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ALEXANDRIA FLUIDIZED BED PROCESS DEVELOPMENT UNIT
COLD MODE TESTING: TOPICAL REPORT

February 1981

Work Performed Under Contract No. AC21-80MC14110

Burns & Roe Services Corporation
Oradell, New Jersey

and

Benmol Corporation
Alexandria, Virginia

TECHNICAL INFORMATION CENTER
UNITED STATES DEPARTMENT OF ENERGY

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February 1981

By
Burns & Roe Services Corporation
Oradell, N.J.
and
Benmol Corporation
Alexandria, VA.

Morgantown Energy Technology Center
U.S. Department of Energy
Under Contract No. DE-AC21-80MC14110

ABSTRACT

The objectives of the current test program include: (1) validation of predictions from the Massachusetts Institute of Technology (MIT) Coal Atmospheric Fluidized Bed Combustor System Model, (2) experimental studies supporting AFBC process development, and (3) the collection of transient data for process control design. This topical report summarizes results from cold mode testing, i.e. experiments performed without combustion for MIT Model verification. During these tests, sulfated limestone (generated from normal AFBC operations) was fluidized with air at temperatures ranging from 80 to 500°F in the 3' x 3' (nominal) size PDU at Alexandria, VA.

MIT Model predictions tested include: slumped bed height, minimum fluidization velocity, and expanded bed height. In all cases, there were large discrepancies between the Model predictions and corresponding experimental results. Other results obtained included solids size distribution and particle size profiles in the bed. Size distribution was adequately modeled by the Rosin-Rammler equation. No transient process data was collected due to hardware problems with the Data Acquisition System.

A series of tests were also performed to determine the effect of maldistribution of air, caused by leaks in the air distributor, on experimental results. The data indicated that effects of these leaks seemed to be undetectable.

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1.0 INTRODUCTION

Since the early 1960's, a number of small to medium-sized combustors have been installed at the Alexandria facility for atmospheric fluid-bed combustion (AFBC) research. The current unit, installed in 1978, is a 0.5 MWe Process Development Unit (PDU).

The current program consists of experimental operations intended to support overall AFBC process development. The specific goals of the current test plan include:

- experimental studies to verify various submodels of the MIT Coal AFBC System Model
- experimental studies supporting AFBC process development, i.e. fly ash reinjection and the effects of varying feed materials
- studies of AFBC process control systems, involving the collection of transient data describing fluid dynamic and combustion responses.

These goals are approached through a test program containing three major tasks and 14 subtasks. Generally, each subtask consists of a set of experimental runs utilizing the PDU. The overall test plan is summarized in Table 1.1.

This topical report contains results obtained from cold mode testing during the period from June 1980 to January 1981. During cold mode testing, no coal is burned in the PDU. All testing takes place at temperatures ranging from ambient up to 500°F. Test results contained within this report include:

- Task 1.1 - Test Series 20
measurement of the fluidization characteristics of a fixed amounts of spent bed material with no solids flow and without the in-bed tube bundle
- Task 1.2 - Test Series 30
measurement of the fluidization characteristics of spent bed material with solids flow, but without the in-bed tube bundle
- Task 1.3 - Test Series 40
measurements to determine the effects of the in-bed tube bundle on fluidization characteristics.

In addition to test results, a description of various auxiliary tasks supporting the test program are presented. Finally, some significant preliminary conclusions are presented regarding the validity of various MIT Model predictions.

Table 1.1

ALEXANDRIA PDU - TEST PLAN

<u>TASK NUMBER</u>	<u>TEST SPECIFICATION</u>
<u>Task 1.0: MIT MODEL</u>	
1.1: Fluid Dynamics I	No solids flow
1.2: Fluid Dynamics II	With solids flow
1.3: Fluid Dynamics III	With in-bed tube bundle
1.4: Combustion Efficiency	Without fly ash reinjection
1.5: Heat Transfer	Horizontal tubes
1.6: Sulfur Retention	One coal/limestone
<u>Task 2.0: AFBC DEVELOPMENT</u>	
2.1: Fly Ash Reinjection	Fly Ash Injection System
2.2: Effects of Feed Materials	Additional coal/limestone
2.3: MIT Model Validity Check	Prediction check
<u>Task 3.0: PROCESS CONTROL</u>	
3.1: Fluid Dynamic Responses	Bed height, gas interchange, ΔP
3.2: Combustion Responses	Combustion efficiency, heat transfer, sulfur retention
3.3: Reinjection Responses	Fly ash reinjection
3.4: Turndown Tests	Combination of all responses to follow steam demand
3.5: Data Automation	Data handling and reduction

2.0 TEST FACILITY

2.1 GENERAL FEATURES OF THE PDU

The Alexandria AFBC PDU is an integrated coal combustion system based upon a single-cell, waterwall fluidized-bed combustor. The PDU's general characteristics are outlined in Table 2.1, while Figure 2.1 presents an overall process flow diagram.

As shown in Figure 2.2, the tube walls are membrane type construction and the heat exchanger tubes are removable. The heat exchangers include in-bed and freeboard region bundles. The solids feed line is common for both limestone and coal, protruding 18" into the bed with a needle-type feed point. The solids feed line receives coal and limestone from silos equipped with load cells and rotary valves for rate control. The flue gas leaving the combustor passes through a gas cooler before entering a multi-cone dust collector and baghouse.

Fluidizing air, supplied by a forced draft fan, passes upward through the drilled-bolt air distributor. An underbed burner is installed to preheat the fluidizing air up to 500°F. The main air is mixed with the burner combustion products in the plenum before entering the bed through the distributor. The distributor drilled bolts are placed into a $\frac{1}{4}$ " steel plate at equilateral triangular pitch of 0.724" spacing, and protrude 1-3/4" above the plate surface. With this arrangement, a 1-3/4" thick stagnant bed layer protects the distributor plate from excessive temperatures. The spent bed material drain is installed at the lowest side point of the bed, 10" from the inside back corner of the PDU. Discharged spent bed material is pneumatically transported to storage. Other features of the unit include fly ash reinjection, solids size screening and a closed water circuit absorbing combustion heat and rejecting to atmosphere via an air-cooled condenser.

The facility is supported by an analytical laboratory and a general maintenance workshop.

2.2 TEST FACILITY FOR COLD MODE TESTING

The cold mode tests did not require the use of coal handling equipment, fly ash reinjection, or the water circuit. However, some modifications and additional equipment were necessary. The actual equipment arrangement for cold mode testing is shown in Figure 2.3.

The coal and limestone silo lines were modified to become part of the closed solids circuit. One of the silos was emptying for bed feeding while the other was filling with the returning solid. Fly ash reinjection pipes were disconnected, and the reinjection needles were removed from the bed.

Water circulation was not required and is not shown in Figure 2.3. The in-bed tube bundle was used to study its effect on the fluid dynamics of the fluidized bed. It was removed for those tests that did not require in-bed tubes. The bed was equipped with pressure taps for vertical pressure profile measurements, sampling probes to traverse the bed for solid and tracer gas sampling, and a tracer injection probe for introducing a constant flow of tracer gas (CO_2) into the bed. Sampling probe designs are presented in Section 7.0.

The pressure taps were located at the center line of the walls and were 1.5, 3.5, 4.5, 12, 15, 18, 24, 28, 32, 39, 44 and 73 inches above the tips of the air distributor nozzles. Configurations of the air distributor nozzles and in-bed heat exchangers are described as follows:

- o Air Distributor: The air distributor is a drilled-bolt type. The distributor has two sections of 3'-2" by 1'-7" each and has a total of 2712 nozzles. The configuration of the distributor and the dimensions of a typical nozzle are shown in Figure 2.4.
- o In-Bed Heat Exchanger: The in-bed heat exchanger consists of nine tubes bent into four passes. The tubes are arranged to form a 3 7/8" x 5" pitch bundle, with inlet and outlet headers located on the same side. Each tube is 1 1/2" OD and 29" long. The total external surface area of the tubes is 48 ft². A schematic diagram of the bundle in the PDU boiler is shown in Figure 2.5.

2.3 MEASUREMENT METHODS DURING COLD MODE TESTING

The vertical pressure profile of the bed is measured relative to the pressure tap 1.5 inches above the nozzles. Fluidizing air and solids transport air are measured by orifice plates. Orifice pressure drop and upstream gauge pressure are measured with water manometers. Upstream orifice air temperature is recorded on a chart recorder.

While solids flow is controlled by rotary valves, it is measured by recording load cell readings. The rate is calculated by dividing the difference in two load cell readings by the elapsed time. The solids flow rate into the bed is determined from the silo being emptied. The solids drain rate is determined from the silo being filled.

Solids entrainment is determined by: (1) turning off the induced draft fan to bring the pressure inside the cyclone dust collector close to atmospheric, and (2) then opening the gate valve at the bottom of the cyclone to empty the collected solids. The collected solids were less than the amount entrained from the bed because of solids loss in the cyclone overflow. The collected solids determined during the test runs were rather small because of the short durations of the test runs. High errors would be expected. For this reason, data on collected solids are available, but would not be sufficiently accurate for elutriation study. Therefore, these data will not be reported.

Solids and gas samples are collected with probes arranged in pairs. One pair of probes traverses the bed from the side wall of the PDU. The other pair traverses the bed from the front wall. Only one pair of probes was utilized during tests covered in this report. Solids samples were collected and analyzed for size distribution, bulk and particle densities. Gas samples were analyzed for tracer gas concentration. The gas sampling and analysis system is further described in Section 7.0.

Bed temperatures are measured by thermocouples at various locations in the bed. The mean bed temperature was determined to be the arithmetic mean of the thermocouple readings if these readings are not uniform. Ambient conditions are determined using a barometer and mercury thermometer. Relative humidity is recorded, but was not used as a variable in analyzing test results.

Bed height under fluidizing conditions (expanded bed height) was determined from the vertical pressure profile. Slumped bed height was measured with a marked rod after defluidizing the bed and through the middle manhold located above the bed level (See Figure 2.2).

Table 2.1

ALEXANDRIA PDU GENERAL FEATURES

Nominal Steam Flow	5000 lbs/hr
Outlet Temperature/Pressure	Saturated/300 psig
Bed Dimensions, Nominal	3' x 3'
Bed Dimensions, Actual	3' - 4-½" x 3' - 4-½"
Design Coal	Flexible
Heating Value	≈12,500 BTU/LB
Sulfur Content	>3%
Mean Size	Flexible
Design SO ₂ Absorbent	Limestone, Flexible
Mean Size	≈1000 Microns
Operational Features	
Bed Depth (expanded)	2 ½ ft
Bed Temperature	1550° F
Fluidizing Velocity	6-12 ft/s
Freeboard Height	9-½ ft
Bed Tube, OD	1-½ in., 4 passes, Horizontal
Bed Tube, Pitch	3-7/8" x 5", rectangular
Bed Tube, External Surface	48 ft ²
Bed Tube, Height	15 in. Center to Center
Ca/S Molar Ratio	3:1, Flexible
Solids Feed	1 point, Needle Type
Fly Ash Reinjection	2 points, Needle Type
Water/Steam Circuit	Closed Circuit, Forced Circulation, heat rejected to atmosphere
Final Dust Cleaning	Baghouse

FIGURE 2.1: Alexandria PDU Flow Diagram

FIGURE 2.2 - Fluid-Bed Combustor Cut-Away View

FIGURE 2.3 - Cold Mode Testing Equipment

FIGURE 2.4 - Air Distributor Configuration

FIGURE 2.5 - In-Bed Heat Exchanger

3.0 MIT MODEL VALIDATION APPROACHES

3.1 MIT MODEL DESCRIPTION

The MIT Energy Laboratory has developed a comprehensive mechanistic model of Atmospheric Fluidized-Bed Combustors. The objective is an analytical framework for the design and scale-up of fluidized-bed combustors.

The MIT modeling effort was initially focused on evaluation and application of state-of-the-art models. The initial system model was divided into five basic components: fluid dynamics, combustion, sulfur capture, heat transfer and emissions. Due to the technical complexity of modeling AFBC operations, MIT anticipated evolutionary improvements in understanding and correlating AFBC phenomena. Therefore, the Model is designed in a modular form; each submodel can be replaced or modified without reprogramming of the entire system model.

Upgrading efforts to date have taken two approaches:

The First Path Improvement: MIT discovered that the behavior of relatively large particles (e.g., 1000 μm) differs significantly from the behavior of relatively small particles (e.g., 60 μm). Hence, the initial system model required substantial modifications to both concepts and detailed mechanisms. During the first path improvement, major refinements included:

- Fluid Dynamics: the state-of-the-art fast bubble model was replaced by a two-zone model consisting of a "slow bubble regime" in the lower zone of the bed and a "fast bubble regime" in the upper zone.
- Combustion: the classical two film combustion mechanism was replaced by a one film mechanism in which oxygen reacts directly with carbon on the carbon surface. The role played by chemical kinetic resistance was also incorporated.
- Desulfurization: a new model which takes into account the pore plugging phenomena during the sulfation process was developed to replace the Borgwardt model.
- Heat Transfer: The Vreedenberg model was replaced with a new model which was able to correlate data on large particles as well as on small particles.

The Second Path Improvement: The second path improvement extends the domain of model predictions. The improvements achieved included:

- A bubble growth model developed previously was replaced by a new fluid dynamics model based upon potential flow theory and a more realistic treatment of bubble coalescence.

- An improved two-phase theory was developed and incorporated into the system model.
- In addition to transition between the slow and fast bubble regimes, two criteria were tentatively developed to compute the transition between the fast bubble regime and the turbulent flow regime.
- An interim model for carbon density distribution in the freeboard was structured, based upon the functional relation for solid loading along the freeboard height, and the surface splashing flux. This model is used to compute parameters of interest in freeboard combustion.
- New models were developed for heat transfer to the immersed wall tubes, to the tubes in the splashing zone, and to the tubes in the top section of the freeboard near the gas exit. In the design mode, these models together with the heat transfer model for the horizontal immersed tubes can be used to calculate required heat transfer areas. In the operational mode, they can be used to calculate the bed and exit gas temperature.
- Models for NO_x emission from the bed and freeboard were developed.

With the aid of these new developments, current computational capabilities of the MIT Model are listed in Table 3.1.

Currently, the MIT Model consists of a systematic combination of the following related areas: fluid mechanics and bubble growth, char combustion and associated kinetics for particle burnout, sulfur capture, NO_x formation and reduction, freeboard reactions, and heat transfer.

Figure 3.1 shows a block diagram of the Model program. The overall structure of the program is illustrated in terms of the various submodels. A more detailed outline of the system program can be found in relevant MIT reports.

3.2 MIT MODEL APPLICATION

Input variables supplied by the user are inspected to ensure that they lie inside the allowed range of values. The necessary physical and fluid mechanic properties are calculated and utilized in estimating char combustion and sulfur capture in the bed and the freeboard. NO_x and CO emissions are estimated by taking into account all relevant chemical reactions. A material and energy balance is made around the bed. The Model outputs are then stored and can be printed when required.

The Model contains a number of input and output variables. All input variables have a default value and a specified range of applicability. Tables 3.2 and 3.3 list the input and the output variables of the Model. Output variables do not have default values or limits.

3.3 MODEL VALIDATION APPROACH

The general approach used in Model validation includes the steps shown in Figure 3.2. Essentially, the Model is treated as a "black box" which receives required input values and generates values for output variables. Validation consists of comparing these generated values, or predictions, with experimental data obtained from tests.

Test design begins with a sensitivity analysis to determine those variables or parameters which noticeably affect the predictive results. These variables or parameters are then considered in the validation test program. Only those parameters which fall within the capability of the PDU are included in the test plan.

To illustrate this approach, consider the minimum fluidizing velocity, U_{mf} , defined as:

$$U_{mf} = \text{Function (Particle diameter, density, and sphericity, bed temperature, and voidage at slumped bed condition.)}$$

All independent variables, except bed temperature, are kept constant at typical values and the bed temperature varies arbitrarily from 1000 to 2000°F to compute the sensitivity factor, C. The sensitivity factor, C, is obtained from the following equation:

$$C = \frac{\Delta U_{mf} / U_{mf}}{\Delta T / T}$$

$$\text{Where } T \approx 1550^{\circ}\text{F}$$

$$U_{mf} = \text{typical value at } T$$

If C turns out to be less than 0.10, then U_{mf} is considered insensitive to temperature. Therefore, bed temperature will be maintained constant throughout the test. However, if C is large, then U_{mf} is sensitive to bed temperature and the test program will include the bed temperature as a variable.

During sensitivity analysis, if test parameters are highly sensitive to certain test variables, then these variables are implemented and used as test variables during the actual tests. However, if previous experimental test results are found to be significantly different from the equivalent predicted values of the Model, then subsequent tests are designed independent of the sensitivity analysis.

FIGURE 3.1 - AFBC Systems Model Program Block Diagram

FIGURE 3.2 - Approach to Validation of the MIT Model

TABLE 3.1
COMPUTATIONAL CAPABILITIES OF
THE MIT MODEL

FREEBOARD AREA

1. Heat Transfer
2. Gas Exit Temperature
3. Carbon Fraction
4. Stone Fraction
5. NO_x Emission
6. CO Emission
7. Combustion Fraction
8. Sulfur Retention

SPLASHING ZONE

1. Heat Transfer

IN-BED

1. Heat Transfer, Immersed Tubes
2. Heat Transfer, Wall Tubes
3. Temperature
4. Carbon Loading, Profile
5. Carbon Loading, Overall
6. NO_x Emission
7. CO Emission
8. Combustion Efficiency, Profile
9. Combustion Efficiency, Overall
10. Sulfur Retention, Overall
11. Stone Utilization, Overall
12. Expanded Bed Height
13. Bubble Fraction
14. Void Fraction

TABLE 3.2

INPUT VARIABLES FOR THE MIT MODEL-2**

VARIABLE	DEFAULT VALUE	UNIT*	EXPLANATION	LIMITS	
				LOWER	UPPER
BED PROPERTIES					
1. BEDA	3.355	m ² (ft ²)	Bed Cross-Sectional Area	0	-
2. CONDT	30.0	J/m.s. ^{°K} (BTU/ft-s-°F)	Tube Thermal Conductivity	0	-
3. IDTUB	0.04	m (ft)	Tube Inside Diameter	0	-
4. ODTUB	0.0508	m (ft)	Tube Outside Diameter	0, IDTUB	0.153(0.502)
5. PITCH	0.25	m (ft)	Tube Spacing	0	BEDA ^{1/2} 1.1 (3.6)
6. PSTEM	12156000	Pascals (atmospheres)	Steam Pressure	101300(1.0)	(BEDA) ² 20260000(200.0)
7. TCODE	-1.0	-	Negative: Small Particle Positive: Large Particle		
8. TUBEA	15.4625	m ² (ft ²)	Heat Transfer Area	0	-
9. WIERT.	2280.0	kg (lb)	Mass of Bed Material	10(22)	-
OPERATING CONDITIONS					
1. EXAIR	0.15	-	Excess Air-Mole Basis	0.01	2.0
2. PGAS	101300	Pascals (atmospheres)	Bed Pressure	90000(0.9)	101300(1.0)
3. TACPF	298	°K (°F)	Acceptor Feed Temperature	290(62)	1200(1700)
4. TAIR	298	°K (°F)	Inlet Air Temperature	290(62)	1200(1700)
5. UO	2.0	m/s (ft/s)	Superficial Velocity - Hot	0	8.8(26.2)
6. TGAS	1117.0	°K (°F)	Gas Temperature In-Bed	290(62)	1400(2060)
ACCEPTOR PROPERTIES					
1. CASRT	2.0	-	Ca/S Molar Feed Ratio	1.0	10.0
2. DPART	0.0008	m (ft)	Acceptor Mean Particle Size	-	-
3. RHOAL	2250.0	kg/m ³ (lb/ft ³)	Acceptor Feed Density	-	-
4. SACC	0.008353	-	Molar Ratio of Mg/Ca in Acceptor	0	1.0
5. TAUCH	0.001905	s	Characteristic Sulfation Time	-	-
6. PPRIME	2520000	ppm-s	Hyperbolic Constant	-	-

TABLE 3.2 (Cont'd)
INPUT VARIABLES FOR THE MIT MODEL-2**

VARIABLE	DEFAULT VALUE	UNIT*	EXPLANATION*	LIMITS	
				LOWER	UPPER
CYCLONE PROPERTIES					
1. DCYCLN	10 ⁶	m (ft)	Cyclone Diameter	-	-
2. DPCYC	0	Pascals (atm)	Cyclone Pressure Drop	-	101300 (1.0)
CONVERGENCE & OTHER CODES					
1. CODOUT	-1.0	-	Negative: Ignore TGAS Loop, Cal- culate HT Area	-	-
2. CODJIM	1.0	-	Code to Decide Initial Carbon Loading	-	-
3. HCODE	-1.0	-	-	-	-
4. CODDES	-1.0	-	-	-	-
COAL PROPERTIES					
1. ALPHB	0.402	-	Weight Fraction of Volatiles	-	-
2. HCOMB	6344	kcal/kg (BTU/lb)	Upper Heating Value of Coal	0	-
3. RHOC	1200	kg/m ³ (lb/ft ³)	Coal Apparent Density	750(46.8)	2000(124.9)
4. RHOCT	2500	kg/m ³ (lb/ft ³)	Coal True Density	1000(62.4)	4000(249.7)
5. RMED	8 x 10 ⁻¹⁰	m (ft)	Coal Pore Radius	-	-
6. ROS2	0.00114	m (ft)	Coal Mean Particle Radius	0.0	0.005(0.016)
7. ROS1	1.50	-	Standard Deviation for Distribution	-	-
8. THETA	0.50	-	Porosity of Coal Feed	-	-
9. XASH	0.158	-	Weight Fraction of Ash	0	1.0
10. XC	0.695	-	Weight Fraction of Carbon	0	1.0
11. XH	0.043	-	Weight Fraction of Hydrogen	0	1.0
12. XN	0.014	-	Weight Fraction of Nitrogen	0	1.0
13. XO	0.045	-	Weight Fraction of Oxygen	0	1.0
14. XS	0.045	-	Weight Fraction of Sulfur	0	1.0

*; The hyphen, -, means dimensionless or not reported by MIT

**; English units are included in parentheses for convenience

TABLE 3.3
OUTPUT VARIABLES FOR THE MIT MODEL-2

VARIABLES	UNIT	EXPLANATION
<u>BUBBLE PROPERTIES</u>		
1. DBUB	m	Characteristic Bubble Diameter
2. DEL	-	Characteristic Voids Fraction
3. DELB	-	Characteristic Bubble Fraction
4. H	m	Characteristic Height
5. HT	m	Transition Height
6. KBP	s ⁻¹	Mass Exchange Coefficient
7. UB	m/s	Characteristic Bubble Velocity
8. UF	m/s	Emulsion Gas Velocity
9. DELBMN	-	Average Bubble fraction
<u>COMBUSTION CHARACTERISTICS</u>		
1. COALF	kg/s	Coal Feed Rate
2. COCON	kgmole/s	Oxygen Consumption Rate
3. COE	kgmole/m ³	Effective Oxygen Concentration
4. COMEF	-	Combustion Efficiency
5. COVFL	kg/s	Carbon Overflow Rate
6. ELTOT	kg/s	Elutriation Rate
7. ELTOTE	kg/s	Fines Escape Rate
8. ELTOTR	kg/s	Fines Recycle Rate
9. WCOAL	kg	Carbon Loading
<u>FLUID MECHANIC PROPERTIES</u>		
1. BEDH	m	Expanded Bed Height
2. SBEDH	m	Slumped Bed Height
3. UMF	m/s	Minimum Fluidizing Velocity
4. UT	m/s	Terminal Velocity
<u>HEAT TRANSFER PROPERTIES</u>		
1. Q	kcal/s	Total Heat Absorption by Tubes
2. QCON	kcal/s	Convective Heat Absorption
3. QRAD	kcal/s	Radiative Heat Absorption
4. TUBEN	-	Number of Tubes
5. U	J/s.m. ² k.	Overall Heat Transfer Coefficient

TABLE 3.3 (Cont'd)

OUTPUT VARIABLES FOR THE MIT MODEL-2

VARIABLES	UNIT	EXPLANATION
<u>EFFLUENT GAS STREAM</u>		
1. AIRF	kgmole/s	Inlet Air Flow Rate
2. GASF	kgmole/s	Effective Gas Flow Rate
3. RHOG	kg/m ³	Gas Density-Hot
4. YCO	-	CO Effluent Gas Mole Fraction
5. YCO ₂	-	CO ₂ " " " "
6. YH ₂ O	-	H ₂ O " " " "
7. YNO	-	NO " " " "
8. YN ₂	-	N ₂ " " " "
9. YO ₂	-	O ₂ " " " "
10. YSO ₂	-	SO ₂ " " " "
11. XSO ₂	ppm	ppm of SO ₂ in Effluent Gas
<u>CALCULATED STONE PROPERTIES</u>		
1. ACPTE	kg/s	Effluent Stone Flow Rate
2. ACPTF	kg/s	Feed Stone Flow Rate
3. RHOA	kg/m ³	Effluent Stone Density
4. XSULF	-	Stone Utilization
<u>STEAM PROPERTIES</u>		
1. TSTEM	°K	Steam Temperature
2. WATRF	kg/s	Steam Production Rate
<u>MISCELLANEOUS VARIABLES</u>		
1. PSRET	%	Percent Sulfur Retention
2. COVEL	kg/s	-
3. SO ₂ F	kg/s	-
4. VAR	-	-
5. TAUGAS	s	Characteristic Time
6. FRACT	-	-
7. ATRIT	-	Char Attrition Constant

*: The hyphen means dimensionless or not reported by MIT.

Having determined the scope of the required tests, other test specifications such as procedures, schedule, test parameters, and test conditions will then be formulated and included in the Operating Plan.

Each test series usually lasts about a day. Data are reduced immediately after each series has been completed. Model validation begins as soon as sufficient data is available.

This approach was used during cold mode testing and seemed to provide the most cost-effective approach.

4.0 COLD MODE TEST PROCEDURES AND RESULTS

The primary objective of the cold mode tests was empirical validation of MIT Model predictions. The scope of each test series was defined as described in Section 3.0. The testing was divided into three subtasks with the following designations:

Fluid Dynamics I : Batch tests, Test Series 20
Fluid Dynamics II : Test with solids flow, Test Series 30
Fluid Dynamics III : Test with in-bed tubes, Test Series 40

Prior to the cold mode tests, orifice plates for measuring the flow rate of fluidizing air and coal/limestone conveying air were calibrated with a static-pitot tube. The flow rates calculated from the orifice equation were compared with those calculated from the pitot tube equation. The differences were found negligible, and in the worst cases, did not exceed 5%. Before each test series began, all the instruments (e.g., load cells, Magnehelic gauges, gas analyzer, etc.) were checked and calibrated. In addition, the standard start-up operating procedure for the PDU was followed for each test.

In all the tests, the bed weight, W , is calculated from the experimental values of both the slumped bed heights, H_s , and the bulk density of the bed ρ_B , as follows:

$$W = H_s \cdot A \cdot \rho_B$$

where, A is the cross section area of the bed.

The accuracy of bed weight determination is approximately ± 200 lbs. This is primarily due to: (1) bed surface not being level during the data collection via a measuring stick ($\sim \pm 2$ inches for the measurement), (2) non-uniformities in the bed density from one location to the next within the bed; and (3) random pattern of filling furnace cavities with bed material upon slumping bed.

The objectives, procedures, and results of these tests are described in the sections which follow:

4.1 FLUID DYNAMICS I

4.1.1 Test Objective

The fluidization characteristics of a fixed amount of spent bed material in the absence of solids flow and the in-bed tube bundle were studied. The fluidization characteristics include slumped bed height, expanded bed height, minimum fluidization velocity, and gas dispersion coefficient.

Test conditions and parameters are listed in Table 4.1. Note that the test series started with test 20-4. The preceding three tests, 20-1, 2 and 3, studied the pressure drop across the air distributor and are discussed in Appendix A.

4.1.2 Test Description

As shown in Table 4.1, the tests performed can be divided into two groups. The first group included the minimum fluidizing velocity, expanded bed height and slumped bed heights, which were obtained during Tests 20-4, 5, 6, 7 and 8. The second group included gas dispersion, solids attrition and elutriation. These parameters were examined during Tests 20-9 and 10.

4.1.3 Test Procedures

During Test 20-4 through 20-10, data collected once every fifteen minutes included: all pressures, drafts, temperatures, pressure drops and flow rates. Solids samples were collected once every half-hour. In addition, at least 15 minutes were allowed after a new air flow rate was set for the bed to become stabilized.

Standard log sheets (Ref: Quality Assurance Plan) were used for collection of data. These logs were modified, when necessary, to meet the data collection requirements for each test. The data collection period for these tests varied from four to eight hours.

Test 20-4

An initial amount of 638 lbs. of spent bed material was fluidized in the PDU using ambient temperature air. The slumped bed height before the test series was 9.0 inches. Testing began with 0.15 inches of water pressure drop across the main air orifice plate and later was increased in steps to 0.3, 0.5, 0.7, 1.0, 2.1, 3.0, 4.0, and 5.1. The pressure drop was then decreased in steps to 5.1, 3.1, 1.5, 1.0, 0.8, 0.6, and 0.4 inches of water. At this time, air flow was cut off and the slumped bed height was measured to be 9.0 inches. Up to this point, dust collected from cyclone underflow was negligible. The test series was continued by setting the pressure drop across the main air orifice plate at 5.0 inches of water immediately after slumped bed height was measured. Pressure drop was then increased in steps to 6.0, 6.8, 8.0, 8.8, 9.8, 11.0, 12.0, 13.2, 14.0, 15.8, 18.1, 20.1 and 21.9 inches of water. At the conclusion of the test the slumped bed height was measured and found to be 7.5 inches.

Test 20-5

An initial amount of 1135 lbs. of spent bed material was fluidized. The slumped bed height before the test was 15 inches. The test was started with 0.3 inches of water pressure drop across the main air orifice plate, and later was increased in steps to 0.5, 0.8, 1.05, 2.0, 3.0, 4.0, 4.8, and 7.0. The air orifice pressure drop was then decreased in steps to 6.0, 4.1, 2.0, 0.8, 0.5, and 0.3 inches of water. The final slumped bed height was 14.5 inches.

Test 20-6

An initial amount of 1312 lbs. of spent bed material was fluidized in the PDU. The fluidizing air was pre-heated up to a maximum temperature of 500°F by the under-bed burner. The slumped bed height prior to the test series was 18.5 inches. The test began with 2 inches of water pressure drop across the main air orifice plate, and later was changed to 0.9, 0.3, 1.0, 2.0, 3.0, 3.9, 4.8, 6.0, and 6.7. Further increase in air flow was not possible without losing excessive amounts of solids. The air orifice pressure drop was then decreased in steps to 5.3, 3.2, 1.0, 0.7, 0.5, and 0.4 inches of water. At the conclusion of the test, the slumped bed height was 16 inches.

Test 20-7

An initial amount of 2200 lbs. of spent bed material was fluidized in the PDU. The test began at the maximum air flow rate, which was equivalent to about 13.6 inches of water pressure drop across the main air orifice plate. The air orifice pressure drop was then reduced in steps to 9.5, 8.7, 7.8, 5.9, 3.8, 1.9, 0.6, 0.4, and 0.1 inches of water with the last 3 steps being in the slumped bed region. At the end of the test, the slumped bed height was found to be 23 inches.

The test was continued with the air orifice pressure drop being increased in steps to 1.8, 4.9, 8.7, 11.4, and finally 15.0 inches of water. The final slumped bed level was 19 inches, indicating a total loss of about 850 lbs. of bed material.

This test started out with high air flow so that most of the fines would be removed from the bed in the first one or two runs. However, it appeared from a brief examination of the data that fines were generated in the bed as far as they could be carried out of the PDU by entrainment.

This test was designed to run under ambient temperature but due to the friction loss in the blower, actual test temperatures were slightly higher than the ambient temperature.

Test 20-8

This test involved the same test parameters as Test 20-7 except that the air was heated by the under-bed burner to 450-520°F.

Because of losses in the previous test, 1200 lbs. of spent bed material was added to the PDU before the test began. The slumped bed level was 33.0 inches above the air distributor nozzles before the test began. The maximum air flow rate corresponded to 7 inches of water pressure drop across the main air orifice plate. This pressure drop was later reduced in steps to 5, 3.6, 3, 2.1, 0.8, 0.8, 0.4, and 0.3 inches of water without entering the slumped bed region. Further decrease in air flow caused instability in the under-bed burner flame and consequently uncontrollable bed temperature. After several attempts to enter the slumped bed region failed, the air orifice pressure drop was increased in steps to 0.6, 1.6, 2.7, 5, 7, and 0.3 inches of water.

In a final attempt to achieve the slumped bed conditions, the burner and one of the ID fans were turned off to reduce the cell draft. With one of the ID fans off, the burner was started up and the slumped bed conditions were achieved at an air orifice pressure drop of 0.15 inches of water. One set of data was taken (Run No. 14) when the burner was temporarily turned off to prevent unsafe conditions due to a temperature run-away situation. As a result the bed temperature dropped (at this point the bed was at or near slumped conditions). The final slumped bed height, measured the next day when the bed was cooled, was 21 inches.

Tests 20-9 and 10

The purpose of these tests was to study the gas dispersion in the bed and the elutriation losses at a constant fluidizing velocity. Carbon dioxide was used as a tracer gas for measuring the gas dispersion in the bed. It was injected into the bed at the same velocity as the in-situ velocity of the fluidizing air. The tracer gas injector was located at the center of the PDU, 8.0 inches above the air distributor nozzles. The vertical distance between the injector and the gas sampling probe, at the center of the PDU, was 16 inches. Gas samples were taken at different locations along the sample probe axis, and were analyzed for CO₂ concentration.

During Test 20-9, an initial amount of 2518 lbs. of spent bed material was fluidized with ambient temperature air. The air flow rate was kept constant at 7490 lb/hr. At this air flow rate, the expanded bed height was determined from vertical pressure profile measurements, and the bed voidage, ϵ_f , was calculated for determining the in-situ velocity of air. From the in-situ air velocity, the CO₂ flow rate was determined and set at 8.0 SCFH during the test period. Gas samples were then taken at 0, 5, 10 and 18 inches from the center of the PDU. The slumped bed height at the start and end of the test was 27.0 and 23.8 inches, respectively.

During Test 20-10, an initial amount of 2880 lbs. of spent bed material was fluidized with ambient temperature air. The air flow was kept constant at 10676 lb/hr and the CO₂ flow rate was set at 11.5 SCFH. Gas samples were taken at 1, 1.5, 5 and 10 inches from the center of the PDU. The slumped bed height at the start and end of the test was 27.0 and 23.0 inches, respectively.

4.1.4 Results

The reduced data for these tests are presented in Tables 4.2 through 4.10. An explanation for the terms used in these Tables is given in Section A-4 of the Appendix.

4.2 FLUID DYNAMICS II

4.2.1 Test Objective

Fluid Dynamics II consisted of three tests: Number 30-1, 2, and 3. These tests were performed to study the fluidizing characteristics of spent bed material, with solids flow into the PDU. This involved determining the bed pressure drop, and expanded bed height characteristics of the PDU under various air and solids flow rates, in absence of the in-bed tube bundle.

4.2.2 Test Description

Table 4.11 summarizes the duration, operating conditions, and parameters for Tests 30-1, 2 and 3. Spent bed material was used as feed material. The bed weight was controlled at the nominal values of 1000 and 2000 lbs. The test series was performed at a temperature of about 110°F. Solids samples, as well as data for determining the gas dispersion coefficient, were obtained.

4.2.3 Test Procedures

During Tests 30-1 and 30-2, bed pressure drop, expanded bed height, elutriation and attrition, and fluid dynamic responses were studied. During Test 30-3, minimum fluidizing velocity and gas dispersion were also studied.

Pressure, draft and temperature data were recorded every fifteen minutes. Data for spent bed material feed and discharge rates were recorded every five minutes. Solids samples were taken periodically.

Test 30-1

Air at ambient temperature and at a constant flow rate was used to fluidize the spent bed material in the PDU. There were four test runs of varying solids flow rates. Each test run consisted of two periods. During the first period, spent bed material feed and discharge rates were set to maintain a constant bed height. During the second period, the feed rate was kept the same, but the discharge was varied to determine the rise of expanded bed height with respect to time.

Prior to the test, the slumped bed height was 19.3 inches. The first test run started with spent bed material feed and discharge rates of 612 and 606 lbs/hr respectively. For the subsequent three test runs, feed and discharge rates were changed to 449 and 544, 40 and 9.6, and 10 and 0 lbs/hr. The discharge rate for the second test run was higher than the feed rate due to the relative insensitivity of the solids flow control system. The solids flow rates for the last two runs were much smaller than the preceding run due to false display on the control setting gauge. This was later corrected.

The fluidizing air flow rate was intended to be held constant; however, as the bed height fluctuated, the air flow varied slightly. The range of air flow variation was from 6766 lbs/hr to 6891 lbs/hr, which is quite acceptable. The slumped bed height after the completion of the test was 21.0 inches.

Test 30-2

During this test, the same procedure was used as in Test 30-1. The test started with a slumped bed height of 18.5 inches and spent bed material feed and discharge rates of 611 and 532 lbs/hr, respectively. The feed and discharge rates were 454 and 474; 357 and 390; and finally 187 and 222. The fluidizing air flow rate was to be kept constant, but as the bed height fluctuated, the air flow varied from 8996 lbs/hr to 9926 lbs/hr. The slumped bed height after completion of the test was measured to be 20.3 inches.

When the discharge rate was set to zero by closing the discharge block valve during the second half of each test run period, small accumulations were still recorded for the discharge storage bin as shown in Table 4.13. These accumulations resulted from incomplete valve closure. This was later corrected.

Test 30-3

This test consisted of five runs. During run numbers one through four, pressure drop, expanded bed height, elutriation and attrition, gas dispersion, and fluid dynamic responses were studied.

During these runs, pressure, draft, and temperature data were recorded every fifteen minutes. Solids samples collected in the cyclone were taken every 30 minutes. Samples of spent bed material were collected at the beginning and end of each run. Samples were taken from the center of the PDU at 12 and 21 inches above the grid plate, and from the rear right-hand side corner of the PDU at 2 inches above the grid plate. The slumped bed height was also measured at the beginning and end of each run. The gas dispersion was measured in the same manner as for Tests 20-9 and 20-10.

Prior to the test series, the feed and discharge rates were set at 600 lbs/hr. However, because of the relative insensitivity of the control system, rates deviated slightly from 600 lbs/hr. Air flow rate was intended to be held constant, but actually fluctuated during each run. The variations in these flow rates were small and within acceptable ranges (See Table 4.14).

During run number five, the minimum fluidization velocity was studied. Prior to this run, the slumped bed height was 33.8 inches. This run started with 1.7 inches of water pressure drop across the main air orifice plate. This pressure drop was later changed in steps to 1.2, 0.6, 0.3, 2.6, 3.2, and 4.2 inches of water. During this run, no spent bed material was fed or discharged. Pressure, draft and temperature data were recorded after each change. Solids and gas samples were not collected. The slumped bed height was 33.3 inches at the end of the test.

4.2.4 Results

The reduced data for this test series are presented in Tables 4.12 to 4.16. An explanation for the terms used in these tables is given in Section A-4 of the Appendix.

4.3 FLUID DYNAMICS III

4.3.1 Test Objectives

Fluid Dynamics III consists of Test Numbers 40-1, 2, 3, and 4. These tests studied the effects of the in-bed tube bundle on fluid dynamics.

4.3.2 Test Description

Table 4.17 summarizes the duration, operating conditions, and parameters for these tests. The bed pressure drop, expanded bed height, elutriation and attrition characteristics of the PDU under various air flow and solid flow rates were determined with the in-bed bundle installed. Emphasis was placed on determining the effects of the bundle on the pressure drop, expanded bed height and gas dispersion coefficient. Spent bed material was used as feed material. The bed weight was controlled.

These tests were performed at ambient temperature. Solids samples were collected from different locations in the bed to determine solids size distribution profile.

4.3.3 Test Procedures

Pressure, draft, and temperature measurements were recorded every 15 minutes. Flow rates of spent bed material, feed and discharge were recorded every five minutes. The slumped bed height was measured at the beginning and at the end of each test series. The solids sample collected in the cyclone was taken at the beginning and at the end of each run.

During each test run, the feed and discharge rates were to be kept constant. However, because of the relative insensitivity of the control system, the flow rates fluctuated slightly.

Test 40-1

Solids samples were collected during each run at different locations from the center of the PDU to the wall. The two solids sampling probes, when at the center of the PDU, were 12.5 and 27.5 inches above the grid plate. A detailed description of these probes is presented in Section 7.1.

Prior to run number 1, the slumped bed height was measured at 21.0 inches. Air at ambient temperature was used to fluidize the bed, and the test series started with 8.0 inches pressure drop across the main air orifice plate. This pressure drop was later reduced in steps to 6.0, 4.0, 2.0, 1.6, 1.0, 0.5, 0.2, and 0.15 inches of water. No spent bed material was fed or discharged when the bed was at slumped conditions. The slumped bed height at the completion of the test was measured to be 20.5 inches.

Test 40-2

During this test, measuring the in-bed gas dispersion was emphasized. The measuring method for gas dispersion has been described in Tests 20-9 and 20-10 Section.

Prior to this test, the slumped bed height was 20.5 inches. Air at ambient temperature fluidized the bed and the test started with 8.0 inches of water pressure drop across the main air orifice plate. This pressure drop was later reduced in steps to 4.0, 1.7, 0.8 and 0.4 inches of water. The slumped bed height at the completion of the test was 20.5 inches. Spent bed material samples were taken at the beginning and at the end of each run. Sampling locations are listed in Table 4.27.

Test 40-3

Prior to this test, the slumped bed height was 14.0 inches. The first run started with 8.1 inches of water pressure drop across the main air orifice plate. This pressure drop was later changed in steps to 6.1, 5.2, 4.0, 3.0, 2.0, 1.5, 0.7 and 0.3 inches of water. Bed samples were collected at different locations from the center of the bed to the wall via the two solids sampling probes described in Section 7.1. At the end of the test, the slumped bed height was measured to be 14.0 inches.

Test 40-4

This test started with 6.0 inches of water pressure drop across the main air orifice plate. The air flow and temperature were kept constant throughout the test. Air at ambient temperature fluidized the bed. No solids were fed or discharged. This test consisted of two runs. During run number one, bed samples were collected at different locations from the center of the PDU to the wall via two solids samples probes described in Section 7.1. During the second run, carbon dioxide was used as the tracer gas for measuring the gas dispersion in the bed. The measuring method for gas dispersion has been described in the Tests 20-9 and 20-10 Section. Slumped bed height prior to initiation and at the completion of the test was measured to be 25 inches.

4.3.4 Results

The reduced data of these tests are presented in Tables 4.18 through 4.30. An explanation for the terms used in these tables is given in Section A-4 of the Appendix.

4.4 Air Distributor Leak Test

A series of unscheduled cold mode tests were performed to evaluate the effects of maldistribution of distributor air, caused by leaks, on the expanded bed height and minimum fluidizing velocity. Testing is performed with shallow bed ($\approx 15"$) and normal bed ($\approx 21"$). Tests were also performed to quantify the air distributor leakage and characterize the distributor.

Experimental results and conclusions are in Appendix C and Section 5.2.5, respectively.

TABLE 4.1

FLUID DYNAMICS I TEST MATRIX

Test Numbers	Test Duration (hrs)	TEST CONDITIONS ⁽¹⁾			Test Parameters ⁽³⁾	In-Bed Measurements
		Solids Weight In-Bed (lbs)	Air ⁽²⁾ Flow (SCFM)	In-Bed Air Temperature (°F)		
20-4	4	600	0-3330	Ambient	U _{mf} , H, H _s , and responses	No
20-5,6	4,8	1200	0-2500	Ambient,500	same as above	No
20-7,8	4,8	1500	0-2500	Ambient,500	same as above	No
20-9	10	1850	2xQ _{mf}	Ambient	Gas dispersion, U _{mf} , H, and responses	Gas Sampling
20-10	10	2000	5xQ _{mf}	Ambient	same as above	Gas Sampling

(1) These values are nominal settings. Actual values during the test are somewhat different.

(2) Q_{mf} denotes air volumetric flow at minimum fluidization.

(3) U_{mf} denotes minimum fluidization velocity; H, expanded bed height; and H_s, slumped bed height.

TABLE 4.2
TEST 20-4 DATA⁽¹⁾

Time	Run No.	Air Flow (lb/hr)	Bed		Bed DP ⁽²⁾ (in. H ₂ O)
			Temp. (°F)	Press. (atm)	
0	1	1281	125	0.891	3.0
20	2	1806	115	0.894	3.5
40	3	2329	110	0.896	3.9
55	4	2759	112	0.898	4.9
65	5	3299	112	0.900	5.6
85	6	4789	120	0.903	6.9
100	7	5740	120	0.903	6.9
115	8	6627	120	0.903	6.9
130	9	7504	120	0.903	6.9
145	10	5825	122	0.903	6.9
155	11	4033	120	0.903	6.4
170	12	3286	115	0.901	5.4
185	13	2932	120	0.900	4.7
195	14	2426	120	0.898	4.0
205	15	2344	118	0.897	3.5
0 ⁽³⁾	NA	0.0	NA	NA	NA
15	16	7384	120	0.895	6.7
30	17	8088	125	0.902	6.7
45	18	8620	120	0.903	6.6
60	19	9353	120	0.902	6.6
75	20	9813	128	0.902	6.6
90	21	10347	120	0.902	6.6
105	22	10973	128	0.902	6.5
120	23	11464	119	0.902	6.4
135	24	12029	114	0.902	6.4
150	25	12257	117	0.902	6.3
165	26	13158	120	0.902	6.3
180	27	14085	111	0.902	6.3
195	28	14846	112	0.902	6.3
210	29	15512	115	0.902	6.0

- (1) Expanded bed height could not be determined because of insufficient vertical pressure profile data due to shallowness of the bed.
- (2) Pressure drop from 1.5 inches above the grid plate to the surface of the bed. Pressure drop for the lower 1.5 inches of the bed could not be determined because of the lack of vertical pressure profile (see Note 1). As a result, total bed DP and U_{mf} could not be determined.
- (3) Restarted the Test.
NA: Not Applicable

TABLE 4.3
TEST 20-5 DATA

Run No.	Air Flow ($\frac{\text{lb}}{\text{hr}}$)	Bed		Total Bed ΔP (in H_2O)	Expanded Bed Height(in)
		Temp ($^{\circ}\text{F}$)	Press (atm)		
1	1655	119	0.916	8.2	15.2
2	2134	120	0.921	11.1	15.2
3	2700	120	0.927	14.1	16.6
4	3095	121	0.930	15.3	15.4
5	4279	124	0.932	16.4	15.2
6	5253	125	0.932	16.3	15.5
7	6075	124	0.932	16.2	15.4
8	6660	125	0.932	15.8	15.7
9	8043	125	0.931	15.8	15.6
10	7440	125	0.932	15.8	15.6
11	6136	125	0.931	16.1	15.4
12	4266	125	0.931	15.9	15.3
13	2696	125	0.926	13.4	14.6
14	2121	127	0.920	10.0	14.6
15	1636	125	0.914	7.8	14.5

The minimum fluidization velocity for this test is graphically determined to be 1.2 ft/s.

TABLE 4.4
TEST 20-6 DATA

Run No.	Air Flow (lb/hr)	Bed		Total Bed DP (in H ₂ O)	Expanded Bed Height (in.)
		Temp (°F)	Press. (atm)		
1	5364	438	0.906	17.3	20.5
2	3864	325	0.909	17.3	20.4
3	2480	318	0.907	16.8	19.0
4	4060	400	0.906	17.3	19.8
5	5332	392	0.907	17.3	19.8
6	6358	440	0.906	17.3	19.7
7	7152	500	0.904	17.3	20.5
8	7870	380	0.908	17.6	20.5
9	8704	292	0.909	16.8	19.8
10	9168	261	0.911	17.3	20.8
11	8009	315	0.908	16.9	20.0
12	6258	365	0.907	16.8	19.8
13	3984	355	0.906	16.3	18.6
14	3344	370	0.906	16.2	19.0
15	2990	358	0.905	16.0	18.1
16	2639	307	0.903	15.8	17.4

The minimum fluidization velocity for this test is graphically determined to be 1.1 ft/s.

TABLE 4.5
TEST 20-7 DATA

Run No.	Air Flow ($\frac{\text{lb}}{\text{hr}}$)	Bed		Total Bed DP (in H ₂ O)	Expanded Bed Height (in)
		Temp. (°F)	Press. (atm)		
1	11828	130	0.955	23.4	26.4
2	NR	130	0.959	23.1	26.4
3	9867	130	0.954	23.1	26.0
4	9434	130	0.953	23.2	25.8
5	8898	135	0.953	22.7	26.5
6	7734	140	0.953	23.1	25.4
7	6190	145	0.950	23.3	25.0
8	4365	158	0.949	23.0	25.2
9	2444	170	0.944	19.2	23.4
10	1984	155	0.940	16.0	24.0
11	989	170	0.918	8.0	24.0
12	4236	180	0.950	22.4	24.6
13	7029	165	0.951	22.3	25.4
14	9388	150	0.952	21.7	25.2
15	10762	150	0.913	21.5	25.2
16	12407	130	0.950	20.6	25.2

The minimum fluidization velocity for this test is graphically determined to be 1.0 ft/s.

NR - Not Recorded

TABLE 4.6
TEST 20-8 DATA

Run No.	Air Flow ($\frac{\text{lb}}{\text{hr}}$)	Bed		Total Bed DP (in H_2O)	Expanded Bed Height (in)
		Temp ($^{\circ}\text{F}$)	Press (atm)		
1	9252	490	0.920	26.5	26.0
2	7947	490	0.919	26.4	26.0
3	6800	486	0.919	26.5	25.4
4	6250	480	0.920	25.5	25.4
5	5111	505	0.918	25.9	25.8
6	3188	512	0.912	25.1	23.8
7	3230	510	0.917	26.3	24.5
8	2360	520	0.915	25.7	22.5
9	2079	510	0.908	25.2	23.5
10	3101	480	0.920	26.6	25.0
11	4487	502	0.919	26.5	25.8
12	5767	510	0.919	26.5	25.5
13	7812	490	0.893	26.6	25.5
14	9276	475	0.893	26.5	26.4
15	1793	240*	0.916	16.1	22.5
16	1654	450	0.912	21.4	23.0

*- The underbed burner flame was lost.

The minimum fluidization velocity for this test is graphically determined to be 1.2 ft/s.

TABLE 4.7
TEST 20-9 DATA

Time (min.)	Bed		Expanded Bed Height (in.)	Total Bed DP (in. H ₂ O)
	Temp (°F)	Press (atm)		
0	101	NR	NR	NR
15	102	NR	NR	NR
30	105	0.913	25.7	27.5
45	107	0.913	25.5	26.3
60	107	0.913	26.0	26.4
75	100	0.910	25.9	26.5
90	107	0.912	25.9	26.5
105	107	0.911	27.2	26.5
120	107	0.912	27.7	26.0
135	107	0.911	25.7	26.3
150	109	0.912	25.5	26.0
165	109	0.912	26.8	26.0
180	101	0.911	27.8	25.8
195	101	0.911	26.3	25.7
210	100	0.910	25.0	25.7
225	100	0.910	25.3	25.7
240	107	0.910	25.3	25.5
255	106	0.909	24.7	25.3
270	105	0.910	25.0	25.0
285	102	0.910	25.0	25.0
300	102	0.910	25.0	25.0
315	106	0.909	25.0	25.0
330	107	0.909	25.2	24.5
345	109	0.907	24.7	24.8
360	101	0.907	24.5	24.8
375	101	0.907	25.2	25.0
390	109	0.907	24.0	24.5
405	109	0.907	24.5	24.5

Notes:

Air Flow, (lbs/hr)

7490

NR - Not Recorded

TABLE 4.8

TEST 20-9 IN-BED CO₂ CONCENTRATION
MEASUREMENTS

Sample No.	Sampling Location from the center of the PDU (in.)	CO ₂ Concentration Reading (Volume %)				
		Zero Gas	Air	Sample Gas	Air	Zero Gas
1	0	0	0.018	0.082	0.018	0
2	0	0	0.018	0.082	0.018	0
3	0	0	0.018	0.082	NR	0
4	10	0	NR	0.034	NR	0
5	10	0	0.018	0.036	NR	0
6	10	0	NR	0.034	0.018	0
7	18	0	NR	0.021	0.018	0
8	18	0	0.018	0.020	0.018	0
9	5	0	0.019	0.064	0.019	0
10	5	0	0.019	0.060	0.019	0
11	0	0	0.019	0.074	0.019	0
12	0	0	0.020	0.070	0.019	0
13	0	0	0.022	0.084	0.024	0

Notes: The CO₂ injector was located at the center of the PDU, at 8.0 inches above the air distributor nozzles. The vertical distance between the injector and the gas sampling probe, when at the center of the PDU, was 16 inches.

NR: Not Recorded

TABLE 4.9
TEST 20-10 DATA

Time (min)	Bed		Expanded Bed Height (in.)	Total Bed DP (in. H ₂ O)
	Temp (°F)	Press (atm)		
0	102	NR	NR	NR
30	102	NR	NR	NR
45	102	0.909	28.6	27.4
60	102	0.907	28.5	27.4
75	102	0.907	28.0	26.4
90	103	0.907	27.8	26.4
105	103	0.905	27.4	26.4
120	103	0.905	27.0	25.9
135	104	0.905	26.8	26.0
150	104	0.906	26.2	25.5
165	104	0.905	26.4	25.5
180	102	0.904	26.4	25.5
195	105	0.903	25.2	25.5
210	105	0.903	24.2	25.3
225	105	0.903	24.6	25.5
240	105	0.903	24.0	25.2
255	102	0.904	24.0	25.0
270	102	0.904	25.7	25.0
285	101	0.904	24.0	25.3
300	101	0.904	24.6	25.0
315	102	0.903	25.5	25.0
330	102	0.903	25.4	25.0
345	103	0.903	23.8	25.0
360	103	0.903	25.0	24.5
375	103	NR	NR	NR
390	103	NR	NR	NR

Notes:

Air Flow, (lbs/hr)

10676

NR - Not Recorded

TABLE 4.10

TEST 20-10 IN-BED CO₂
CONCENTRATION MEASUREMENTS

Sample No.	Sampling Location from the center of the PDU (in.)	CO ₂ Concentration Reading (Volume %)				
		Zero Gas	Air	Sample Gas	Air	Zero Gas
1	1	0	0.126	0.161	0.126	0
2	1.5	0	0.126	0.144	0.126	NR
3	0	NR	0.126	0.250	0.123	NR
4	0	NR	0.123	0.240	0.126	NR
5	10	NR	0.124	0.120	0.124	NR
6	5	NR	0.125	0.162	0.110	NR
7	5	NR	0.110	0.145	0.126	NR
8	5	NR	0.126	0.148	0.126	NR
9	0	NR	0.126	0.200	0.126	NR
10	0	NR	0.126	0.202	0.126	NR
11	0	NR	0.126	0.200	0.126	NR

Notes: The CO₂ injector is located at the center of the PDU, at 8.0 inches above the air distributor nozzles. The vertical distance between the injector and the gas sampling probe, when at the center of the PDU, was 16 inches.

NR: Not Recorded

TABLE 4.11

FLUID DYNAMICS II TEST MATRIX

Test Nos.	Test Duration (hrs)	Test Conditions*				Test Parameters	In-Bed Measurements
		Bed Weight (lb)	Air Flow (SCFM)	Solids Flow (lb/hr)	Bed Temp. (°F)		
30-1	10	1000	1400	0-600	Ambient	Pressure drop, bed height, fluid dynamic responses	Solid Sampling
30-2	10	1000	2000	0-600	Ambient	Same as above	Solid Sampling
30-3	8	2000	0-2100	600	Ambient	Same as above, plus gas dispersion, minimum fluidization velocity	Solid and Gas Sampling

*These values are nominal settings. Actual values during the tests are somewhat different, see text.

TABLE 4.12

TEST 30-1 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Solids Feed Rate (lbs/hr)
			Temp. (°F)	Press. (atm)			
1	0	7328	100	1.020	16.7	17.5	611
	15	7286	100	1.020	16.3	16.8	
	30	7286	100	1.023	18.0	18.8	
	45	7286	100	1.024	19.4	20.6	
2	0	7286	108	1.021	18.5	18.5	449
	15	7286	108	1.022	18.0	18.0	
	35	7286	108	1.027	20.4	21.3	
	50	7286	108	1.029	21.0	22.0	
3	0	7286	100	1.027	20.2	21.8	40
	16	7328	100	1.028	20.4	22.2	
	35	7328	100	1.028	20.4	22.2	
	50	7328	100	1.028	20.4	22.2	
4	0	7411	86	1.026	18.7	20.4	10
	15	7411	87	1.026	18.8	19.5	
	45	7411	87	1.026	18.8	21.6	

TABLE 4.13

TEST 30-2 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Solids Feed Rate (lbs/hr)
			Temp. (°F)	Press. (atm)			
1	0	10,013	99	0.999	18.8	15.6	611
	15	10,013	99	0.999	18.8	15.6	
	30	10,013	99	0.999	18.8	15.6	
	45	10,013	99	0.999	18.8	15.6	
	60	9,705	99	0.999	20.2	17.0	
	75	9,322	99	0.999	22.7	20.2	
2	0	9,768	100	0.999	18.5	19.0	454
	15	9,768	100	0.999	17.8	18.6	
	30	9,891	100	0.999	17.7	18.8	
	45	9,891	100	0.999	17.9	18.9	
	60	9,579	100	1.004	19.4	21.7	
	75	9,387	100	1.005	21.4	22.5	
3	0	10,073	95	1.000	17.5	17.8	357
	13	10,252	96	0.999	18.4	18.0	
	28	10,252	96	1.000	18.5	18.4	
	43	10,073	97	0.999	18.1	18.0	
	58	9,891	99	1.001	19.3	19.0	
	73	9,705	100	1.004	20.7	20.8	
4	0	9,893	99	1.000	18.5	18.0	187
	15	10,014	99	0.998	18.2	17.5	
	30	9,954	99	0.999	18.2	17.5	
	45	9,954	98	0.999	18.6	18.0	
	60	9,830	98	1.000	18.7	18.2	
	75	9,768	98	1.002	19.5	19.2	

TABLE 4.14

TEST 30-3 DATA

Fun No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Average Solids Feed (lbs/hr)
			Temp. (°F)	Press. (atm)			
1	0	10,154	80	1.040	31.3	37.2	611
	15	10,141	82	1.042	31.3	37.2	
	30	10,133	85	1.042	31.3	36.6	
	45	10,131	85	1.039	31.3	37.2	
	60	10,120	88	1.039	31.3	36.0	
	75	10,412	88	1.039	31.3	37.5	
	90	10,418	89	1.040	31.3	37.8	
2	0	8,975	89	1.039	31.3	36.5	583
	15	8,977	89	1.039	31.3	36.5	
	30	8,552	90	1.043	33.3	39.0	
	45	8,625	90	1.043	31.3	35.8	
3	0	5,198	90	1.042	32.4	35.5	611
	15	5,060	90	1.042	31.4	34.5	
	30	5,060	90	1.044	32.4	35.8	
	45	5,189	90	1.033	31.9	34.5	
	60	5,194	90	1.044	33.4	36.2	
4	0	7,420	90	1.043	32.4	35.8	581
	15	7,417	90	1.041	31.3	36.5	
	30	7,421	88	1.041	31.3	36.5	
	45	7,431	84	1.044	31.3	37.2	
	60	7,433	84	1.042	31.3	35.5	
	75	7,456	82	1.044	31.3	36.8	
	90	7,472	82	1.040	31.3	35.8	
	105	7,476	81	1.042	32.0	36.0	
	120	7,476	80	1.042	31.3	35.8	

TEST 4.14 (Cont'd)

TEST 30-3 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Average Solids Feed (lbs/hr)
			Temp. (°F)	Press. (atm)			
5	0	4499	89	1.038	31.3	35.5	0
	4	3785	100	1.036	30.3	35.0	
	13	2674	100	1.030	27.2	35.5	
	19	1892	100	1.013	23.0	34.0	
	39	5592	96	1.037	31.4	34.8	
	44	6207	96	1.037	31.4	35.4	
	59	7176	94	1.037	31.9	35.0	

Minimum Fluidizing Velocity = 0.8 ft/s.

TABLE 4.15

TEST 30-3 IN-BED CO₂
CONCENTRATION MEASUREMENTS

Run No.	CO ₂ Flow Rate (SCFH)	Air Flow Rate (SCFH)	Sampling Location From The Center Of The PDU (in.)	CO ₂ Concentration Reading (Volume %)				
				Zero Gas	Air	Sample Gas	Air	Zero Gas
1	11.0	121680- 129560	0	0	NR	0.059	0.030	0
			2	0	NR	0.049	0.032	0
			4	0	0.032	0.039	0.032	0
			5.5	0	0.032	0.034	0.032	0
2	9.0	106430- 111760	0	0	0.032	0.045	0.032	0
			2	0	NR	0.039	0.032	0
			5	0	NR	0.032	0.032	0
3	5.5	62960- 64650	0	0	0.032	0.039	0.032	0
			3	0	0.032	0.032	NR	0
			1.5	0	0.032	0.034	0.032	0
4	8.5	92360- 93060	0	0	0.032	0.062	0.032	0
			3	0	0.032	0.032	0.032	0
			4	0	0.032	0.034	0.032	0
			6	0	0.032	0.032	NR	0
			7.5	0	0.032	0.032	NR	0
			3	0	0.032	0.041	NR	0

Notes: The CO₂ injector was located at the center of the PDU, at 8.0 inches above the air distributor nozzles. The vertical distance between the injector and the gas sampling probe, when at the center of the PDU, was 16 inches.

NR: Not Recorded

TABLE 4.16

THE ROSIN-RAMMLER CONSTANTS FOR PARTICLE SIZE DISTRIBUTION
AT VARIOUS BED LOCATIONS, TEST 30-3

Run No.	Air Flow (SCFH)	CO ₂ Flow (SCFH)	Time (min.)	Sampler Location Above the Grid Plate								
				R.F. Corner @ 2"			Center @ 12"			Center @ 21"		
				n	drr (mic.)	dps (mic.)	n	drr (mic.)	dps (mic.)	n	drr (mic.)	dps (mic.)
1	126180-129560	11.0	0	2.36	1584	1014	2.15	1465	878	2.24	1053	928
			90	2.40	1521	985	2.14	1434	854	2.14	1457	870
2	106430-111760	9.0	0	2.27	1488	927	2.15	1436	859	2.14	1462	873
			45	2.40	1502	971	2.14	1444	860	2.14	1433	855
3	62960-64650	5.5	0	2.35	1460	931	2.13	1410	838	2.33	1409	894
			60	2.33	1485	941	2.41	1506	976	2.39	1504	971
4	92360-93060	8.5	0	2.35	1401	1011	2.06	1389	803	2.04	1583	805
			120	2.74	1731	1214	2.10	1384	815	2.10	1435	843

R.F. - Right Front corner of the PDU

n and drr are exponent and Rosin-Rammler mean diameter respectively.
dps is the surface mean diameter calculated from Rosin-Rammler equation.

TABLE 4.17

FLUID DYNAMICS III TEST MATRIX

Test Nos.	Test Duration (hrs)	Test Conditions*				Test Parameters	In-Bed Measurements
		Bed Weight (lb)	Air Flow (SCFM)	Solids Flow (lb/hr)	Bed Temp. (°F)		
40-1	6	1800	0-2000	600	Ambient	Pressure drop, bed height, fluid dynamic responses and minimum fluidization velocity	Solid Sampling
40-2	6	1800	0-2000	600	Ambient	Same as above, and gas interchange	Solid and Gas Sampling
40-3	6	1200	0-2000	300	Ambient	Pressure drop, bed height, fluid dynamic responses, and minimum fluidization velocity	Solid Sampling
40-4	6	2000	1700	0	Ambient	Same as above, and gas interchange	Solid and Gas Sampling

*These values are nominal settings. The actual values during the tests are somewhat different.

TABLE 4.18

TEST 40-1 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Average Solids Feed (lbs/hr)
			Temp. (°F)	Press. (atm)			
1	0	10,268	89	1.021	26.5	31.5	536
	15	10,266	89	1.022	26.5	30.0	
	30	10,274	89	1.022	26.5	30.0	
	45	10,378	89	1.021	26.5	29.8	
2	0	9,007	90	1.021	26.5	29.0	633
	15	9,079	91	1.021	26.0	28.5	
	30	8,951	91	1.021	26.0	28.5	
	45	8,938	91	1.021	26.5	29.4	
3	0	7,554	92	1.020	26.5	28.5	0
	15	7,385	92	1.020	26.0	28.5	
	30	7,383	92	1.020	26.0	28.5	
4	0	5,478	94	1.020	26.0	26.0	692
	15	5,473	94	1.020	26.2	26.7	
	30	5,477	95	1.020	26.2	26.0	
	45	5,597	95	1.020	25.8	26.5	
5	0	4,850	95	1.019	26.6	26.6	876
	15	4,851	95	1.019	26.8	26.5	
	45	4,855	94	1.021	26.8	26.5	

TABLE 4.18 (Cont'd)

TEST 40-1 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Average Solids Feed (lbs/hr)
			Temp. (°F)	Press. (atm)			
6	0	3,773	94	1.020	26.8	26.0	230
	15	3,951	95	1.020	26.9	25.8	
	30	3,954	95	1.020	26.4	25.5	
7	0	2,942	95	1.012	26.1	22.2	0
8	5	2,055	95	1.004	23.4	20.8	0
9	0	1,851	95	1.004	24.1	20.5	NC
	5	1,851	95	1.000	20.3	20.0	

The minimum fluidizing velocity was determined graphically to be 0.6 ft/sec.

NC - Not Collected

TABLE 4.19

TEST 40-2 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Average Solids Feed (lbs/hr)
			Temp. (°F)	Press. (atm)			
1	0	10,357	73	1.023	26.5	29.6	223.8
	15	10,364	73	1.023	26.5	29.8	
	30	10,297	73	1.024	26.5	30.0	
	45	10,298	74	1.024	26.5	29.6	
	60	10,361	74	1.023	26.5	29.8	
	75	10,421	73	1.023	26.5	30.0	
	90	10,481	75	1.021	26.5	28.5	
	105	10,479	76	1.021	26.5	28.5	
	120	10,342	73	1.023	26.5	28.5	
2	0	7,500	78	1.022	26.0	25.5	457
	15	7,581	79	1.017	26.0	26.5	
	30	7,842	78	1.019	25.8	26.0	
	45	7,587	78	1.020	26.5	28.5	
	60	7,487	77	1.020	26.6	28.0	
	75	7,584	80	1.022	26.5	27.5	
	90	7,498	80	1.022	26.5	27.3	
	105	7,399	79	1.022	28.0	29.4	
3	0	5,057	78	1.016	27.0	25.4	566
	15	5,052	79	1.016	27.0	25.8	
	30	5,052	80	1.016	27.0	25.8	
	45	5,056	81	1.014	27.0	26.0	
	60	5,054	80	1.016	27.0	25.0	

TABLE 4.19 (Cont'd)

TEST 40-2 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Average Solids Feed (lbs/hr)
			Temp. (°F)	Press. (atm)			
4	0	3,626	80	1.014	27.0	24.5	0
	5	3,626	80	1.013	27.0	25.0	
	10	3,626	80	1.014	27.0	25.0	
5	0	2,718	81	1.007	27.0	25.0	0
	5	2,717	81	1.007	27.0	26.0	

TABLE 4.20

TEST 40-2 IN-BED CO₂
CONCENTRATION MEASUREMENTS

Run No.	CO ₂ Flow Rate (SCFH)	Air Flow Rate (SCFH)	Sampling Location From The Center of the PDU (in.)	CO ₂ Concentration Reading Volume %				
				Zero Gas	Air	Sample Gas	Air	Zero Gas
1	9.0	128000-129700	0	0	0.040	0.099	0.040	0
			2	0	0.035	0.079	0.035	0
			4	0	0.035	0.081	0.035	0
			6	0	0.030	0.078	0.035	0
			10	0	0.035	0.064	0.035	0
			14	0	0.035	0.050	0.035	0
			20	0	0.035	0.036	0.035	0
2	7.0	92830-97050	0	0	0.036	0.096	0.036	0
			2	0	0.036	0.140	0.036	0
			4	0	0.036	0.082	0.036	0
			6	0	0.036	0.079	0.036	0
			9	0	0.036	0.070	0.036	0
			11	0	0.036	0.061	0.036	0
			13	0	0.036	0.056	0.036	0
			16	0	0.036	0.046	0.036	0
			18	0	0.036	0.038	0.036	0
3	4.5	62520-62580	0	0	0.036	0.098	0.036	0
			2	0	0.036	0.090	0.036	0
			4	0	0.038	0.090	NR	0
			7	0	0.038	0.080	0.038	0
			10	0	0.036	0.070	0.036	0
			13	0	0.036	0.061	0.036	0
			15	0	0.036	0.038	0.036	0
			18	0	0.036	0.036	0.036	0
4	3.0	44870	0	0	0.036	0.079	0.036	0

Notes: The CO₂ injector was located at the center of the PDU, at 8 inches above the air distributor nozzles. The vertical distance between the injector and gas sampling probe, when at the center of the PDU, was 16 inches.

NR: Not Recorded

TABLE 4.21

TEST 40-3 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Average Solids Feed (lbs/hr)
			Temp. (°F)	Press. (atm)			
1	0	10,504	78	0.995	17.6	15.4	363
	15	10,568	78	0.994	17.4	16.0	
	30	10,506	78	0.994	17.9	15.4	
	45	10,502	78	0.994	17.7	16.0	
	60	10,502	78	0.994	17.4	15.7	
	75	10,491	79	0.993	17.3	15.6	
2	0	9,150	80	0.993	17.6	15.5	350
	15	9,056	83	0.993	17.5	15.6	
	30	9,125	84	0.993	17.7	15.9	
3	0	8,435	87	0.993	17.9	15.6	305
	15	8,430	88	0.993	18.0	15.1	
	30	8,425	89	0.993	17.4	15.5	
4	0	7,439	90	0.995	18.8	15.6	136
	15	7,436	92	0.994	18.2	15.3	
	30	7,430	93	0.995	18.7	15.5	
5	0	6,504	95	0.992	18.4	15.4	312
	15	6,498	95	0.992	18.1	15.4	
	30	6,495	96	0.992	18.0	15.1	
	45	6,491	97	0.992	19.0	15.5	

TABLE 4.21 (Cont'd)

TEST 40-3 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)	Average Solids Feed (lbs/hr)
			Temp. (°F)	Press. (atm)			
6	0	5,391	99	0.990	18.9	15.4	320
	15	5,386	100	0.990	18.7	15.3	
7	0	4,755	100	0.988	19.0	15.0	310
	15	4,611	99	0.988	18.7	15.0	
8	0	3,407	100	0.983	18.6	14.7	270
	15	3,407	101	0.983	18.4	14.6	
9	0	2,417	104	0.974	16.7	14.3	0
	15	2,450	105	0.974	16.2	14.1	

Note: The minimum fluidizing velocity was determined graphically to be 0.9 ft/sec.

TABLE 4.22
TEST 40-4 DATA

Run No.	Time (min.)	Air Flow (lbs/hr)	Bed		Total Bed DP (in. H ₂ O)	Expanded Bed Height (in.)
			Temp. (°F)	Press. (atm)		
1	0	8552	73	1.036	32.3	34.3
	15	8615	73	1.036	32.3	34.0
	30	8544	73	1.036	32.4	33.6
	45	8542	73	1.036	31.9	33.2
	60	8535	73	1.036	31.8	33.8
2	0	8531	73	1.036	32.4	33.6
	15	8531	73	1.036	32.0	32.3
	30	8514	75	1.037	32.5	32.8
	45	8518	76	1.037	32.0	32.4
	60	8514	77	1.035	31.9	33.2
	75	8514	77	1.035	32.0	32.6
	90	8518	77	1.033	32.0	32.5
	105	8502	77	1.035	31.7	30.3
	120	8499	78	1.033	31.6	30.9
	135	8504	78	1.035	31.0	31.5
	150	8500	78	1.033	31.1	30.4

TABLE 4.23
TEST 40-4 IN-BED CO₂
CONCENTRATION MEASUREMENTS

CO ₂ Flow Rate (SCFH)	Air Flow Rate (SCFH)	Sampling Location From The Center of The PDU (in.)	CO ₂ Concentration Reading (Volume %)				
			Zero Gas	Air	Sample Gas	Air	Zero Gas
9.5	105900	0	0	0.036	0.075	0.036	0
		3	0	0.033	0.070	0.033	0
		6	0	0.033	0.064	0.033	0
		9	0	0.033	0.050	0.033	0
		12	0	0.033	0.046	0.033	0
		15	0	0.033	0.040	0.033	0
		18	0	0.033	0.033	0.033	0

Notes:

The CO₂ injector was located at the center of the PDU, at 8.0 inches above the air distributor nozzles. The vertical distance between the injector and the gas sampling probe, when at the center of the PDU, was 16 inches.

TABLE 4.24

THE ROSIN-RAMMLER CONSTANTS FOR PARTICLE
SIZE DISTRIBUTION FOR TEST 40-1

Run No.	Time (min.)	Air Flow (lbs/hr)	n	drd (mic.)	Surface Mean Diameter (mic.)
1	0	10,297	1.641	1283.68	615.80
	45		1.755	1309.01	665.40
2	0	8,994	1.738	1262.48	637.45
	45		1.665	1218.78	594.12
3	0	7,469	NC	NC	NC
	30		1.544	1097.90	505.86
4	0	5,506	NC	NC	NC
	15		1.541	1121.23	514.79
5	0	4,853	NC	NC	NC
	15		1.599	1079.63	513.02
6	0	3,893	1.515	1078.32	489.87
	30		1.513	1068.16	485.34

The particles were collected from the right front corner
of the PDU at 2 inches above the grid plate.

NC - Not Collected

TABLE 4.25

THE ROSIN-RAMMLER CONSTANTS FOR PARTICLE SIZE DISTRIBUTION AT
VARIOUS BED LOCATIONS, TEST 40-1

Run No.	Air Flow (lbs/hr)	Rosin- * Rammler Constants	Sampling Location from Center of PDU (in.) Along the Axis of the Lower Solids Sampling Probe								
			21	17	16	13	9	8	4	5	0
1	10,297	n drr (mic.) drs (mic.)	1.637 1371.22 653.96	1.722 1365.75 681.60		1.671 1347.68 655.39		1.693 1359.05 668.55	1.672 1308.01 637.35		1.665 1290.87 627.10
2	8,994	n drr (mic.) drs (mic.)	1.617 1399.40 659.18	1.612 1286.09 607.46		1.614 1206.59 572.98	1.604 1196.99 565.75			1.598 1244.89 584.80	1.663 1382.37 668.49
3	7,469	n drr (mic.) drs (mic.)	1.470 1344.56 581.74		1.498 1242.63 551.22	1.456 1196.77 519.82		1.502 1169.55 523.08	1.536 1283.55 580.60		1.535 1339.34 603.28
4	5,506	n drr (mic.) drs (mic.)	1.155 946.46 355.36			1.237 1025.63 398.03	1.252 1011.55 397.49		1.234 1137.91 431.74		1.262 1251.29 475.72
5	4,853	n drr (mic.) drs (mic.)	1.431 1060.49 460.54		1.602 1242.52 584.97	1.537 1124.76 514.99		1.470 1138.97 501.44		1.601 1303.22 611.33	1.571 1311.06 604.23
6	3,893	n drr (mic.) drs (mic.)	1.412 1017.92 439.63	1.557 1168.15 539.15		1.437 1021.28 447.80	1.485 1156.89 513.08			1.452 1151.31 500.98	1.486 1225.75 540.73

Samples were collected at approximately 12 inches above the grid plate..

*drs denotes surface mean diameter that is calculated from but is not a part of the Rosin-Rammler equation.

TABLE 4.26

THE ROSIN-RAMMLER CONSTANTS FOR PARTICLE
SIZE DISTRIBUTION FOR TEST 40-1

Run No.	Air Flow (lbs/hr)	Rosin- Rammler Constants	Sampler Location from Center of PDU (in.)*	
			17	13
1	10,297	n	1.593	1.535
		drr (mic.)	1287.49	1381.65
		drs (mic.)	601.47	621.00

*Samples were collected at approximately 21 inches above the grid plate and along the axis of the (upper) sampling probe.

TABLE 4.27

THE ROSIN-RAMMLER CONSTANTS FOR PARTICLE SIZE DISTRIBUTION
AT VARIOUS BED LOCATIONS, TEST 40-2

Run. No.	Air Flow (lbs/hr)	Time (min.)	R.F. Corner @ 2"			Center @ 12"			Center @ 21"		
			n	dr (mic.)	dr (mic.)	n	dr (mic.)	dr (mic.)	n	dr (mic.)	dr (mic.)
1	10,377	0	1.565	1042.41	488.06	1.402	1068.56	455.68	1.308	1168.13	463.61
		120	1.481	1086.10	483.98	1.412	1155.17	490.41			
2	7,560	0	1.457	1048.68	462.79	1.429	1179.45	504.68			
		105	1.470	1044.30	451.53	1.448	1148.01	498.44			
3	5,055	0	N.A.	N.A.	N.A.	N.A.	N.A.	N.A.			
		60	1.440	921.11	409.90	1.382	1116.35	467.61			

R.F. - Right front corner.

N.A - Samples not available.

TABLE 4.28

THE ROSIN-RAMMLER CONSTANTS FOR PARTICLE SIZE
DISTRIBUTION FOR TEST 40-3

Run No.	Air Flow (lbs/hr)	Time (min.)	n	drr (mic.)	drs (mic.)
1	10,484	0	1.358	996.23	418.49
2	9,082	0	1.323	946.59	393.11
3	8,402	0	1.272	929.64	376.22
4	8,074	0	1.287	958.03	388.74
5	6,469	0	1.277	941.51	381.21
6	5,360	0	1.273	967.42	388.66
7	4,655	0	1.269	902.31	366.75
8	3,379	0	1.330	969.68	402.61
9	2,406	0	1.506	982.99	449.02

Particles were collected from the right front corner
of the PDU at 2 inches above the grid plate.

TABLE 4.29

THE ROSIN-RAMMLER CONSTANTS FOR PARTICLE SIZE DISTRIBUTION
AT VARIOUS BED LOCATIONS, TEST 40-3

Run No.	Air Flow (lbs/hr)	Rosin-Rammler Constants	Sampling Location from Center of PDU (in.) Along the Axis of the Lower Solids Sampling Probe						
			21	13	11	9	7	3	0
1	10,484	n drr (mic.) drs (mic.)	1.562 1152.16 533.72	1.423 1268.06 535.80		1.383 1165.73 485.67		1.364 1174.24 482.93	1.350 1144.78 468.35
2	9,082	n drr (mic.) drs (mic.)							1.270 1154.91 453.68
3	8,402	n drr (mic.) drs (mic.)	1.254 1123.40 437.02				1.166 1162.27 418.75	1.222 1061.07 405.14	
4	6,469	n drr (mic.) drs (mic.)	1.317 1193.45 474.72						
5	5,360	n drr (mic.) drs (mic.)	1.535 1148.41 524.31						
6	4,655	n drr (mic.) drs (mic.)	1.300 847.37 355.40						
7	3,379	n drr (mic.) drs (mic.)	1.215 730.52 303.02						

Samples were collected approximately 12 inches above grid plate.

TABLE 4.30

THE ROSIN-RAMMLER CONSTANTS FOR PARTICLE SIZE DISTRIBUTION
AT VARIOUS BED LOCATIONS, TEST 40-4

Run No.	Air Flow (lbs/hr)	Rosin- Rammeler Constants	Sampling Location from Center of PDU (in.) Along the Axis of the Lower Solids Sampling Probe					
			21	17	13	9	3	0
1	8557	n	1.246	1.320		1.306	1.289	1.283
		drr (mic.)	1117.47	1019.98		1018.03	1011.87	1007.74
		drs (mic.)	428.79	417.40		413.14	406.83	403.91
		n	1.375	1.305	1.288	1.349	1.352	1.323
		drr (mic.)	1195.55	1074.45	1061.76	1108.67	1101.95	1109.03
		drs (mic.)	493.66	431.31	422.80	455.50	454.07	448.02

Samples were collected approximately 12 inches above the grid plate.

5.0 MODEL VALIDATION AND ANALYSIS

5.1 Initiation of Model Validation

The familiarization process with the MIT Model consisted of two steps. First, a meeting was held with the Model developers to become acquainted with the Model usage. Later, trial validation runs were made using test data.

A sample output for the Model is shown in Figure 5.1, where the linkage after the "CALL FREEBD" statement was terminated. The run was aborted before the main program reached the subroutine which retains all the predictions, including those which were already printed. Hence, predictions such as the minimum fluidizing velocity, slumped bed height, etc. could not be obtained. The abnormal output was brought to the attention of the MIT staff and subsequently resolved. As explained by the MIT staff the termination of the linkage during each computer run occurred because the subroutine "FREEBD" had not been completed and debugged.

5.2 VALIDATION RESULTS

Spot checking of the MIT Model was performed during the course of this effort with the objective of covering the widest possible range of operating conditions. Experimental data from the Alexandria PDU were obtained and used for validating the Model. The results are reported in this section.

The values of input parameters listed in the tables are in American Engineering units. However, the Model only accepts SI units. Unit conversion must therefore be made before validation runs can be performed. Since SI units may not be convenient for most users of the Model, the Model should be upgraded to accept American Engineering units as well as SI units.

Among the listed input variables, the surface mean particle diameter requires some clarification. The surface mean particle diameter changes with position in the bed, as shown in Figures 5.2 through 5.4. Hence, values for the bed particle diameter in the Model validation should be determined from numerical integration of the profiles. The integration should take into account not only the surface mean value at each position, but also the spread or distribution of sizes at that position. In addition, the profiles should include vertical and radial distributions. The integration process is therefore tedious and requires substantially more data points than could be provided. To expedite the results, the arithmetic mean of all the data for in-bed size distribution profiles was used. The use of the arithmetic mean should not effect the validation results significantly.

FIGURE 5.1: MIT Model Sample Outputs

```

tpsa: run main
run main
      CALL INPUT
      OUTER LOOP
      CALL PRECALL
      MIDDLE LOOP
      CALL FLUIDM
bedh= 0.8886E+00    ht= 0.2966E+00
      CALL MASSEX
      INNER LOOP
      CALL COMB
ICCC=  0
ICC=   1
ICC=   2
ICC=   3
      sumpbc= 0.9999E+00
      fines by attrition= 0.1222E-02
wcoal= 0.1782E+02    wcoal1= 0.0000E+00
      CALL DESUIF
      CALL MATBAI
      CALL CO
edh= 0.889    ht= 0.297
ycot= 0.7197    yccif= 1.3086    ycodef= 1.2686    ycoef= 1.2968
      CALL FREEED
Error: linkage error by >udd>FBC>library>executable>freebd|5375 (line 593)
referencing merf|merf
Segment not found.
r 11:37 11.698 203 level 2,19 $1.69

pi
pi

```

The solids mean size varied both radially and axially. Figure 5.2 shows the size profiles for Test 40-1, determined at 12 inches above the air distributor nozzles. At 2.5 ft/sec fluidizing velocity, this profile indicates that coarser particles were accumulated at the walls and center of the PDU. At a higher fluidizing velocity, 3.0 ft/sec, the amount of coarser particles at the center of the bed was reduced compared with other regions in the bed. At a still higher fluidizing velocity, 3.4 ft/s, the peak at the center of the bed vanished, leaving the coarser particles to collect near the bed walls. These findings are reinforced by data obtained, at the same distance above the air distributor nozzles, from Test 40-3 (shown in Figure 5.3 and the upper curve in Figure 5.4).

The lower curve in Figure 5.4 shows the particle size profile at 21 inches above the air distributor nozzles. This profile does not have a center peak as was the case when data was collected at 12 inches above the air distributor. Two possible causes for this occurrence may be: (1) the presence of the in-bed tubes and (2) the larger distance above the air jets emerging from the distributor nozzles. Exact determination of the cause is not possible because of a limited scope of activity and insufficient data.

For low fluidizing velocities, the center peaks in the profiles obtained from 12" above the distributor were probably caused by the absence of air nozzles in the 2 inch area between the two air distributor sections. The distributor was built in two sections for easier installation in the PDU.

Thus, it appears that particle size distribution in the PDU generally varies with location and is a function of fluidizing velocity. Non-uniform size is probably influenced by the non-uniform fluidizing velocity and air distributor design.

During some tests, the fluidizing velocity did not exceed 4 ft/sec. In these tests, solids were circulated in a closed cycle. Due to repeated handling, particles attrited to smaller sizes rather rapidly. Increasing the air velocity above 4 ft/sec would have caused unacceptable bed material losses.

A summary of input and validation results for Test Series 20 through 40 are presented in Tables 5.1 and 5.2. Further discussion of specific test results is in sub-sections 5.2.1 through 5.2.4.

As shown in tables 5.1 and 5.2, data for solids feed rate, fluidizing velocity, and expanded bed height are listed as "variable" for most tests. This is to indicate that there is more than one run and the validation results will be presented as graphs in figures which follow. Experimental data for Bed Weight in table 5.1 have been selected as the representatives of test conditions. This value may be equal to the arithmetic mean of the beginning and ending of test values for some tests, and may be closer to the ending value than the beginning value for some other tests. Bed Temperature data have been taken as the arithmetic mean of temperatures measured for individual

runs within the corresponding tests. Particle Density and Diameter data are selected in the same manner as the Bed Weight data.

In Table 5.2, experimental data for Slumped Bed Height, like the Bed Weight data, have been selected as representative values from the beginning and ending of test results. The Minimum Fluidization Velocity has been determined from plots of Bed ΔP versus fluidizing velocity. The Expanded Bed Height has been determined from the pressure profile as discussed in Section A-4.3 of the Appendix A.

It is noted that the bed weight data have been calculated from the measured values of slumped bed height and solids bulk density (not listed). But representative values for bed weight and slumped bed height have been selected independently.

The bulk density was measured by taking solids samples from the bed. A number of samples were measured for each test run and an arithmetic mean value was determined from these measurements for use in calculating the bed weight for that run. Since there are many runs in each test, a second level of reduction was necessary to arrive at single value for each test as described earlier. The particle density was measured for selected samples only. The variations of measurements for bulk and particle densities were $\pm 30\%$ about the mean values. Because of the two levels of reduction, the final value of the bulk density is rather close to the particle density for some tests (e.g., Serves 40). This will indicate some unreasonably small voidages for these tests.

Figure 5.2 - Test 40-1 Bed Solids Surface Mean Diameter Profiles

Figure 5.3 - Test 40-3 Bed Solids Surface Mean Diameter Profiles

Figure 5.4 - Test 40-4 Bed Solids Surface Mean Diameter Profiles

TABLE 5.1
INPUT VALUES FOR THE MODEL VALIDATION

Test No.	Tube* Pitch (in.)	Bed Weight (lbs)	Bed Temp. (°F)	Particle		Solids Feed (lbs/hr)	Fluidizing Velocity (ft/s)
				Density (g/cc)	Diameter (microns)		
20-5	40 1/2	1123	124	2.09	1110	0	var.
20-6	40 1/2	1224	364	2.09	1110	0	var.
20-7	40 1/2	1496	148	2.09	1110	0	var.
20-8	40 1/2	1880	478	2.09	1110	0	var.
20-9	40 1/2	1886	105	2.09	570	0	2.4
20-10	40 1/2	2093	103	2.09	740	0	3.3
30-1	40 1/2	1413	99	1.41	869	var.	2.5
30-2	40 1/2	1335	99	1.41	906	var.	3.4
30-3	40 1/2	2315	89	1.44	879	var.	var.
40-1	37/8	1915	93	1.51	547	var.	var.
40-2	37/8	1943	78	1.50	465	var.	var.
40-3	37/8	1312	91	1.71	412	300	var.
40-4	37/8	2062	75	1.75	432	0	2.7

* Bed width is 40½ inches, hence inputting 40½ inches for tube pitch indicates that no tubes were present in the bed.

Note : Bed weights presented are not always equal to bed weights at the end of each test. Bed weights were based on review of bed data recorded at regular intervals. Most representative data were subsequently selected.

TABLE 5.2

VALIDATION RESULTS

Test No.	Slumped Bed Height (in.)		Min. Fluidization Velocity (ft/s)		Expanded Bed Height (in.)	
	Meas.	Pred.	Meas.	Pred.	Meas.	Pred.
20-5	15.7	15.98	1.2	1.71	var.	var.
20-6	17.3	17.43	1.1	1.61	var.	var.
20-7	23.7	21.32	1.0	1.71	var.	var.
20-8	22.8	21.09	1.2	1.57	var.	var.
20-9	23.8	26.60	-	0.58	25.00	43.94
20-10	23.0	29.81	-	0.95	24.70	49.17
30-1	20.1	20.14	-	1.22	20.20	25.96
30-2	19.8	19.02	-	1.30	18.70	28.43
30-3	33.3	32.70	0.8	1.24	var.	var.
40-1	20.8	27.28	0.6	0.57	var.	var.
40-2	20.5	27.68	-	0.42	var.	var.
40-3	14.0	18.69	0.9	0.33	var.	var.
40-4	25.0	29.35	-	0.37	32.59	58.37

5.2.1 Slumped Bed Height

Theoretically, the slumped bed height can be calculated if bed weight, void fraction, and particle density are known. Among these variables, bed weight is the most significant variable because it indicates the quantity of bed present. Hence, it is logical to plot the slumped bed height against bed weight for comparison. This plot for both measured and predicted values is shown in Figure 5.5. The predicted curve in Figure 5.5 is a straight line for all particle sizes and densities, suggesting that the slumped bed height as computed by the Model depends on the bed weight alone. Particle size and density are inputs to the Model, but are apparently not used to calculate this parameter.

It is obvious that the Model will require improvements in computing the slumped bed height. The bed height is most definitely a function of physical properties of the solids in the bed.

5.2.2 Minimum Fluidization Velocity

In general, the minimum fluidization velocity is a function of particle physical characteristics and, to a lesser extent, gas properties. The predominant variable defining the particle characteristics is the surface mean particle size and particle density. Hence, it is reasonable to plot the minimum fluidization velocity versus particle size with particle density as a parameter. Air properties change very little for the temperature ranges covered by these experiments.

This plot is shown in Figure 5.6 and only includes four data points because these points have a reasonable particle size spread. The predicted curve is almost a straight line and is higher than experimental values for particles larger than 550 microns. Below 550 microns the predictions are lower than experimental values, however.

A comparison of the results with published predictions for minimum fluidization velocity indicates that a straight line relationship between the minimum fluidization velocity and particle size, as predicted by the Model here, is highly unusual. Moreover, the Model results closely resemble results predicted by Kunii and Levenspiel's Equation (1) which is derived from Ergun's Equation (2). This correlation is for systems with uniform particle size and does not accurately represent most fluidized bed combustion systems, including the Alexandria PDU.

5.2.3 Expanded Bed Height

The expanded bed height is strongly dependent on the fluidizing velocity. Plots of expanded bed height versus fluidizing velocity for Test Series 20 are in Figures 5.7 through 5.11. It should be noted in these figures that the expanded bed heights are presented for increasing and decreasing the fluidizing air velocity. The fact that solids were flowing into the PDU during Test Series 30 did not exhibit any change from the trends detected in the plots for Test Series 20. Therefore, plots for Test Series 30 are not included. Since Test Series 40 was performed in the presence of the in-bed bundle, some effects on the fluid dynamics of the bed are expected. Plots for Test Series 40 are shown in Figure 5.12 through 5.14.

In examining the curves in Figures 5.7 through 5.14, one can clearly see large discrepancies between predicted and experimental data. It is also evident that the predicted expanded bed height is more sensitive to the fluidizing velocity than experimental values. The presence of the in-bed bundle apparently produces more expansion in both the predicted and the experimental curves as shown in Figure 5.12. The reason for the Model prediction being much higher than the actual data could be partially caused by the fact that the Model does not accurately account for variations in bundle geometries.

5.2.4 Radial Diffusion Coefficient

Three dimensional concentration profiles of carbondioxide are determined experimentally. A typical three-dimensional profile of CO₂ concentration in the bed is shown in Figure 5.15.

Carbon dioxide was injected into the center of the bed in the direction of air flow to determine its concentration profile. Carbon dioxide concentrations were measured at different locations in the bed, from the center of the PDU to the wall. The concentration of the tracer gas (CO₂) was 100% at the injection point and decreased with increasing distance in the upward and radial directions.

The experimental radial diffusion coefficients can be determined from the measured concentration profile of the tracer gas (CO₂).

Because of shortage of time and the fact that the radial diffusion coefficient is an intermediate model parameter and not an end result, the derivation was not completed. However, some additional information is presented in Appendix A.

5.2.5 Conclusions of Air Distributor Leak Tests

From the results of the air distributor leak tests performed, the following conclusions are drawn:

- o There were four leak areas in the air distributor where air flow was concentrated to the extent that uneven distribution of bubbles in a shallow bed (2" - 8" deep) was visually apparent.
- o The four leak areas are calculated to be equivalent to a total of 0.9 to 1.1 in.² or 3.6 to 4.2% of total nozzle flow area.
- o Testing before and after repairing the leak areas shows no detectable differences in expanded bed height.
- o Minimum fluidizing velocity determined from tests after the repair appears to be higher than that determined before the repair. However, this might be explained by the differences in the size distribution of particles.

Additional information on these tests is presented in the Appendix C.

FIGURE 5.5 - Predicted/Experimental Data for Slumped Bed Height vs. Bed Weight

FIGURE 5.6 - Experimental/Predicted Values for Minimum Fluidization Velocity vs. Particle Surface Mean Diameter

FIGURE 5.7 - Test 20-5 Experimental/Predicted Expanded Bed Height
(Experimental U_{mf} has been determined separately)

FIGURE 5.8 - Test 20-6 Experimental/Predicted Expanded Bed Height

FIGURE 5.9 - Test 20-7 Experimental/Predicted Expanded Bed Height (Experimental U_{mf} has been determined separately)

FIGURE 5.10 - Test 20-8 Experimental/Predicted Expanded Bed Height

FIGURE 5.11 - Tests 20-9 and 20-10 Experimental/Predicted Expanded Bed Height for Different Particle Sizes and Bed Weights

FIGURE 5.12 = Test 40-1 Experimental/Predicted Expanded Bed Height

FIGURE 5.13 · Test 40-2 Experimental/Predicted Expanded Bed Height

FIGURE 5.14 - Test 40-3 Experimental/Predicted Expanded Bed Height

FIGURE 5.15 - Test 20-9 Three-Dimensional Plot of Tracer Gas (CO₂) Concentration Profile

5.3 MIT MODEL VALIDATION CONCLUSIONS

MIT Model validation primarily consists of comparing cold mode test results with corresponding Model predictions. In almost every case, the comparison revealed discrepancies between the Model's fluid dynamic predictions and corresponding experimental data. There may be a number of possible causes for these discrepancies, including:

- incorrect use of the Model to generate predictions
- incorrect comparison of Model predictions with experimental results
- inaccurate Model predictions
- physical incompatibility between the Alexandria PDU and the AFBC data upon which the MIT Model was based

It is highly unlikely that the Model was used incorrectly, or that results were incorrectly compared. During the Model validation process, there were frequent contacts with the MIT staff to ensure correct Model useage. In addition, a review of the results indicate that proper comparisons were made.

A few apparent factors which may have contributed to the discrepancies between the predicted and experimental results stem partially from the fact that the Model, in its present form, does not account for variations in the in-bed bundle geometries. The bundle configuration in the Model is an output parameter and cannot be independently varied to represent the exact arrangement of the bundle. Moreover, the Model appears to treat the solids in the bed as uniform particles, where in fact there is a large spread of particle sizes in the bed at all times.

It is difficult to completely separate the issues of inaccurate Model predictions from possible physical incompatibilities. The MIT fluid dynamic sub-model appears to have been developed from theoretical considerations coupled with mathematical manipulation of fluidization correlations obtained from certain fluidized bed units. If there are significant physical differences between the Alexandria PDU and these other beds, then Model predictions would differ from Alexandria experimental data. For example, if the Alexandria PDU had significantly more or less channeling than the fluidized beds on which the Model correlations were based, then expanded bed height predictions from the Model would differ considerably from measurements at Alexandria. However, it is possible to present considerable evidence supporting the conclusion that the Alexandria PDU is highly representative of most AFB combustors. Therefore, it is quite plausible to discount physical incompatibility as the primary cause of differences between predicted and experimental values.

In summary, it appears that Model predictions do not accurately characterize the fluid dynamics and associated parameters of a fluidized bed combustor at this scale. The investigation of the specific causes of these inaccuracies requires considerable effort and is beyond the scope of current effort.

6.0 FLUID DYNAMIC RESPONSE

This effort will increase our understanding of AFBC control systems by accurately characterizing fluid dynamic response of the bed. The basic approach involves collecting transient data in response to step changes in test variables. For example, the rate of change in expanded bed height, pressure drop, and gas dispersion coefficient would be recorded in response to step changes in air flow rates and solids feed rate.

The specific tasks required to implement this portion of the test plant included:

- rehabilitation of the existing computer-based data acquisition system hardware
- rewiring of the 63 channel cable run from the PDU to the Data Acquisition System
- development of new software for conversion of the recorded raw data into more easily understood engineering units, plus the addition of some simple computational capabilities

The procedure for data collection and reduction consists of:

- collect data on a sampled basis for all test runs
- record all channels as digital data on magnetic tape; record six-channels on an analog chart recorder
- reduce recorded data by converting digital information to engineering units, with the capability for simple computational processing

Current plans call for the collection and reduction of data during all test runs. However, detailed analysis and evaluation are beyond the scope of the current effort. Rather, a substantial database will be accumulated for subsequent evaluation.

Although the original test plan involved data collection during the cold mode tests, no reliable data was actually collected. There were a number of data acquisition system hardware failures. Repairs were difficult and lengthy due to the age of the equipment. By the time the system was fully operational, cold mode testing had been completed. However, now that the system is fully operational, transient data will be collected during the remainder of the test program.

A block diagram of the Data Acquisition System is shown in Figure 6.1. A more detailed description of the Data Acquisition System can be found in Support Task 7.6.

Figure 6.1 Data Acquisition System Block Diagram

7.0 SUPPORT TASKS

7.1 DESIGN, FABRICATION AND INSTALLATION OF IN-BED PROBES

The in-bed sampling probes were primarily designed and fabricated for use during cold mode testing. Figures 7.1 through 7.3 show the detailed probe design drawings as well as their orientation in the PDU. If the economizer/superheater tube bundle is not installed, then four sample probes can be used. Otherwise, only two probes can be installed at two locations in the front of the PDU.

In the design of the probes, certain factors had to be considered to ensure proper operation. First of all, gaseous reactions taking place within the fluidized bed must not continue within the sample probe. It is impossible to obtain representative PDU gas concentrations unless these reactions are quenched when the gas sample enters the probe. Therefore, the probe design includes a double water jacket to rapidly cool the gas sample and quench the reaction. A second design consideration involved the need for gas concentration profiles, as well as single point measurements, within the bed. To obtain these profiles, the probes were designed to move in and out of the bed as required. A third design consideration involved the need to avoid condensation within the gas samples. To avoid this problem, the gas sampling probe is inclined downward at a 5° angle in the bed to allow condensation to drain out of the probe. Conversely, the solids probe is inclined upwards at a 10° angle to catch the solids as they fall due to gravity.

The gas probe was later modified to include a 5µm-ceramic filter attached to the end of the probe, as shown in Figure 7.4. The gas sampling system now includes a 5µm-ceramic filter, followed by a porous metal-media filter placed before the gas analysis system entrance. The solids probe is purged with nitrogen. This stops particle reactions in the probe and enables the collection of representative solid samples.

7.2 REDESIGN OF GAS SAMPLING SYSTEM

Schematic drawings for the gas sample conditioning system, gas sample analyzers and header system are shown in Figures 7.5 and 7.6. Symbols used in these Figures are explained in Table 7.1.

Several problems required a redesign of the gas sampling system. These problems included:

- Condensation in the sampling lines
- Uncontrolled pressure and flow through the analyzers
- Insufficient number of gas sampling locations
- High leakage rate

To prevent condensation, all sample lines were heat traced. Uncontrolled pressure and flow through the analyzers is undesirable because most of the analyzers are sensitive to both pressure and flow variations. The current system maintains constant pressure and flow through the analyzers.

Originally, only one gas sample was taken at the top of the PDU. The current system can accept samples from four different locations: at the top of the PDU, two locations from the bed, and a spare location. The original filter to the analyzer was designed to operate under pressure while the system is under vacuum. A filter suitable for vacuum service was ordered and installed in the system. Finally, the original system was not leak tight. The current system meets the desirable minimum leakage requirements. Figure 7.7 shows details of gas sampling system intakes.

7.3 GENERATION OF SPENT BED MATERIAL

In the Fluid Dynamics I batch test, spent bed material, generated from burning Sewickley coal with Greer limestone, was used. After these tests, the remaining spent bed was insufficient for subsequent tests. The PDU was operated continuously to generate 4500 pounds of spent bed material for Fluid Dynamics II and III tests. Redstone coal was burned with Greer limestone to generate the sulfated stone. Because of the new coal and new operating conditions, the size distribution and characteristics of this spent bed material were determined. The results were used in the data analysis for the subsequent tests.

The PDU was again operated continuously to generate 4500 pounds of spent bed material for use during various tests. Redstone coal was burned with Mosheim (Shenandoah Valley, Winchester, VA) limestone. During this run, bed samples were collected every two hours for determining spent bed material characteristics. This sampling ensured consistent spent bed material physical properties. The average particle and bulk densities of this spent bed material were found to be relatively constant, and were used in the data analysis for the remaining test.

7.4 REMOVAL, MODIFICATION & INSTALLATION OF THE IN-BED TUBE BUNDLE

The original in-bed boiling tube bundle was removed prior to Test Series 20 (Fluid Dynamics I). Both Test Series 20 and Series 30 required the absence of in-bed tubes. After examining the original tube bundle, it became apparent that a new bundle would be required.

A new tube bundle was fabricated. The design of the new bundle was essentially similar to the original bundle. The only significant modification was the addition of several thermal wells for thermocouple temperature measurements during upcoming heat transfer tests. The new tube bundle was installed before the starting Test Series 40 (Fluid Dynamics III), which required the presence of in-bed tubes.

For details of the modification see Figure 7.8

7.5 INSTALLATION OF DATA ACQUISITION SYSTEM

63 transducers can be wired to the computerized Data Acquisition System (DAS). However, new transducer leads were required between the PDU and the DAS. The available instruments at the facility were used to expedite this task. Table B.1 in Appendix B, lists the instruments which were initially connected.

PDU data is collected on a magnetic tape at the site and then transferred to an off-site computer for processing. An analog chart recorder is used for simultaneously recording any six output channels of the DAS. Operability of the system was partially checked during Tests 20-9 and 10 (Fluid Dynamics I).

7.6 CALIBRATION OF THE AIR LINES

In any test, the air lines must be calibrated to ensure accurate results. The three air streams that normally provide total combustion air to the PDU are:

- (1) Main Air: Supplied by a forced draft (FD) fan (40 hp motor, 3000 cfm capacity) through a 12" duct.
- (2) Coal and Limestone Injection Air: Supplied by a Schwitzer blower, capable of 250 cfm at 2 psi head, through a 2" schedule 40 pipe.
- (3) Fly Ash Reinjection Air: Supplied by the same blower as for coal and limestone through a 1½" schedule 40 pipe. (Not in use)

The air flow rates for these streams are measured with orifice plates. The flow rates are controlled by valves installed upstream of the orifice plates for streams 2 and 3, and by a downstream damper for Stream 1.

Orifice plates for main air, coal and limestone conveying air, and fly ash reinjection air were calibrated by determining air mass flow rate by a static pitot tube and comparing it with mass flow rate calculated by orifice readings.

The flow of compressible fluids through orifices can be expressed by the following equation:

$$W = 359.47 Y \cdot C \cdot d^2 (\Delta P \rho)^{\frac{1}{2}} \quad (7.1)$$

Where:

W = mass flow rate through the orifice (lb/hr)

d = orifice diameter (inches)

ΔP = differential pressure across the orifice (inches of water)

ρ = density of fluid at upstream of orifice (lb/ft³)

Y = expansion factor

C = flow coefficient

Values for C and Y are available in the literature. The values of C and Y for main air, coal and limestone injection air, and fly ash reinjection air are constant for the normal operating range of the PDU. Values for C, Y and d are as follows:

<u>Stream</u>	<u>C</u>	<u>Y</u>	<u>d, inches</u>
Main air	0.71	0.98	6.915
Coal/Limestone Transportation	0.80	0.99	1.438
Fly Ash Rein- jections Air	0.767	0.99	1.156

Substituting these values of C, Y and d into Equation (7.1) for stream 1, 2, and 3, the following equations are obtained:

$$W = 11960 (\Delta P_p)^{\frac{1}{2}} \text{ for main air} \quad (7.2)$$

$$W = 588.72 (\Delta P_p)^{\frac{1}{2}} \text{ for coal + limestone injection air} \quad (7.3)$$

$$W = 364.77 (\Delta P_p)^{\frac{1}{2}} \text{ for fly ash reinjection air} \quad (7.4)$$

Pressure difference of a static-pitot tube is related to the air velocity at the point of measurement as follows:

$$V_m = C \left(2gR \left(\frac{\rho_w}{\rho_a} - 1 \right) \right)^{\frac{1}{2}} \quad (7.5)$$

Where:

C = pitot tube coefficient

g = gravitational acceleration = 32.2 ft/s²

R = pressure difference in feet of water

ρ_w = density of water used in manometer

ρ_a = air density

Since the flow of air is well within the turbulent region, the one-seventh power law for velocity distribution applies. The mean velocity V is determined by:

$$V = \frac{49}{60} V_m \quad (7.6)$$

The mass flow rate of air is therefore calculated by:

$$W = \rho_a V \frac{\pi D^2}{4} \quad (7.7)$$

Where D is the pipe inside diameter, substituting Equations (7.6) and (7.7) into Equation (7.5) and applying the appropriate conversion factors yield:

$$W = 37.15 D^2 [R' (\rho_w - \rho_a) \rho_a]^{\frac{1}{2}} \quad (7.8)$$

Where:

W = air mass flow in lb/hr

D = pipe diameter in inches

R' = pressure difference in inches of water

ρ = density in lbm/ft³

The results are graphically presented in Figures 7.9 through 7.11. Orifice plates installed in the air lines have been in service for many years. Wearing out of their square edges are therefore expected, and, as a consequence, mass rate calculated from orifice readings would be more erroneous than that calculated from pitot tube readings. For this reason, wherever variance occurs, mass rate calculated from orifice readings will be corrected, in reducing test data, to match the corresponding pitot tube mass rate.

Figure 7.1 Gas Sampling Probe

REVISIONS		ITEM	REQD	PART NO.	DESCRIPTION
DESCRIPTION	DATE				
PRINTING RE DONE	12-16-80				
					SCALE 1" = 1"
					DATE 8-5-80
					DRAWN J.B.
					CHECKED
					APPROVED
					ALEXANDRIA, FBC-PDU
					HEAT EXCHANGER
					TEES DETAIL
					DRAWING NO. C-57
					SHEET NO. 2 of 4

Figure 7.1 (Cont'd)

REVISIONS		ITEM	REQD	PART NO.	DESCRIPTION
DESCRIPTION	DATE				
PRINTING REDONE	12-19-80				
SCALE: 1"=1'		ALEXANDRIA, F.B.C.-PDU <u>GAS SAMPLING</u> <u>PROBE PORT</u>			
DATE: 7-30-80					
DRAWN: JLB					
CHECKED:					
APPROVED: 		DRAWING NO. C-58			
		SHEET 3 of 4			

NOTE: SEE DWG. #D-25
FOR PARTS LIST

PLUG FOR
UNUSED PORTS

Figure 7.1 (Cont'd)

15	2	FCS(ONE END)	1/4" FEMALE SWEAT FITTING (FLEXONICS)
14	120"	0025X6000 BZ-NB	1/4"-361/BRONZE HOSE (FLEXONICS)
13		DETERMINE IN FIELD	FLEXIBLE H ₂ O OUTLET HOSE
12	2	DETERMINE IN FIELD	3/16" THK .834 O.D. X .250 I.D. SS BUSHING
11	90"	1/4".035-316 S	STAINLESS STEEL PROCESS TUBE (TIME)
10	84"	5/8".049-316 S	STAINLESS STEEL JACKETING TUBE (TIME)
9	72"	1".083-316 S	STAINLESS STEEL JACKETING TUBE (TIME)
8	8	SS 1614-1	BACK FERRULE (SWAGELOK)
7	8	SS 1613-1	FRONT FERRULE (SWAGELOK)
6	2	SS 1610-P	PORT CONNECTOR PLUG (DITTO)
5	2	SS 1610-1-16 BT	SAMPLE PROBE PORT CONNECTOR (DITTO)
4	2	SS 400-R-10-BT	1/4" BORED THRU REDUCER (DITTO)
3	2	SS 1010-3	5/8" UNION TEE (DITTO)
2	2	SS 1010-R-16-B	5/8" BORED THRU REDUCER (DITTO)
1	2	SS 1610-3T	1" UNION TEE (DITTO)
ITEM	REQ'D	PART NO.	DISCRIPTION

GAS SAMPLING PROBE PARTS LIST

ALEXANDRIA FBC-PDU

DRAWN _____
APPROVED _____

DRAWING NO D-2
DATE 12-6-80

Figure 7.1 (Cont'd)

Figure 7.2 Solids Sampling Probe

ITEM	QTY.	PART NO	DESCRIPTION	
REVISIONS DESCRIPTION Date			DATE 8-19-80	
			DRAWN DAT	ALEXANDRIA FBC-PDU
			CHECKED	SOLID SAMPLE PROBE ASSEMBLY SECTION 2
			APPROVED BM	DRAWING NO C-64 SCALE: 1"=1"
			SHEET NO	2 of 6

Figure 7.2 (Cont'd)

ITEM		QTY.	PART NO	DESCRIPTION	
REVISIONS	DESCRIPTION	DATE	DATE 8-25-80		
			DRAWN
			CHECKED		
			APPROVED
			ALEXANDRIA FBC-PDU SOLID SAMPLE PROBE ASSEMBLY SECTION 4 DRAWING NO C-66 SCALE: 1"=1' SHEET NO 4 of 6		

Figure 7.2 (Cont'd)

ITEM	QTY.	PART NO	DESCRIPTION	
REVISIONS DESCRIPTION Date	<p>PROB</p> <p>13/80</p>		DATE	8-26-80
			DRAWN	JAT
			CHECKED	
			APPROVED	BM
			ALEXANDRIA FBC-PDU SOLID SAMPLE PROBE ASSEMBLY DETAIL	
		DRAWING NO	C-67	SCALE: $\frac{5}{16} = 1"$
		SHEET NO	5	of 6

Figure 7.2 (Cont'd)

1	2	SS-1610-1-16 BT	SWAGELOCK - MALE CONNECTOR
2	6'-0"	1" - .083 - 316S	S.T. & P. - STAINLESS STEEL TUBE
3	7'-0"	3/4" - .065 - 316S	" - STAINLESS STEEL TUBE
4	2	SS-601-PC	SWAGELOCK - PORT CONNECTOR
5	2	SS-1610-6-12	" - 3/4" BT REDUCING UNION
6	6'-0"	DETERMINE IN FIELD	WIRE
7	4	" "	EYEHOOKS
8	2	SS-1210-R-16	SWAGELOCK - REDUCER
9	2	SS-1610-7-16	" - FEMALE CONNECTOR
10	5'-0"	1" - 316S - SCH'D 40	S.T. & P. - STAINLESS STEEL PIPE
10a	2	" "	FORM IN LAB - 6" OAL, 45° BEND, MALE D/E
10b	1	" "	" " - DETERMINE LENGTH IN FIELD
11	3	1", B-36-DT	JAMESBURY - 1" BALL VALVE, ALL 316 SS
12	3	SS-16-CN	CAJON - CLOSE NIPPLE
13	2	1" - 316SS - UNION	S.T. & P. - STAINLESS STEEL UNION
13a	1	" "	" - FEMALE END OF ONE UNION
14	2	SS-16 DKM4-F4	WHITEY - FORGED BODY SHUT-OFF VALVE
15	1	304 L-HDF4-1000	" - SAMPLING CYLINDER
16	1	SS-400-2-4	SWAGELOCK - MALE ELBOW
17	2	SS-1610-P	" - PLUG, SEE DWG. NO D-24
ITEM	QTY.	PART NO	DESCRIPTION
REVISIONS DESCRIPTION DATE			DATE 8-28-80
			DRAWN DAT
			CHECKED
			APPROVED BAC
			ALEXANDRIA FBC - PDU SOLID SAMPLE PROBE PARTS LIST DRAWING NO D 35 SCALE: = SHEET NO 6 of 6

Figure 7.2 (Cont'd)

FIGURE 7.3: Location and Orientation of the Gas and Solid Sampling Probes in the PDU.

ITEM	REQD	PART NO.	DESCRIPTION
REVISIONS DESCRIPTION DATE PRINTING REDONE RO			SCALE: 1/8" = 1'-0"
			DATE: 8-1-80
			DRAWN: C.B.
			CHECKER:
			APPROVED: BM
			ALEXANDRIA, F.B.C. - PDU. <u>L.H.S. ELEVATION</u> <u>LOCATION OF PROBE ORIFICE</u> DRAWING NO. C-61 SHEET NO. 3 OF 4

FIGURE 7.3 (Cont'd)

FIGURE 7.3 (Cont'd)

FIGURE 7.4 - Revision to Gas Probe Filter System

Figure 7.5 Gas Sample Conditioning System

FIGURE 7.6

TABLE 7.1
LIST OF SYMBOLS TO FIGURES
7.5 and 7.6

<u>Symbol</u>	<u>Meaning</u>
AUX	Auxiliary
FI	Flow Indicator
HCV	Hand Control Valve
PI	Pressure Indicator
TI	Temperature Indicator

FIGURE 7.8 - Modification of Lower Boiling Bundle

FIGURE 7.8 (Cont'd)

FIGURE 7.9: Pitot tube mass flow rate vs. orifice plate mass flow rate for main air.

FIGURE 7.10: Pitot tube mass flow rate vs. orifice plate mass flow rate for Coal and Limestone Feed Air.

FIGURE 7.11: Pitot tube mass flow rate vs. orifice plate mass flow rate for Fly Ash Reinjection Air.

8.0 REFERENCES

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2. Ergun, S. Chemical Engineering Program, (48), 89 (1952)

APPENDIX A

A-1 DISCUSSION OF IN-BED DIFFUSION COEFFICIENT

A-1.1 Steady-State Diffusion Equation

Determining the axial and radial diffusion coefficient, D_a and D_r , makes it possible to calculate the mass transport of the constituents within the PDU.

The in-bed axial and radial diffusion coefficient, D_a and D_r , are related to the gas concentration C under steady-state conditions by:

$$u \frac{\partial C}{\partial z} = D_r \left(\frac{\partial^2 C}{\partial r^2} + \frac{1}{r} \frac{\partial C}{\partial r} \right) + D_a \frac{\partial^2 C}{\partial z^2} \quad (A-1.1)$$

where r is the radial direction, z is the axis of symmetry in the vertical direction, and u is the local air velocity.

A-1.2 Gas Interchange Coefficient of MIT Model

The gas interchange coefficient is the coefficient of mass transfer across a boundary where a concentration gradient exists. In the "bubbling bed" model, on which the MIT Model is based, there are three gas-interchange coefficients whose definitions depend on the mass transfer boundary. They are defined as follows:

$$- \frac{dN_{Ab}}{V_b dt} = K_{bp} (C_{Ab} - C_{Ae}) \quad (A-1.2)$$

$$= K_{bc} (C_{Ab} - C_{Ac}) \quad (A-1.3)$$

$$\approx K_{ce} (C_{Ac} - C_{Ae}) \quad (A-1.4)$$

where:

N_{Ab} : Number of moles of gas A in bubble.

V_b : Bubble volume.

C_{Ab} : Concentration of gas A in bubble.

C_{Ac} : Concentration of gas A in bubble cloud.

C_{Ae} : Concentration of gas A in emulsion phase.

K_{bp} : Gas interchange coefficient between bubble and emulsion phase.

K_{bc} : Gas interchange coefficient between bubble and bubble cloud.

K_{ce} : Gas interchange coefficient between bubble cloud and emulsion phase.

If Equations (A-1.3) and (A-1.4) are divided by K_{bc} and K_{ce} respectively, summed, and the result compared with Equation (A-1.2), the following relationship results:

$$\frac{1}{K_{bp}} \approx \frac{1}{K_{bc}} + \frac{1}{K_{ce}} \quad (A-1.5)$$

Note that subscript "bp" in K_{bp} should have been "be" to be consistent with other K subscripts in taking b for bubble and e for emulsion. However, bp is used to correspond to the same notation used in the MIT Model.

Following the same definitions as above, the MIT Model calculated K_{bc} and K_{ce} as follows:

$$K_{ce} = \text{very large} = 10^8 \quad (\text{assumed}) \quad (A-1.6)$$

$$K_{bc} = 4.5 \frac{U_{mf}}{d_b} + 5.85 \left(\frac{D^2 g}{d_b} \right) \quad (A-1.7)$$

where:

U_{mf} : Minimum fluidization velocity.

d_b : Bubble diameter.

D : Diffusivity of O_2 in N_2 .

g : Gravitational acceleration.

A-1.3 Diffusion Coefficient and Gas Interchange Coefficient

The interchange of gas between bubbles and the emulsion phase disperses the gas from one location to another in the bed. Hence, gas interchange coefficients must be related to the diffusion coefficients in the bed. Indeed, Kunii and Levenspiel reported that the radial and axial diffusion coefficients D_r and D_a could be computed from K_{bp} as follows:

$$D_r = 0.2 \frac{d_p^2 K_{bp}}{\delta} \quad (A-1.8)$$

$$D_a = \frac{b}{1-b} \frac{U_o U_b}{K_{bp}} \quad (A-1.9)$$

$$b = \alpha \epsilon_{mf} - \frac{U_{mf}(1-\delta-\alpha\delta)}{U_o - U_{mf}} \quad (A-1.10)$$

where:

U_b : Bubble velocity.

δ : Bubble volume fraction of the bed.

α : Volume ratio between bubble wake and bubble.

ϵ_{mf} : Bed voidage at minimum fluidization.

The coefficient K_{bp} , together with other bubble variables, are predicted by the MIT Model. D_r and D_a are neither predicted nor used as inputs by the Model, but they can be measured experimentally. Thus, the K_{bp} prediction from the Model can be verified through the use of Equation (A-1.8) and D_r measurements, or the use of Equation (A-1.9) and (A-1.10) and D_a measurements. The former method also requires the prediction from d_b and δ as shown in Equation (A-1.8). The later method involves many predictions such as U_o , U_{mf} , U_b , ϵ_{mf} , α and δ as shown in Equations (A-1.9) and (A-1.10), and therefore involves more sources of discrepancies than the former method. Hence, to verify the prediction of K_{bp} by the Model, the former method would be preferred over the later method.

Measurements leading to the experimental determination of D_r have been reported earlier in this report.

A-2 SOLIDS CHARACTERISTICS

A-2.1 Particle Size Distribution Data

The need for an analytical description of the particle size distribution has been lessened by the use of digital computers. It is, however, helpful if the data can be reduced to a more compact and manageable form. Many attempts have been made to find an analytical expression that would contain just two variable parameters; one characterizing the mean fineness and one characterizing the dispersion of sizes about the mean. It is also desirable that the expression should permit conversion between weight, surface area and number distribution. The expression should also allow easy calculation of the specific surface from any one of the distributions.

The most frequently applicable distributions include the normal, the log-normal, and the Rosin-Rammler.

Normal Distributions: If the distribution obeys the normal distribution, a plot of particle size d_p , in linear scale, versus the undersize P_u , or oversize P_o in probability scale should produce a straight line. Such a plot for Greer limestone is shown in Figure A-2.1. The non-linearity is apparent.

Log-Normal Distribution: This represents a skewed normal distribution. If such distribution applies, the plot of d_p , in log scale, versus P_u in probability scale should result in a straight line. Such plots for five solids are shown in Figure A-2.2. The non-linearity for all the solids is apparent.

Rosin-Rammler Distribution: The Rosin-Rammler relation is given by:

$$P_o = 100 \exp [-(d_p/d_{rr})^n] \quad (A-2.1)$$

where d_{rr} is the Rosin-Rammler mean diameter, n is a factor indicating the spread of the distribution, and P_o is expressed in percents. Rearranging Equation (A-2.1) and taking the log gives:

$$\begin{aligned} \log (100/P_o) &= (d_p/d_{rr})^n \log (e) \\ \log (100/P_o) &= 0.43 (d_p/d_{rr})^n \end{aligned} \quad (A-2.2)$$

Thus, if the distribution obeys the Rosin-Rammler distribution, a plot of $\log (100/P_o)$ versus d_p in a log-log graph should give a straight line having the slope equal to n . The d_p corresponding to a value of 0.43 for $\log (100/P_o)$ is d_{rr} .

Figure A-2.3 shows the Rosin-Rammler plot for Greer limestone data. The curve is linear for $d_p > 200$ microns. The deviation from linearity for $d_p < 200$ microns is quite sudden and appears to follow no defined trend. This deviation suggests that sieve screening results are often erroneous for fines. Pulverized coal measurements show that mechanical sieving is unreliable below 75 microns. The limit below which sieve screening becomes unreliable seems to be dependent on the type of solid.

Therefore, for Green limestone, this limit seems to be 200 microns. Assuming a limit of 200 microns, Greer limestone apparently obeys the Rosin-Rammler distribution with $n = 3.33$ and $d_{rr} = 1650$ microns.

Data for two spent bed materials samples are plotted according to the Rosin Rammler distribution in Figure A-2.4, and show good linearity for $d_p > 200$ microns. Sewickley coal, ashed Sewickley coal, and fly ash data are also plotted according to the Rosin-Rammler distribution in Figure A-2.5. They also show good linearity.

However, for coal and ashed coal, deviations from linearity (see Figure A-2.5) are apparent. This type of deviation shows a clear trend representing the effects of removing the coarser particles. These effects should exist in any distributions which only contain two characteristic constants indicating the mean and the spread.

This analysis has produced the following conclusions:

- The particle size distribution, of various solid streams in a coal AFB combustor can be expressed by the Rosin-Rammler equation better than the log-normal distribution equation.
- Mechanical sieving for size determination of Greer limestone and spent bed materials from an AFBC combustor is likely to be unreliable below 200 microns.

A-2.1.1 Weight Mean Particle Diameter

The weight mean particle diameter is calculated by:

$$\bar{d}_p = \frac{\int_{d_p=0}^{d_p=\infty} d_p d(P_o)}{\int_{d_p=0}^{d_p=\infty} d(P_o)} \quad (\text{A-2.3})$$

If the Rosin-Rammler distribution is assumed, Equation (A-2.1) can be substituted into the above equation which would then yield:

$$\bar{d}_p = d_{rr} \Gamma\left(\frac{1}{n} + 1\right) \quad (\text{A-2.4})$$

where Γ is the gamma function of n .

A-2.1.2 Surface Mean Particle Diameter

In fluidizations, the surface mean diameter is more representative of solid size than the weight mean diameter, and therefore has been used in this data analysis. The surface mean diameter is calculated by:

$$\bar{d}_{p,s} = \frac{1}{\sum_{i=1}^m \frac{x_i}{d_{pi}}} \quad (\text{A-2.5})$$

where x_i is the mass fraction of i th interval having the average diameter of d_{pi} . $\bar{d}_{p,s}$ can be calculated from the Rosin-Rammler equation via numerical method.

A-2.2 Solid Densities

A-2.2.1 Particle Density

Average values of particle apparent and true densities of Greer limestone and a typical spent bed material are presented below. These values were calculated as the mean of all measurements. This assumes that particle densities (apparent and true densities) are independent of particle size.

- For limestone:

$$\rho_p \text{ (Apparent)} = 2.62 \quad (\text{A-2.6})$$

$$\rho_p \text{ (True)} = 2.65 \quad (\text{A-2.7})$$

- For spent bed material:

$$\rho_p \text{ (Apparent)} = 2.09 \quad (\text{A-2.8})$$

$$\rho_p \text{ (True)} = 2.71 \quad (\text{A-2.9})$$

A-2.2.2 Bulk Density

The bulk density ρ_B is related to the particle density (apparent) as follows:

$$\rho_B = \rho_p(1-\epsilon) \quad (\text{A-2.10})$$

where ϵ is the void fraction. Hence:

$$\epsilon = 1 - \frac{\rho_B}{\rho_p} \quad (\text{A-2.11})$$

Values of ϵ , as calculated from Equation (A-2.11), are plotted versus particle size for Greer Limestone and a spent bed material in Figures A-2.6 and A-2.7, respectively. As shown in these Figures, ϵ seems to be a function of particle size. However, since this trend is unconfirmed, ϵ is assumed constant with respect to particle size and has a (average) value of:

- For Greer limestone:

$$\epsilon \text{ (unpacked)} = 0.44 \quad (\text{A-2.12})$$

$$\epsilon \text{ (packed)} = 0.41 \quad (\text{A-2.13})$$

Figure A-2.1 Greer Limestone Normal Distribution Plot

Figure A-2.2 Log-Normal Distribution Plots P_u (%)
for Greer Limestone, Sewickley Coal, Spent Bed Material, Ashed Sewickley Coal, and Fly Ash

Figure A-2.3 Greer Limestone Rosin-Rammler Plot

Figure A-2.4 Rosin-Rammler Plot of Two Spent-Bed Materials

Figure A-2.5 Rosin-Rammler Plots for Sewickley Coal, Ashed Sewickley Coal, and Fly Ash

Figure A-2.6 Greer Limestone Voidage vs. Particle Size

Figure A-2.7 Spent Bed Material Voidage vs. Particle Size

TABLE A-2.1

PARTICLE CHARACTERISTICS OF AFB SOLIDS

Solid	Rosin-Rammler Constants		Weight Mean Particle Dia. (Microns)		Specific Surface (CM ² /g)	
	n	d _{rr} (microns)	(1)	Eqn. 19	Eqn. 20	Literature
Greer Limestone	3.33	1650	1063 1667 1332	1480	18.63	409.0** (EPRI)
Spent Bed Materials	2.44 3.00	1680 1674	1160 1381	1490 1494	25.02	
Sewickley Coal	1.25	2360	2294	2198	--	
Ashed Coal	0.47	1040	1470	2347*	--	
Fly Ash	1.21	157.5	156	148	--	

(1) Calculated from $\bar{d}_p = \sum P_i d_{pi}$

* The gamma function solution is not appropriate because $(\frac{1}{n} + 1)$ is greater than 2.0

** This includes surface area of the particle pores.

- For spent bed material:

$$\epsilon \text{ (unpacked)} = 0.46 \quad (\text{A-2.14})$$

$$\epsilon \text{ (packed)} = 0.43 \quad (\text{A-2.15})$$

Note: Void fraction for each test performed is different from above and has its own specific value.

A-3 AIR DISTRIBUTOR CHARACTERISTICS

Studies of the air flow through the distributor plate nozzles and of the corresponding pressure drop across the plate can yield data describing the flow regime close to the distributor plate. This provides a better understanding of the fluid dynamics within the PDU.

Air flow through distributor nozzles can be regarded as frictionless adiabatic, or isentropic, because of the short distance traveled. Also, the flow through the nozzle can be approximated as one-dimensional (1-D) flow.

For one dimensional, isentropic flow, it can be proved from Euler's Equation that:

$$\frac{P_1}{P_2} = \left(1 + \frac{k-1}{2} M^2\right)^{k/(k-1)} \quad (\text{A-3.1})$$

where P_1 and P_2 are pressures at inlet (air plenum) and outlet of the nozzle, k is the adiabatic exponent in the gas equation of state, and M is the Mach number defined by:

$$M = \frac{\text{Air Velocity } U}{\text{Sound Velocity } C} \quad (\text{A-3.2})$$

The sound velocity C is calculated by:

$$C = \sqrt{\frac{kP}{\rho} g_c} = \sqrt{kRT g_c}, \text{ for a perfect gas} \quad (\text{A-3.3})$$

where ρ is air density, T is air absolute temperature and R is the gas constant.

Test series 20-1 and 20-2 were intended to investigate existing air nozzle design to develop a correlation between air flow rate and nozzle pressure drop (for the existing air nozzle design). The objective of Test 20-3 was to correlate the pressure drop across the distributor plate with the air flow rate through the plate.

The location of the nozzles during Tests 20-1 and 20-2 is shown in the following sketch:

A small bench scale apparatus was designed and fabricated for these tests. This apparatus was used to obtain pressure drop data for a single nozzle and data on the effects of 6 additional surrounding nozzles. The apparatus arrangement and the container for air distributor nozzle test are shown in Figures A-3.1 and A-3.2. The data for Tests 20-1 and 20-2 are given in Table A-3.1.

In all the tests, the air velocity is so low that the flow regime is in the laminar range.

During Test 20-3 the pressure drop across the air distributor was measured as a function of the mass flow rate of air. The reduced data for Test 20-3 is shown in Table A-3.2.

The analysis presented here will examine three aspects of air distributor pressure drop:

- (1) Pressure drop, $P_1 - P_2$, of a single nozzle compared with theoretical prediction from Equation (A-3.1).
- (2) Effects on the nozzle pressure drop due to flow through adjacent nozzles, and

- (3) The pressure drop of the PDU air distributor compared to the pressure drop of a single nozzle.

Each of the above topics is discussed in the following sections.

A-3.1 Pressure Drop of Single Nozzle

To compare the pressure drop data in Table A-3.1 with predictions from Equation (A-3.1), C, U, and M must be calculated at the exit temperature and pressure conditions. Equation (A-3.3) can be rearranged and the exit sound velocity C and air velocity U can be calculated as follows:

$$C = \sqrt{\frac{g_c k P_2}{\rho_1} \left(\frac{P_1}{P_2}\right)^{1/k}} \quad (A-3.4)$$

$$U = \frac{W}{\rho_2 A} = \frac{W}{\rho_1 A} \left(\frac{P_1}{P_2}\right)^{1/k} \quad (A-3.5)$$

Substituting into Equation (A-3.2):

$$M = \frac{U}{C} = \frac{W}{A} \sqrt{\frac{(P_1/P_2)^{1/k}}{g_c \rho_1 k P}} \quad (A-3.6)$$

where W is the mass flow rate of air, and A is the flow area at the exit of the nozzle.

The values of M determined for the data in Table A-3.3, and their corresponding pressure drops, are compared with the predictions from the theory of isentropic, one-dimensional flow in Figure A-3.3. The results are a series of curves, one for each nozzle. The data for the center nozzle is the only curve shown in Figure A-3.3. These curves are almost parallel to the theoretical curves, i.e. the solid lines. The scattering of the data points may be due to the variation of nozzle geometry. Experimental errors can also be an additional factor contributing to the scattering of data.

A-3.2 Effects of Adjacent Nozzles

The effect of flow through adjacent nozzles on the center nozzle is determined by plotting pressure drop, ΔP , versus air mass flow rate, W for the following arrangements:

1. six nozzles, at 3/4" from each other and from the center
2. the six nozzles in (1), plus the nozzle at the center

3. six nozzles at $1\frac{1}{2}$ " from each other and from the center
4. the six nozzles in (3), plus the center nozzle
5. all 13 nozzles.

If there is no effect on the center nozzle, the difference in W for a particular value of ΔP between (1) and (2), (3) and (4), and (5) and (1) + (3) would yield the mass flow rate of air through the center nozzle. Also this mass flow rate would lie on the curve of ΔP versus W for the flow through the center nozzle alone. The values of W for the center nozzle determined from the above setups are compared with the measured values in Table A-3.3. The curves for these setups are shown in Figure A-3.4.

An analysis of Table A-3.3 and Figure A-3.4 yields the following conclusions:

- The assumption that there are no effects caused by flow through adjacent nozzles does not appear true because of large deviations from measured values in the last three columns in Table A-3.3.
- The data indicates that multiple nozzle flow produces a pressure drop somewhat lower than the pressure drop of an individual nozzle operated at the same velocity.
- The effects also seem to be dependent on the distance between the nozzles. However, because of limited data, it is uncertain whether pressure drop would be reduced when the nozzle spacing is increased.

A-3.3 Overall Air Distributor Pressure Drop

The air distributor plate in the PDU has 2712 nozzles installed 0.724" apart. The total flow area, in Equation (A-3.6), is 2712 times the area of a single drilled bolt. With an approximate hole size of $7/64$ ", the area of a single drilled bolt is about 0.00933 in^2 . Therefore, total flow area is about 25.31 in^2 . The plot of $\Delta P/P_2$ versus M^2 is shown in Figure A-3.5. This figure clearly shows that the plate pressure drop deviates from the isentropic, 1-D flow.

A-3.4 Conclusions

Multiple nozzles behave differently from a single nozzle, and from isentropic, one-dimensional flow, when subjected to multiple nozzle flow.

LOCATION OF
1/4"Ø HOLE FOR
THERMOMETER
(SEE DWG. 3B
FOR ENDPLATE
DETAIL)

WELD 1/4"Ø TUBING
TO ONE SIDE OF CONTAINER.
TUBE TO BE FLUSH TO THE
INSIDE AND 2" IN LENGTH

SAFETY VALVE
SET AT 10 P.S.I.

WELD 1"Ø X 12"
LONG PIPE ALL
AROUND, FLUSH TO
INSIDE ON SIDE
OPPOSITE 1/4"Ø TUBING

TOP VIEW
N.T.S.

NOZZLE TEST CONTAINER DETAILS

NOTE:
FABRICATE THIS PART,
BEFORE WELDING TO WALL
FROM A BRASS COUPLING
USED FOR JOINING 1/4"
COPPER TUBING.

Figure A-3.1 Apparatus Arrangement for Tests 20-1 and 20-2

Figure A-3.2 Nozzle Plate of the Apparatus for Tests 20-1 and 20-2

TABLE A-3.1
DATA FOR TESTS 20-1 & 20-2

Run No.	Air Flow (SCFH)	Temp. (°F)	Press. Drop (in H ₂ O)	Number of Nozzles	Nozzle Location	Distance from Center (in.)
1	32.5	80.0	5.60	1	center	0
2	39.8	80.0	8.68	1	center	0
3	68.0	78.0	23.00	1	center	0
4	95.2	78.0	45.40	1	center	0
5	86.0	72.0	55.40	1	L.M.	3/4
6	73.0	72.0	41.30	1	L.M.	3/4
7	56.2	72.0	11.56	1	L.M.	3/4
8	29.5	72.0	4.10	1	L.M.	3/4
9	28.5	72.0	4.75	1	L.B.	3/4
10	53.0	72.0	16.05	1	L.B.	3/4
11	76.5	72.0	31.30	1	L.B.	3/4
12	95.0	72.0	47.80	1	L.B.	3/4
13	89.0	72.0	55.80	1	R.B.	3/4
14	78.0	72.0	48.50	1	R.B.	3/4
15	61.0	72.0	29.70	1	R.B.	3/4
16	42.0	72.0	12.60	1	R.B.	3/4
17	24.0	72.0	3.40	1	R.B.	3/4
18	22.0	72.0	2.15	1	R.M.	3/4
19	49.4	72.0	14.20	1	R.M.	3/4
20	69.8	72.0	28.80	1	R.M.	3/4
21	85.0	72.0	43.75	1	R.M.	3/4
22	87.0	71.0	45.80	1	R.T.	3/4
23	71.8	70.5	30.30	1	R.T.	3/4
24	45.6	70.5	12.70	1	R.T.	3/4
25	25.8	70.5	3.90	1	R.T.	3/4

Barometric Pressure: 30.34 in Hg @ 50.0°F

Notes:

R.T. - Right Top

R.B. - Right Bottom

R.M. - Right Middle

L.T. - Left Top

L.B. - Left Bottom

L.M. - Left Middle

TABLE A-3.1 (Cont'd)

DATA FOR TESTS 20-1 & 20-2

Run No.	Air Flow (SCFH)	Temp. (°F)	Press Drop (in H ₂ O)	Number of Nozzles	Nozzle Location	Distance from Center(in.)
26	25.0	70.5	3.70	1	L.T.	3/4
27	52.0	70.5	13.60	1	L.T.	3/4
28	74.0	70.0	31.80	1	L.T.	3/4
29	88.0	70.0	56.10	1	L.T.	3/4
30	100.0	70.0	2.40	6	corners	3/4
31	182.5	70.0	5.15	6	of	3/4
32	356.0	70.0	20.00	6	hexagonal	3/4
33	430.5	70.0	26.70	6		3/4
34	533.0	70.0	44.90	6		3/4
35	28.0	60.0	4.70	1	R.M.	1 1/2
36	50.0	60.0	15.60	1	R.M.	1 1/2
37	68.0	60.0	38.20	1	R.M.	1 1/2
38	75.8	60.0	48.90	1	R.M.	1 1/2
39	71.8	60.0	54.70	1	R.T.	1 1/2
40	57.0	60.0	32.30	1	R.T.	1 1/2
41	40.5	60.0	17.30	1	R.T.	1 1/2
42	22.8	60.0	5.90	1	R.T.	1 1/2
43	24.0	60.0	5.25	1	L.T.	1 1/2
44	43.8	60.0	17.90	1	L.T.	1 1/2
45	63.6	60.0	38.80	1	L.T.	1 1/2
46	71.0	60.0	56.70	1	L.T.	1 1/2
47	75.0	60.0	54.80	1	L.M.	1 1/2
48	56.2	60.0	29.50	1	L.M.	1 1/2
49	38.4	60.0	15.70	1	L.M.	1 1/2
50	22.4	60.0	5.40	1	L.M.	1 1/2

TABLE A-3.1 (Cont'd)
DATA FOR TESTS 20-1 & 20-2

Run No.	Air Flow (SCFH)	Temp. (°F)	Press Drop (in H ₂ O)	Number of Nozzles	Nozzle Location	Distance from Center(in.)
51	69.5	60.0	24.25	1	R.B.	1 1/2
52	73.0	60.0	25.40	1	R.B.	1 1/2
53	55.0	60.0	15.10	1	R.B.	1 1/2
54	26.2	60.0	3.75	1	R.B.	1 1/2
55	25.0	60.0	3.40	1	L.B.	1 1/2
56	43.0	60.0	9.60	1	L.B.	1 1/2
57	62.0	60.0	19.40	1	L.B.	1 1/2
58	76.0	60.0	34.00	1	L.B.	1 1/2
59	87.5	60.0	0.30	13	13 central nozzles	0, 3/4 and 1 1/4
60	193.0	60.0	1.60	13		
61	298.0	60.0	2.20	13		
62	351.0	60.0	5.20	13		
63	516.0	60.0	9.40	13		
64	528.0	70.0	28.35	7	corners of hexa- gonal & center	3/4, 0
65	363.0	70.0	14.60	7		
66	256.0	70.0	8.50	7		
67	89.0	70.0	1.10	7		
68	94.8	70.0	1.95	7	corners of hexa- gonal & center	1 1/2, 0
69	228.0	70.0	7.70	7		
70	319.0	70.0	13.60	7		
71	433.5	70.0	39.65	7		
72	445.7	70.0	47.20	6	corners of hexagonal	1 1/2
73	339.9	70.0	28.90	6		
74	224.9	70.0	11.55	6		
75	140.1	70.0	5.30	6		
76	87.0	70.0	1.60	6		

TABLE A-3.2
PRESSURE DROP DATA FOR GRID PLATE

Run No.	Air Flow Rate (lbs/h)	Upstream		Grid ΔP (in. H ₂ O)	Down-stream Temp. (°F)
		Temp. (°F)	Pressure (in. H ₂ O abs.)		
1	15896.5	108.5	433.4	30.5	98.4
2	14918.5	114.8	432.4	27.5	106.3
3	13193.8	115.8	425.9	21.5	108.4
4	11840.2	116.9	422.4	17.5	109.5
5	10921.2	117.7	419.4	15.0	109.9
6	10136.3	118.5	416.9	13.0	110.1
7	8820.5	118.5	413.9	10.0	109.5
8	7196.7	118.5	410.7	6.7	109.2
9	5680.7	118.3	408.9	4.5	109.2
10	3263.8	118.4	402.9	1.75	108.9
11	4729.8	118.9	405.9	3.0	108.2
12	6546.1	120.0	410.4	5.5	108.9
13	8102.2	120.3	411.4	8.0	109.8
14	9288.9	120.0	414.4	10.5	110.0
15	10350.1	119.7	416.4	13.0	109.9
16	11401.0	119.5	418.9	16.0	109.7
17	12584.2	119.3	422.4	20.0	109.7
18	13978.4	118.8	426.4	24.0	109.7
19	15169.6	118.6	430.9	28.5	109.6
20	15777.4	118.8	432.9	30.5	109.7

TABLE A-3.3
AIR FLOW RATE IN LB/H THROUGH
THE CENTER NOZZLE

<u>ΔP (in. H₂O)</u>	<u>Central Nozzle Measured</u>	<u>Single Nozzle Theoretical</u>	<u>Determined From Set Ups*</u>		
			<u>1 & 2</u>	<u>3 & 4</u>	<u>5 & (1 + 3)</u>
5.60	2.5	2.8	2.6	3.4	5.2
8.68	3.0	3.5	3.4	4.5	6.6
23.00	5.2	5.9	6.0	8.0	8.8
45.40	7.1	9.0	9.7	11.2	--

* For explanation see text

FIGURE A-3.3: Pressure Drop vs. Flow Through the Center Air Distributor Nozzle.

Figure A-3.4 Pressure Drop vs. Flow Through Various Arrangements of Air Distributor Nozzles

FIGURE A-3.5: Pressure Drop vs. Flow Through the PDU Distributor Compared to Nozzle.

Therefore, design of a multiple nozzle distributor plate may not be based on data from single nozzle tests.

The difference between the pressure drop of multiple nozzles and that of single nozzle, or of the ideal flow, is probably due to the flow mixing at the outlets of these nozzles. If this is true, then the effects of the fluidizing solids at the outlets of the nozzles can also become a significant factor.

Thus, future tests need to be done. These tests would include measuring the pressure drop across the distributor plate and the air flow rate while solids are being fluidized by the air flow. These tests would help define PDU fluidization and relate all of the previous information more closely to cold mode testing.

A-4 DATA REDUCTION

A-4.1 Air Flow

Air flow is the calculated mass flow rate of air through the main air orifice plate, as described by the following equation:

$$W = 11960 \sqrt{\Delta P \rho} \quad (A-4.1)$$

where:

W: Mass flow rate

ΔP : Pressure drop across main air orifice plate

ρ : Density of air at upstream pressure and temperature of the orifice plate

Note: Whenever transport air and burner were in service the flow rates were added to the main air flow.

A-4.2 Temperature and Bed Pressure

Thermocouples TC-H3 and TC-H4 monitor the bed temperature at two different locations. The arithmetic mean of these temperatures is presented within relevant Tables.

Bed pressure is calculated as follows:

$$\left(\begin{array}{c} \text{Bed} \\ \text{Mean} \\ \text{Pressure} \end{array} \right) = \left(\begin{array}{c} \text{Barometric} \\ \text{Pressure} \end{array} \right) + \left(\begin{array}{c} \text{Gauge} \\ \text{Pressure} \\ \text{at Lowest} \\ \text{Point in} \\ \text{Bed} \end{array} \right) - 1/2 \left(\begin{array}{c} \text{Bed} \\ \text{DP} \end{array} \right) \quad (A-4.2)$$

Thermocouples TC-H3 and TC-H4 are located at 11 and 12 inches above the air distributor nozzles, respectively.

A-4.3 Determination of Expanded Bed Height and Bed DP

The pressure drop in the PDU is measured at 4.5, 12.0, 15.0, 18.0, 24.0, 28.0, 32.0, 39.0 and 45.0 inches above the air distributor nozzles in reference to a pressure tap located at 1.5 inches above the air nozzles. Expanded bed height is determined graphically by plotting the height above the nozzles versus the pressure drop. A typical plot of the height versus pressure drop is presented in Figure A-4.1 for the case without an in-bed bundle. The pressure drop for this case from 1.5 inches above the air nozzles to the top of the bed is determined from Figure A-4.1. Also, see Figure A-4.2 for the determination of the expanded bed height and the bed DP for the case with an in-bed bundle. Note that in Figure A-4.2, the reference pressure tap was at 3.5" above the nozzle tips.

There are a number of methods that can be used to determine the expanded bed height of a fluidized bed. However, a pressure profile plot in Figure A-4.1 has been chosen here because it is the most practical and accurate method for the Alexandria PDU. Other methods have been attempted without encouraging results. The accuracy of the pressure profile method depends on the accuracy of pressure profile data and also on the merit of one basic assumption, namely that the bulk density under fluidizing conditions is constant in the vertical direction of the bed. This assumption results in a straight line with the slope being an indication of the bed bulk density. This assumption is made because of the complete lack of the information on the acceleration of air or gas passing through the bed. Note that bubble growth has no relation to the residence time of total air or gas in the bed.

A different procedure as indicated in Figure A-4.2 was used only for Test Series 40 in which the in-bed tube bundle was installed. When there is a tube bundle in the bed, bubbles are retarded and broken up as they pass through the tube arrays. This action would result in a higher residence time for the air or gas, although they do not affect the mass throughput rate of the air or gas. The outcome would therefore be a less dense bed with lower bulk density under fluidizing conditions. The slope of the pressure line in the tube region would therefore be different from the pressure line below the bundle to indicate lower "fluid" density.

In terms of the accuracy of pressure measurements near the distribution jets, both pressure taps located at 1.5 and 3.5 inches above the nozzles are not in the jet stream, thus, the jets from the nozzles do not affect the accuracy of the readings from these pressure taps.

The total pressure drop across the bed is determined in reference to Figures A-4.1 and A-4.2 as:

$$\text{Total Bed DP} = \text{Bed DP} + \Delta P'$$

A-4.4 Collected Solids Determination

To ensure that the collected bed samples were representative of the material in the PDU, bed material deposited in the sample drain pipe was emptied prior to actually taking a sample. The sample was weighed and documented at the completion of each test series. The entrained solids collected in the cyclone hopper was also weighed. Collected solids data were obtained for material balance checks only. They are not reported in this document.

A-4.5 Determination of Minimum Fluidization Velocity

The minimum fluidization velocity is determined by plotting the total bed pressure drop vs. the superficial velocity. A typical graph is presented in Figure A-4.3. Under fluidizing conditions, the superficial air velocity is the same as the fluidizing velocity, and is calculated by:

$$\text{Superficial Velocity} = \frac{\text{Air Flow (lb/hr)}}{\rho_{\text{air}}(\text{lb/ft}^3) \cdot 3600 (\text{s/hr}) \cdot 11.4 (\text{ft}^2)} \quad (\text{A-4.6})$$

where:

ρ_{air} - The density of air at bed temperature and pressure

A-4.6 Grid ΔP Determination

The pressure drop across the grid plate is the differential pressure between plenum and a pressure tap located 1.5" above the air distributor nozzles in case the bed is empty.

Figure A-4.1 Test 20-8, Run No. 1 Expanded Bed Height

Figure A-4.2 Test 40-2, Run No. 2, at Time 105 Minutes
Expanded Bed Height

Figure A-4.3 Test 40-1 Minimum Fluidization Velocity

APPENDIX B

B-1 DESCRIPTION OF THE COMPUTERIZED DATA ACQUISITION SYSTEM

Transient process data is sampled and recorded for later analysis through the use of a real-time computerized data acquisition system. Primary hardware consists of:

- DEC PDP 11/04 Computer
- NEFF System 620/Series 400 Multiplexer
- 6 NEFF D/A Converters
- Kennedy Model 9000 Tape Drive
- Gould Model 2000 6-channel Strip Chart Pen Recorder
- System Interface Panel

The software consists of a monitor/control program which allows system initialization, modifications to data scan lists and rates, changes to calibration channel alarm limits, etc. The monitor also outputs alarm and diagnostics, active channel lists and current measured values of selected channels.

The primary output is tape recorded data, which is not accessible until after test completion when the tape is removed for processing in a separate computer facility. Additional output hardware can be added should more rapid access or data recall during operation be required.

Other DAS outputs include single or multiple printout of current measured values, and analog strip chart records of up to 6 channels selected from among 63 active channels in the scan list shown in Table B-1.

B-2 INPUT SIGNALS

All inputs to the DAS must be voltage signals. In many cases, a voltage signal is readily available. For example, temperature signals from thermocouple junctions are voltages whose range is typically 0 to 40 millivolts. Other instruments that measure pressure, flow, signals from load cells, gas composition, etc., can have voltage outputs in various ranges such as 0 to 10 millivolts, 0 to 1 volt, 0 to 5 volts, etc. These signals are fed directly into the system.

However, process signals are often represented by a direct current. Typically, a 4 to 20 milliamp current would represent a process variable such as 0 to 300 psia, or 3,000 to 5,000 lbm/hr. Current signals are measured by inserting a precision 110 ohm resistor in the current loop, and detecting the voltage drop across it with the DAS. The 110 ohm resistor is well within the series resistance limit of most current output signal transmitters.

For existing current output signal transmitters, the current measuring resistor has been mounted inside the transmitter enclosure. Power for the transmitters is provided by existing instrumentation. In addition, the system is equipped with a regulated voltage source in the system Interface Panel, which can also supply transmitters with the power necessary for their operation. In this case, the current measuring resistor is mounted in the DAS. Certain channels are available for this purpose.

B-3 MULTIPLEXING AND AMPLIFICATION

Input voltage signals are multiplied into a single programmable gain amplifier chain. Multiplexing occurs before amplification, using low-noise solid-state, CMOS-FET computer controlled switching elements. Following amplification, the analog voltage is sampled, held, and digitized in a 12-bit A/D converter. The full scale converter acceptance range is -10.24 v to +10.24 v, corresponding to -2047 counts to +2047 counts. The gain factor applied to each input signal is selected to produce an output value that utilizes the available input range of the A/D converter.

The amplifier chain has 12 gain values to choose from: 1, 2, 4, 8, 16, 32, 64, 128, 256, 512, 1,024 and 2,048. These correspond, respectively, to full scale voltages of 10.24 v, 5.12 v, 2.56 v, 640 mv, 320 mv, 160 mv, 80 mv, 40 mv, 20 mv, 10 mv, 5mv. Thermocouple channels typically have assigned gain values of 256. "Current" channels, in which a 4-20 mamp signal passes through 110 ohms producing a 440-2200 mv signal, typically have assigned gain values of 4.

Each channel has a 3-wire input: +, -, and guard shield. The + inputs are separately multiplied, and the amplifier chain is differential, providing excellent common mode rejection.

B-4 DATA CHANNEL SCANNING

The basic sampling rate of the DAS multiplexer-amplifier-A/D converter is 5000 samples/second. The sequence in which channels are scanned is manually entered into the DAS, and can be changed at any time. In a single 200 microsecond interval, the multiplexer switches are set to feed the signal from a selected channel to the programmable gain amplifier, the gain is set to the selected value for that channel, the signal is sampled, held, digitized, and the signed number of digital counts is loaded into the output data buffer prior to being stored on tape. In the next 200 microsecond interval, the DAS moves to the next scan list channel, and the steps are repeated. Upon scan list completion, the DAS waits until the designated time to initiate the succeeding scan, then proceeds with another scan of every channel in the scan list.

The period is the time interval between initiation of successive data scans. It can be less than a second, or many seconds. Typical values are 1, 2 or 5 seconds. The period can be changed at any time, even during test operation.

When proceeding through a scan list, it is possible for signal inputs in successive channels to be alternatively very large (close to 10 v) or very small (less than 10 mv), in which case the programmable gain would switch from 1 to 2048, back to 1, etc. When switching from high to low gain, the system proceeds at the basic sampling rate for each channel in the list. When switching from low to high gain, however, the system changes gain by no more than 3 factors of 2 in any 200 microsecond interval. This gives the switching network time to discharge and reduces any possible channel cross-talk. If a gain change of more than 3 factors of 2 is called for (for example, from 4 to 1024), the system automatically adds dummy sampling intervals, until the targeted gain value is reached before sampling that channel (in the example, 4 to 32 to 256 to 1024, or two extra intervals).

In most cases, the scan is so rapid that the sampling time for each channel can be assumed equal to scan initiation time. If necessary, the exact sampling time can be computed from the initiation time by keeping track of all the 200 microsecond sampling intervals in going through the scan list.

The same channel can also be entered repeatedly in the scan list, for more frequent sampling.

B-5 FREQUENCY RESPONSE, CLAMPLING, OVER-VOLTAGE PROTECTION

Input circuitry for every channel on the multiplexer circuit boards includes RC-networks to limit frequency response above 10 cycles/second. Attenuation of 60 cycles/sec signals is greater than 30dB.

Clamping circuits are provided to limit amplifier saturation by overvoltage signals. This guards against channel cross-talk and speeds amplifier recovery.

Protection is provided in each input channel for inadvertent connection to 100 v ac or dc signals.

B-6 CALIBRATION

The same amplifier gain is used for every channel; gain can be adjusted under digital control to one of 12 possible values. (Two amplifiers in series have gains of: 1 and 64; and 1, 2, 4, 8, 16 and 32). Gain is set by a 4-bit binary control word, equivalent to the gain factor two's exponent. It is assumed that actual value of amplifier gain is independent of channel number assignment, and that possible variations in gain value associated with component drift or failure are uncorrelated with channel number or scan list position. To guard against possible variations in gain value, 16 channels are dedicated to gain calibration. A precision voltage source equipped with calibrated resistive dividers provides 12 voltages corresponding to 7/8ths full scale signal for each gain value. The voltages are available in plus-and-minus pairs, to enable amplifier gain calibration for positive and negative going signals. Therefore, only 8 gain values can be calibrated by the 16 channels dedicated to that purpose during system operation. Typically, fewer than 8 gain values are needed in any data-gathering operation.

Calibration channels are scanned once per minute to test that A/D converter outputs corresponding to the amplified calibration voltages fall within specified low and high limits. An alarm bell sounds and a message is typed out if calibration is out of range. Typical gain stability is better than 1-part-in-2,000.

Calibration channels are customarily included in the data scan list. Therefore, updated calibration correction signals are available on the data tapes for each scan. Even though a particular gain value may have varied considerably, its effect is removed when the Data Reporting Tape is prepared from the Primary Data Tape.

TABLE B-1
SCAN LIST OF INSTRUMENTS CONNECTED TO THE
DATA ACQUISITION SYSTEM

<u>Instrument ID</u>	<u>Units</u>	<u>Description</u>
F2	In-H ₂ O	Feed Water DP
F6	In-H ₂ O	Steam DP
F9	In-H ₂ O	DP betw. pressure taps 1/5
F11	In-H ₂ O	Coal transport air DP
F12	In-H ₂ O	DP betw. pressure taps 1/15
G1	PPM	SO ₂ concentration
G2(Westinghouse)	%	O ₂ concentration
G3(Taylor)	%	O ₂ concentration
G4	PPM	NO concentration
G6	PPM	CO concentration
G7	%	CO ₂ concentration
G8	PPM	HxCy concentration
M2	RPM	Coal feeder speed
M3	RPM	Limestone feeder speed
M5	Lb	Baghouse bin mass
M6	Lb	Coal hopper mass
M7	Lb	Limestone hopper mass
L1	In-H ₂ O	Drum level DP
P4	In-H ₂ O	Boiler bundle DP
P11	In-H ₂ O	Boiler bundle DP
P13	In-H ₂ O	Boiler bundle DP
P12	In-H ₂ O	Baghouse DP
P7	In-H ₂ O	Low bed press
P9B	In-H ₂ O	Bed level DP
P10	In-H ₂ O	Cell draft
P1	In-H ₂ O	Main air DP
P2	In-H ₂ O	Plenum air press.
G9	%	Coal feeder setting
G10	%	Limestone feeder setting

TABLE B-1 (Cont'd)

Thermocouple	LOCATION		
	From Front of PDU	From Grid Plate	Into the Wall
TC-1	18"	+2"	10"
TC-2	18"	+6"	10"
TC-3	"	+10"	"
TC-4	"	+13"	"
TC-5	"	+17"	"
TC-6	"	+25"	"
TC-7	"	+36"	"
TC-8	"	+60"	"
TC-9	"	+82"	"
TC-10	"	+94"	"
TC-11	"	+105"	"
TC-12	"	+127"	"
TC-13	"	-6"	"
TC-14	Main air orifice temp (same location as existing one)		
TC-15	Dust collector gas outlet temp.		
TC-16	Stack gas temp.		
TC-17	Between steam drum and gas cooler gas temp.		
TC-18	Gas cooler water outlet temp.		
TC-19	Gas cooler water inlet temp.		
TC-20	FD fan inlet temp.		
TC-21	Gas temp. between cooler and cyclone		
TC-22	Feed water temp. (after pump)		
TC-23	Circulation water temp. (after pump)		
TC-24	Lower bundle water inlet temp.		
TC-25	Lower bundle water outlet temp.		
TC-26	L.H. upper bundle water inlet temp.		
TC-27	L.H. upper bundle water outlet temp.		
TC-28	R.H. upper bundle water inlet temp.		
TC-29	R.H. upper bundle water outlet temp.		
TC-30	Bed drain temp.		
TC-31	Dust collector cyclone hopper fly ash temp.		
TC-32	Fly ash reinjection temp.		
TC-33	(midway between cyclone and PDU)		
TC-34	Steam leaving steam drum		

APPENDIX C

C-1 Effect of Air Distributor Leakage on Test Results

C-1.1 Objective

To determine the impact of air distributor leakage on the experimental results reported so far.

C-1.2 Approach

The approach to the preceeding test objective was to sequentially:

- (1) Identify leakages
- (2) Repeat two tests that represent the cold mode test data previously reported, without fixing-up any of the leaks
- (3) Repair all the leaks
- (4) Visually inspect a shallow bed (8" maximum) to determine if any leaks exist
- (5) Repeat tests in Step 2 above
- (6) Compare the experimental results collected in Step 2 to results collected in Step 4 and,
- (7) Visually inspect a shallow bed (8" maximum) to determine if any leaks exist at end of Step 5 above

Six tests were performed during this effort. Four tests were quantitative whereas, the remaining two consisted of visual observation for leaks. The tests were designed to check the expanded bed height data as well as minimum fluidization velocities in the range reported previously.

C-1.3 Test Plan & Procedure

Six tests designated C-1A, C-1B, V-1, C-2A, C-2B and V-2 were performed in cold mode with the spent bed material on hand. These tests are described in the sections which follow and are listed in Table C-1.1 in a summary form. To insure consistent results from the four quantitative tests C-1A, C-1B, C-2A and C-2B, the same bed material was used before and after the grid plate was repaired. However, to enable repairs to be made, the bed material was removed and stored in the bed

TABLE C-1.1

SUMMARY OF
OPERATING CONDITIONS

Test No.	Test Duration (hrs)	Slumped Bed Height (in.)	Test Parameters	Material Samples	Air Orifice Plate ΔP , (in. H ₂ O)
C-1A	2	15	Minimum Fluidizing Velocity and Expanded Bed Height	Bed	12,10,8,6, 4,2,1.6,1.0, 0.5,0.2,0.5, 1.0,2,4,10,15, I7
C-1B	2	20.5		Bed	Same as above
V-1	1/2	8	Observation for Leaks	Bed	~7.0
C-2A	2	20.5	Same as for Test C-1 (A & B)	Bed	Same as for Test C-1 (A & B)
C-2B	2	15		Bed	Same as above
V-2	1/2	8	Observation for Leaks	Bed	~7.0

material bin during the repair work. Pressures and drafts were recorded for each run and temperatures were recorded every 5 minutes throughout the tests. The slumped bed height was measured at the beginning, middle and end of each test. Solids samples were collected from the side port of the PDU at the beginning and end of each run. No spent bed material was fed or discharged throughout the tests.

Test C-1A

Prior to initiation of run number 1 in C-1A, the slumped bed height was measured to be 15.0 inches. The bed was then fluidized with air from the coal transport line and the main air orifice plate with a pressure drop of 12.0 inches of water. The main air orifice pressure drop was later changed sequentially to 10.0, 8.0, 6.0, 4.0, 2.0, 1.6, 1.0, 0.5, 0.2, 0.5, 1.0, 2.0, 4.0, 10.0, 15.1 and 17.0. After run number 10 the slumped bed height was measured at 14.5 inches, and the slumped bed height at the completion of C-1A, which consisted of 17 runs, was measured to be 14.0 inches.

Test C-1B

The slumped bed height was measured to be 20.5 inches before the initiation of run number 1 in C-1B. The bed was fluidized using the coal transport air and the main air with the main air orifice pressure drop at 12.4 inches of water. The main air orifice pressure drop was changed sequentially in later runs to 9.5, 7.0, 4.0, 1.9, 1.0, 0.5, 1.5, 3.0, 5.6, 8.6, 11.4 and 13.4 inches of water. At the completion of the test, the slumped bed height was measured to be 19.8 inches.

Test C-2A

At the beginning of C-2A, the slumped bed height was measured at 21.5 inches. The bed was then fluidized using coal transport air and the main air with the main air orifice pressure drop 11.0 inches of water in run number 1. The main air pressure drop was then sequentially changed in later runs to 8.0, 5.0, 3.0, 1.0, 0.5, 0.5, 1.5, 2.5, 3.0, 6.5, 10.0 and 12.5 inches of water. After run number 6 the slumped bed height was measured at 20.5 inches. At the end of the test, the slumped bed height was 19.5 inches.

Test C-2B

At the beginning of C-2B, the slumped bed height was measured to be 15.0 inches. Coal transport air and the main air with a pressure drop of 15.3 inches of water across the

main air orifice was then used to fluidize the bed. The main air pressure drop was then sequentially changed for later runs to 12.5, 9.0, 7.0, 4.0, 2.0, 1.0, 0.5, 0.5, 1.5, 2.5, 5.0, 9.0, 11.0, 16.3 inches of water. After run number 8 the slumped bed height was 15.0 inches. At the end of the test (15 runs total), the slumped bed height was 14.3 inches.

Visual Observation, V-1 and V-2

In order to identify leak areas, visual observations of the fluidized bed were made and a motion picture was taken before the air distributor was repaired. After completing Tests C-1A and C-1B the distributor was repaired and visual test V-1 was performed with 2, 4, 6 and finally 8" of bed depth to insure that the leaking areas had been properly repaired. After completing tests C-2A and C-2B, visual test V-2 was performed also with 2, 4, 6 and 8" of bed depth to insure that the repair was still in good condition.

C-2 Air Distributor Tests

Two tests were performed in order to characterize the air distributor. One was performed before the leaks were repaired, and the other followed completion of the repair work.

The purpose of these tests was to determine the nozzle equivalent of the leakage area. During these tests, distributor plate pressure drop was measured as a function of air flow rate with an empty bed. Temperature and pressure at various locations along the air flow stream were recorded. The data for these test was reduced and presented in the Data Reduction section of this Appendix.

C-3 Repair of Air Distributor Leaks

From the motion picture taken before the repair, four locations where the air flow was concentrated were observed. These locations are identified as leak areas and are shown in Figure C-3.1. A check into the history of the PDU operation revealed that four oil nozzles had been installed in these four areas for test burning of high sulfur oil. The oil nozzles were later removed and the holes left behind were patched up.

A close examination of the patched holes indicated that the patches were faulty as shown in Figure C-3.2 (a, b, and c). The patches were tack-welded onto the distributor plate leaving gaps around the perimeter. Air flowing through these gaps resulted in high localized concentration of air flow and resulted in formation of four jets leaving the plate from four different locations. Thus, these areas needed repair.

The repair was done by inserting a disc of the same thickness as the plate into the hole and solidly welding this disc to the plate. After welding, a hole was drilled through the center of the disc and a new nozzle was installed to the same height as the adjacent nozzles. Figure C-3.2(d) shows the finished repair. Same repair was made to all four leak areas. Figure C-3.3 shows the dimensions of a typical distributor nozzle.

FIGURE C-3.1: LOCATIONS OF LEAK AREAS ON THE AIR DISTRIBUTOR.

TOP VIEW OF FRONT LEFT LEAK-
AGE AREA ON DISTRIBUTOR PLATE

(a)

UNDERSIDE OF DISTRIBUTOR
PLATE SHOWING PREVIOUS
FAULTY REPAIR

(b)

LEAK REPAIRED AND NEW
NOZZLE INSTALLED

(d)

FRONT-LEFT LEAKAGE AREA

(c)

FIGURE C-3.2: Photograph Showing
One of the Leak Areas Before
(a,b,c) and After (d) Repair

NOZZLE DETAILS

6/28/81

TOTAL NOZZLES: 2712
 EXIT AREA: $25.31 \text{ IN}^2 + 2.37$
 $- 2.27$

HOLE SIZE: $0.109" \pm 0.005"$

FIGURE C-3.3: Nozzle Dimensions and Flow Areas

C-4 Data Reduction

C-4.1 Tests C-1A, C-1B, C-2A and C-2B

The data from Tests C-1A, C-1B, C-2A and C-2B are presented in Tables C-4.1 through C-4.4. The data reduction method and an explanation of the data are given below:

- Main air and coal feed air flow rates were determined from orifice pressure drops, upstream pressures and upstream temperatures for each test run.
- Total air flow rate is the sum of the main air and coal feed air flow rates.
- Fluidizing velocity is determined by the total air flow rate, the air density at the bed pressure and temperature, and the cross-section area of the bed.
- Bed pressure is the mean absolute pressure of the bed determined by adding the gauge pressure 1.5" above the grid plate to the barometric pressure and then subtracting one-half of the pressure drop across the bed.
- Bed temperature is determined as the average value of readings from two thermocouples installed in and below the in-bed bundle.
- Total pressure across the bed and the expanded bed height are determined from plots of the vertical bed pressure profiles.
- Slumped bed height was determined using a measuring stick while the bed was completely slumped.
- Particle densities and bulk density were determined for samples from the indicated test runs.
- Surface mean particle diameter was calculated from size distribution data for samples from the indicated test runs.
- Expanded bed height was determined graphically by plotting the PDU pressure readings versus height from the tip of the nozzles. And the bed level was the height beyond which pressure no longer changes.
- Minimum fluidization velocity was determined by plotting total bed pressure drop (excluding air distributor

pressure drop) versus fluidizing velocity. The minimum fluidization velocity is the velocity beyond which bed pressure drop no longer changes.

C-4.2 Air Distributor Pressure Drop Tests

The data for these tests are presented in Tables C-4.5 and C-4.6. The data reduction method and an explanation of the listed data are given below.

- Air Flow rate is based on the main air only and is calculated from orifice pressure drop, upstream temperature and pressure.
- Plenum temperature was read off the Doric recorder. Plenum pressure was the sum of barometric pressure and the logged plenum gauge pressure. Pressure data listed have been corrected for the density of the fluid in the manometers.
- Grid ΔP was read as the pressure difference between the plenum and the pressure tap 3.5" above the distributor nozzles.
- Bed temperature is in fact the downstream temperature of air flowing through the air distributor. No bed material was present.
- $W \sqrt{T_0} / (P_1 A_1)$ is flow parameter. W is the mass flow rate of air, T_0 is the air temperature before the distributor, P_1 is the bed absolute pressure and A_1 is the nozzle flow area.
- $\Delta P / P_1$ is the ratio of air distributor pressure drop over the absolute pressure of the bed (empty).
- W / A is the superficial mass velocity of air and is equal to the air mass flow rate divided by the bed cross-sectional area.
- K is the dimensionless constant that characterizes the flow of air through the distributor. It is defined by:

$$K = \frac{W/A}{\sqrt{\Delta P \rho}}$$

where ρ is the air density at plenum conditions.

TABLE C-4.1

REDUCED DATA FOR TEST C-1A (Before Repair)

Run No.	Main Air Flow Rate (lbs/hr)	Coal Feed Air Flow Rate (lbs/hr)	Total Air Flow Rate (lbs/hr)	Fluidizing Velocity (ft/s)	Bed					Physical Properties of Bed Material			
					Press. (atm.)	Temp. (°F)	ΔP Total (in. H ₂ O)	Expanded Height (in.)	Slumped Height (in.)	Particle Density		Bulk Density (g/cc)	Surface Mean Particle Diameter (microns)
										True (g/cc)	Apparent (g/cc)		
1	11483.4	472.0	11955.4	4.18	1.0038	113.4	18.3	16.5	15	-	-	-	1124.8
2	10464.0	450.5	10914.5	3.84	1.0085	120.6	17.4	16.3	-	2.53	2.05	1.35	-
3	9345.3	450.0	9795.3	3.45	1.0095	121.4	16.6	16.1	-	-	-	-	-
4	8071.0	444.2	8515.2	3.00	1.0100	122.4	16.2	15.7	-	-	-	-	-
5	6579.5	437.4	7016.9	2.46	1.0189	123.8	15.0	15.5	-	-	-	-	-
6	4628.5	436.4	5064.9	1.80	1.0085	126.6	14.4	15.1	-	-	-	1.37	-
7	4310.1	443.8	4753.9	1.73	0.9910	129.0	15.2	14.9	-	-	-	-	-
8	3264.9	443.4	3708.3	1.36	0.9897	131.2	12.7	15.1	-	-	-	-	-
9	2300.9	443.3	2744.2	1.00	0.9918	132.0	8.0	15.2	-	-	-	-	-
10	1452.6	439.5	1892.1	-	-	134.6	-	-	-	-	-	-	-
11	2294.8	452.4	2747.2	1.00	0.9924	128.4	6.5	15.8	14.5	-	-	-	-
12	3246.1	440.3	3686.4	1.34	0.9907	130.2	10.9	15.0	-	-	-	1.39	-
13	4607.3	440.6	5047.9	1.84	0.9896	129.2	15.8	14.2	-	-	-	-	-
14	6539.9	443.5	6983.4	2.49	1.0074	127.8	14.3	15.3	-	-	-	-	-
15	10375.3	446.7	10822.0	3.87	1.0077	128.9	15.6	15.8	-	-	-	-	1208.6
16	12774.8	438.6	13213.4	4.73	1.0073	129.6	15.9	15.4	-	-	-	-	-
17	13546.5	437.9	13984.4	4.99	1.0099	129.4	16.3	15.4	14.0	2.50	1.93	1.36	-

The minimum fluidization velocity (U_{mf}) is 1.64 ft/s.

TABLE C-4.2
REDUCED DATA FOR TEST C-1B (Before Repair)

Run No.	Main Air Flow Rate (lbs/hr)	Coal Feed Air Flow Rate (lbs/hr)	Total Air Flow Rate (lbs/hr)	Fluidizing Velocity (ft/s)	Bed					Physical Properties of Bed Material			
					Press. (atm.)	Temp. (°F)	ΔP Total (in. H ₂ O)	Expanded Height (in.)	Slumped Height (in.)	Particle Density		Bulk Density (g/cc)	Surface Mean Particle Diameter (microns)
										True (g/cc)	Apparent (g/cc)		
1	11609.6	463.0	12072.6	4.28	1.0135	127.9	26.0	25.5	20.50	-	-	-	1196.9
2	10130.7	451.4	10582.1	3.76	1.0134	127.9	26.1	26.0	-	2.46	2.00	1.36	-
3	8673.2	449.1	9122.3	3.08	1.0148	129.8	25.0	24.5	-	-	-	-	-
4	6540.2	444.3	6984.5	2.33	1.0150	131.0	24.8	23.5	-	-	-	-	-
5	4495.4	443.9	4939.3	1.61	1.0141	134.8	23.5	20.0	-	-	-	1.42	-
6	3255.2	444.3	3699.5	1.18	1.0059	136.2	18.2	19.0	-	-	-	-	-
7	2297.0	444.2	2741.2	0.85	0.9958	144.4	14.4	19.6	-	-	-	-	-
8	3979.9	444.0	4423.9	1.45	1.0120	141.6	21.2	19.8	-	-	-	-	-
9	5654.6	443.8	6098.4	2.04	1.0145	138.3	23.2	20.2	-	-	-	1.35	-
10	7734.8	443.5	8178.3	2.79	1.0120	136.2	25.2	24.0	-	-	-	-	-
11	9628.9	443.1	10072.0	3.46	1.0133	134.8	24.2	24.0	-	-	-	-	-
12	11110.4	443.4	11553.8	3.97	1.0129	132.2	24.5	25.9	-	-	-	-	1139.1
13	12034.6	443.6	12478.2	4.30	1.0123	131.0	25.0	26.4	19.75	2.49	1.99	1.41	-

The minimum fluidization velocity (U_{mf}) is 1.42 ft/s.

TABLE C-4.3

REDUCED DATA FOR TEST C-2A (After Repair)

Run No.	Main Air Flow Rate (lbs/hr)	Coal Feed Air Flow Rate (lbs/hr)	Total Air Flow Rate (lbs/hr)	Fluidizing Velocity (ft/s)	Bed					Physical Properties of Bed Material			
					Press. (atm.)	Temp. (°F)	ΔP Total (in. H ₂ O)	Expanded Height (in.)	Slumped Height (in.)	Particle Density		Bulk Density (g/cc)	Surface Mean Particle Diameter (microns)
										True (g/cc)	Apparent (g/cc)		
1	10971.5	448.3	11419.8	3.96	1.0175	116.4	25.2	24.8	21.5	-	-	-	-
2	9356.6	444.6	9801.2	3.40	1.0160	116.4	25.4	25.3	-	-	-	-	1009.2
3	7366.5	434.9	7801.4	2.71	1.0170	117.1	24.6	22.2	-	-	-	-	-
4	5695.6	434.5	6130.1	2.13	1.0150	117.4	24.2	20.0	-	-	-	-	-
5	3276.5	438.4	3714.9	1.31	1.0057	118.8	17.9	20.7	-	-	-	-	-
6	2311.1	438.8	2749.9	0.98	0.9966	120.8	13.2	20.3	20.5	2.63	1.95	1.44	-
7	2312.9	450.4	2763.3	0.98	0.9989	121.4	13.7	20.0	-	-	-	-	1380.1
8	4013.0	447.8	4460.8	1.57	1.0109	121.4	19.5	19.9	-	-	-	-	-
9	5186.5	448.0	5634.5	1.98	1.0130	121.2	23.8	19.9	-	-	-	-	-
10	5689.5	442.6	6132.1	2.15	1.0135	120.9	23.4	20.0	-	-	-	-	-
11	8405.5	435.3	8840.8	3.10	1.0136	120.0	23.8	25.5	-	2.53	2.03	1.40	-
12	10439.2	438.4	10877.6	3.81	1.0117	119.7	24.9	24.8	-	-	-	-	-
13	11680.6	433.4	12114.0	4.24	1.0128	119.1	24.0	25.1	19.5	-	-	-	-

The minimum fluidization velocity (U_{mf}) is 1.90 ft/s.

TABLE C-4.4
REDUCED DATA FOR TEST C-2B (After Repair)

Run No.	Main Air Flow Rate (lbs/hr)	Coal Feed Air Flow Rate (lbs/hr)	Total Air Flow Rate (lbs/hr)	Fluidizing Velocity (ft/s)	Bed					Physical Properties of Bed Material			
					Press. (atm.)	Temp. (°F)	ΔP Total (in. H ₂ O)	Expanded Height (in.)	Slumped Height (in.)	Particle Density		Bulk Density (g/cc)	Surface Mean Particle Diameter (microns)
										True (g/cc)	Apparent (g/cc)		
1	12969.8	453.4	13423.2	4.68	1.0079	114.8	18.0	17.0	15.00	-	-	-	1037.5
2	11735.7	445.3	12181.0	4.25	1.0093	115.6	17.8	17.2	-	2.56	2.08	1.39	-
3	9947.1	439.8	10386.9	3.59	1.0109	115.9	16.5	16.8	-	-	-	-	-
4	8768.6	438.7	9207.3	3.21	1.0104	116.2	16.9	16.0	-	-	-	-	-
5	6434.1	437.9	6872.0	2.40	1.0103	117.2	16.0	15.4	-	-	-	-	-
6	4661.9	437.8	5099.7	1.79	1.0084	118.0	15.6	15.0	-	-	-	-	-
7	3289.1	438.0	3727.1	1.32	1.0007	119.6	10.8	15.1	-	-	-	-	-
8	2321.4	438.1	2759.5	0.99	0.9915	121.5	7.3	15.1	15.00	-	-	-	-
9	2340.1	458.1	2798.2	0.99	0.9934	114.3	7.8	15.0	-	-	-	-	1344.7
10	4037.3	449.7	4487.0	1.57	1.0063	116.6	12.8	15.2	-	-	-	-	-
11	5217.4	445.8	5663.2	1.98	1.0081	117.0	15.8	14.8	-	2.54	1.64	1.40	-
12	7390.9	443.4	7834.3	2.74	1.0085	116.5	15.5	15.0	-	-	-	-	-
13	9945.4	440.7	10386.1	3.63	1.0098	116.6	15.9	16.0	-	-	-	-	-
14	11011.3	439.5	11450.8	4.00	1.0097	116.7	16.0	16.0	-	-	-	-	-
15	13447.3	438.4	13885.7	4.93	1.0070	116.2	16.7	16.1	14.25	-	-	-	-

The minimum fluidization velocity (U_{mf}) is 1.96 ft/s.

TABLE C-4.5

REDUCED DATA FOR GRID PLATE TEST
(Before Plate Repair)

Run No.	Air Flow Rate (lbs/h)	Plenum		Grid ΔP (in. H ₂ O)	Bed Temp. (°F)	$\frac{W}{P_1} \frac{\sqrt{T_0}}{A_1}^*$ $\left(\frac{\text{lbm}\sqrt{R}}{\text{lb}_f \text{S}^2}\right)$	$\frac{\Delta P^*}{P_1}$
		Temp. (°F)	Pressure (in. H ₂ O abs.)				
1	-	-	-	-	-	-	-
2	4584.7	110.8	404.1	2.7	102.8	0.0823	0.0067
3	6118.6	111.8	407.1	4.5	105.0	0.1097	0.0112
4	7302.8	111.7	409.1	6.4	106.6	0.1308	0.0159
5	8754.5	111.2	411.6	8.9	106.9	0.1568	0.0221
6	9930.0	110.3	413.9	11.5	106.8	0.1778	0.0286
7	10980.9	110.0	416.4	13.9	107.0	0.1965	0.0345
8	11992.7	110.2	418.8	16.6	107.3	0.2149	0.0413
9	12813.8	110.2	421.1	18.8	107.5	0.2295	0.0467
10	15019.0	110.0	427.6	25.7	107.7	0.2692	0.0639

* Explanation in the text.

TABLE C-4.6

REDUCED DATA FOR GRID PLATE TEST
(After Plate Repair)

Run No.	Air Flow Rate (lbs/h)	Plenum		Grid ΔP (in. H ₂ O)	Bed Temp. (°F)	$\frac{W \sqrt{T_0}}{P_1 A_1}$ $\left(\frac{\text{lbm} \sqrt{R}}{\text{lb f s}}\right)$	$\frac{\Delta P^*}{P_1}$	$\frac{W}{A}$ (lb/h ft ²)	K*
		Temp. (°F)	Pressure, P ₀ (in. H ₂ O abs.)						
1	15896.5	108.5	433.4	30.5	98.4	0.2839	0.0757	1391	0.0198
2	14918.5	114.8	432.4	27.5	106.3	0.2666	0.0679	1305	0.0197
3	13193.8	115.8	425.9	21.5	108.4	0.2362	0.0532	1154	0.0199
4	11840.2	116.9	422.4	17.5	109.5	0.2119	0.0432	1036	0.0199
5	10921.2	117.7	419.4	15.0	109.9	0.1959	0.0371	955	0.0199
6	10136.3	118.5	416.9	13.0	110.1	0.1821	0.0322	887	0.0199
7	8820.5	118.5	413.9	10.0	109.5	0.1585	0.0248	772	0.0198
8	7196.7	118.5	410.7	6.7	109.2	0.1293	0.0166	630	0.0198
9	5680.7	118.3	408.9	4.5	109.2	0.1019	0.0111	497	0.0191
10	3263.8	118.4	402.9	1.75	108.9	0.0590	0.0044	286	0.0178
11	4729.8	118.9	405.9	3.0	108.2	0.0852	0.0074	414	0.0196
12	6546.1	120.0	410.4	5.5	108.9	0.1175	0.0136	573	0.0199
13	8102.2	120.3	411.4	8.0	109.8	0.1460	0.0198	709	0.0204
14	9288.9	120.0	414.4	10.5	110.0	0.1671	0.0260	813	0.0204
15	10350.1	119.7	416.4	13.0	109.9	0.1864	0.0322	906	0.0204
16	11401.0	119.5	418.9	16.0	109.7	0.2056	0.0397	997	0.0201
17	12584.2	119.3	422.4	20.0	109.7	0.2271	0.0497	1101	0.0198
18	13978.4	118.8	426.4	24.0	109.7	0.2522	0.0596	1223	0.0200
19	15169.6	118.6	430.9	28.5	109.6	0.2736	0.0708	1327	0.0198
20	15777.4	118.8	432.9	30.5	109.7	0.2846	0.0758	1380	0.0199

* Explanation in the Text.

C-5 Discussion of Results

C-5.1 Effects of Air Distributor Leakage

Effects of air distributor leakage on the expanded bed height and minimum fluidizing velocity are presented here. Effects on the gas mixing are also of interest but could not be studied because of the limited scope that can be afforded for these unplanned activities.

The data obtained for the tests are presented in the Data Reduction of this Appendix. Expanded bed height for Tests C-1A and C-1B are plotted versus fluidizing velocity in Figure C-5.1 for shallow (nominal 15" slumped height) and normal (nominal 21" slumped height) beds. The horizontal portions of the curves are extended at the slumped bed height from zero velocity to the minimum fluidization velocity point and are extended, as best-fit lines, through the remainder of experimental data points. Note that the minimum fluidization velocity has been determined separately as described in the Data Reduction section of this Appendix.

Figure C-5.2 shows an overlay of Figure C-5.1 onto the expanded bed height results obtained for Tests C-2A and C-2B. No curves are drawn through the C-2 data points to avoid possible bias in comparisons.

In examining Figure C-5.2, one can see that there might be discrepancies between the results of Tests C-1 and C-2. However, the scattering of data points in both types of tests is generally larger than any identifiable discrepancy. Discrepancies if found are therefore unwarranted. In other words, we can say that the effects represented by the discrepancies of air distributor leakage on the expanded bed height and minimum fluidization velocity are undetectable.

C-5.2 Air Distributor Pressure Drop

Pressure drop data for the air distributor before and after the leakage repair are presented in the Data Reduction section of this Appendix. In general, pressure drop of gas flowing through a nozzle is a function of the Mach number, which, in turn is a function of the flow parameter:

$$\frac{W \sqrt{T_0}}{P_1 A_1}$$

Where W is mass flow rate, T_0 upstream temperature, P_1 downstream pressure and A_1 flow area. If flow is treated as

FIGURE C-5.1: Bed Level Versus Fluidizing Velocity for Tests Before Distributor Repair

FIGURE C-5.2: Comparison of Test Results Between Before and After Distributor Repair.

isentropic, one-dimensional flow, a mathematical relationship can be derived. This relationship represents the theoretical conditions. In practical situations, friction and other losses cause deviations from such relationships.

Data for air distributor pressure drop are plotted against the flow parameter mentioned above in Figure C-5.3. The deviations from theoretical curve of the data are apparent and expected. The discrepancies between before and after repair curves indicate that there was unaccounted flow area before the repair. To bring the before-repair curve close to the after-repair curve, area A_1 for the before-repair data must be increased. Such increase is the nozzle-equivalent area of the leaks.

Since the before and after-repair curves are not exactly proportional, there is a lower and upper limits for the increase in A_1 that would bring the two curves together. It was determined that this increase is from 0.9 to 1.1 in.² or 3.6 to 4.2% of the total nozzle flow area.

In designing air distributor, the following equation is generally of interests:

$$\frac{W}{A} = K \sqrt{\Delta P \rho} \quad (C-5.1)$$

Where W/A is the superficial mass velocity of air based on the bed cross-sectional area, ΔP is the pressure drop across the distributor, ρ is the upstream air density, and K is the flow coefficient. Data for W/A , ΔP and ρ are presented in Table C-4.6. Calculated value of K is plotted against W/A in Figure C-5.4.

It is apparent from Figure C-5.4 that K varies with W/A but seems to have a value of about 0.020 for a W/A above 700 lb/h.ft².

FIGURE C-5.3: Pressure Drop vs. Flow Parameter for the "Drilled-Bolt" Air Distributor Before and After Repair.

K

FIGURE C-5.4: K-Value for the "Drilled-Bolt" Air Distributor as the Function of Air Mass Velocity Based on Bed Cross-Sectional Area.

C-6 Conclusions

For conclusions see section 5.2.5 of this report entitled "Conclusions of Air Distributor Leak Tests".