					
low carbon fly ash collection rate					
bed removal rate	0	0	0	0	
classifier removal rate	0	0	0	0	
dust loading					
AIR AND GAS RATES (lb/hr)					
main air flow rate					
light-off burner air rate	0	0	0	0	
light-off burner gas rate	0	0	0	0	
duct burner air rate	0	0	0	0	
duct burner gas rate (CFH)	0	0	0	0	
fly ash injection air rate					
coal-feed air rate					
classifier air rate	0	0	0	0	
fly ash removal air rate	0	0	0	0	
stack gas flow rate	-	-	-	-	
WATER RATES (GPM)					
steam rate	-	-	-	-	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	5.4	5.4	5.4	5.4	
duct burner rate	0	0	0	0	

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FBM OPERATING CONDITIONS LOG

Test No. 650-5

Date June 2, 1977

Time of Day (Hours)	10:45	11:15	12:15	12:30	
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	94	92	97	97	
air @ FD fan discharge	121	124	125	125	
air after preheater	-	-	-	-	
air @ plenum					
gas below steam drum	1389	1365	1382	1382	
gas before economizer	1150	1155	1170	1160	
gas after economizer	1030	1020	1020	1020	
gas after preheater	645	650	655	650	
gas before ID fan	520	520	530	530	
bed (average)	1580	1585	1575	1585	
inlet to classifier	0	0	0	0	
return from classifier	0	0	0	0	
economizer inlet	0	0	0	0	
economizer outlet	0	0	0	0	
water door inlet	0	0	0	0	
water door outlet	119	117	117	120	
air distributor (average)	-	-	-	-	
Bailey O_2 in flue gas (vol %)	1.2	1.0	0.8	0.7	
Mean Particle Size (Microns)	-	-	-	-	

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Test No. 650-6

Date June 10, 1977

Time of Day (Hours)	14:15	15:00	10:30		
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	18.9	18.9	18.8		
coal feed rate	499	486	480		
limestone feed rate	0	0	0		
high carbon fly ash feed rate	0	0	0		
low carbon fly ash collection rate	-	-	126		
bed removal rate	0	0	0		
classifier removal rate	0	0	0		
dust loading	27.1	28.1	-		
AIR AND GAS RATES (lb/hr)					
main air flow rate	4358	4350	4412		
light-off burner air rate	0	0	0		
light-off burner gas rate	0	0	0		
duct burner air rate	0	0	0		
duct burner gas rate (CFH)	0	0	0		
fly ash injection air rate	133	134	123		
coal-feed air rate	402	402	408		
classifier air rate	0	0	0		
fly ash removal air rate	0	0	0		
stack gas flow rate	-	-	-		
WATER RATES (GPM)					
steam rate	-	-	-		
make-up (steam drum) rate	-	-	-		
economizer rate	-	-	-		
water door rate	-	-	-		
duct burner rate	0	0	0		

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Test No. 650-6

Date June 10, 1977

Time of Day (Hours)	14:15	15:00	10:30		
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	75	76	66		
air @ FD fan discharge	110	110	103		
air after preheater	-	-	-		
air @ plenum	441	444	419		
gas below steam drum	1383	1394	1373		
gas before economizer	1190	1200	1175		
gas after economizer	1155	1075	1060		
gas after preheater	660	670	645		
gas before ID fan	525	530	520		
bed (average)	1610	1625	1610		
inlet to classifier	0	0	0		
return from classifier	0	0	0		
economizer inlet	0	0	0		
economizer outlet	0	0	0		
water door inlet	0	0	0		
water door outlet	112	113	110		
air distributor (average)	-	-	-		
O_2 in flue gas (vol %) Bailey	2.3	1.7	3.0		
O_2 in flue gas (vol %) W.H.	2.8	2.0	2.8		
O_2 in flue gas (vol %) Taylor	4.4	4.2	5.0		
Mean Particle Size (Microns)	-	-	-		

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FBM OPERATING CONDITIONS LOG

Test No. 650-7

Date June 15, 1977

Time of Day (Hours)	9:30	11:15	12:30	14:00	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	19.4	21.9	21.5	21.5	
coal feed rate	520	503	488	503	
limestone feed rate	288	288	173	173	
high carbon fly ash feed rate	0	0	0	0	
low carbon fly ash collection rate	143	113	90	90	
bed removal rate	280	228	163	220	
classifier removal rate	0	0	0	0	
dust loading	19.9	32.8	31.3	28.7	
AIR AND GAS RATES (lb/hr)					
main air flow rate	4464	4430	4415	4415	
light-off burner air rate	0	0	0	0	
light-off burner gas rate	0	0	0	0	
duct burner air rate	0	0	0	0	
duct burner gas rate (CFH)	0	0	0	0	
fly ash injection air rate	132	138	136	137	
coal-feed air rate	416	411	405	405	
classifier air rate	0	0	0	0	
fly ash removal air rate	0	0	0	0	
stack gas flow rate	-	-	-	-	
WATER RATES (GPM)					
steam rate	-	-	-	-	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	5.4	5.4	5.4	5.4	
duct burner rate	0	0	0	0	

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FBM OPERATING CONDITIONS LOG

Test No. 650-7

Date June 15, 1977

Time of Day (Hours)	9:30	11:15	12:30	14:00	
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	71	74	81	79	
air @ FD fan discharge	104	109	116	114	
air after preheater	-	-	-	-	
air @ plenum	430	443	449	446	
gas below steam drum	1394	1419	1412	1385	
gas before economizer	1170	1180	1180	1190	
gas after economizer	-	-	-	-	
gas after preheater	665	680	670	670	
gas before ID fan	475	490	490	490	
bed (average)	1595	1580	1570	1550	
inlet to classifier	0	0	0	0	
return from classifier	0	0	0	0	
economizer inlet	0	0	0	0	
economizer outlet	0	0	0	0	
water door inlet	0	0	0	0	
water door outlet	115	112	113	115	
air distributor (average)					
O_2 in flue gas (vol %) Bailey	1.3	0.5	1.0	1.5	
W.H.	2.4	3.0	2.7	3.4	
Taylor	4.1	4.0	4.0	4.2	
Mean Particle Size (Microns)	-	-	-	-	

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FBM OPERATING CONDITIONS LOG

Test No. 650-8

Date June 17, 1977

Time of Day (Hours)	11:15				
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	15.6				
coal feed rate	473				
limestone feed rate	266				
high carbon fly ash feed rate	0				
low carbon fly ash collection rate	0				
bed removal rate	113				
classifier removal rate	0				
dust loading	37				
AIR AND GAS RATES (lb/hr)					
main air flow rate	4369				
light-off burner air rate	0				
light-off burner gas rate	0				
duct burner air rate	0				
duct burner gas rate (CFH)	0				
fly ash injection air rate	137				
coal-feed air rate	411				
classifier air rate	0				
fly ash removal air rate	0				
stack gas flow rate	-				
WATER RATES (GPM)					
steam rate	-				
make-up (steam drum) rate	-				
economizer rate	-				
water door rate	5.4				
duct burner rate	0				

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FBM OPERATING CONDITIONS LOG

Test No. 650-8

Date June 17, 1977

Time of Day (Hours)	11:15				
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	86				
air @ FD fan discharge	118				
air after preheater	-				
air @ plenum	420				
gas below steam drum	1387				
gas before economizer	1195				
gas after economizer	1095				
gas after preheater	685				
gas before ID fan	500				
bed (average)	1635				
inlet to classifier	0				
return from classifier	0				
economizer inlet	0				
economizer outlet	0				
water door inlet	0				
water door outlet	118				
air distributor (average)	-				
O_2 in flue gas (vol %) Bailey W.H. Taylor	2.8 3.3 4.4				
Mean Particle Size (Microns)	-				

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FBM OPERATING CONDITIONS LOG

Test No. 650-9

Date June 21, 1977

Time of Day (Hours)	10:15	13:00	14:15	15:15	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	14.3	13.9	14.0	14.0	
coal feed rate	533	533	548	528	
limestone feed rate	256	255	272	272	
high carbon fly ash feed rate	0	260	408	690	
low carbon fly ash collection rate	150	360	375	465	
bed removal rate	110	119	119	119	
classifier removal rate	0	0	0	0	
dust loading					
AIR AND GAS RATES (lb/hr)					
main air flow rate	4959	4874	4874	4848	
light-off burner air rate	0	0	0	0	
light-off burner gas rate	0	0	0	0	
duct burner air rate	0	0	0	0	
duct burner gas rate (CFH)	0	0	0	0	
fly ash injection air rate	142	153	153	155	
coal-feed air rate	404	402	400	400	
classifier air rate	0	0	0	0	
fly ash removal air rate	0	0	0	0	
stack gas flow rate	-	-	-	-	
WATER RATES (GPM)					
steam rate	-	-	-	--	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	5.4	5.4	5.4	5.4	
duct burner rate	0	0	0	0	

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FBM OPERATING CONDITIONS LOG

Test No. 650-9

Date June 21, 1977

Time of Day (Hours)		10:15	13:00	14:15	15:15	
Time Duration (Hours)						
TEMPERATURES (°F)						
air @ FD fan inlet		93	85	96	86	
air @ FD fan discharge		121	119	126	121	
air after preheater		-	-	-	-	
air @ plenum		415	450	450	455	
gas below steam drum		1391	1459	1474	1481	
gas before economizer		1205	1255	1255	1265	
gas after economizer		-	-	-	-	
gas after preheater		700	740	735	745	
gas before ID fan		540	530	580	545	
bed (average)		1670	1690	1700	1675	
inlet to classifier		0	0	0	0	
return from classifier		0	0	0	0	
economizer inlet		0	0	0	0	
economizer outlet		0	0	0	0	
water door inlet		0	0	0	0	
water door outlet		129	125	135	129	
air distributor (average)		-	-	-	-	
O_2 in flue gas (vol %)	Bailey	4.0	2.2	2.0	1.2	
	W.H.	3.2	3.4	3.6	3.6	
	Taylor	5.2	4.9	4.0	3.7	
Mean Particle Size (Microns)		-	-	-	-	

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Test No. 650-10

Date June 23, 1977

Time of Day (Hours)	10:30	12:00	14:00	15:00	
Time Duration (Hours)					
MATERIALS RATES (lb/hr)					
bed depth (inches)	13.8	13.5	13.0	13.0	
coal feed rate	480	504	496	495	
limestone feed rate	300	288	288	288	
high carbon fly ash feed rate	0	409	353	565	
low carbon fly ash collection rate	320	339	340	450	
bed removal rate	139	139	140	140	
classifier removal rate	0	0	0	0	
dust loading	18.0	31.7	29.2	41.7	
AIR AND GAS RATES (lb/hr)					
main air flow rate	4478	4422	4422	4398	
light-off burner air rate	0	0	0	0	
light-off burner gas rate	0	0	0	0	
duct burner air rate	0	0	0	0	
duct burner gas rate (CFH)	0	0	0	0	
fly ash injection air rate	153	154	153	153	
coal-feed air rate	408	405	408	405	
classifier air rate	0	0	0	0	
fly ash removal air rate	0	0	0	0	
stack gas flow rate	-	-	-	-	
WATER RATES (GPM)					
steam rate	-	-	-	-	
make-up (steam drum) rate	-	-	-	-	
economizer rate	-	-	-	-	
water door rate	5.4	5.4	5.4	5.4	
duct burner rate	0	0	0	0	

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FBM OPERATING CONDITIONS LOG

Test No. 650-10

Date June 23, 1977

Time of Day (Hours)	10:30	12:00	14:00	15:00	
Time Duration (Hours)					
TEMPERATURES (°F)					
air @ FD fan inlet	89	81	94	93	
air @ FD fan discharge	121	117	126	126	
air after preheater	-	-	-	-	
air @ plenum	-	-	-	-	
gas below steam drum	1422	1432	1432	1459	
gas before economizer	1180	1215	1220	1235	
gas after economizer	-	-	-	-	
gas after preheater	675	690	695	710	
gas before ID fan	500	510	520	530	
bed (average)	1700	1685	1700	1710	
inlet to classifier	0	0	0	0	
return from classifier	0	0	0	0	
economizer inlet	0	0	0	0	
economizer outlet	0	0	0	0	
water door inlet	0	0	0	0	
water door outlet	131	125	130	133	
air distributor (average)	-	-	-	-	
Bailey	3.4	1.6	1.8	0.9	
W.H.	3.4	3.5	3.7	3.9	
O_2 in flue gas (vol %) Taylor	5.2	3.9	4.1	3.8	
Mean Particle Size (Microns)	-	-	-	-	

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