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INDUSTRIAL APPLICATION OF FLUIDIZED BED COMBUSTION
PHASE I, TASK 4—SUB-SCALE UNIT TESTING AND DATA ANALYSIS

Final Report. Volume 1

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U. S. DEPARTMENT OF ENERGY

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INDUSTRIAL APPLICATION OF
FLUIDIZED BED COMBUSTION
PHASE I, TASK 4 - SUB-SCALE
UNIT TESTING AND DATA ANALYSIS
FINAL REPORT
VOLUME I

PREPARED BY

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WINDSOR, CONNECTICUT

FOR

U. S. DEPARTMENT OF ENERGY
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DIVISION OF FOSSIL FUEL UTILIZATION
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SECTION 1 SUMMARY

1.1 Introduction

Fluidized Bed Combustion (FBC) offers industry an efficient and environmentally sound technology for utilizing high sulfur coal in generating steam. The U.S. Department of Energy has encouraged the commercialization of this technology so that high sulfur coal will become a practical fuel in widespread use for industrial applications. Combustion Engineering, under contract with the Department of Energy, has developed, designed, and is constructing a 50,000 lbs steam/hr Industrial FBC Demonstration Plant. The plant will provide steam for space heating at the Great Lakes Naval Base in North Chicago, Illinois. Its operation will enable industry to objectively appraise the performance, reliability, and economics of FBC technology.

A hot sub-scale unit (SSU), simulating the operating conditions of the demonstration plant, has been constructed and operated at Combustion Engineering's Kreisinger Development Laboratory (KDL) in Windsor, Connecticut. The SSU facility has served as a valuable developmental tool in establishing the performance characteristics of the FBC process and equipment as used in the larger Demonstration Plant. Experience gained during more than 2000 hours of operation, including the analytical results derived from an extensive test program of 1500 hours operation, has defined problems and identified solutions in engineering the larger FBC Demonstration Plant. This report presents documentation of the results of the SSU test program.

The SSU is a fully instrumented, operational plant designed to similar operating conditions and utilizing the same coal and limestone sorbent as selected for the FBC Demonstration Plant. As such, it fills a necessary step in the development of FBC technology from theory to practice. In supporting the design of the Demonstration Plant the SSU satisfied the following program objectives:

- Verification of FBC process design process design procedures and the suitability of water side cooling by natural circulation.
- Quantification of performance data for the coal and limestone sorbent selected for the Demonstration Plant.
- Identification of anticipated problems in the operation of process equipment components similar to those used in the Demonstration Plant design.

Beyond satisfying the direct needs of the Demonstration Plant development and design, the results serve to broaden the general knowledge of FBC technology. The specific impact of the test results on the demonstration plant has included boiler modifications based on revised design criteria for heat transfer rates, heat release distribution, and changes in the distribution of material flows. Overall, the data indicated that the larger demonstration plant will efficiently generate steam from the selected high-sulfur coal while effectively controlling the emission of environmental pollutants.

The SSU, with its extensive analytical capabilities, will be available during the Demonstration Plant operational test programs which will begin in the first quarter of 1981. It will serve as an experimental facility to resolve problems that may surface in the Demonstration Plant operation.

Small differences in performance between the large scale and the sub-scale are expected. Understanding these differences will increase the ability to predict large scale FBC behavior from controlled conditions of a small scale unit. The unit provides the opportunity to test design modifications to boiler components and adjustments of operating conditions before committing to the expense of modifying the large scale Demonstration Plant.

The design of the Sub-Scale Unit began immediately after the award of the Demonstration Project contract with DOE in June, 1976. Designed in parallel with the preliminary engineering of the Demonstration Plant, the SSU configuration incorporated features to test the assumptions used to establish the preliminary design basis of the Demonstration Plant. Construction of the SSU facility began at KDL in June, 1977. The facility was ready for preoperational checks of systems and components in February, 1978.

1.3 Test Facility Description

The Sub-Scale Unit test facility consists of a fluidized bed boiler with all auxiliary systems needed for generating superheater steam. Figure 1 is a schematic of the entire unit. The fluidized bed itself has a square cross section of 26 inches and a nominal expanded depth of three feet. Design full load is achieved at 7 feet per second superficial gas velocity with a coal feed rate of 230 pounds per hour. Typical sizing for the coal is 1/4" x 0

C-E-DOE SSU FLUIDIZED BED STEAM GENERATOR

1-4

while the limestone, purchased pre-screened, is 1/8" x 40 mesh. Further details of the facility are discussed below.

1.3.1 Boiler

The fluidized bed boiler is the central part of the facility. Fin welded waterwalls, using 2 inch diameter tubes on 4 inch centers, form the furnace enclosure. One wall is removable to allow access to the interior. Natural circulation is used for both the wall tubes and the heat transfer surface immersed in the fluidized bed. A boilerwater circulating pump was installed although it was operated only during initial shakedown operations. A superheater and an economizer are located in the upper section of the furnace. A photograph of the boiler, Figure 2, with the wall removed shows the boiler interior.

The air distributor plate is 1/8 inch thick, 310 stainless steel with a drilled pattern of 0.1 inch holes. A corrugated stainless steel-asbestos composite gasket is used to form a positive seal around the edges. Although some minor warpage has occurred, this arrangement has retained its integrity for 1500 hours of operation.

The waterwalls in the bed and lower freeboard region are covered with refractory to prevent excessive heat loss from the bed that occurs in small scale units because of their large surface to volume ratio. The resulting bed area is 4.7 square feet. Heat is extracted from the bed by a series of immersed tubes. Three tubes are part of the steam generating circuitry of the boiler. Although a double loop is shown in Figure 2, this was modified to a single loop arrangement. Two additional tubes are supplied with water directly from the feedpump and do not generate steam. These tubes are individually instrumented to permit calculation of heat transfer coefficients from the bed to the tube.

1.3.2 Combustion Air System

Combustion air is supplied by a forced draft blower. The air passes through a control valve and into a tubular air heater. The air is then ducted through a flow measurement venturi and into the boiler plenum. An automatic controller maintains a set air flowrate.

1.3.3 Flue Gas System

Flue gases leaving the boiler are given a primary cleaning in a cyclone to remove elutriated solids. The cyclone consists of three parallel tubes with slide dampers to allow operation with one, two, or three tubes in service. The gases then pass through the tubular air heater and into a bag filter for final cleanup. An induced draft fan and control valve complete the system. An automatic controller maintains the freeboard pressure slightly below atmospheric pressure.

1.3.4 Coal/Limestone Feed System

The coal/limestone feed system consists of equipment to size, store, and meter fuel and sorbent to the boiler. Raw material is loaded into a receiving hopper and transported to a screener. The screener separates over sized material and directs it to a crusher. Properly sized material is collected in live bottom bins. Two gravimetric feeders meter the coal and limestone. The composite stream drops through a rotary valve and is transported pneumatically through a 1-1/2 inch pipe to the boiler. Transport air is supplied from the FD fan discharge. The feed needle extends downward into the bed ending about 10 inches above the center of the distributor plate.

1.3.5 Bed Drain System

Bed material is drained through a penetration in the side of the boiler to maintain a controlled active bed level. The material drops into a water cooled screw where it is cooled from bed temperature down to about 150°F. The material then goes through a rotary air lock and into a collection hopper.

1.3.6 Recycle System

Particulate material collected by the cyclone is fed back into the bed for additional carbon burn up. A rotary air lock is used to seal the bottom of the cyclone. Material is then pneumatically transported to the bed through a penetration in the lower corner of the unit.

A section of the recycle line was water jacketed. Measurements of the solids inlet and outlet temperatures and the water flow and temperatures allowed calculation of the recycle flow rate by heat balance. Calibration of the system was performed by flowing hot material through the system and collecting it in a weighted drum. This check showed good agreement between measured and calculated solids flow.

1.3.7 Start-up Burners

Two natural gas fired burners are used to heat the bed to over 1000°F for start-up. One burner fires into the combustion air duct to preheat the fluidizing air. The other burner generates a stream of hot gases which are ducted into the bed above the air distributor. Although the burners have the capacity to heat up the bed fairly quickly, start-ups were performed over a six to eight hour period to avoid cracking the refractory lining the bed area.

1.3.8 Instrumentation and Controls

Measurements of more than 80 temperatures, pressures, flows, and composition are made throughout the system. These measurements are scanned and converted to digital form by a data logger. The data logger in turn is controlled by micro-computer. The micro-computer performs two functions. First, it converts measured voltages to reduced engineering units, eg flows in pounds per hour or pressures in pounds per square inch, etc. The results are immediately available as an aid in unit operation. Second, it manages data recording activities. In most cases, data scans were performed at one minute intervals. Five of these consecutive scans are then averaged to smooth out short term fluctuations and the average values logged for later analysis. The micro-computer will also compare consecutive averages and provide a separate log of values which deviated by more than a preset margin from the initial values. This was done to insure that steady state test periods were truly steady state.

Automatic controls are installed to hold a number of parameters at constant conditions. The parameters are: coal feedrate, limestone feedrate, combustion air flow, combustion air temperature, furnace pressure, drum pressure, and drum water level.

Initial operation of the SSU during shakedown revealed several problems which limited performance. The valuable experience gained in solving these problems led to a clearer understanding of FBC plant design requirements. Major modifications of the fluidized bed air distributor and boiler heat transfer surface were required. As a result of these corrections, bed temperatures and combustion efficiencies were improved to meet the criteria set for starting the formal test program.

In its original configuration, the air distributor for fluidizing the bed consisted of three separate smaller plates designed to allow for easy installation and removal. However, restrictions in the plate edge seals and stiffening structure caused distortion and buckling of the plate, thus leading to air leakage at seals between plate sections and plate edges. With this leakage, the unit was unable to sustain uniform combustion. The air distributor was replaced with a one-piece plate with a stiffening structure which provided less restriction of thermal expansion. At the plate edges, effective sealing was achieved through the use of a composite corrugated stainless steel asbestos gasket which also enabled the full thermal expansion of the plate.

The redesigned plate has shown minor distortion, but is still servicable after 1500 hours of operation. After its installation, stable and uniform combustion in the SSU was possible. Data gained from the operation of the air distributor have provided detailed information for the development of an improved configuration used in the Demonstration Plant.

While the SSU performance was greatly improved following the replacement of the air distributor, bed temperatures were still below the acceptable range of 1500-1600°F. Data gained from

further shakedown testing defined the problem as requiring changes in heat transfer surface. Heat transfer rates in the bed was approximately 50 percent higher than design values and the heat release in the freeboard was about double the design value of ten percent. Combustion efficiency was limited to 85 percent at best. Heat balances for the SSU were recalculated using the measured values of heat transfer coefficients and heat release distribution. Based on the results, bed heat transfer surface was reduced and refractory was extended higher on the waterwalls.

Shakedown testing was completed in January, 1979. All shakedown performance criteria were met. The SSU achieved better than 90 percent combustion efficiency, bed temperatures in the range of 1500-1600°F, expanded bed depth of 36 inches and acceptable control of pollutant emissions. The formal test matrix, planned in an earlier program task, was started in January and completed in May, 1979.

The various process equipment components of the SSU performed well from a mechanical standpoint during the 2000 hours of operation accumulated in shakedown and testing. Other minor problems were encountered which, while slowing progress, did not significantly affect performance. Early difficulties with poor collection efficiency in the bag filter were attributed to bags torn during installation. After replacement of these bags, no further maintenance was necessary. The recycle rotary airlock was a problem in that close clearances, set to prevent excessive gas leakage at high temperatures, would increase when operating at lower temperatures and gas flow would prevent material flow. Setting the clearance for low temperature operation resulted in binding of the rotor at high temperature. This problem was handled by grouping test points together with similar anticipated recycle conditions and setting clearances appropriately. The coal/limestone handling system could not handle excessively wet materials. This

required air drying the coal and limestone at times. Coal can be successfully handled and fed at total moisture levels up to 12 percent. Bed pressure taps, originally 3/8 inch tubing entering horizontally, would plug frequently and were modified to 3/4 inch pipe sloping downward at 45°. These taps were provided with a periodic back purge of air and performed acceptably.

The SSU was instrumented to provide mass and energy balances, zonal heat absorption and heat transfer data. The exhaust gas stream was monitored to evaluate gaseous and particulate pollution emissions.

All solid waste streams were sampled for constituent analyses, carbon loss, and bed particle size. Coal and limestone feed streams were analyzed on a regular basis to provide sensible, and chemical energy.

Two data reduction and analyses programs were developed and executed on the CDC Cyber 172 Computer. One program performed data conversion, overall mass and energy balances, and results retrieval. Manual input of solid stream analyses is required. Results are based on five minute averaging of raw data scans. The second program performs zonal energy balances on the bed, freeboard and convective sections of the SSU.

The most significant performance results which have a major bearing on large size unit design are discussed below.

1.5.1 Bed Expansion

The expanded bed height is one of the few controllable parameters available to the fluid bed operator. The expanded bed height is normally set so that the in bed tubes are fully covered, which assures the maximum utilization of the submerged tube surface. During operation, the expanded bed height can be adjusted through changes in limestone feed rate and bed drain rate.

Bed expansion data is used during the design phase to determine the expected pressure drop of a bed needed to cover a given tube

bundle design when the bed is in operation. This information is required in selecting fans for the unit. Further it is customary to set the air distribution design pressure drop in proportion to the bed pressure drop.

In the SSU, the expanded bed height was determined using a series of four static pressure taps on four inch vertical spacing, beginning eight inches above the air distributor. The intersection of a best fit pressure-elevation line with the freeboard pressure was considered the expanded bed height. Static bed level was determined from overall bed pressure drop measurements with particle density values determined from evaluation of the limestone utilization.

Bed expansion is primarily due to bubble gas holdup. A measure of bubble voidage is given by the excess gas velocity ratio. Figure 3 shows a plot of bed expansion ratio as determined from active bed height and bed pressure drop measurements versus excess gas velocity ratio. The low bed expansion even at high velocity ratios is a result of the small bed particle sizes obtained with this limestone. The harmonic mean particle size ranged from 400 to 750 microns. The particle size infers that the bed is in the fast bubble regime of small bubble size and in which larger amounts of gas are accommodated in the bed at low voidage, hence the low bed expansion.

1.5.2 Heat Transfer

One of the most important advantages of fluidized bed combustion is the high heat transfer coefficient from the bed to an immersed surface. The ability to accurately predict heat transfer coefficients is essential in designing commercial units. A large amount of literature is available on the prediction of heat transfer in fluidized beds. However, caution is required in

applying these analyses to coal-fired fluidized beds, since most analyses are based on bed operating at low temperatures (less than 500°F). In addition, most of the work has been performed on relatively small bench scale beds.

The total heat transfer to the immersed bed tubes is composed of a direct radiation component and a combined solids-gas convection component.

Direct radiation is a function of bed and tube surface temperatures. The convective heat transfer component in a fluidized bed has been shown to depend on the thermophysical properties of the fluidizing gas and bed solids, bed temperature and to a lesser extent on the bed particle size.

The SSU data was compared with a number of existing correlations. The data did not appear to correlate with these empirical equations, which include Reynolds number as a correlating parameter. Figure 4 is a curve of the Nusselt Number vs. the Archimedes Number representing SSU the data.

The curve in Figure 4 is the line representing the following correlation:

$$Nu = \frac{hcdp}{kg} = .31 (AR)^{.27} (Pr)^{.33} \quad (1)$$

where

dp = Harmonic mean particle diameter, $ft = \frac{1}{\sum x_i dp_i}$

g = Gravitational constant, ft/hr^2

hc = Total overall heat transfer coefficient (excluding radiation) $BTU/hr\ ft^2\ ^\circ F$

Nu = Nusselt Number

Pr = Prandtl Number

Ar = Archimedes Number

The correlation of Equation 1 was obtained using multiple regression analysis and has a correlation coefficient of 0.87. It is evident that the present data can be correlated quite well in a form as shown in Equation 1 suggesting Archimedes Number as a correlating parameter.

Equation 1 shows that the overall convective heat transfer coefficient is a weak function of particle size. Moreover, Equation 1 does not show any dependence of the heat transfer coefficient upon the superficial gas velocity. The functional relationship expressed by the equation is also independent of the type of technique used to estimate the mean particle diameter. For comparison purpose, B&W data for a 36 square foot bed are also plotted in Figure 4.

The data shows an eight percent reduction in the convective heat transfer component over a three-fold range of superficial velocity. This is due to two competing effects. Increase in gas velocity increases particle mixing and the faster the particles move, the less their temperature drops below the bulk bed temperature while in the vicinity of the heat transfer surface. This results in increased heat transfer to the immersed surface. At the same time, as gas velocity is increased, the bed voidage increases resulting in less solids contact and, less heat being transferred to the immersed surface, thus decreasing the heat transfer coefficient. At low velocities, the effect of increased particle mixing far outweighs the effect of increasing voidage. The effects of increasing voidage becomes predominant with increases in gas velocity. The net result is effectively a constant heat transfer coefficient over a relatively large range of velocities.

1.5.3 Combustion Efficiency

One of the main objectives of the SSU test program was to develop an understanding of the interrelationship of critical operating variables affecting combustion efficiency. The factors that influence the overall combustion efficiency are: (1) loss of carbon in the elutriated solids (governed by bed carbon loading, bed particle size and superficial gas velocity); (2) loss of carbon in the overflow stream (which is determined by bed carbon loading and inerts feed rate); and (3) loss of carbon as CO due to incomplete combustion.

The overall combustion efficiency in the SSU was determined from a series of tests in which excess air, bed temperature, and active bed height were varied over the range of expected demonstration unit operating conditions. As can be seen from Figure 5 the overall combustion efficiency remains constant with increase in excess air level ranging from -10% to +15%. This is in contrast to the experimental results reported in the literature which shows an increase in overall combustion efficiency with excess air level. The combustible loss for each run was estimated from the carbon loss due to CO in the exit flue gas, and the elutriated combustible loss due to carryover of unburnt carbon. Total combustible loss is expressed as a weight percentage of the total carbon in the feed. In general, if the particle combustion is purely diffusion-limited, then the overall combustion efficiency should show little or no temperature dependence.

Results from the test series show virtually no dependence of overall combustion efficiency on bed temperature over a range from 1400F to 1650F. This is indicative of the diffusion controlled char combustion occurring in the SSU combustor.

Figure 5

Since coal combustion is basically a chemical reaction involving oxygen transport, then the gas residence time or bed height should be an important parameter. It may be inferred that improved overall combustion efficiency should be achieved by resorting to increasing bed heights. The SSU data shows no significant improvement in overall combustion efficiency with increasing bed height over a threefold range of height.

Recycle operation results in a 3% to 4% increase in the overall combustion efficiency value over that for no recycle runs. These results show that the average value of overall combustion efficiency without recycle is about 93% whereas that for recycle runs is about 96%.

In order to investigate the influence of different coals on the performance of SSU fluidized bed combustor, two recycle runs were carried out using Old Ben coal. Under the SSU operating conditions used in this investigation, Old Ben coal is comparatively less reactive than the Freeman coal. The average value of overall combustion efficiency for two recycle runs with Old Ben coal is about 91% compared with about 96% for the Freeman coal. In addition, approximately 50% of the total input fuel energy is being released in the region above the active bed for two recycle runs with Old Ben coal.

Thermogravimetric and flammability tests were conducted on both Old Ben coal and Freeman coal. The flammability point for Old Ben coal was 200°F higher than for the Freeman coal. In addition, burnout time in the TGA Test for the Old Ben coal was about 15 percent longer. These results demonstrate that different coals can perform differently in a fluidized bed.

1.5.4 Combustion in Bed

In order to design a fluidized bed combustor it is essential to be able to predict the amount of in-bed heat release so that the heat transfer surface can be accurately designed for the heat release expected in each zone.

Conditions for the burnout of char in the fluidized bed zone are favorable because of the good solids mixing and long residence time. However, conditions for the combustion of volatile products in the bed may be less favorable than those for solids because of the potential for release near the top of the bed. Unburned CO volatile products and fine char particles may escape the bed and reduce the proportion of combustion in the bed. It is highly desirable to optimize the proportion of combustion occurring in the bed in order to take advantage of the high heat transfer rates attainable in the bed and at the same time utilize to the fullest, the above bed volume for convective heating.

The SSU furnace was divided into several zones: including the fluidized bed combustion zone or the active bed zone, the freeboard zone, and the superheater economizer zone.

In order to determine the relative proportion of combustion taking place in each zone, a detailed mass and energy balance calculation of individual zones was performed.

The percent of the total input fuel energy being released in the active bed zone as a function of excess air, active bed height, bed temperature, and superficial gas velocity was investigated in the SSU test program. The percent of the heat release occurring in the active bed zone is a weak function of these four parameters.

Figure 6 shows the percent heat released in the bed as a function of bed temperature. At the design bed temperature of 1550°F about 65 percent of the feed coal chemical energy is released in the bed under recycle operation. The small increase in heat release with increasing temperature further confirms the diffusion controlled bed combustion. Similar weak trends were seen with excess air and bed height.

The effect of gas residence time on bed performance, was also investigated. The percent heat released in the bed is more strongly dependent on the bubble residence time, L_f/U_b . The bed performance is quite dependent on the oxygen interchange rates between bubble and emulsion phases and oxygen transfer rate from emulsion phase to char particle surface.

In small particle systems, the bubble velocity is greater than the interstitial gas velocity in the dense phase and gas bypassing occurs because some of the gas recirculates inside the bubbles and does not contact the particles. The harmonic mean particle size of the bed material obtained from SSU test data varied from 400 μ to 750 μ and the superficial to fluidizing velocity ratio criteria suggests that the SSU is in the fast bubble regime.

Gas interchange ratios between bubble and emulsion phases can be interpreted as the number of times the bubble is replaced on passage through the bed. In order to obtain an estimate of the extent of oxygen bypassing occurring in the SSU fluid bed, value of the interphase gas exchange was calculated. These values of gas interchange ratio varied from 0.4 to 1.8 implying the inter-phase gas exchange rate is very low, hence the high level of oxygen bypassing of the bed. It is believed that high level of oxygen bypassing from the bed is one of the principal causes of low level of heat release in the bed.

Approximately 58% of the input air is being utilized in the bed implying that combustion in the bed is being limited by mass transfer considerations.

1.5.5 Freeboard Combustion

The SSU freeboard zone is defined as the region between the top of the bed and the superheater. The primary function of this zone is to allow the larger particles ejected from the bed to decelerate and fall back into the bed. Another important function of freeboard section is to provide additional gas/gas and or gas/solid contact so that combustion of CO, volatiles and char particles can take place.

The proportion of combustion in the freeboard is about one-third of that in the bed. Splashing of the large particles which are too large to be elutriated and which eventually fall back in the bed provide good solids mixing and hence help promote char combustion in the freeboard. With low levels of combustion in the SSU bed, considerable oxygen is available in the freeboard section to support combustion.

The experimental measurements of this investigation show that the average flue gas temperature in the freeboard is about the same as the bed temperature despite convective heat loss to the walls and radiation to the bed and superheater. This

result clearly indicates that freeboard combustion, of CO, char and volatile combustion and to a lesser extent SO_2 absorption and NO_x reduction are taking place. Figure 7 shows the percent heat released in the freeboard region as a function of excess air level. The combustion reactions also depends on freeboard temperature. The heat release in the freeboard region is approximately between 25 to 30% of the coal input energy. As can be seen, somewhat higher heat release is occurring above the bed for recycle runs indicating that additional heat release is due to combustion of some of the fine recycle char.

The extent of char combustion in the freeboard depends on the solids hold up in the freeboard. Solids hold up in turn depends on the amount of solids thrown up into the freeboard by the bursting bubbles at the bed surface.

However, the char elutriation rate from the SSU furnace is independent of superficial gas velocity implying that the extent of char combustion in the freeboard may be very nearly constant.

In order to estimate the contribution of volatiles/CO and recycle char combustion to the overall heat release in the freeboard region, it was assumed that the amount of coal feed carbon elutriated during recycle operation was the same as that for no recycle operation. Under steady state operating conditions, the recycle carbon content will remain constant only if an amount of carbon equivalent to the feed carbon being elutriated is burnt out from the recycle stream. With this assumption approximately 30-35% of the input volatiles and 20-25% of the recycle char must be combusted in the freeboard region in order to achieve the indicated freeboard heat release rates.

1.5.6 Environmental Considerations

The major environmental pollutants potentially discharging from the stack of a fluidized bed boiler 1) sulfur dioxide, 2) nitrogen oxides, and 3) particulates. Trace metal and organic compounds are also under consideration as potential pollutants emitted from FBC units. The government has an extensive program in progress to identify and quantify these environmentally harmful discharges. As a result, it was specifically not a goal of this program to examine emissions from the SSU in great detail. Rather, the objective was to monitor the major pollutants to insure that conventional FBC emission controls would allow the Demonstration Plant to meet the applicable air quality standards. This objective also included the determination of limestone feed requirements for equipment design purposes.

Sulfur Dioxide

FBC boilers provide integral control of sulfur dioxide emissions when operated with a limestone bed. The chemical reactions involved in the sulfur removal process essentially calcination and subsequent sulfation of the limestone, are temperature dependent. A design bed temperature of 1550°F has typically been identified as the temperature where optimum sulfur capture is achieved. Westinghouse has examined sulfur removal process in detail, leading to the development of a predictive model. This model allows determination of the required limestone feedrate for a given sulfur removal, based on laboratory thermogravimetric analysis (TGA) of the limestone used.

In order to take advantage of this information, three candidate limestones for use in the Demonstration Plant were submitted to Argonne National Laboratory for TGA testing. Argonne applied the Westinghouse model to the results and generated predictions of limestone usage. The limestone supplied by Medusa Aggregates Co. was chosen as the most economical (delivered cost x usage) limestone. While this was the most expensive limestone, its predicted sulfur removal performance more than compensated for its cost.

Results from the SSU test matrix are presented in Figure 8 where sulfur removal is related to the major variables of bed temperature and calcium to sulfur molar ratio (Ca/S). Design sulfur removal of 90% can be achieved with a Ca/S ratio as low as 2.2 under optimum conditions. In practice, the actual Ca/S ratio employed is likely to be somewhat higher to account for the variations in bed temperature which will be encountered during normal operation. Overall performance was better than expected reflecting the fact that this particular limestone is especially active in removing sulfur.

Further data reduction indicates that the test results obtained match the model predictions rather closely. Figure 9 shows a comparison of SSU data obtained at a bed depth of 30 to 35 inches bracketed by model predictions for bed depths of 20 and 40 inches. Similar results were obtained with respect to superficial velocity.

The Westinghouse model is designed to predict sulfur capture in the bed. There are definite indications that additional sulfur removal can take place in the freeboard region. Sulfur dioxide concentration profiles in the SSU show somewhat lower concentration vertically above the recycle reinjection point. The recycled limestone appears to be capable of absorbing more

SULFUR REMOVAL vs BED TEMPERATURE

SULFUR REMOVAL vs CALCIUM/SULFUR
MOLE RATIO

Figure 9

sulfur dioxide although the SSU does not take full advantage of this because of maldistribution. Several tests were performed with the recycled material mixed with the coal limestone feed with improved sulfur removal resulting.

Particulate

FBC units generate large amounts of particulate matter. This material comes from the coal ash char and limestone feed. Depending on whether or not material collected in the cyclone was reinjected back to the bed, flue gases at the boiler outlet contained as high as 150 grains/cubic foot. The cyclone operated at about 90% collection efficiency, reducing the dust loading by a factor of 10. After passing through a tubular air heater, the flue gases received final particulate cleanup in a bag filter.

A determination of the particulate emissions of the SSU was made using procedures outlined in EPA Method 5 - Determination of Particulate Emissions from Stationary Sources. Average particulate loading the stack gas was 0.006 grains per cubic foot or 0.012 pounds per million Btu fired. Bag filter collection efficiency for this test was 99.96 percent.

While these results confirm that successful control of particulate emissions can be achieved, they also point to the need for highly efficient bag filter operation. This was not the case during initial operation of the bag filter. Poor collection efficiency was traced to several torn bags, damaged during installation. Once these bags were replaced, the units operated successfully for the remainder of the test program (2,000 hours). Although the ultimate life of the bags was not determined, periodic bag replacement will be required to maintain collection efficiency.

Nitrogen Oxides

Nitrogen oxide emissions from conventional coal fired boilers are generated from two sources, fuel nitrogen and atmospheric nitrogen. The atmospheric nitrogen is converted to NO_x by thermally controlled equilibria. In fluidized bed boilers, combustion temperatures are too low for any significant thermal NO_x generation. The NO_x emissions from an FBC boiler are therefore governed by the release of fuel nitrogen compounds. The coal burned in the SSU contained slightly over 1% nitrogen and would have resulted in emissions on the order to two pounds per million Btu if all of the nitrogen were converted to NO_x . Fortunately, the fluidized bed converts fuel nitrogen to NO_x with low efficiency. Actual emissions from the SSU were in the 0.2 to 0.5 pound per million Btu range.

Almost all of the fuel nitrogen is released in gaseous form during the combustion process. Less than 10% of the fuel nitrogen was accounted for in the solids streams leaving the unit. A complex set of formation and destruction reactions takes place that governs the amount of NO_x emitted. An in depth examination of these reactions was beyond the scope of this investigation; however, an empirical correlation was developed. This correlation is shown in Figure 10.

The presence of a term including carbon monoxide in the correlation is caused by an NO_x destruction reaction. In this reaction, NO_x is consumer by a reducing species, i.e. carbon or carbon monoxide to yield nitrogen. While cabon monoxide was chosen for the correlation, there is a strong interrelationship between carbon and carbon monoxide in the SSU. It was not possible to identify the individual roles of these materials in controlling NO_x emissions.

NO_x MEASURED vs NO_x CALCULATED WITH RECYCLE

A comprehensive and detailed treatment of the SSU results is provided in this final test report submitted to the Department of Energy. The foregoing is, necessarily, a brief summary of this report. It will contribute to the expanding body of knowledge that is FBC technology. Of primary value, however, is the direct contribution of SSU results to the development and design confirmation of the Demonstration Plant.

Design changes to the Demonstration Plant based upon SSU results significantly improve the prospect for the rapid shakedown and start-up of the Great Lakes unit. The major changes that have been implemented include:

- o Removal of in-bed heat transfer surface to compensate for changes in the design heat transfer rates and heat release distribution.
- o Addition of provisions to vary reinjection rates during operation over the range of 0 to 90% to achieve optimum flow rates and control dust loadings.
- o Redesign of the bed air distributor, allowing for free thermal expansion and using improved high temperature materials.
- o Improvements in the design of instrumentation probes to provide reliable sensing of control values.

In addition, SSU operation has verified that the FBC process design is appropriate for achieving the required performance. Direct reinjection for carbon recovery appears successful in achieving acceptable combustion efficiency at a moderate, 7

ft/sec, bed superficial velocity. The selected limestone can be expected to attain effective SO_2 removal at an economical limestone feedrate. Also, where possible, the SSU utilized equipment components similar to those of the Demonstration Plant providing valuable practical experience in their operation. The Demonstration Plant, beyond its objectives in proving the practical application of FBC technology, will serve as follow-on for refining the predictive capabilities of sub-scale testing.

SECTION 2 DISCUSSION OF TEST RESULTS

2.1 Bed Dynamics

2.1.1 Two important design variables for fluidized beds are minimum fluidization velocity and pressure drop. Minimum fluidization velocity not only sets a lower limit on the operating conditions of a fluidized bed, but also is useful for prediction of bed expansion, heat transfer rates, and pressure drop. Several correlations⁽¹⁻⁵⁾ have been published for estimating the minimum fluidization velocity, U_{mf} , based on the physical properties of the fluidizing medium and solids. Some of these correlations⁽³⁻⁵⁾ require information on the density and viscosity of the fluidizing medium, and on the density and the average diameter of the fluidized solids, while other need additional information on the shape factor and minimum fluidization voidage^(1,2). The most rigorous correlation to predict the minimum fluidization velocity is the Ergun Correlation. The general form of Kunii and Levenspiel's modification of the Ergun Correlation is given by the following equation:

$$\frac{1.75}{\phi_s \epsilon_{mf}^3} \left(\frac{\bar{d}_p U_{mf} \rho_g}{\mu_g} \right)^2 + \frac{150(1-\epsilon_{mf})}{\phi_s^2 \epsilon_{mf}^3} \left(\frac{\bar{d}_p U_{mf} \rho_g}{\mu_g} \right) = \frac{\bar{d}_p^3 \rho_g (\rho_s - \rho_g) g}{\mu_g^2} \quad (1)$$

where

$\bar{d}_p$	=	Mean particle size, cm
ϕ_s	=	Particle shape factor
μ_g	=	Viscosity of gas, gm/cm-sec
ρ_s	=	Solids density, gm/cc
ρ_g	=	Fluid density, gm/cc
U_{mf}	=	Superficial gas velocity above the bed at minimum fluidization, cm/sec
ϵ_{mf}	=	Bed void fraction at minimum fluidization

Data pertaining to densities and fluid viscosity are known from process specifications. Throughout this study data pertaining to mean particle size, particle shape factor, ϕ_s , and bed voidage

at minimum fluidization, ϵ_{mf} , were obtained from the experimental determinations⁽⁶⁾. During our cold flow test results analysis, a comparison of the calculated (Equation 1) and experimental values of U_{mf} showed a standard deviation ranging from 32.8% to 64.4% for the topsizes of 1/4" and 1/8"⁽⁶⁾. Therefore, the Ergun Correlation as given by Equation 1 was modified to agree with our drop time method used to determine ϕ_s . The following modified Ergun Correlation was obtained⁽⁶⁾.

$$1.75 \frac{\phi_s}{\epsilon_{mf}^3} \left(\frac{dp}{\mu_g} U_{mf} \rho_g \right)^2 + 150 \frac{(1-\epsilon_{mf})}{\epsilon_{mf}^3} \left(\frac{dp}{\mu_g} U_{mf} \rho_s \right) = \frac{\sim^3 dp g \rho_g (\rho_s - \rho_g)}{\mu_g^2} \quad (2)$$

In our cold flow test results analysis, the agreement between the calculated (Equation 2) and experimental values of U_{mf} was within $\pm 9\%$. Therefore, throughout this study we have used Equation 2 to estimate minimum fluidization velocity, U_{mf} . The estimated values of U_{mf} using Equation 2 are presented in Table I for Test Series K through M.

The general form of Equation 2 is:

$$K_1 (Re_{mf})^2 + K_2 (Re_{mf}) - Ar = 0 \quad (3)$$

where

$$Re_{mf} = \frac{\sim dp_{mf} \rho_g}{\mu_g}$$

$$Ar = \frac{\sim^3 dp g \rho_g (\rho_s - \rho_g)}{\mu_g^2}$$

It should be noted that K_1 and K_2 are functions of shape factor, ϕ_s , and voidage at minimum fluidization, ϵ_{mf} . Using $\phi_s = 0.74$ and $\epsilon_{mf} = 0.42$, the constants K_1 and K_2 were determined to be:

$$K_1 = 17.48$$

$$K_2 = 1174.28$$

Substituting these values of K_1 and K_2 in Equation 3 and simplifying, the following correlation for U_{mf} can be obtained:

$$U_{mf} = \frac{\mu_s}{d_p \rho_g} \left[\{(33.59)^2 + 0.057 Ar\}^{\frac{1}{2}} - 33.59 \right] \quad (4)$$

This correlation is shown in Figure 11, as a plot of estimated values (Equation 2) of U_{mf} vs Ar .

2.1.2 Bubbling Fluidized Bed

A large number of models have been proposed to describe the fluid mechanical behavior of a gas-solid fluidized bed^(1,2,7-10).

Currently, none of the proposed models completely explain the basic phenomena occurring in the bed. One of the best existing models for freely bubbling beds was developed by Kunii and Levenspiel⁽²⁾. It should be pointed out that the bubbling bed models, in general, assume the bubble size as an adjustable parameter in the absence of bubble size formation. For example, to simplify the overall representation of the fluidized bed, Kunii and Levenspiel assumed the bubble size to be uniform throughout the bed⁽²⁾.

In gas fluidized beds, with gas velocities beyond minimum fluidization, the bed expands and gas bubbles are seen to rise with resulting nonhomogeneity in the bed. The so-called two-phase theory with Davidson's bubble model forms the basis for much of the fluid-dynamic analysis⁽¹⁾. The following assumptions are generally made to account for the physical events occurring in a gas fluidized bed^(1,2,7-10):

TABLE 1

TABLE I
Summary of Fluidization and Bed Expansion Data
Freeman Coal

Run	T _B , °F	U _o , ft/sec	RECY	Lf, inches	ΔB Bed in. Hec	$\bar{n}$ dp, μ	ρ _s , lbs/ft ³	Lmf, inches	R=Lf/Lnt	E _{bed}	Umf, ft/sec	U _o /Umf	U _o -Umf/Umf	U _o -Umf
K-1	1525	7.58	-	39.35	30.35	433	128.72	25.39	1.57	0.63	0.242	31.27	30.37	7.34
K-2	1634	7.90	-	30	31.10	438	129.45	25.87	1.16	0.50	0.236	33.52	32.52	7.66
K-3	1662	7.92	-	30.13	23.80	469	141.78	18.07	1.67	0.65	0.453	17.47	16.47	7.47
K-5	1470	5.40	-	43.55	32.42	499	161.26	21.45	2.01	0.71	0.392	13.78	12.78	5.01
K-6	1432	5.95	-	35.72	35.91	538	158.52	24.40	1.46	0.61	0.451	13.19	12.19	5.50
K-7	1469	6.32	-	29.86	40.57	618	153.58	28.44	1.05	0.49	0.578	10.93	9.93	5.74
K-8	1420	6.64	-	35.23	36.09	711	136.77	28.42	1.24	0.53	0.089	9.64	8.64	5.95
K-9	1561	7.02	-	39.18	38.47	600	153.49	26.99	1.45	0.60	0.532	13.19	12.19	6.49
K-10	1468	7.51	-	37.38	35.52	665	153.25	24.96	1.50	0.61	0.667	11.26	10.26	6.84
K-11	1457	8.05	-	37.18	35.52	335	157.28	24.31	1.53	0.62	0.177	45.51	44.51	7.87
K-12	1477	8.65	-	31.30	37.00	638	156.31	25.52	1.46	0.60	0.626	13.82	12.82	8.02
K-13	1663	8.83	-	37.42	34.20	612	146.19	28.19	1.49	0.61	0.511	17.28	16.28	8.32
K-14	1607	9.20	-	37.10	35.02	621	160.07	23.56	1.58	0.63	0.555	15.72	14.72	8.62
K-15	1691	9.97	-	35.06	35.90	631	142.75	27.08	1.30	0.55	0.527	18.94	17.94	9.44
K-16	1500	9.13	YES	45.05	39.01	592	123.94	33.09	1.33	0.56	0.426	21.44	20.44	8.70
K-17	1545	5.70	YES	31.06	20.52	606	142.65	15.49	2.01	0.71	0.506	11.27	10.27	5.19
K-18	1478	5.98	YES	28.41	20.84	502	143.65	15.62	0.18	0.35	0.355	16.86	15.86	5.13
K-19	1483	6.77	YES	28.49	25.10	455	142.87	18.91	1.51	0.62	0.306	22.14	21.14	6.46
K-20	1498	7.31	YES	25.62	21.17	613	142	16.19	1.58	0.63	0.522	14	13	6.79
K-21	1487	7.87	YES	21.72	12.16	417	141.10	9.28	2.32	0.75	0.244	32.31	31.31	7.63
K-22	1470	8.12	YES	29.97	21.41	381	144.36	15.97	1.88	0.69	0.208	38.98	37.98	7.92
K-23	1470	8.10	YES	31.38	21.51	333	144.01	16.08	1.95	0.70	0.159	51	50	7.94
K-24	1511	7.33		33.52	35.85	530	131.41	29.36	1.14	0.49	0.362	20.27	19.27	6.94

TABLE I Continued

Run	T _B , °F	U _o , ft/sec	RECY	Lf, inches	ΔB Bed in. Hec	$\sim$ dp, lb	ρ_s , lbs/ft ³	Lmf, inches	R=Lf/Lmt	E _{bed}	Umf, ft/sec	U _o /Umf	U _o -Umf/Umf	U _o -Umf
L-1	1521	7.81	-	41.02	37.26	466	137.33	29.21	1.40	0.59	0.291	26.30	25.80	7.52
L-2	1452	7.35	YES	37.75	39.91	459	137.57	31.24	1.21	0.52	0.290	25.36	24.36	7.06
L-3	1466	6.47	YES	40.09	35.82	500	137.43	28.06	1.43	0.59	0.341	18.97	17.97	5.13
L-4	1389	5.71	YES	38.08	43.02	502	135.10	34.29	1.11	0.48	0.346	16.48	15.48	5.36
L-5	1598	8.98	YES	41.15	34.94	513	136.77	27.51	1.51	0.62	0.483	15.58	17.58	8.50
L-6	1439	9.15	YES	45.79	46.68	360	137.79	36.48	1.26	0.53	0.179	51.09	50.09	8.47
L-7	1602	10.49	YES	33.54	28.68	486	150.79	20.48	1.64	0.65	0.340	30.83	29.83	10.15
L-8	16.35	9.39	YES	33.04	26.86	465	148.32	19.50	1.70	0.66	0.303	30.97	29.97	9.09
L-9	1417	7.38	YES	37.14	44.72	575	123.78	38.09	0.956	0.39	0.410	18.00	17.00	6.97
L-10	1414	6.59	YES	36.24	41.16	567	135.40	32.73	1.11	0.48	0.432	15.26	14.26	6.16
L-11	1423	5.94	YES	34.85	42.49	508	132.03	34.65	1.0	0.42	0.342	17.36	16.36	5.60
L-12	1449	5.05	YES	34.54	36.06	476	134.84	28.79	1.20	0.52	0.305	16.56	15.56	4.75
L-13	1458	4.76	YES	31.18	29.12	544	137.58	22.79	1.37	0.58	0.405	11.77	10.77	4.36
L-14	1456	4.59	YES	41.02	37.26	466	137.59	28.66	1.32	0.56	0.451	10.18	9.18	4.14
L-15	1400	5.21	YES	25.60	26.07	421	124.65	22.52	1.14	0.49	0.225	23.19	22.19	4.99
L-16	1433	5.75	YES	25.63	28.90	415	128.92	21.63	1.19	0.51	0.654	8.80	7.80	5.10
L-17	1458	6.31	YES	26.63	22.14	649	135.30	17.62	1.51	0.62	0.564	11.19	10.19	5.75
L-18	1571	6.16	YES	26.119	22.57	484	137.81	17.63	1.50	0.61	0.311	19.79	18.79	5.85
L-19	1455	6.28	YES	27.65	24.11	434	135.27	19.19	1.44	0.60	0.255	24.68	23.68	6.03
L-20	1400	6.66	YES	26.22	25.34	367	119.51	22.83	1.15	0.50	0.146	45.77	44.77	6.52
L-21	1442	7.42	YES	28.67	23.37	570	125.25	26.09	1.28	0.55	0.406	18.28	17.28	7.01
L-22	1716	8.42	-	26.83	23.01	558	150.97	16.41	1.64	0.65	0.445	18.94	17.94	7.98
L-23	1654	6.43	-	27.33	26.55	-	162.94	17.54	1.56	0.63	-	-	-	-
L-24	1645	6.15	-	28.25	24.66	500	156.87	16.93	1.67	0.65	0.309	16.66	15.66	5.78
L-25	1589	10.08	YES	27.57	20.31	483	135.83	16.10	1.71	0.66	0.304	33.16	32.16	9.78
L-26	1341	9.49	YES	27.35	19.79	426	136.88	15.60	1.75	0.67	0.242	39.26	38.26	9.25
L-27	1554	8.98	YES	27.01	21.46	406	135.68	17.03	1.59	0.63	0.217	40.46	39.46	5.76
L-28	1406	7.28	-	42.78	43.04	439	127.50	36.35	1.18	0.51	0.249	29.23	28.23	7.03
L-29	1439	5.0	YES	41.68	42.51	607	134.07	34.14	1.22	0.53	0.491	10.18	9.18	4.51
L-30	1407	7.38	YES	37.59	41.24	649	144.97	30.63	1.23	0.53	0.611	12.00	11.00	6.72
L-31	1498	6.75	YES	37.27	27.21	549	167.58	23.91	1.56	0.63	0.495	13.63	12.63	6.26
L-32	1446	5.70	YES	37.53	40.75	518	151.23	28.08	1.34	0.57	0.418	13.64	12.64	5.28
L-33	1446	5.22	YES	35.44	40.20	589	169.03	25.61	1.38	0.58	0.571	9.05	8.05	4.64
L-34	1555	5.22	YES	30.07	27.55	525	157.95	17.66	1.70	0.66	0.447	11.67	10.67	4.77
L-35	1425	5.63	YES	35.67	24.75	439	161.47	23.17	1.54	0.62	0.313	18	17	5.32
L-36	1562	6.43	YES	30.55	20.18	573	171.40	18.96	1.61	0.64	0.541	11.89	10.89	5.89

TABLE I Continued

Run	$T_B, ^\circ F$	$U_o, \text{ft/sec}$	RECY	L_f, inches	$\Delta B \text{ Bed}$ $\text{in. H}_2\text{O}$	$\sim$ dp, psi	$\rho_s, \text{lbs/ft}^3$	L_{mf}, inches	$R=L_f/L_{mf}$	E_{bed}	$U_{mf}, \text{ft/sec}$	U_o/U_{mf}	U_o-U_{mf}/U_{mf}	U_o-U_{mf}
L-37	1479	7.47	YES	33.14	33.71	542	115.37	21.95	1.51	0.62	0.478	15.62	14.62	7.00
L-38	1682	9.01	YES	33.98	30.90	625	146.36	22.73	1.50	0.61	0.531	16.97	15.97	8.48
L-39	1619	9.45	YES	33.13	31.78	557	165.32	20.70	1.60	0.64	0.486	19.44	18.44	8.97
L-40	1645	9.01	YES	30.71	27	505	156.35	18.59	1.65	0.65	0.377	23.92	22.92	9.63
L-41	1628	10.06	YES	32.03	29.55	497	161.57	19.69	1.63	0.64	0.378	26.60	25.60	9.68
L-42	1622	9	YES	32.23	28.50	493	137.51	22.32	1.44	0.60	0.321	32.26	30.26	9.71
L-43	1598	9	YES	35.35	27.56	433	136.91	21.67	1.63	0.64	0.246	36.62	35.62	8.75
L-44	1443	8.85	YES	42.80	39.18	500	134.92	31.27	1.37	0.53	0.337	26.25	25.25	8.51
L-45	1660	10.07	-	35.02	27.91	518	163.90	18.34	1.91	0.70	0.411	24.48	23.48	9.66
L-46	1649	7.84	-	30.81	88.38	466	184.16	19.82	1.55	0.63	0.316	24.83	23.83	7.52
L-47	1643	6.91	-	33.39	31.59	500	152.37	22.32	1.50	0.61	0.359	19.28	18.28	6.55
L-48	1545	5.91	-	35.10	29.49	477	165.34	19.35	1.80	0.68	0.365	16.17	15.17	5.55
L-49	1528	5.21	-	34.33	40.58	445	167.65	26.06	1.32	0.56	0.324	16.09	15.09	4.89
L-50	1510	7.70	-	41.02	36.53	416	136.63	28.79	1.43	0.59	0.232	33.15	32.15	7.47

TABLE I Continued

Run	T _B , °F	U _o , ft/sec	RECY	L _f , inches	ΔB Bed in. Hec	$\frac{\mu}{dp_{s,u}}$	ρ_s , lbs/ft ³	L _{mf} , inches	R=L _f /L _{mt}	E _{bed}	U _{mf} , ft/sec	U _o /U _{mf}	U _o -U _{mf} /U _{mf}	U _o -U _{mf}
M-1	1489	7.22	-	42.79	38.29	610	138.06	29.86	1.43	0.60	0.504	14.33	13.33	6.72
M-2	1614	8.67	YES	53.36	29.71	449	139.22	22.98	2.32	0.75	0.267	32.47	31.47	8.40
M-3	1481	9.76	YES	34.16	34.83	568	139.44	26.89	1.27	0.54	0.444	22	21	9.32
M-4	1405	7.07	YES	41.01	40.68	471	123	35.08	1.17	0.50	0.217	25.55	24.55	6.79
M-5	1465	6.20	YES	37.47	42.49	599	135.06	33.87	1.11	0.48	0.480	12.91	11.91	5.72
M-6	1342	5.17	YES	38.24	47.45	546	122.32	41.77	0.92	0.37	0.376	13.76	12.76	4.73
M-7	1383	4.59	YES	46.09	45.49	623	117.93	41.53	1.11	0.48	0.464	9.90	8.90	4.13
M-8	1539	9.68	YES	28.89	25.05	487	141.65	19.04	1.52	0.62	0.326	29.68	28.68	9.35
M-9	1592	8.42	YES	28.71	26.20	439	141.80	19.96	1.44	0.60	0.261	32.30	31.30	8.16
M-10	1417	6.96	YES	32.21	31.13	417	137.61	24.36	1.32	0.56	0.242	28.81	27.81	6.72
M-11	1450	6.14	YES	31.87	34.80	425	140.37	26.69	1.19	0.51	0.254	24.19	23.19	5.89
M-12	1404	5.39	YES	31	36.14	325	126.38	30.79	1.0	0.42	0.136	39.63	38.63	5.25
M-13	1415	4.53	YES	33.90	38.74	726	122.74	33.98	0.99	0.42	0.645	7.02	6.02	3.88
M-14	1573	5.88	YES	27.34	23.89	443	160.75	16	1.70	0.66	0.304	19.36	18.36	5.58
M-15	1510	6.15	YES	26.64	22.40	478	165.21	14.60	1.83	0.68	0.370	16.61	15.61	5.78
M-16	1443	6.42	YES	27.85	23.07	463	152.28	16.31	1.71	0.66	0.32	20.07	19.07	6.10
M-17	1313	6.62	YES	26.55	27.85	476	116.58	25.72	1.03	0.44	0.276	24	23	6.34
M-18	1531	7.59	-	36.31	34.55	594	143.05	26.23	1.83	0.58	0.49	15.50	14.50	7.10
M-19	1617	6.43	YES	27.06	28.84	539	154.97	17.95	1.51	0.62	0.427	15.06	14.06	6.0
M-20	1634	5.96	YES	28.61	24.67	586	144.79	18.35	1.56	0.63	0.468	12.75	11.75	5.49
M-21	1639	5.57	YES	27.23	25.16	488	138.65	19.54	1.39	0.58	0.311	17.90	16.90	5.26
M-22	1631	5.17	YES	27.68	25.30	480	137.09	19.87	1.39	0.58	0.299	17.27	16.27	4.88
M-23	1600	4.66	YES	26.81	23.77	477	148.01	17.29	1.55	0.63	0.321	14.50	13.50	4.34
M-24	1370	6.04	YES	35.56	39.02	463	119.23	35.24	1.07	0.43	0.262	23.05	22.05	5.78
M-25	1371	6.49	YES	36.27	39.03	555	116.01	36.22	1.01	0.42	0.365	17.17	16.77	6.13
M-26	1524	8.14	YES	38.11	34.15	541	138.46	26.56	1.44	0.60	0.487	16.72	15.72	7.65
M-27	1466	9.17	YES	36.76	34.90	527	-	-	-	-	-	-	-	-
M-28	1426	6.97	YES	32.46	31.17	527	137.21	24.46	1.33	0.56	0.383	18.19	17.19	6.59
M-29	1445	6.15	YES	36.26	30.16	544	142.38	22.81	1.59	0.64	0.42	14.64	13.64	5.73
M-30	1523	8.13	YES	35.52	28.91	435	142.24	21.88	1.82	0.64	0.263	30.91	29.91	7.87
M-31	1437	8.92	YES	36.82	27.76	527	-	-	-	-	-	-	-	-
M-32	1382	6.36	YES	37.43	37.21	396	124.02	32.30	1.16	0.50	0.199	31.96	30.96	6.16
M-33	1393	5.71	YES	39	35.16	483	120.04	31.54	1.24	0.53	0.285	20.07	19.07	5.43
M-34	1440	6.62	YES	37.18	38.74	471	125.65	33.20	1.12	0.48	0.279	23.09	22.69	6.34
M-35	1361	4.56	YES	37.91	35.24	467	120.63	31.45	1.21	0.52	0.271	16.81	15.81	4.29
M-36	1472	6.75	YES	45.38	31.38	534	137.85	24.49	1.85	0.69	0.389	17.37	16.37	6.36
M-37	1551	7.54	YES	37.54	34.05	488	142.611	25.71	1.41	0.59	0.328	22.96	21.96	7.22

- o The properties of the emulsion phase, such as gas velocity, void fraction, etc., remain at minimum fluidization conditions.
- o All the gas over that required for minimum fluidization passes through the bed in the bubble phase partly as visible bubble flow and partly as bubble throughflow and/or bubble gas recirculation.

The so-called bubble wake sets up a pattern of solids recirculation within the bed. The bubble phase can exchange both gas and solids (through the wake) with the emulsion phase.

Therefore, bubble properties play a fundamental role in determining mass and heat transfer properties of the bed. An understanding of the phenomena of bubbling and the effects it produces is the key to understanding many of the relevant properties of gas fluidized beds.

One of the fundamental differences between the fluid dynamics of large particle and small particle systems is the nature of the flow regimes. For example, in large particle systems ($\bar{d}_p > 1000\mu$), the rise velocity of the bubbles will be less than the gas velocity through the dense phase. This is so-called the slow-bubble regime in which most of the gas flowing through the bed is associated with the dense phase. In this regime the bubbles may serve primarily as a short cut for gas flowing through the emulsion phase. In small particle systems, the bubble velocity is greater than the interstitial gas velocity in the dense phase and gas bypassing occurs because some of the gas recirculates inside the bubbles and does not contact the particles. As can be seen from Table I, the harmonic mean particle size ($\bar{d}_p = \sum X_i / \sum X_i / d_{p_i}$) of the bed material obtained from SSU Test Series K through M varies from 330μ to 700μ . The Kunii and Levenspiel's criteria⁽²⁾ suggests SSU bed operating regime to be in the fast bubble regime.

In bubbling fluidized bed models bubble diameter is one of the important parameters which affect the performance of a gas-fluidized bed. It is well-known that bubble size increases as the gas flow rate is increased and height above the distributor is increased. Moreover, for a given flow rate and bed depth a wide spectrum of bubble diameters could be observed passing a given level. Various corrections for estimating bubble diameters in fluidized beds have appeared in the literature⁽¹¹⁻¹⁴⁾. Many of these correlations are based on the data obtained using porous plates as the distributor. Recent correlations of Darton et al⁽¹⁵⁾ for estimating the bubble diameter is given by the following relation:

$$d_B = 0.54 (U_o - U_{mf})^{0.4} (\bar{h} - \bar{h}_j + 4 \sqrt{A_o})^{0.8} / g^{0.2} \quad (5)$$

where

- d_B = Diameter of the sphere having the same volume as the bubble, cm
- U_o = Superficial gas velocity, cm/sec
- U_{mf} = Minimum fluidization velocity, cm/sec
- $\bar{h}$ = Height above the distributor, cm
- $\bar{h}_j$ = Jet penetration distance, cm
- A_o = Catchment area for a bubble stream at the distributor plate
- g = Acceleration due to gravity, cm/sec²

For the SSU distributor plate, $A_o = 5.51 \text{ cm}^2/\text{hole}$ and for a 1600 lbs/hr of air flow rate, $\bar{h}_j = 7 \text{ cm}$. Substituting values of A_o and $\bar{h}_j$ in Equation 5, we have:

$$\begin{aligned} d_B &= 0.136 (U_o - U_{mf})^{0.4} (\bar{h} - 7 + 9.34)^{0.8} \\ &= 0.136 (U_o - U_{mf})^{0.4} (\bar{h} + 2.39)^{0.8} \\ d_B &\cong 0.136 (U_o - U_{mf})^{0.4} (\bar{h})^{0.8} \end{aligned} \quad (6)$$

Recently, Bar-Cohen et al⁽¹⁶⁾ presented gas-flow continuity relation for the fast-bubble regime.

$$U_o = [1 - \delta (1 + \beta + \alpha)] \epsilon_{mf} U_e + \delta [1 + (\alpha + \beta) \epsilon_{mf}] U_b \quad (7)$$

where

δ = Bubble fraction

U_e = Interstitial gas velocity, cm/sec

α = Bubble wake fraction, V_w/V_b

β = Bubble cloud fraction, V_c/V_b

U_b = Bubble rise velocity for a freely bubbling bed, cm/sec

In the case of the fast bubble, the relative cloud volume, β , is defined as⁽¹⁶⁾:

$$\beta = \frac{3 (1+\alpha) U_f}{(U_{br} - U_f)} \quad (8)$$

where

$$U_f = \text{Absolute velocity of gas in the emulsion phase at minimum fluidization, cm/sec}$$

In a freely bubbling bed occupied by a significant bubble fraction, the bubble rise velocity relative to the distributor, U_b , is assumed to be given by⁽¹⁾:

$$U_b = (U_o - U_{mf}) + U_{br} \quad (9)$$

where

$$\begin{aligned} U_{br} &= \text{Rise velocity of an isolated bubble, cm/sec} \\ &= 0.711 \quad g d_B \end{aligned}$$

Assuming $U_e = U_f = U_{mf}/\epsilon_{mf}$ and on rearranging, Equation 7 can be written as:

$$\delta = \frac{(U_o - U_{mf})}{U_b (1 + \alpha \epsilon_{mf} + \beta \epsilon_{mf}) - U_{mf} (1 + \alpha + \beta)} \quad (10)$$

We can now substitute Equation 9 in the Equation 10 and obtain:

$$\delta = \frac{(U_o - U_{mf})}{[U_b (1 + \alpha \epsilon_{mf} + \beta \epsilon_{mf}) + U_{br} (1 + \alpha \epsilon_{mf} + \beta \epsilon_{mf}) - U_{mf} (2 + \alpha \epsilon_{mf} + \beta \epsilon_{mf} + \alpha \beta)]} \quad (11)$$

where

$$\beta = \frac{3 (1 + \alpha)}{[\frac{U_{br} \epsilon_{mf}}{U_{mf}} - 1]}$$

The relationship between total gas flow and that through the two phases expressed by Equation 11 is the so-called "fast bubble" equation. The following assumptions have been made in writing Equation 11:

The emulsion phase void fraction remains at minimum fluidizing conditions.

Interstitial gas velocity in the emulsion phase is given by U_{mf} .

2.1.3 Bed Expansion

Since the bed expansion above the static bed height is primarily due to bubble holdup, the bed bubble fraction, δ , can in principle be used to predict bed expansion at various gas flow rates. The average bed voidage, ϵ_{bed} , is related to the voidage in bubbles and emulsion by:

$$\epsilon_{bed} = \delta \epsilon_b + (1 - \delta) \epsilon_e \quad (12)$$

Assuming $\epsilon_b = 1$ and $\epsilon_e = \epsilon_{mf}$, we have:

$$\epsilon_{bed} = \delta + (1 - \delta) \epsilon_{mf} \quad (13)$$

In gas fluidized beds, beyond the point of minimum fluidization, the pressure drop across the bed remains constant, i.e.,:

$$L_f (1-\epsilon_{bed}) (\rho_s - \rho_g) \frac{g}{g_c} = L_{mf} (1-\epsilon_{mf}) (\rho_s - \rho_g) \frac{g}{g_c} \quad (14)$$

Combining Equations 12 and 14 we have:

$$\frac{L_f}{L_{mf}} = R = \frac{1}{1-\delta} \quad (15)$$

Where

$$\begin{aligned} L_f &= \text{Expanded bed height} \\ L_{mf} &= \text{Static bed height} \\ R &= \text{Bed expansion ratio} \end{aligned}$$

Throughout this study, the expanded bed height (L_f) was calculated from pressure drop measurements. The bed was instrumented with a set of four pressure probes. A numerical procedure was used to fit best straight line through four points. Only those runs in which all four points lay on a straight line were considered. Figure 12 gives a plot of R against $(U_o - U_{mf})/U_{mf}$ for Test Series Runs K through M.

In order to obtain a theoretical fit to the data, Equation 6 along with the point bubble frequency relation of Cranfield and Geldard⁽¹⁷⁾:

$$f_{pt.} = 16.7 h^{-0.72} \quad (16)$$

were used to obtain average bubble diameter in the bed. Assuming $\alpha=0.2$ and $\epsilon_{mf}=0.4$, Equation 17 can be used to predict δ , i.e., R :

$$\delta = \frac{(U_o - U_{mf})}{U_o(1.084 + 0.42\beta) + U_{br}(1.084 + 0.42\beta) - U_{mf}(2.284 + 1.42\beta)} \quad (17)$$

where

$$\beta = \frac{3.6}{\left(\frac{0.42 U_{br}}{U_{mf}} - 1 \right)}$$

$$U_{br} = \frac{71.12 (U_o - U_{mf})^{0.2}}{(1-\delta)^{0.41}} \quad \text{cm/sec}$$

A detailed derivation of Equation 17 is presented in Appendix I. The calculated values of R are compared with the experimental values of R in Figure 12. The agreement between calculated and experimental values of R is within $\pm 10\%$.

One of the most important advantages of fluidized bed combustion is the high heat transfer coefficient from the bed to an immersed surface. A large amount of literature is available on the prediction of heat transfer in fluidized beds^(1,2,18). However, caution is required in applying these analyses to coal-fired fluidized beds, since most analyses are based on beds operating at very low temperatures (less than 500°F). In addition, most of the work has been performed on relatively small bench scale beds. Therefore, when using the vast majority of available correlations, careful consideration must be given to the applicability of these correlations to fluidized bed boilers where combustion and other chemical process are taking place.

The heat transfer rate in a fluidized bed has been shown to depend on the fluidizing gas velocity and on the thermal conductivity, the size and density of the solid particles, their thermophysical properties, bed temperature and to a lesser extent on the surface geometric design parameters^(1,2,18). In various correlations which have been developed for fluidized beds in general, the dependence of heat transfer coefficient on the effective particle diameter varies from⁽¹⁹⁾:

$$h \propto d_p^{-0.17} \quad \text{to} \quad h \propto d_p^{-1.26}$$

Figure 13 shows the plot of convective component of the heat transfer coefficient vs. harmonic mean particle diameter. As can be seen from Figure 13, the convective component of heat transfer coefficient obtained in this study is essentially independent of effective particle diameter.

The effect of superficial gas velocity on the heat transfer coefficient can be broken down into three regions⁽¹⁾. From zero

velocity to the onset of minimum fluidization, the heat transfer coefficient is relatively small and increases gradually with the gas velocity. At the onset of minimum fluidization, the heat transfer coefficient increases very rapidly with velocity until a maximum is reached. Two mechanisms affecting heat transfer take place in this flow regime. Increase in gas velocity increases particle mixing. The faster the particles move, the less their temperature drops below the bulk bed temperature while in the vicinity of the heat transfer surface. This results in increased heat transfer to the immersed surface. At the same time, as gas velocity is increased, the voidage increases resulting in less total energy being transferred to the immersed surface, thus decreasing the heat transfer coefficient. In the emulsion region, the effect of increased particle mixing far outweighs the effect of increasing particle voidage. The increasing voidage becomes the predominant effect with further increases in gas velocity. The net results is a gradual decrease in heat transfer coefficient giving the effect of a constant heat transfer coefficient over a relatively large range of velocities. Figure 14 shows the effect of superficial gas velocity on the convection component of the heat transfer coefficient obtained in this study.

The effect of bed temperature on the heat transfer in a fluidized bed may be divided into: 1) changes in the radiation component of the heat transfer coefficient, 2) changes in the temperature gradient between the fluidized bed and the outside tube wall, and 3) changes in the bed material properties. It has been shown that the thermal conductivity of the fluidizing gas and the specific heat of the bed solids are the bed material properties which are most significant in fluidized bed heat transfer⁽¹⁾. Table II summarizes important heat transfer results obtained in this study.

2.2.2 Physical Model

To a first approximation, the overall bed to immersed surface heat transfer coefficient for a bubbling gas fluidized bed can be thought to consist of the following additive components:

- o The particle convective component, which is dependent on packets of solid particles exchange between the region adjacent to the heat transfer surface and the rest of the bed.
- o The interphase gas convective component, the gas filtering between the particles of the packet and the surface.
- o By gas bubbles.
- o The direct radiation component.

The total heat transfer coefficient is given by the following equation:

$$h = h_{pc} (1 - f_o) + h_{gc} (1 - f_o) + h_b f_o + h_r \quad (18)$$

where

- h_{pc} = Overall average particle convective coefficient, BTU/hr ft² °F
- h_{gc} = Overall average gas convective coefficient, BTU/hr ft² °F
- h_b = Overall average coefficient between the tube and a bubble, BTU/hr ft² °F
- h_r = Overall average radiation coefficient, BTU/hr ft² °F
- f_o = Fraction of heat transfer surface occupied by bubbles
- $1-f_o$ = Fraction of heat transfer surface occupied by emulsion phase.

Particle Convective Heat Transfer

It is generally accepted that particle convective heat transfer is due to surface renewal, i.e., stirring due to bubbles causes packets of particles near an immersed surface to be replaced by other particles. This model was first proposed by Mickley and Fairbanks⁽²⁰⁾. In their model of a packet of particles at bed temperature is swept close to the immersed surface and unsteady conduction takes place. The packet is represented as a homogeneous medium with conductivity and heat capacity based on the average values of gas-solid mixture in the packet.

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TABLE II
Summary of SSU Heat Transfer Coefficient Results
Freeman Coal

Run	T_B °F	U_o , ft/sec	RECY	T.L.I. BTU H_o , m, ft ² , °F	T.L.II H_o	T.L.I. H_o , conv	T.L.I. H_o , Rad	T.L.II. H_o , conv	T.L.II. H_o , Rad
J-1	1415	8.71	-	55.26	56.64	40.60	14.66	41.98	14.66
J-2	1443	8.94	-	53.75	55.47	38.56	15.19	40.28	15.99
J-3	1506	8.18	-	56.46	57.43	39.94	16.51	40.91	16.51
J-4	1405	5.83	-	54.98	44.09	40.48	14.50	29.62	14.47
J-5	1389	6.22	YES	54.60	43.93	40.39	14.21	29.74	14.19
J-6	1339	6.52	YES	50.09	40.43	36.84	13.25	27.20	13.24
J-7	1413	7.32	YES	51.68	41.19	37.04	14.64	26.57	14.12
J-8	1358	7.31	YES	47.99	37.68	34.40	13.59	24.11	13.57
J-9	1433	8.37	YES	51.80	42.77	37.01	15.02	27.77	15
J-10	1391	8.69	YES	47.58	38.67	33.41	14.17	24.52	14.16
J-11	1474	7.65	YES	52.47	44.91	36.64	15.83	29.09	15.81

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TABLE II. Continued

Run	T _B °F	U _o , ft/sec	RECY	T.L.I. BTU ₂ Ho, m, ft ² , °F	T.L.II Ho	T.L.I. Ho, conv	T.L.I. Ho, Rad	T.L.II. Ho, conv	T.L.II. Ho, Rad
K-1	1525	7.58	-	57.97	54.58	41.01	16.96	37.64	16.94
K-2	1634	7.30	-	65.63	64.02	46.15	19.46	44.56	19.44
K-3	1662	7.32	-	66.75	65.32	46.62	20.12	45.21	20.10
K-5	1470	5.40	-	56.11	11.43	40.30	15.81	20.35	15.78
K-6	1482	5.95	-	56.86	46.59	40.80	16.06	30.56	16.04
K-7	1419	6.32	-	55.14	45.02	35.36	15.78	29.26	15.76
K-8	1420	6.64	-	52.30	42.49	37.55	14.78	27.73	14.76
K-9	1561	7.02	-	55.78	47.20	38.01	17.76	29.46	17.74
K-10	1468	7.51	-	53.76	45.17	37.94	15.82	29.32	15.80
K-11	1457	8.05	-	52.48	44.06	36.88	15.60	28.48	15.58
K-12	1477	8.65	-	51.53	43.03	35.52	16.01	27.04	15.99
K-13	1663	8.83	-	56.43	48.35	31.20	20.22	28.15	20.20
K-14	1607	9.20	-	57.87	49.86	38.48	18.88	30.99	18.86
K-15	1691	9.97	-	59.68	51.92	38.72	20.91	31.04	20.89
K-16	1500	9.13	-	57.44	51.04	41.00	16.44	34.60	16.43
K-17	1545	5.70	YES	54.54	46.72	37.33	17.51	29.23	17.49
K-18	1478	5.98	YES	55.46	46.98	39.40	16.06	30.93	16.05
K-19	1483	6.77	YES	54.40	45.62	38.24	16.16	29.48	16.15
K-20	1498	7.31	YES	53.80	46.41	37.34	16.46	29.96	16.45
K-21	1487	7.87	YES	51.14	45.49	34.92	16.22	29.27	16.22
K-22	1470	8.12	YES	55.58	47.78	39.83	15.85	31.94	15.84
K-23	1470	8.10	YES	58.02	49.30	42.16	15.85	33.45	15.85
K-24	1511	7.33		58.11	53.18	41.35	16.76	36.43	16.75

TABLE II Continued

Run	T _B °F	U _o , ft/sec	RECY	T.L.I. BTU ₂ Ho, m, ft ² , °F	T.L.II Ho	T.L.I. Ho, conv	T.L.I. Ho, Rad	T.L.II. Ho, conv	T.L.II. Ho, Rad
L-1	1521	7.81		59.04	49.37	41.89	17.15	32.23	17.14
L-2	1452	7.35	YES	58.03	47.60	42.33	15.70	31.93	15.67
L-3	1466	6.47	YES	58.68	48.19	42.71	15.97	32.25	15.94
L-4	1389	5.71	YES	58.59	47.49	44.47	14.42	33.09	14.39
L-5	1598	8.98	YES	58.86	48.13	39.76	19.10	29.06	19.07
L-6	1439	9.15	YES	55.19	41.10	39.80	15.39	28.74	15.35
L-7	1102	10.49	YES	56.15	42.47	37.19	18.97	23.54	18.92
L-8	1135	9.39	YES	58.94	45.64	39.15	19.78	25.91	19.74
L-9	1417	7.38	YES	56.57	42.71	41.61	14.96	27.79	14.92
L-10	1464	6.59	YES	55.77	42.84	39.78	15.99	26.89	15.95
L-11	1428	5.94	YES	55.92	42.57	40.74	15.17	27.42	15.14
L-12	1449	5.25	YES	56.16	41.83	40.57	15.38	26.28	15.55
L-13	1458	4.76	YES	56.28	42.75	40.49	15.79	26.99	15.76
L-14	1456	4.59	YES	55.35	42.75	39.59	15.76	27.01	15.73
L-15	1400	5.21	YES	53.93	39.93	39.31	14.62	25.14	14.59
L-16	1433	6.31	YES	51.59	38.25	36.34	15.26	23.02	15.23
L-17	1458	6.16	YES	49.69	37.76	33.43	15.76	22.02	15.74
L-18	1571	6.28	YES	53.99	41.54	35.73	18.26	23.31	18.23
L-19	1455	6.28	YES	55.08	41.81	39.35	15.73	26.10	15.71
L-20	1400	6.66	YES	50.55	37.92	35.96	14.60	23.35	14.58
L-21	1442	7.42	YES	47.32	36.37	31.88	15.44	20.95	15.42
L-22	1716	8.42	-	52.91	46.15	31.12	21.79	24.37	21.79
L-23	1654	6.43	-	55.05	47.75	34.81	20.25	27.56	20.22
L-24	1645	6.15	-	55.88	47.50	35.85	20.04	27.49	20.01
L-25	1589	10.08	YES	57.95	49.41	39.15	18.71	30.72	18.69
L-26	1541	9.49	YES	59.95	49.94	42.64	17.10	32.86	17.58
L-27	1554	8.98	YES	60.62	51.21	43.31	17.30	33.92	17.28
L-28	1406	7.28	YES	58.85	47.35	44.39	14.45	32.94	14.42
L-29	1439	5.00	YES	54.66	42.02	39.44	15.22	26.83	15.19
L-30	1407	7.33	YES	52.85	39.77	38.34	14.51	25.28	14.49
L-31	1498	6.75	YES	52.48	40.08	36.10	16.39	23.72	16.36
L-32	1446	5.70	YES	52.55	38.96	37.20	15.35	23.63	15.37
L-33	1482	5.22	YES	52.27	39.68	36.55	16.12	23.08	16.10
L-34	1555	5.22	YES	55.89	42.32	39	16.59	25.34	16.87
L-35	1425	5.63	YES	54.37	41.86	39.92	14.45	27.43	14.44
L-36	1562	6.43	YES	57.27	44.58	39.82	17.47	26.63	17.45
L-37	1479	7.47	YES	56.52	43.19	40.86	15.67	27.56	15.64

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TABLE II Continued

Run	T_B °F	U_0 , ft/sec	RECY	T.L.I. BTU H_0 , m, ft ² , °F	T.L.II H_0	T.L.I. H_0 , conv	T.L.I. H_0 , Rad	T.L.II. H_0 , conv	T.L.II. H_0 , Rad
L-38	1682	9.01	YES	59.60	48.60	39.33	20.27	21.31	20.23
L-39	1619	9.45	YES	59.95	49.68	41.24	15.17	31.00	18.68
L-40	1645	9.01	YES	61.15	51.50	42.10	19.15	32.38	19.12
L-41	16.28	10.06	YES	60.77	51.04	42.21	18.76	32.31	18.73
L-42	16.22	10.03	YES	60.44	50.53	41.56	18.58	31.98	18.55
L-43	15.98	9	YES	13.38	54.65	45.22	18.17	36.50	18.19
L-44	1443	8.85	YES	58.01	47.78	43.28	14.75	33.07	14.76
L-45	1660	10.07	-	59.12	50.66	39.54	19.58	31.10	19.55
L-46	1649	7.84	-	59.27	52.02	40.65	19.22	32.82	19.20
L-47	1643	6.91	-	59.66	52.98	40.03	19.63	33.37	19.62
L-48	1545	5.91	-	58.21	51.00	40.81	17.40	33.62	17.38
L-49	1528	5.21	-	57.27	49.60	40.26	17.00	32.42	16.98
L-50	1510	7.70	-	57.57	47.83	40.94	16.57	31.28	16.55

TABLE II Continued

Run	T _B °F	U _o , ft/sec	RECY	T.L.I. BTU Ho, m, ft ² , °F	T.L.II Ho	T.L.I. Ho, conv	T.L.I. Ho, Rad	T.L.II. Ho, conv	T.L.II. Ho, Rad
M-1	1459	7.22	-	57.03	46.39	40.50	16.27	30.17	16.21
M-2	1614	8.67	-	61.17	50.80	42.17	19	31.83	18.97
M-3	1481	9.76	YES	58.27	47.94	42.35	16.02	31.95	15.99
M-4	1425	7.07	YES	56.38	44.45	41.87	14.51	29.96	14.49
M-5	1465	6.20	YES	57.81	43.77	42.07	15.74	26.05	15.71
M-6	1342	5.17	YES	54.73	41.13	41.38	13.35	27.80	13.33
M-7	1383	4.59	YES	54.43	39.65	40.30	14.14	25.54	14.11
M-8	1539	9.68	YES	59.40	49.74	42.12	17.29	32.48	17.26
M-9	1592	8.42	YES	62.45	53.85	43.95	18.50	35.38	18.47
M-10	1417	6.96	YES	61.97	52.06	47.30	14.77	37.32	14.75
M-11	1450	6.14	YES	62.42	51.33	46.98	15.45	35.90	15.43
M-12	1404	5.39	YES	62.02	48.91	47.46	14.55	34.39	14.53
M-13	1415	4.53	YES	61.70	47.98	41.93	14.77	33.24	14.74
M-14	1573	5.88	YES	60.87	50.88	42.75	18.12	33.79	18.09
M-15	1510	6.15	YES	60.58	49.71	43.86	16.72	33.02	16.69
M-16	1443	6.42	YES	59.92	48.89	44.62	15.30	33.61	15.27
M-17	1313	6.62	YES	56.68	43.92	43.88	12.80	31.15	12.77
M-18	1531	7.59	-	56.41	45.67	39.25	17.16	28.54	17.13
M-19	1617	6.43	YES	59.82	50.76	41.29	18.52	32.26	18.50
M-20	1634	5.96	YES	60.94	52.06	42.01	18.94	33.15	15.91
M-21	1639	5.17	YES	61.73	52.92	42.74	18.99	33.96	18.97
M-22	1631	5.17	YES	61.82	52.41	42.31	19.51	32.92	19.48
M-23	1600	4.66	YES	62.36	51.93	43.48	18.77	33.18	18.74
M-24	1370	6.04	YES	57.59	57.25	43.73	13.86	43.38	13.87
M-25	1371	6.49	YES	54.76	54.83	40.89	13.88	40.94	13.89
M-26	1524	8.14	YES	57.35	56.28	40.38	16.97	39.30	16.98
M-27	1466	9.17	YES	57.51	56.22	41.79	15.73	40.48	15.73
M-28	1426	6.97	YES	58.04	48.97	43.09	14.95	34.03	14.94
M-29	1445	6.15	YES	58.12	49.12	42.78	15.34	33.71	15.33
M-30	1523	8.13	YES	59.67	50.56	42.76	16.91	33.67	18.59
M-31	1439	8.92	YES	58.06	48.75	42.95	15.11	33.65	15.09
M-32	1382	6.36	YES	60.72	48.95	41.13	14.10	34.87	14.08
M-33	1393	5.71	YES	58.50	45.39	44.19	14.31	31.10	14.29
M-34	1440	6.62	YES	57.94	44.67	42.70	15.24	29.47	15.22
M-35	1361	4.56	YES	57.31	43.83	43.56	13.74	30.10	13.72
M-36	1472	6.75	YES	55.53	42.42	39.65	15.89	26.65	15.87
M-37	1551	7.54	YES	56.47	43.54	38.87	17.57	26	17.54

TABLE II Continued

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Run	T _B °F	U _o ,ft/sec	RECY	T.L.I. BTU ₂ , °F Ho,m,ft ² , °F	T.L.II Ho	T.L.I. Ho, conv	T.L.I. Ho, Rad	T.L.II. Ho, conv	T.L.II. Ho, Rad
N-1	1410	4.73	YES	56.76	40.69	41.32	14.84	25.59	14.79
N-2	1417	5.21	YES	55.14	39.42	40.18	14.96	24.51	14.91
N-3	1581	6.32	YES	55.64	43.29	37.13	18.51	24.82	18.46
N-4	1566	7.36	YES	55.51	44.61	37.34	18.16	26.49	18.12
N-5	1625	8.58	YES	57.42	48.00	37.37	19.56	28.49	19.52
N-6	1594	9.51	YES	58.33	48.65	39.51	18.82	29.87	18.75
N-7	1434	4.68	YES	58.52	50.34	43.21	15.31	25.05	15.28
N-8	1411	5.46	YES	57.17	49.14	42.33	14.54	34.32	14.81
N-9	1483	6.22	YES	56.48	48.86	40.56	16.33	32.57	16.30
N-10	1505	7.16	YES	55.78	49.30	38.98	16.80	32.53	16.77
N-11	1559	8.36	YES	57.87	51.63	39.16	18.70	32.96	18.67
N-12	1516	9.43	YES	57.87	51.85	40.81	17.06	34.82	17.03
N-13	1464	7.82	yes	59.54	51.28	43.69	15.94	35.35	15.92
N-14	1457	7.82	YES	57	48.30	41.23	15.78	32.54	15.76
N-15	1446	7.75	YES	56.22	41.38	40.99	15.22	31.18	15.20
N-16	1420	7.51	YES	56.27	46.28	41.60	14.67	31.62	14.65
N-17	1440	8.91	YES	51.54	43.03	36.50	15.03	28.02	15.01
N-18	1489	7.88	YES	54.69	45.92	38.59	16.10	29.85	16.07
N-19	-	-	YES	-	-	-	-	-	-
N-20	-	-	YES	-	-	-	-	-	-
N-21	1437	6.21	YES	57.70	51.47	42.52	15.18	36.31	15.16
N-22	1426	7.02	YES	56.56	46.12	41.62	14.94	31.19	14.93
N-23	1566	8.13	YES	57.62	50.91	39.73	17.90	33.03	17.88
N-24	1485	8.94	YES	57.85	51.35	41.72	16.13	35.24	16.11
N-25	1348	8.04	YES	62.48	54	49.03	13.46	40.56	13.44
N-26	1415	5.93	YES	59.88	50.64	45.14	14.73	35.93	14.72
N-27	1461	7.09	-	54.42	46.37	38.73	15.69	30.69	15.68
N-28	1581	6.39	YES	61.42	58.81	43.13	18.29	40.52	18.29
N-29	1541	6.32	YES	62.19	59.34	44.80	17.39	41.94	17.39
N-30	1539	6.34	YES	62.93	60.37	45.57	17.36	43.01	17.36
N-31	1497	7.11	YES	62.03	59.80	45.19	16.44	43.36	16.44
N-32	1585	8.19	YES	63.14	60.75	44.81	18.33	42.42	18.33
N-33	1512	9.37	YES	60.84	58.07	44.18	18.66	41.41	16.66

Each time a bubble sweeps by, the packet is replaced and the process repeats itself. The packet to surface heat transfer is viewed in terms of two resistances in series; namely, the gas film resistance between the packet and the tube wall and the resistance within the packet. Baskakov, et al⁽²¹⁾, suggested that for the case where the transfer surface temperature remains constant during the period of contact with the pack,

$$h_{pc} = \frac{1}{R_c + 0.5 R_e} \quad (19)$$

where

R_c = Contact thermal resistance due to gas film between heat transfer surface and packet.

R_e = Thermal resistance due to emulsion packet.

The value of R_e is given by⁽²¹⁾:

$$R_e = \sqrt{\frac{\pi \tau}{k_e \rho_s (1 - \epsilon_{mf}) C_{ps}}} \quad (20)$$

where

k_e = Effective thermal conductivity of the emulsion packet, BTU/hr ft °F

ρ_s = Density of solid, lbs/ft³

C_{ps} = Heat capacity of solid, BTU/lb °F

ϵ_{mf} = Minimum fluidization void fraction

τ = Average emulsion packet residence time, sec

The effective thermal conductivity can be estimated from⁽²²⁾:

$$k_e = k_e^0 + 0.1 \rho_g C_{pg} d_p U_{mf} \quad \text{and} \quad (21)$$

$$R_c = 1/h_{film} = \delta'/k_g \quad (22)$$

where

h_{film} = Gas film heat transfer coefficient

k_g = Gas thermal conductivity, BTU/hr ft °F

δ' = Average gas film thickness

The average gas film thickness lies in the range of $dp/6 < \delta' > dp/10$ ⁽²³⁾ Recently, Xavier and Davidson⁽²²⁾ fitted their fine and intermediate size particle data by using $\delta' = dp/10$ for single vertical surface and $\delta' = dp/4$ for horizontal surfaces in tube arrays. Assuming $\delta' = dp/4$ and substituting Equations 20 and 22 into Equation 19, we have:

$$h_{pc} = \frac{1}{\frac{dp}{4kg} + 0.5 \sqrt{\frac{\pi \tau}{k_e P_s (1-\epsilon) C P_s}}} \quad (23)$$

The value of τ may be calculated from⁽²⁴⁾:

$$\tau = \frac{1}{15} \frac{\rho_s C_p}{kg} dp^2 \quad (24)$$

2.2.3 Gas Convective Heat Transfer

Since in the present study it is difficult to measure gas convective heat transfer coefficient, h_{gc} , at incipient fluidization condition, it was assumed that Baskakov and Suprun⁽²⁵⁾ correlation given below could be used to calculate h_{gc} under our existing bed conditions:

$$Nu_{conv}^* = 0.0175 (Ar)^{0.46} Pr^{0.33} \quad (25)$$

where

$$\begin{aligned} Nu_{conv}^* &= \text{Convective Nusselt Number at incipient fluidization condition} \\ &= h_{gc} \cdot dp / kg \\ Pr &= \text{Prandtl Number} \\ &= C_{pg} \cdot \mu_g / kg \\ Ar &= \text{Archimedes Number} \\ &= g \cdot dp^3 \cdot \rho_g (P_s - P_g) / \mu_g^2 \\ g &= \text{Acceleration due to gravity} \\ \mu_g &= \text{Gas viscosity, lb/hr ft} \\ \rho_g &= \text{Gas density, lbs/ft}^3 \end{aligned}$$

For air, under existing fluidized bed conditions, $Pr \cong 0.65$

$$Nu_{conv.}^* = 0.0152 (Ar)^{0.46}$$

or

$$h_{gc} = \frac{0.0152 \text{ kg}}{dp} (Ar)^{0.46} \quad (26)$$

2.2.4 Bubble Convective Heat Transfer

In order to determine the bubble induced heat transfer, in Figure 15 we have plotted convective Nusselt Number based on tube diameter, Nu_D , as a function of tube Reynolds Number at $U_o = 3U_{mf}$, Re_D . This implies that in our case h_b is essentially negligible

$$h_b \cong 0 \quad (27)$$

Substituting Equations 23, 26, and 27 into Equation 18, we have

$$h = \frac{(1-f_o)}{\frac{dp}{4kg} + 0.5 \sqrt{\frac{\pi \tau}{keP_s(1-\epsilon_{mf})Cp_s}}} + \frac{(1-f_o)}{dp} \cdot 0.0152 \text{ kg} (Ar)^{0.46} + h_r \quad (28)$$

Therefore, conduction-convection component of the overall outside heat transfer coefficient can be written as follows:

$$h_{cond/conv} = \frac{(1-f_o)}{\frac{dp}{4kg} + 0.5 \sqrt{\frac{\pi \tau}{keP_s(1-\epsilon_{mf})Cp_s}}} + \frac{(1-f_o)}{dp} \cdot 0.0152 \text{ kg} (Ar)^{0.46} \quad (29)$$

Theoretical (Equation 29) and experimental values of the conduction-convection component of the overall outside heat transfer coefficient, for several randomly selected test runs, are given in Table III. The agreement between calculated (Equation 29) and experimental values is within $\pm 16\%$. It should be noted that in the literature^(1,21-23) different values of the effective thickness of the gas film, δ' , adjacent to the immersed tube have been used to fit the experimental results.

TABLE III
Comparison of Heat Transfer Data With Equations (29) and (29A)

RLM	T.L.I. hc,exp.	T.L.I. hc,cal Eq.(29)	T.L.II hc,exp.	T.L.II hc,cal Eq.(29A)
K-3	46.62	50.71	45.21	39.57
K-6	40.80	45.72	30.56	33.85
K-7	39.36	39.72	29.26	29.63
K-8	37.53	34.98	27.73	26.26
K-12	35.52	37.02	37.04	29.28
K-17	37.33	42.19	29.23	30.89
K-20	37.34	39.35	29.96	30.22
L-2	42.33	49.41	31.93	38.20
L-4	44.41	47.09	33.09	34.71
L-9	41.61	40.19	27.79	31.00
L-16	36.34	35.73	23.02	26.11
L-38	39.33	39.07	28.31	31.12
M-5	42.07	40.88	28.05	30.48
M-11	46.98	54.52	35.90	40.86
M-13	46.93	36.77	33.24	25.61
M-19	41.29	46.33	32.26	34.82
M-26	40.38	39.40	39.30	30.86
M-36	39.65	44.26	26.65	33.59

Since Equation 29 requires the knowledge of the fraction of time which the immersed tube spends in contact with the emulsion phase ($1-f_o$), this quantity gives an approximate indication of average voidage in the vicinity of the tube. To a first approximation, $1-f_o \cong 1-\delta^{(1,24)}$. In order to obtain a general correlation for $1-\delta$, we have replotted $1-\delta (= Lmf/Lf)$ as a function of the excess gas velocity, U_o-U_{mf} . Figure 16 presents such a plot. The following equation

$$1-\delta = 0.49 + \frac{1.6}{(U_o-U_{mf}) + 3.1373}$$

gives the best fit line through the data points. Therefore,

$$(1-f_o) = 0.49 + \frac{1.6}{(U_o-U_{mf}) + 3.1373} \quad (30)$$

As can be seen from Table III, the experimentally obtained heat transfer coefficient values for Test Loop II is always lower than that for Test Loop I. It should be noted that Test Loop II was located in the center of the bed whereas Test Loop I was located near the wall. This means that on the average Test Loop II spends less time in contact with the emulsion phase and more time with the bubble phase. This is equivalent to assuming that frequency of the bubble passage through the center of the bed is higher than that near the wall. Assuming $(1-f_o) \cong 0.5$ for Test Loop II, Equation 29 becomes:

$$h_{conv} = \frac{0.5}{\frac{dp}{4kg} + 0.5} \sqrt{\frac{\pi - \tau}{kgPs(1-\epsilon_{mf})\zeta ps}} + \frac{0.0076 \text{ kg}}{dp} (Ar)^{0.46} \quad (29A)$$

Table III compares calculated (Equation 29A) and experimental values of heat transfer coefficient for Test Loop II. In most cases, the agreement between the two values is within $\pm 20\%$.

2.2.5 Correlation of Data

Several correlations for heat transfer in a fluidized bed exists^(1-2,18-19,26). These correlations may be broken down into three groups: correlations for single horizontal tubes immersed in the bed, correlations for vertical tubes immersed in the bed, and correlations for the containment walls. It should be pointed out that most of these correlations do not cover the temperature range and entire particle size range found in fluidized bed combustors. Furthermore, the work of the various authors is scattered over wide ranges of the variables affecting fluidized bed heat transfer. Therefore, the general applicability of many of these correlations to design of a large fluidized bed is limited.

Heat transfer to horizontal tubes immersed in a fluidized bed has had the least amount of experimental work of the above mentioned three groups. The three empirical correlations for calculating heat transfer coefficients between horizontal immersed tubes and fluidized bed are: 1) Vreedenberg⁽²⁷⁾, 2) gel'perin et al⁽¹⁹⁾, and 3) Petrie, et al⁽¹⁹⁾.

Vreedenberg Correlation: $\frac{U_o d_t \rho_p}{\mu_g} > 2550$

$$\frac{h_c d_t}{\text{kg}} = 420 \frac{U_o d_t \rho_p}{\mu_g} \frac{\mu_g^2}{d_p^3 \rho_p^2 g} \text{Pr}^{0.3} \quad (31)$$

where

- h_c = Total heat transfer coefficient (excluding radiation component), BTU/hr ft² °F
- d_t = Tube diameter, ft
- kg = Thermal conductivity of gas, BTU/hr ft °F
- Pr = Prandlt Number, dimensionless
- = $C_p \mu_g / \text{kg}$

Figure 17 shows SSU test series results plotted according to Equation 31*. It is quite clear that our results do not correlate at all with Equation 31. Vreendenberg's correlation, Equation 31, was modified by Andeen and Glicksman⁽²⁹⁾:

$$\frac{h_c d_t}{kg} = 900(1-\epsilon) \frac{U_o d_t \rho}{\mu g} P^{0.326} \frac{\mu g^2}{d_p^3 g \rho_p^2} Pr^{0.3} \quad (32)$$

Figure 18 shows SSU test series results plotted according to the modified Vreendenberg correlation, Equation 32. This indicates that our experimental data does not fit the modified Vreendenberg correlation of Equation 32. Finally, SSU test series results were compared with:

A. Gel'perin et al correlation⁽¹⁹⁾:

$$\frac{h_c d_p}{kg} \left(\frac{\epsilon}{6(1-\epsilon)} \right) = 0.815 \left(\frac{U_o d_p \rho_g}{\mu g} \right)^{0.32} Pr^{0.33} \left(\frac{1}{6(1-\epsilon)} \right)^{0.32} \quad (33)$$

B. Ainshtein's correlation modified by Andeen and Glicksman⁽²⁹⁾:

$$\frac{h_c d_p}{6(1-\epsilon)kg} = 0.487 \left(\frac{G \cdot d_p}{\mu g \epsilon} \right)^{0.277} Pr^{0.33} \quad (34)$$

where

ϵ = Bed voidage
 G = Gas mass flux

Figures 19 and 20 show SSU test series results plotted according to the correlation of Equations 33 and 34, respectively. It is quite evident that SSU experimental heat transfer results do not fit any of the tested correlations for calculating heat transfer coefficient between immersed tubes and a fluidized bed.

Figure 17

Figure 18

2-41

Figure 20

One of the most commonly used dimensionless groups to correlate heat transfer data is the particle Reynolds Number. All the SSU heat transfer data of the present investigation are plotted in Figure 21 it appears that Nusselt Number vs. particle Reynolds Number. From Figure 21 it appears that Nusselt Number may be proportional to some power of the Reynolds Number. However, due to narrow range of Reynolds Number investigated, no attempt was made to find a line of the best fit through the data points. Another dimensionless group which was tested for correlation was the Archimedes Number,

$$\frac{\bar{h} dp}{k_g} \rho_g (\rho_p - \rho_g) / \mu_g^2$$

which can be defined as the modified friction factor or drag coefficient. Figure 22 is a plot of the Nusselt Number vs. the Archimedes Number for all the data of the present investigation. The solid line of Figure 22 is the line representing the following correlation:

$$\frac{\bar{h} dp}{k_g} = Nu_{conv} = 0.27 \frac{\bar{h} dp}{k_g} \rho_g (\rho_p - \rho_g) / \mu_g^2 \quad (35)$$

where

$$\frac{\bar{h} dp}{k_g} = 1 / (\sum_i \frac{1}{dp_{ai}})$$

= Harmonic mean particle diameter, ft

g = Gravitational constant, ft/hr²

hc = Total overall heat transfer coefficient (excluding radiation) BTU/hr ft² °F

Nu_{conv} = Nusselt Number

The correlation of Equation 35 was obtained using multiple regression analysis with a correlation coefficient $r = 0.87$. It is quite evident that the present data can be correlated quite well in a form as shown in Equation 35 suggesting Archimedes Number as a correlating parameter.

2-43

Figure 21

2-44

Figure 22

At $T = 1550^{\circ}\text{F}$, we have $\rho_g = 0.02 \text{ lbs/ft}^3$, $k_g = 0.046 \text{ BTU/hr ft } ^{\circ}\text{F}$,
 $\rho_p = 145 \text{ lbs/ft}^3$, $\mu_g = 0.11 \text{ lb/ft hr}$, $g = 4.173 \times 10^8 \text{ ft/hr}^2$.
 Substituting values of various parameters in Equation 35 and
 simplifying, we have

$$h_c = 65.733 \text{ kg } \rho_s^{0.27} / d_p^{0.19}$$

or

$$h_c = 11.591 / d_p^{0.19} \quad (36)$$

This means that the total heat transfer coefficient (excluding radiation component) is a very weak function of particle diameter. This agrees quite well with the experimental heat transfer coefficient data plotted in Figure 13. Moreover, Equation 35 does not show any dependence of the heat transfer coefficient upon the superficial gas velocity (see Figure 14). Figure 13 is a plot of the Nusselt Number vs. the Archimedes Number using arithmetic mean particle diameter ($\bar{d}_p = \sum X_i d_{p_i}$). It is quite evident that the functional relationship expressed by Equation 35 is independent of the type of technique used to estimate the mean particle diameter. For comparison purpose, B&W data obtained from Reference 28 are also plotted on Figure 22.

According to two-phase theory of fluidization, the excess gas above that required to fluidize a bed of particles passes through the bed in a bubble phase. This means that the heat transfer coefficient (excluding radiation component) should increase as the fluidizing gas rate is increased. This is quite possible due to increased contribution of the bubble induced particle convective component into the heat transfer mechanism while assuming, to a first approximation, the gas convective component to be independent of the fluidizing gas rate. Figure 24 shows a plot of the total heat transfer coefficient (excluding radiation

Figure 23

Figure 24

component) vs. U_o/U_{mf} . It is quite possible that both the interphase gas convective and particle convective components of heat transfer are little influenced by gas flow rate beyond the point of fluidization in the present investigation.

2.3 Combustion Efficiency

2.3.1 General Theory

One of the main objectives of the fluidized bed combustion technology is to develop an understanding of the interrelationship of critical operating variables affecting combustion efficiency. Theoretically, carbon combustion efficiency can be defined as follows:

$$\eta_{\text{carbon}} = 1 - \frac{X_D \cdot W_D + X_E \cdot W_E}{X_f \cdot W_f} \quad (37)$$

where

- X_D = wt. fr. of carbon in overflow stream
- W_D = overflow stream rate, lbs/hr
- X_E = wt. fr. of carbon in elutriated stream
- W_E = Elutriated stream flow rate, lbs/hr
- X_f = wt. fr. of carbon in coal feed stream (on a dry basis)
- W_f = Coal feed stream rate (on a dry basis), lbs/hr
- η_{carbon} = Carbon combustion efficiency

Alternately, the overall combustion efficiency can be defined as follows:

$$\eta_{\text{C(overall)}} = 1 - \frac{(C_D \times 14086) + (C_E \times 14086) + (CO \times 4344)}{(W_C \cdot \text{HHV})} \quad (38)$$

where

- C_D = solid carbon loss rate in overflow stream, lbs/hr
- C_E = solid carbon loss rate in elutriated stream, lbs/hr
- CO = carbon monoxide flow rate in flue gas (due to incomplete combustion), lbs/hr
- W_C = coal feed rate, lbs/hr
- HHV = higher heating value, BTU/lb

Equation (38) assumes that there is no loss of volatile hydrocarbons in the flue gas due to incomplete combustion. The factors that influence the overall combustion efficiency are: (1) loss of carbon in the elutriated solids (governed by bed carbon loading, bed particle size distribution and superficial gas velocity); (2) loss of carbon in the overflow stream (which is determined by bed carbon loading and overflow withdrawal rate); and (3) loss of carbon as CO due to incomplete combustion.

The important factors which affect bed carbon loading are:

Gas exchange rate between bubble and emulsion phase

Diffusion of oxygen through the emulsion phase to the surface of a combustible particle

Particle shrinkage rates due to combustion

It should be noted that an increase in bed carbon loading is a direct result of the reduction in the burning rates of coal particles. A particle combustion model depends on such factors as excess air, bed fluid mechanics, and feed particle size distribution, all of which play an important role in determining the overall combustion efficiency.

In order to consider the fluidized combustion of coal it is necessary to consider the processes by which a single particle of coal is burned. A number of modeling studies relating to the fluidized bed combustion of coal particles have been reported⁽³⁰⁻³⁶⁾. However, due to its complexity very little is understood about fluid bed combustion.

When coal is heated to temperatures in excess of 1200°F, it undergoes a rapid devolatilization during which it loses most of its volatiles. Little is known about the detailed processes of volatile evolution and combustion in a fluidized bed. It can be estimated that under fluidized conditions a 100 μ particle will take approximately 0.01 second to reach 95% completion of the devolatilization process, whereas a 1 mm particle will take approximately 1 second to devolatilize. The volatile release rate and its subsequent combustion plays an important role in determining the oxygen and temperature profiles in the bed and the proportion of the volatile heat released in the bed and in the freeboard.

Several theories have been presented to explain the combustion of residual char particles in a fluidized bed⁽³⁰⁻³⁶⁾. In general, a carbon particle can be assumed to be surrounded by a boundary layer of stagnant gas. Oxygen diffuses through this layer from the free stream and reacts at the particle surface to form combustion products which diffuse back out. The following reactions are possible:

It should be noted that reactions (39) and (40) are highly exothermic, whereas reaction (41) is endothermic. Whether and where the above reactions take place is not completely understood. In general, they should depend on temperature, coal type, particle diameter, etc.

2.3.2 Oxygen Mass Transfer

The rate of reaction of a spherical particle of char involves two mechanisms: (1) the diffusion of oxygen through the boundary layer surrounding the particle, and (2) the rate of chemical reaction between oxygen and char on the particle surface. In a diffusionally controlled mechanism, it is generally assumed that the chemical reactions occurring at the surface of the char particle are sufficiently rapid that the combustion is controlled by the rate of diffusion of oxygen through the boundary layer. Whereas a chemically controlled particle combustion mechanism assumes that the chemical reactions occurring at the surface of the char particle are slow compared to the rate of diffusion of oxygen through the boundary layer. Furthermore, the particle can be assumed to burn either at constant density and decreasing diameter, or at constant and decreasing density.

Under fluidization conditions, the range of particle sizes and bed temperatures (1400°F - 1600°F), are such that either diffusion or chemical rate control can be significant. Moreover, char has a porous structure and char particles can burn internally within the pores as well as on the free surface. Therefore, it is quite possible that a true mechanism lies somewhere in between the two mechanisms mentioned above.

In order to develop an overall systems model for fluidized combustion of coal, one must couple an appropriate model of coal-particle combustion with the model of the fluidized bed. The generally accepted picture of fluidized beds postulates a movement of bubbles with no solids in them through an emulsion phase. This emulsion phase, which contains all of the carbon particles, has the same void fraction as a bed at incipient fluidization. In addition, all air that flows through the bed in excess of the minimum fluidization flow is assumed to pass through the bed in

this form of bubbles. Therefore, the rate of transfer of oxygen from the bubble phase to an emulsion phase should have a strong influence on the combustion hence the amount of heat being released in the bed.

A number of correlations have been developed to predict gas interchange rates between bubble and emulsion phases in fluidized bed^(1,2,37). This means that the picture of interphase transfer from bubbles to bed is by no means clear. Defining cross-flow ratio as follows:

$$X_b = \frac{(K_{be})_b L_f}{U_b} \quad (43)$$

where

$(K_{be})_b$ = overall interchange coefficient between the bubble and the emulsion, 1/sec

This has a physical interpretation as the number of times the bubble gas is replaced on passage through the bed. Assuming $(K_{be})_b$ is of the form^(1,2):

$$(K_{be})_b = 4.5 \left(\frac{U_{mf}}{d_b} \right) + 5.85 \left(\frac{Dg^{1/2} g^{1/4}}{d_b^{5/4}} \right) \quad (44)$$

for fast bubbles.

where

Dg = diffusion coefficient of oxygen, um^2/sec .

$$= 0.015 T_B^{1.81} \frac{\left(\frac{M_1 + M_2}{M_1 M_2} \right)^{1/2}}{P (T_{C_1} T_{C_2})^{0.14} (V_{C_1}^{0.4} + V_{C_2}^{0.4})^2}$$

where

P	=	total pressure
M_1	=	Mol. wt. of oxygen
M_2	=	Mol. wt. of inert gas (assumed to be Nitrogen)
T_{C1}	=	Critical temperature of oxygen, ok
T_{C2}	=	Critical temperature of Nitrogen, ok
V_{C1}	=	Critical volume of oxygen, cc/g _{mole}
V_{C2}	=	Critical volume of Nitrogen, cc/g _{mole}
T_B	=	bed temperature, ok

Interchange rates for slow bubbles have also been developed. Davidson and Harrison⁽¹⁾ have derived the following theoretical equation for an average oxygen concentration above the bed:

$$\frac{[O_2]_{L_f}}{[O_2]_o} = \beta e^{\beta' - X_b} + \frac{[1 - \beta e^{\beta' - X_b}]^2}{[k^1 + 1 - \beta e^{\beta' - X_b}]} \quad (45)$$

$[O_2]_{L_f}$ = oxygen concentration in flue gas above bed

$[O_2]_o$ = inlet oxygen concentration

β' = $U_o - U_{mf} / U_o$

k^1 = $\frac{12.m.Sh.Dg.dp}{P_s dp_i^3 AB U_o}$

m = wt. of carbon

from which bed oxygen utilization and hence bed heat release rates can be calculated.

Equation (45) assumes that gas in the emulsion and bubble phases is perfectly mixed. Alternatively, the equation for plug flow of gas through the emulsion phase is more involved. The limiting conditions of Equation (45) may be noted:

$$\begin{aligned}
 \text{(i)} \quad \lim_{K^1 \rightarrow \infty} \frac{[O_2]_{L_f}}{[O_2]_0} &= \beta e^{-X_b} \quad (\text{for slow bubbles}) \\
 \text{(ii)} \quad \lim_{X_b \rightarrow \infty} \frac{[O_2]_{L_f}}{[O_2]_0} &= \frac{1}{1+K^1} \quad (\text{slow and fast bubbles in deep beds}) \\
 \text{(iii)} \quad \lim_{K^1 \rightarrow \infty} \frac{[O_2]_{L_f}}{[O_2]_0} &= 1 \quad (\text{for fast bubbles})
 \end{aligned}$$

According to Equation (45), a proportion of the entering oxygen will by-pass the bed and this proportion depends only on bed fluid-mechanics since X_b depends entirely on the properties of the bubbles and not on the nature of the reaction.

Table IV summarizes important results of the J, K, L, M, and N series runs obtained at various steady periods of operation. In order to evaluate the performance of SSU based on data from Test Series J through N, the variation in overall combustion efficiency with excess air, bed temperature, and active bed height is plotted in Figures 25 through 27. As can be seen from Figure 25 the overall combustion efficiency remains constant with increase in excess air level ranging from -10% to +15%. This is in contrast to the experimental results reported in the literature⁽³⁸⁻⁴¹⁾ which shows an increase in overall combustion efficiency with excess air level. The lack of dependence of overall combustion efficiency on excess air level can be explained by considering the influence of excess air on the combustibles loss (% of feed) as shown in Figure 28. The combustible loss for each run was estimated from the carbon loss due to CO in the exit flue gas, and the elutriated combustible loss due to carryover of unburnt carbon. Total combustible loss is expressed as a weight percentage of the total carbon in the feed. Although there is a

TABLE IV

TABLE IV
Summary of SSU Combustion Results - Freeman Coal

Run	$\frac{W_{O_2}}{lbs}$ m	$\frac{W_{N_2}}{lbs}$ m	$\frac{W_{H_2O}}{lbs}$ m	$\frac{W_{CO_2}}{lbs}$ m	$\frac{W_{CO}}{lbs}$ m	$\frac{W_{H_2}}{lbs}$ m	RECY	T_B °F	h_f inches	U ft/sec	X_{air} %	N_{CHEM}	Total Combust- ibles loss of fuel	Stack Gas CO PPM	$\sim$ dp, μ	$\frac{lbs}{psft^3}$
J-1	250.75	75.81	2951.23	54	49	1264.80	-	1415	29.77	8.71	50.83	91.27		660		132.35
J-2	250.75	75.81	2979.70	52	53	1354.37	-	1443	35.25	8.94	43.26	92.66		680		143.90
J-3	250.75	76.41	2623.85	52	46	1342.65	-	1506	32.24	8.18	34.04	94.15		540		147.46
J-4	184.41	57.77	1980.17	33	72	1006.28	-	1405	34.78	5.83	30.02	91.94		1480		131.69
J-5	186.68	77.01	2134.13	202	49	1082.09	YES	1389	33.06	6.22	38.15	98.10		2000		132.27
J-6	186.68	78.21	2296.48	658	61	1056.29	YES	1339	31.02	6.52	50.86	96.72		1870		120.46
J-7	232.61	95.05	2452	715	90	1322.74	YES	1413	36.15	7.32	34.71	95.87		2080		125.27
J-8	232.61	95.05	2528.92	741	50	1273.06	YES	1358	33.05	7.31	35.11	94.90		2450		120.88
J-9	279.10	113.08	2759.92	910	43	1533.89	YES	1433	34.88	8.37	22.42	94.67		1830		137.18
J-10	279.10	113.08	2973.46	948	70	1540.44	YES	1391	36.61	8.69	33.89	93.73		2200		136.89
J-11	279.10	113.08	2448.30	840	46	1508.14	YES	1474	27.66	7.65	10.19	94.63		2000		138.35

TABLE IV Continued

Run	$\frac{W}{10s}$ m	$\frac{W_2}{10s}$ m	$\frac{W_1}{10s}$ m	$\frac{W_F}{10s}$ m	$\frac{W}{10s}$ m	$\frac{W_2}{10s}$ m	RECY	T_B °F	h_f inches	U_0 ft/sec	X_{air} %	N_{CHEM}	Total Combust- ibles loss of fuel	Stack Gas CO PPM	$\sim$ dp,u	$\frac{lbs}{psft^3}$
K-1	243.68	115.80	2395.36	68	77	1364.10	-	1525	39.86	7.58	28.73	94.76	7.113	690	438	128.72
K-2	243.68	116.40	2376.72	67	55	1432	-	1634	30	7.90	18.92	93.11	9.036	340	438	129.45
K-3	143.68	60	2369.60	53	29	1452.47	-	1662	30.13	7.92	16.46	95.73	5.729	620	469	141.78
K-5	195.78	48	1748.40	42	41	1061.64	-	1470	43.55	5.40	8.66	90.58	12.557	2540	499	161.26
K-6	195.78	47.40	1933.32	28	27	1164.45	-	1482	35.72	5.95	20.18	95.35	5.945	1350	538	158.52
K-7	195.78	47.40	2888.59	36	32	1116.69	-	1469	29.86	6.32	29.96	92.99	9.33	1070	618	153.58
K-8	195.78	46.20	2237.42	34	33	1659.98	-	1420	35.23	6.64	40.90	91.18	11.892	990	711	136.77
K-9	243.68	58.20	2192.28	37	17	1317.41	-	1561	39.18	7.00	15.14	94.66	6.708	1490	600	153.49
K-10	243.68	58.20	2434.06	48	38	1322.73	-	1468	37.38	7.51	26.65	92.96	9.161	1000	665	153.25
K-11	243.68	58.20	2125.82	52	35	1344.05	-	1457	37.18	8.05	28.37	92.99	9.444	900	335	157.28
K-12	243.68	59.42	2830.56	46	36	1364.69	-	1477	37.30	8.65	27.05	94.69	6.974	400	638	156.31
K-13	231.58	69	2626.71	65	27	1623.60	-	1663	37.42	8.83	5.46	95.32	6.16	1070	612	146.19
K-14	291.58	70.20	2882.65	60	29	1624.77	-	1607	37.10	9.20	18.24	92.46	9.95	530	621	160.07
K-15	339.48	79.80	2924.79	81	24	1908.53	-	1691	35.06	9.97	1.33	89.26	13.84	1870	631	142.75
K-16	339.48	120	2904.41	675	32	1944.28	YES	1500	45.05	9.13	2.77	95.81	5.73	3440	592	123.94
K-17	195.78	70.20	1769.64	337	0.19	1100.54	YES	1545	31.06	5.70	10.12	94.66	8.103	4980	606	142.65
K-18	195.78	70.20	1932.49	435	46	1184.74	YES	1478	28.41	5.98	21.74	93.09	8.97	3020	502	143.65
K-19	243.30	87	2170.90	500	40	1378.64	YES	1483	28.49	6.77	3.96	93.96	7.81	4500	465	142.57
K-20	243.30	87	2341.14	754	32	1408.69	YES	1498	25.62	7.31	15	95.29	6.28	2500	612	142
K-21	243.30	87	2543.06	1947	51	1409.73	YES	1487	21.72	7.87	22.67	95.55	6.00	2000	417	141.10
K-22	291.58	103.20	2621.48	665	45	1593.69	YES	1470	29.97	8.12	5.72	95.05	6.69	3000	381	144.36
K-23	291.58	103.20	2613.36	671	55	1610.64	YES	1470	31.38	8.10	3.78	93.36	8.48	2500	333	144.01
K-24	243.68	115.80	2323.51	66	30	1258.20		1511	33.52	7.33	18.80	95.27	6.53	1520	530	131.41

TABLE IV Continued

Run	$\frac{W}{10s}$ m	$\frac{W_2}{10s}$ m	$\frac{W_1}{10s}$ m	$\frac{W_F}{10s}$ m	$\frac{W}{10s}$ m	$\frac{W_2}{10s}$ m	RECY	T_B °F	h_f inches	U ft ² /sec	X_{air} %	N_{CHEM}	Total Combust- ibles loss of fuel	Stack Gas CO PPM	$\sim$ dp, in	$\frac{lbs}{psft^3}$
L-1	248.60	116	2417.03	76	28	1549.79	-	1521	41.02	7.81	19.77	93.07	9.25	780	466	137.33
L-2	229.40	107	2368.14	312	52	1486.20	YES	1452	37.75	7.35	24.59	96.29	5.19	1300	459	137.57
L-3	229.40	107	2046.95	422	52	1448.66	YES	1466	40.09	6.47	9.67	95.14	7.27	3000	500	137.43
L-4	181.40	84	1890.37	297	28	1248.14	YES	1389	38.08	5.71	24.53	95.16	6.73	1950	502	135.10
L-5	296.60	139	2661.76	371	43	1912.94	YES	1598	41.65	8.98	3.99	96.33	5.08	2300	513	136.77
L-6	296.60	90	2947.87	432	78	1833.03	YES	1439	45.79	9.15	15.57	94.97	6.7	2800	360	137.79
L-7	296.60	90	3128.23	393	25	1938.02	YES	1600	33.54	10.49	26.79	96.59	4.43	700	480	150.79
L-8	296.60	90	2728.27	494	66	1905.63	YES	1635	33.04	9.35	8.04	95.61	6.10	2000	465	148.32
L-9	229.40	107	2417.49	257	60	1471.94	YES	1417	37.14	7.38	23.61	96.77	4.53	1500	575	123.78
L-10	229.40	107	2082.91	228	59	1421.43	YES	1464	36.24	6.59	10.70	95.53	6.37	4300	567	135.40
L-11	181.40	84	1939.63	211	59	1218.18	YES	1428	34.85	5.94	27.39	95.63	5.58	1500	508	132.03
L-12	181.40	84	1610.23	181	40	1127.64	YES	1445	34.54	5.05	6.38	95.29	6.52	5000	476	134.84
L-13	181.40	84	1532.57	261	73	1094.70	YES	1458	31.18	4.76	1.02	93.99	8.66	5600	544	137.58
L-14	181.40	84	1474.38	365	40	1082.49	YES	1456	27.56	4.09	-9.82	91.88	10.88	6000	575	137.59
L-15	181.40	84	1751.72	420	56	1136.28	YES	1400	25.62	5.21	16.71	93.21	9.07	4000	421	124.65
L-16	181.40	84	1895.36	312	41	1104.37	YES	1433	25.63	5.75	21.60	96.26	4.89	2000	715	128.92
L-17	181.40	84	2045.50	281	30	1123.66	YES	1458	26.63	6.31	32.62	96.91	4.64	1400	649	135.30
L-18	329.40	107	1867.07	380	33	1388.19	YES	1571	26.49	6.16	-2.85	95.30	7.08	5000	484	137.81
L-19	229.40	107	2043.58	484	46	1445.06	YES	1455	27.65	6.28	7.35	95.10	6.81	3000	434	135.27
L-20	229.40	107	2245.68	373	82	1393.23	YES	1400	26.22	6.66	12.73	95.09	6.97	3500	367	119.51
L-21	229.40	107	2436.49	282	69	1369.28	YES	1442	25.67	7.42	25.27	95.85	5.51	1800	570	125.25
L-22	229.40	53	2454.85	43	18	1397.52	-	1716	26.83	8.42	28.28	96.34	4.68	2500	558	150.97
L-23	181.40	41	1929.37	28	28	1178.73	-	1654	27.33	6.43	31.64	95.38	6.01	550	-	162.94
L-24	181.40	41	1846.50	28	37	1162.81	-	1645	28.25	8.15	23.51	95.81	5.28	1000	500	156.57
L-25	296.60	139	3103.10	574	59	1934.80	YES	1559	27.57	10.08	28.76	95.50	6.20	1500	483	135.83
L-26	296.60	139	2964.65	565	69	1812.44	YES	1541	27.35	9.49	18.02	95.56	5.76	1300	426	136.58
L-27	296.60	139	2771.61	523	64	1786.25	YES	1554	27.01	8.98	8.69	95.24	6.69	2900	406	135.68
L-28	247.07	104.40	2456.52	50	64	1452.14	-	1406	42.78	7.28	19.29	93.59	8.81	2200	439	127.50
L-29	177.70	75.60	1642.21	226	72	1127.65	YES	1439	41.68	5.0	9.08	88.34	15.37	5800	607	134.07
L-30	227.25	47.70	1407.20	466	116	1377.09	YES	1407	37.59	7.33	29.90	94.40	7.37	2200	649	144.97
L-31	227.25	47.70	2102.26	309	22	1351.31	YES	1498	37.27	6.75	10.49	95.30	6.48	3400	549	167.58
L-32	177.70	36.90	1878.74	189	36	1174.19	YES	1446	37.53	5.70	21.57	96.82	4.23	2300	518	156.23
L-33	177.70	36.90	1648.53	168	16	1124.75	YES	1482	38.44	5.22	9.43	96.52	5.29	4000	589	169.03
L-34	177.70	36.90	1620.14	230	26	1093.51	YES	1555	30.07	5.22	8.01	96.35	5.51	3100	525	167.95
L-35	177.70	36.90	1901.07	210	26	1099.46	YES	1425	35.67	5.63	25.41	96.46	4.83	1630	439	101.47

TABLE IV Continued

Run	$\frac{W_{O_2}}{m}$	$\frac{W_{H_2O}}{m}$	$\frac{W_{CO_2}}{m}$	$\frac{W_{H_2}}{m}$	$\frac{W_{CO}}{m}$	$\frac{W_{CH_4}}{m}$	RECY	T_B °F	h_f inches	U ft/sec	X_{air} %	N_{CHEM}	Total Combustibles loss of fuel	Stack Gas CO PPM	$\sim dp_{\mu}$	$\frac{lbs}{psft^3}$
L-36	227.25	47.70	2004.30	266	34	1340.84	YES	1562	30.55	6.43	9.19	97.48	3.87	2700	573	171.40
L-37	227.25	47.70	2446.28	320	44	1392.36	YES	1479	33.14	7.47	42.10	95.64	5.40	550	542	165.37
L-38	296.62	62.10	2663.48	228	36	1477.37	YES	1682	35.98	9.01	10.68	98.19	2.61	1500	625	146.36
L-39	296.62	62.10	2885.57	239	21	1745.95	YES	1619	33.13	9.45	20.06	97.21	4.00	880	557	165.32
L-40	296.62	62.10	2705.52	268	36	1721.54	YES	1645	36.71	9.01	15.83	94.33	7.78	2300	506	156.35
L-41	296.62	62.10	3051.48	256	38	1781.32	YES	1628	32.03	10.06	23.34	96.31	5.11	1000	497	161.57
L-42	296.12	125.10	3040.25	261	79	1766.87	YES	1622	32.23	10.03	23.64	95.82	5.55	800	496	137.51
L-43	296.62	125.10	2733.34	220	46	1719.96	YES	1598	35.25	9.	7.47	97.53	3.24	1750	433	136.91
L-44	296.12	125.10	2895.33	166	40	1535.42	YES	1443	42.80	8.85	14.41	96.65	4.36	1650	500	134.92
L-45	296.62	62.10	3012.72	57	35	1711.36	-	1660	35.02	10.07	1.71	94.56	7.11	600	518	163.95
L-46	227.25	47.70	2381.57	44	25	1362.73	-	1649	30.81	7.84	32.26	96.42	4.98	650	466	154.16
L-47	227.25	47.70	2083.79	34	30	1396.23	-	1643	33.39	6.91	10.44	94.66	7.21	1050	500	152.37
L-48	177.70	36.90	1891.10	34	8	1148.54	-	1545	35.10	5.91	30.70	92.96	9.00	700	477	165.34
L-49	187.85	36.90	1647.94	35	14	1148.28	-	1528	34.33	5.21	12.13	91.44	12.34	2850	445	167.65
L-50	247.07	104.40	2449.72	35	72	1453.98	-	1510	41.02	7.70	22.87	95.57	11.66	700	416	136.63

TABLE IV Continued

Run	$\frac{W_0}{\text{lbs}}$ m	$\frac{W_2}{\text{lbs}}$ m	$\frac{W_a}{\text{lbs}}$ m	$\frac{W_F}{\text{lbs}}$ m	$\frac{W}{\text{lbs}}$ m	$\frac{W_5}{\text{lbs}}$ m	RECY	T_B °F	h_f inches	U ft/sec	X_{air} %	N_{CHEM}	Total Combust- ibles loss of fuel	Stack Gas CO PPM	$\sim$ dp.v	$\frac{\text{lbs}}{\text{psft}^3}$
M-1	246.15	90	2330.58	57	46	1439.79	-	1489	42.79	7.22	16.37	88.97	12.98	1000	610	138.06
M-2	294.30	111	2620.33	294	99	1816.57	YES	1614	53.36	8.67	97.75	97.75	2.97	1100	449	139.22
M-3	294.30	111	3141.66	631	69	1755.43	YES	1481	14.81	34.16	27.76	94.80	6.73	800	568	139.44
M-4	226.89	84	2376.27	264	86	1359.48	YES	1405	41.01	7.07	25.67	96.72	4.76	1600	471	123
M-5	226.89	84	1971.28	175	71	1314.25	YES	1465	37.47	6.20	5.16	96.25	5.81	4800	599	135.06
M-6	178.74	66	1809.60	145	48	1038.84	YES	1342	35.24	5.17	22.55	96.65	4.56	2600	546	122.32
M-7	178.74	66	1533.15	126	44	1025.85	YES	1383	46.09	4.09	8.48	96.16	6.06	5400	623	117.93
M-8	299.80	111	2080.12	375	59	1775.51	YES	1539	28.89	9.68	28.78	97.41	3.56	1000	487	141.65
M-9	299.80	111	2556.53	277	57	1770.91	YES	1592	28.71	8.42	4.37	96.24	5.58	2800	439	141.30
M-10	233.80	84	2344.23	259	43	1395.98	YES	1417	32.21	6.96	27.47	96.54	4.23	700	417	137.61
M-11	233.30	84	1996.16	191	64	1358.87	YES	1450	31.87	6.14	4.03	97.49	3.15	500	425	140.37
M-12	185.80	66	1824.29	158	57	1172.51	YES	1404	31	5.39	20.88	96.33	5.26	2400	325	126.38
M-13	185.80	66	1563.23	122	16	1071.16	YES	1415	33.90	4.53	0.99	94.62	8.03	9500	726	122.74
M-14	233.30	56	1792.21	269	40	1349.87	YES	1573	27.14	5.88	-6.61	95.84	6.00	6000	443	160.75
M-15	233.30	56	1935.02	355	28	1377.85	YES	1510	26.64	6.15	0.87	94.64	7.27	4000	478	165.21
M-16	233.30	56	2097.54	493	38	1444.60	YES	1443	27.85	6.42	10.37	93.47	8.56	2000	463	152.28
M-17	233.30	56	2340.13	621	51	1394.63	YES	1313	26.55	6.62	21.20	90.02	13.29	2700	476	116.58
M-18	252.30	90	2350.79	71	31	1377.08	-	1531	36.31	7.59	8.33	93.80	7.66	1000	594	143.05
M-19	185.80	43	1972.09	162	28	1160.92	YES	1617	27.06	6.43	27.80	98.62	2.20	600	539	154.97
M-20	185.80	43	1801.08	149	28	1133.31	YES	1634	28.61	5.96	16.56	97.84	2.70	1400	585	144.79
M-21	185.80	43	1673.75	141	28	1124.47	YES	1639	27.23	5.57	7.01	97.76	3.40	2700	488	138.65
M-22	185.80	43	1552.61	143	32	1156.45	YES	1631	27.68	5.17	0.45	97.13	4.11	4200	480	137.89
M-23	185.80	43	1395.85	150	36	1054.27	YES	1600	26.81	4.66	-10.11	95.26	6.99	8710	477	148.01
M-24	226.89	84	2061.67	171	80	1239.94	YES	1370	35.56	6.04	10.91	95.28	6.90	5210	463	119.23
M-25	226.89	84	2246.50	185	74	1265.87	YES	1371	36.27	6.49	16.98	95.59	6.20	4100	535	116.01
M-26	294.30	111	2555.03	431	48	1091.88	YES	1524	38.14	8.14	1.96	92.08	9.35	2700	602	138.46
M-28*	226.89	84	2724.53	395	59	1380.92	YES	1426	32.46	8.10	48.27	95.64	6.11	1050	527	137.21
M-29*	226.89	84	2454.67	434	60	1349.13	YES	1445	36.26	7.29	28.51	94.62	7.55	2300	544	142.38
M-30*	294.30	111	2949.24	498	75	1708.36	YES	1523	27.19	9.30	20.60	96.44	4.52	800	435	142.24
M-31*	294.30	84	3381.47	696	75	1740.92	YES	1437	36.82	10.05	33.61	94.28	7.77	1100	527	

* Old Ben Coal

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TABLE IV Continued

Run	$\frac{W_{O_2}}{m}$	$\frac{W_{H_2O}}{m}$	$\frac{W_{CO_2}}{m}$	$\frac{W_{CO}}{m}$	$\frac{W_{H_2}}{m}$	$\frac{W_{CH_4}}{m}$	RECY	T_B °F	h_f inches	U ft/sec	K_{air} %	N_{CHEM}	Total Combust- ibles loss of fuel	Stack Gas CO PPM	$\sim$ dp, μ	$\frac{lbs}{psft^3}$
M-32	226.89	84	2158.46	282	81	1364.21	YES	1382	37.43	6.36	14.81	95.81	5.09	3200	396	124.02
M-33	204.74	64	1910.56	252	64	1272.80	YES	1393	39	5.71	12.86	95.30	6.52	3900	483	120.04
M-34	226.89	84	2161.99	245	70	1355.66	YES	1440	37.18	6.66	18.44	95.59	5.87	3000	471	125.65
M-35	176.43	59	1533.81	170	58	1024.63	YES	1361	37.91	4.56	4.36	95.68	6.59	4800	467	120.63
M-36	226.89	84	2128.60	209	68	1305.09	YES	1472	45.38	6.75	11.10	96.22	5.14	3500	534	137.85
M-37	246.15	90	2342.88	289	59	1542.89	YES	1551	37.54	7.54	16.13	95.26	6.67	1930	488	142.61

TABLE IV Continued

Run	$\frac{W}{10s}$ m	$\frac{W_2}{10s}$ m	$\frac{W}{10s}$ m	$\frac{W_F}{10s}$ m	$\frac{W}{10s}$ m	$\frac{W_5}{10s}$ m	RECY	T_B °F	h_f inches	U ft/sec	X_{air} %	N_{CHEM}	Total Combust- ibles loss of fuel	Stack Gas CO PPM	$\sim$ dp, μ	$\frac{lbs}{psft^3}$
N-1	177.40	66	1586.77	138	41	1089.11	YES	1410	37.09	4.73	12.38	92.31	11.14	8420		129.81
N-2	177.40	66	1772.39	145	40	1138.71	YES	1417	42.32	5.21	20.26	95.08	7.31	4030		131.30
N-3	225.40	84	1923.02	179	34	1406.20	YES	1581	37.73	6.32	2.79	90.89	12.08	4340		142.28
N-4	225.40	84	2291.66	198	60	1494.17	YES	1566	35.62	7.36	20.91	94.84	7.28	2170		143.55
N-5	292.60	111	2569.15	444	65	1838.52	YES	1625	34.70	8.58	5.33	93.55	8.63	3000		140.36
N-6	292.60	111	2917.95	634	51	1914.94	YES	1594	32.38	9.51	21.06	94.51	7.45	1700		142.75
N-7	177.40	66	1540.48	118	92	1040.24	YES	1434	30.01	4.68	6.41	97.71	3.64	3000		126.02
N-8	177.40	66	1853.50	121	40	1067.59	YES	1411	31.12	5.46	23.95	98.16	3.01	1700		128.01
N-9	225.40	84	2295.28	140	105	1254.10	YES	1483	31.89	6.22	2.61	98	5.91	2500		136.52
N-10	225.40	84	2295.28	243	30	1414.61	YES	1505	32.51	7.16	21.97	97.18	4.17	800		142.79
N-11	292.60	111	2564.70	458	40	1523.73	YES	1559	31.51	8.36	6.31	95.05	7.60	1600		135.93
N-12	292.60	111	3018.47	677	50	1895.36	YES	1516	36.18	9.43	23.88	94.89	6.49	500		137.93
N-13	268.41	66	2552.20	373	35	1738.31	YES	1464	30.47	7.82	13.46	94.18	7.94	2960		149.08
N-14	268.41	66	2554.81	401	50	1686.11	YES	1457	28.27	7.82	11.61	94.49	7.63	2740		143.38
N-15	268.41	66	2850.23	458	35	1653.81	YES	1446	32.64	7.75	13.61	94.67	7.10	2590		147.56
N-16	268.41	66	2526.08	621	69	1718.85	YES	1420	31.44	7.51	14.44	93.69	8.83	3300		157.60
N-17*	289.78	76	2927.27	677	50	1677.52	YES	1440	28.62	8.91	24.32	91.36	12.12	2000		130.15
N-18*	289.78	76	2542.10	565	42	1719.42	YES	1489	32.67	7.88	6.19	90.92	13.15	4100		138.59
N-19																
N-20																
N-21	225.40	84	2069.36	198	60	1376.70	YES	1437	33.47	6.21	8.46	97.41	3.87	1650		138.79
N-22	225.40	84	2323.68	212	36	1390.03	YES	1426	35.78	7.02	23.88	96.53	4.53	1250		139.47
N-23	292.60	111	2517.05	376	47	1794.02	YES	1566	40.12	8.13	4.02	93.74	8.44	3830		139.42
N-24	292.60	111	2898.37	680	73	1872.39	YES	1455	38.04	8.94	17.54	95.19	6.62	1600		141.76
N-25	225.40	84	2103.08	382	80	1380.67	YES	1348	33.35	6.04	18.05	96.04	5.65	1500		132.54
N-26	225.40	84	1961.19	290	57	1413.20	YES	1415	34.16	5.93	1.18	90.12	13.32	6210		136.48
N-27	244.60	90	2302.75		37	1483.52	-	1461	35.68	7.09	11.19	98.84	1.96	1000		142.83
N-28	227.63	84	1959.33	181	39	1447.12	YES	1581	34.72	6.39	5.74	97.65	3.69	2700		143.37
N-29	220.34	90	1994.65	195	36	1421.19	YES	1541	32.56	6.32	6.8	97.98	2.58	1900		139.89
N-30	220.34	84	2005.18	164	31	1398.39	YES	1539	34.79	6.34	7.04	98.36	2.17	1200		140.05
N-31	220.34	84	2312.35	194	42	1447.28	YES	1497	34.18	7.11	25.53	98.35	2.47	400		136.86
N-32	292.60	111	2539.58	318	43	1509.88	YES	1585	36.95	8.19	3.16	95.86	5.27	1500		139.76
N-33	292.60	111	3027.65	588	47	1917.17	YES	1512	33.58	9.37	23.23	93.43	8.74	600		137.02

* Old Ben Coal

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Figure 25

Figure 26

considerable scatter in the data points plotted in Figure 28, it is quite evident that an increase in the excess air from sub-stoichiometric up to a level of about 48% essentially has no effect on the combustibles loss, hence the overall combustion efficiency.

Figure 26 shows the influence of bed temperature on the overall combustion efficiency. In general, if the particle combustion is purely diffusion-limited, then the overall combustion efficiency should show little or no temperature dependence. However, for the kinetic limited case, the effect of bed temperature should be quite significant. Chemical kinetics should be governing with small particles as well as at low bed temperatures. As can be seen from Figure 26, results from Test Series J through N show virtually no temperature dependence of the overall combustion efficiency. This is indicative of a diffusion controlled char combustion occurring in the SSU combustor. Gibbs and Beer⁽³⁹⁾ observed significant improvement in the overall combustion efficiency with increasing bed temperature implying a stronger role played by combustion kinetics.

Since coal combustion is basically a chemical reaction involving oxygen transport, then the gas residence time or bed height should be an important parameter. It may be inferred that improved overall combustion efficiency should be achieved by resorting to increasing bed heights. Figure 27 shows no significant improvement in overall combustion efficiency with increasing bed height over a threefold range. Figure 29 shows that superficial gas velocity has no influence on the overall combustion efficiency.

Figures 25 through 27 shows that the recycle operation results in a 3% to 4% increase in the overall combustion efficiency value over that for no recycle runs. These results shows that the average value of overall combustion efficiency without recycle is about 92% whereas that for recycle runs is about 95%.

In order to investigate the influence of different coals on the performance of SSU fluidized bed combustor, two recycle runs (Test Series Runs N-17 and N-18) were carried out using Old Ben coal. Table IV-A summarizes important results of this study. From Table IV-A, it is quite evident that under the operating conditions used in this investigation, Old Ben coal is comparatively less reactive than the Freeman coal. The average value of overall combustion efficiency for two recycle runs with Old Ben coal is about 91% compared with about 96% for the Freeman coal. In addition, because of the longer burnout time for Old Ben coal, approximately 50% of the total input fuel energy is being released in the combined above bed regions for two recycle runs with Old Ben coal. These results demonstrate that different coals perform differently in a fluidized bed.

2.3.3 Carbon Elutriation

Loss by carryover of small carbon particles from the bed is commonly the major source of inefficiency. The problem arises from the fact that the fluidizing gas velocities employed usually exceeds the terminal velocities of the finer particles of the bed, and hence particles become entrained in the gas, and so are removed. Considerable amount of work has been reported in the literature on the development of elutriation rate coefficients^(1-3, 42-44). The rate of elutriation of fine particles is proportional to the concentration of fines in the bed and the gas path velocities. For char particles, the expression is of the form:

$$\frac{dW}{dt} = K^* A_B \frac{W}{W_B} \phi(dp) \quad (46)$$

where

- d_W = elutriation rate of char particles whose diameter is between dp and $dp + d(dp)$
- A_B = bed cross-sectional area, ft^2
- W = wt. of char in the bed, lbs

TABLE IV-A

Comparison of Old Ben Coal and Freeman
Coal Reactivity Results

	<u>Old Ben</u>	<u>Freeman</u>
Flammability ¹	1300°F*	1110°F*
TGA Burnout Time ²	16.1 min*	14.3 min*
SSU Overall Comb. Eff.	91 %	96 %

*Results from ASTM prepared coke buttons.

1 In pure oxygen - particle size -50 to +40 mesh.

2 15 mg sample - 1606°F air combustion.

W_B = wt. of total bed material, lbs
 $\phi(dp)$ = wt. fraction of char based on total wt. of char whose size is between dp and $dp+d(dp)$

The total elutriation rate is simply the integral of Equation 46:

$$W_E = \int_0^{\infty} K^* A_B \frac{W}{W_B} \phi(dp) \quad (47)$$

Several correlations have been published to calculate K^* (1-3, 42-44). Although none of the correlations predict specific elutriation rate constants satisfactorily over the range of operating conditions encountered in fluidized bed combustion, the correlation of Yagi and Aochi is given by the following relation (1,2):

$$K^* = \frac{\mu g}{dp} Fr (0.0015 Re_t^{0.06} + 0.01 Re_t^{1.2}) \quad (48)$$

where

$$\begin{aligned}
 Fr &= \text{Froud Number} \\
 &= (U_o - U_t)^2 g \cdot dp \\
 Re_t &= \text{Terminal Reynold's Number} \\
 &= \frac{U_t \cdot dp \cdot g}{\mu}
 \end{aligned}$$

The correlation of Yagi and Aochi appears to give the best correlation fit of data developed by BCURA and B&W. The terminal velocity of a non-spherical particle can be calculated using the following relation (3):

$$U_t = (0.843 \log_{10} \frac{\phi_s}{0.065}) \cdot U_{t, \text{spherical}} \quad (49)$$

where ϕ_s is the particle shape factor and $U_{t, \text{spherical}}$, the terminal velocity of a spherical particle, can be calculated using relationships given in Kunii and Levenspiel (92).

The dependence of total elutriation rate on superficial gas velocity is shown in Figure 30. A significant effect of superficial gas velocity on the value of total elutriation rates can be seen from Figure 30. A considerable increase in total elutriation rates can be seen for operation with fines recycle. In general, elutriation rates of particles should increase with increasing superficial gas velocity because of the increased range of sizes that can be elutriated. Combination of Equations 47 and 48 suggest $W_E \propto (U_o - U_t)^2$ whereas an approximate value of the slope of both lines in Figure 30 is 1.9. Figure 31 is a plot of char elutriation rates as a function of superficial gas velocity. Figure 2 shows the elutriation carbon loss (% of feed) as a function of superficial gas velocity. Due to considerable scatter in the data plotted in Figure 32, no definite trend can be observed.

The effect of bed temperature on total combustibles loss is shown in Figure 33. In general, increasing the bed temperature should reduce the level of combustibles leaving the bed because of the temperature effect on volatiles, CO, and/or char combustion kinetics. The lack of dependence of total combustibles loss on bed temperature (Figure 33) and excess air level (Figure 29) may indicate that a constant level of oxygen bypassing is occurring in the SSU fluidized bed.

The influence of excess air, freeboard gas velocity, and the average flue gas temperature in the freeboard on stack gas CO level is shown in Figures 34, 35, and 35A, respectively. The trends exhibited in Figures 34 and 35 are in general agreement with the reported results in the literature⁽³⁹⁻⁴¹⁾. However, considerable scatter exists in the data plotted in Figure 35A.

The influence of excess air on the carbon loss in elutriated stream is shown in Figure 35B. Although considerable scatter in the data exists, it is quite evident that increase in excess air

Figure 32

Figure 33

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Figure 35A

substantially reduces the carbon carryover rate. It should be noted that high level of carbon carryover rate with respect to recycle operation may primarily be due to nonmixing of recycle char in the bed.

2.3.4 Combustion in Bed

In order to design a fluidized bed combustor it is essential to be able to predict the amount of in-bed burning versus freeboard performance so that the heat transfer surface can be accurately designed for the heat release expected in each zone. For convenience, the SSU furnace is divided into six zones: namely, (1) the fluidized bed combustion zone or the active bed zone, (2) the splash zone (the zone between the active bed and the design bed level of 40"), (3) the freeboard zone, (4) the superheater zone, (5) the economizer zone, and (6) the furnace exit zone. Figure 36 shows a schematic of the furnace with six zones.

The combustion of coal is complicated chemical process. When this process takes place in a fluidized bed, it is less understood. For example, although it is well known that coal particles admitted to a fluid bed maintained at a temperature above 1400°F are rapidly devolatilized, it is uncertain what proportion of the volatiles burn within the bed and what proportion burn in the freeboard region above the bed surface. Furthermore, neither the chemical reactions that occur during the combustion of devolatilized coal or char nor the actual temperatures reached by the burning particles are known with certainty. Ideally, conditions for the burnout of char in the fluidized combustion zone are favorable because of the good solids mixing and long residence time. However, conditions for the combustion of volatile products in the bed may be less favorable than those for solids because of the potential for release near the top of the bed. Unburned CO and volatile products may escape the bed and reduce the proportion

of combustion in the bed. It is highly desirable to maximize the proportion of combustion occurring in the bed in order to take advantage of the high heat transfer rates attainable in the bed.

In order to determine the relative proportion of combustion taking place in each zone, a detailed mass and energy balance calculation of individual zones was performed. A brief outline of the calculational procedure used is presented in Appendix I. Also presented in Appendix II is a listing of zonal heat balance computer program and zonal heat balance results for Test Series J through N. Table V summarizes the percent heat released (of the coal input energy) for the active bed zone and the combined zone above the bed (splash zone + freeboard zone + SH zone + economizer zone) for Test Series L through N.

For all practical purposes, little or no combustion reactions should take place when the temperature of the flue gas falls below approximately 1000°F. For Test Series Runs J through N, temperature of the flue gas stream leaving the superheater section is approximately 1000°F implying that some combustion is taking place even after the superheater section.

Table V also compares the overall combustion efficiency, $\eta_{\text{SENS}}^{\text{zonal}}$, obtained by combining the percent heat released for each zone with the combustion efficiency calculated from the overall chemical energy balance, η_{CHEM} . In most cases, the agreement between the two values is within 3%. Figures 37 through 40 show the percent of the total input fuel energy being released in the active bed zone as a function of excess air, active bed height, bed temperature, and superficial gas velocity. As can be seen from Figures 37 through 40, the percent of the heat release occurring in the active bed zone remains essentially constant at 64% with respect to all four parameters plotted. In order to determine the effect of gas residence time on bed performance,

TABLE V

TABLE V

Summary of SSU Zonal Heat Balance Results - Freeman Coal

Run	T _B , °F	h _f , inches	U _o ft/sec	X _{air} %	RECY	N _{CHEM}	h _f /U _o Sec	N _{CHEM} ^{Zonal}	N _{SENS} ^{Zonal}	Q _{R,bed}	Q _{R,above bed}	h _f U _o , sec
L-1	1521	41.02	7.81	19.77	-	93.07	0.438	93.08	91.01	66.69	24.32	0.30
L-2	1452	37.75	7.35	24.59	YES	96.29	0.435	96.29	95.30	63.55	31.45	0.292
L-3	1466	40.09	6.47	9.67	YES	95.14	0.528	95.14	95.67	68.37	27.30	0.34
L-4	1389	38.08	5.71	24.53	YES	95.16	0.578	95.16	101.72	71.46	30.26	0.357
L-5	1598	41.65	8.98	3.99	YES	96.33	0.397	96.33	90.70	64.92	25.78	0.277
L-6	1439	45.79	9.15	15.87	YES	94.97	0.426	94.97	92.93	60.34	32.59	0.240
L-7	1602	33.54	10.49	26.79	YES	96.59	0.269	96.59	95.28	63.84	31.44	0.200
L-8	1635	33.04	9.39	8.04	YES	95.61	0.303	95.62	92.83	62.73	30.10	0.215
L-9	1417	37.14	7.38	23.61	YES	96.77	0.432	96.78	92.76	60.68	32.08	0.291
L-10	1464	36.24	6.59	10.70	YES	95.53	0.464	95.54	92.56	62.03	30.53	0.312
L-11	1428	34.85	5.94	27.39	YES	95.63	0.202	95.64	99.44	68.78	30.66	0.322
L-12	1449	34.54	5.05	6.38	YES	95.29	0.580	95.29	91.26	63.81	27.45	0.358
L-13	1458	31.18	4.76	1.02	YES	93.99	0.346	93.99	93.67	65.43	28.18	0.347
L-14	1456	27.56	4.59	-9.82	YES	91.88	0.503	91.89	92.10	63.48	28.62	0.324
L-15	1400	25.60	5.21	16.71	YES	93.21	0.409	93.21	98.39	64.06	34.33	0.268
L-16	1433	25.63	5.75	21.60	YES	96.26	0.378	96.27	91.30	61.95	29.41	0.264
L-17	1458	26.63	6.31	32.62	YES	96.91	0.352	96.91	95.80	67.53	28.27	0.251
L-18	1571	26.49	6.16	-2.85	YES	95.30	0.358	95.30	91.46	61.09	30.37	0.247
L-19	1455	27.65	6.28	7.35	YES	95.10	0.367	95.10	91.82	57.27	34.55	0.250
L-20	1400	26.22	6.66	12.73	YES	95.09	0.355	95.09	87.45	52.54	34.81	0.226
L-21	1442	25.67	7.42	25.27	YES	95.85	0.288	95.85	89.98	56.73	33.24	0.21
L-22	1716	26.83	8.42	28.28	-	96.34	0.266	96.34	92.51	70.87	21.64	0.198
L-23	1654	27.33	6.43	31.64	-	95.38	0.354	95.39	100.52	77.01	23.51	-
L-24	1645	28.25	6.15	23.51	-	95.81	0.303	95.82	100.20	76.72	23.48	0.213
L-25	1589	27.57	10.08	28.76	YES	95.50	0.228	95.51	99.54	64.82	34.72	0.173
L-26	1541	27.35	9.49	18.02	YES	95.56	0.240	95.56	92.49	59.75	32.74	0.179
L-27	1554	27.01	8.98	8.69	YES	95.24	0.251	95.25	91.61	58.91	32.70	0.185
L-28	1406	42.78	7.28	19.29	-	92.59	0.490	93.59	89.70	61.61	28.09	0.326
L-29	1439	41.68	5.0	9.08	YES	88.34	0.695	88.34	99.10	76.08	23.02	0.432
L-30	1407	37.59	7.33	29.90	YES	94.40	0.428	94.46	93.96	64.87	29.09	0.302
L-31	1498	37.27	6.75	10.49	YES	95.30	0.460	95.30	91.80	64.86	26.94	0.315
L-32	1446	37.53	5.70	21.57	YES	96.52	0.549	96.32	96.60	68.01	28.59	0.357
L-33	1482	35.44	5.22	9.43	YES	96.52	0.566	96.52	95.39	67.87	27.52	0.371

TABLE V Continued

Run	$T_B, ^\circ\text{F}$	h_f, inches	$U_o, \text{ft/sec}$	$X_{\text{air}} \%$	RECY	N_{CHEM}	$h_f/U_o, \text{Sec}$	$N_{\text{CHEM}}^{\text{Zonal}}$	$N_{\text{SENS}}^{\text{Zonal}}$	$Q_{R, \text{bed}}$	$Q_{R, \text{above bed}}$	$\frac{h_f}{U_o}, \text{sec}$
L-37	1479	33.14	7.47	42.10	YES	95.64	0.370	95.94	100.49	69.10	31.39	0.263
L-38	1682	33.98	9.01	10.68	YES	98.19	0.309	98.19	90.30	62.86	27.44	0.283
L-39	1619	35.13	9.45	20.06	YES	97.21	0.287	97.21	90.45	61.75	28.70	0.218
L-40	1645	30.71	9.01	15.83	YES	94.33	0.280	94.33	90.63	61.55	29.08	0.210
L-41	1628	32.03	10.05	23.34	YES	96.31	0.261	96.31	91.48	63.00	28.48	0.201
L-42	1122	32.23	10.03	23.64	YES	95.52	0.265	95.82	91.78	63.45	28.33	0.20
L-43	1598	35.25	9.	7.47	YES	97.33	0.320	97.53	89.34	61.73	27.61	0.235
L-44	1443	42.80	8.35	14.41	YES	96.65	0.391	96.65	87.24	59.48	27.76	0.284
L-45	1660	35.02	10.07	1.71	-	94.56	0.264	94.56	89.26	63.52	25.74	0.216
L-46	1649	36.81	7.84	32.26	-	96.42	0.322	96.42	97.74	72.24	25.50	0.234
L-47	1643	33.39	6.91	10.44	-	94.66	0.395	94.66	94.01	68.39	25.62	0.277
L-48	1545	35.10	5.91	20.70	-	92.96	0.484	92.96	101.32	76.16	25.16	0.326
L-49	1528	34.33	5.21	12.13	-	91.44	0.538	91.44	96.14	68.71	27.43	0.349
L-50	1510	41.02	7.70	22.87	-	96.57	0.431	92.57	90.89	67.90	22.99	0.301

TABLE V Continued

Run	T _B , °F	hf, inches	U _o ft/sec	X _{air} %	RECY	N _{CHEM}	hf/U _o Sec	N _{Zonal} CHEM	N _{Zonal} SENS	Q _{R,bed}	Q _{R,above bed}	h _f U _b , sec
M-1	1489	42.79	7.22	16.37	-	88.97	0.494	88.94	91.59	67.51	24.08	0.337
M-2	1614	53.36	8.67	8.29	YES	97.75	0.512	97.74	95.10	71.10	24	0.347
M-3	1481	34.16	9.76	27.76	YES	94.80	0.292	94.79	92.09	62.77	29.32	0.208
M-4	1405	41.01	7.07	25.76	YES	96.72	0.483	96.71	93.34	68.53	24.81	0.322
M-5	1415	37.47	6.20	5.16	YES	96.25	0.504	96.24	89.19	62.71	26.48	0.337
M-6	1342	38.24	5.17	22.55	YES	96.65	0.616	96.64	91.00	66.29	24.71	0.390
M-7	1383	46.09	4.59	8.48	YES	96.16	0.837	96.15	91.69	70.51	21.18	0.486
M-8	1539	28.89	9.68	28.78	YES	97.46	0.249	27.41	90.50	39.51	30.99	0.157
M-9	1592	28.71	8.42	4.37	YES	91.24	0.284	96.24	90.30	59.09	31.21	0.207
M-10	1417	32.21	6.96	27.47	YES	96.54	0.386	96.54	96.98	63.89	33.09	0.264
M-11	1450	31.87	6.14	4.03	YES	97.49	0.435	97.49	88.23	59.01	29.22	0.288
M-12	1404	31	5.39	20.88	YES	96.33	0.479	96.33	100.42	67.64	33.28	0.311
M-13	1415	33.90	4.53	0.99	YES	94.02	0.624	94.62	91.40	62.57	28.53	0.401
M-14	1573	27.14	5.88	-6.61	YES	95.84	0.385	95.84	90.17	61.31	28.86	0.256
M-15	1510	26.64	6.15	0.87	YES	94.64	0.381	94.64	92.60	61.04	31.56	0.25
M-16	1443	27.85	6.42	10.37	YES	93.47	0.362	93.47	100.06	63.77	36.29	0.25
M-17	1313	25.55	6.22	21.20	YES	90.02	0.334	90.02	94.23	56.47	37.76	0.233
M-18	1531	36.31	7.59	8.33	-	93.80	0.399	93.80	86.66	63.50	23.16	0.281
M-19	1617	27.06	6.43	27.80	YES	98.62	0.351	98.12	104.61	79.67	24.94	-
M-20	1634	29.61	5.96	16.56	YES	97.84	0.110	97.84	104.01	79.74	24.27	-
M-21	1639	27.23	5.57	7.01	YES	97.76	0.407	97.76	104.11	77.79	26.32	-
M-22	1631	27.18	5.17	0.45	YES	97.13	0.446	97.13	100.96	74.49	26.47	0.292
M-23	1600	26.81	4.86	-10.11	YES	95.26	0.479	95.26	94.46	69.33	25.13	0.307
M-24	1370	35.56	6.04	10.91	YES	95.28	0.491	95.27	87.32	58.08	29.74	0.32
M-25	1371	36.27	6.49	16.98	YES	95.59	0.468	95.58	87.57	57.93	29.74	0.312
M-26	1524	38.11	8.14	1.96	YES	92.68	0.390	92.66	90.33	62.35	27.98	0.28
M-27	1426	32.46	8.10	48.27	YES	95.64	0.388	94.36	88.31	58.95	29.86	-
M-28	1426	32.46	8.10	48.27	YES	95.64	0.388	95.70	93.31	63.19	30.62	0.27
M-29	1445	36.26	7.29	28.51	YES	94.62	0.491	94.77	91.93	63.19	28.74	0.327
M-30	1523	27.19	9.30	20.60	YES	96.44	0.364	96.47	92.87	61.93	30.94	0.202
M-31	1437	36.82	10.05	33.61	YES	94.28	0.490	94.32	84.75	54.92	29.83	-
M-32	1382	37.43	6.36	14.81	YES	95.81	0.569	95.80	91.78	60.66	31.12	0.32
M-33	1393	39	5.71	12.86	YES	95.30	0.468	95.28	94.22	66.32	27.90	0.301
M-34	1440	37.18	6.62	18.44	YES	95.89	0.693	95.88	93.53	64.27	29.26	0.318
M-35	1361	37.91	4.56	4.36	YES	95.68	0.520	95.67	91.81	64.29	27.52	0.413
M-36	1472	45.38	6.75	11.10	YES	96.22	0.415	96.21	92.97	71.85	21.12	0.368

TABLE V Continued

Run	T _B , °F	h _f , inches	U _o ft/sec	X _{air} %	RECY	N _{CHEM}	h _f /U _o Sec	N _{Zonal} CHEM	N _{Zonal} SENS	Q _{R,bed}	Q _{R,above bed}	$\frac{h_f}{U_o}$, sec
N-1	1410	37.09	4.73	12.38	YES	92.31	0.653	92.31	99.53	68.98	30.58	
N-2	1417	42.32	5.21	20.26	YES	95.08	0.677	95.08	100.09	75.26	24.83	
N-3	1581	37.73	6.32	2.76	YES	90.89	0.498	90.89	95.22	67.57	27.65	
N-4	1566	35.62	7.36	20.96	YES	94.84	0.403	94.85	98.32	69.14	29.18	
N-5	1625	34.70	8.58	5.33	YES	93.55	0.357	93.55	93.42	63.76	29.66	
N-6	1594	32.38	9.51	21.06	YES	94.51	0.284	94.52	100.36	67.21	33.15	
N-7	1434	30.01	4.68	6.41	YES	97.71	0.534	97.71	93.39	65.69	27.70	
N-8	1411	31.12	5.46	23.95	YES	98.16	0.475	98.16	93.89	66.45	27.44	
N-9	1483	31.89	6.22	2.61	YES	98	0.427	98.00	85.27	60.20	25.07	
N-10	1505	32.51	7.10	21.97	YES	97.18	0.378	97.19	91.63	63.21	28.42	
N-11	1589	31.51	8.36	6.31	YES	95.05	0.318	95.05	90.61	59.37	31.24	
N-12	1516	36.18	9.43	23.88	YES	94.89	0.320	94.89	93.19	61.36	31.83	
N-13	1464	30.47	7.82	13.46	YES	94.18	0.325	94.18	90.33	53.96	36.37	
N-14	1457	28.27	7.82	11.61	YES	94.49	0.301	94.49	88.26	53.16	25.10	
N-15	1446	32.04	7.75	13.61	YES	94.67	0.345	94.67	89.27	54.06	35.21	
N-16	1420	31.44	7.51	14.44	YES	93.69	0.349	93.70	92.66	53.78	38.88	
N-17*	1420	28.62	8.91	24.32	YES	91.36	-	91.44	82.29	50.69	31.60	
N-18*	1489	32.67	7.88	6.19	YES	90.92	-	91.01	81.69	49.83	31.86	
N-19	-	-	-	-	-	-	-	-	-	-	-	
N-20	-	-	-	-	-	-	-	-	-	-	-	
N-21	1437	33.47	6.21	8.46	YES	97.41	0.449	97.41	90.69	57.81	32.88	
N-22	1426	35.87	7.02	23.88	YES	96.53	0.425	96.53	92.95	61.23	31.72	
N-23	1566	40.12	8.13	4.02	YES	93.74	0.411	93.74	90.24	61.63	28.61	
N-24	1485	38.04	8.94	17.54	YES	95.19	0.355	95.19	92.20	60.08	32.12	
N-25	1348	33.35	6.04	18.05	YES	96.04	0.463	96.05	97.28	60.54	36.74	
N-26	1415	34.16	5.93	1.18	YES	90.42	0.480	90.13	90.41	56.18	34.23	
N-27	1461	35.68	7.09	11.19	YES	98.84	0.419	98.84	85.74	53.42	32.32	
N-28	1581	34.72	6.39	5.74	YES	97.65	0.453	97.65	92.74	64.17	28.57	
N-29	1541	32.56	6.32	6.80	YES	97.98	0.429	97.98	93.72	63.88	29.84	
N-30	1539	34.79	6.34	7.04	YES	98.36	0.457	98.36	92.51	64.44	28.07	
N-31	1497	34.18	7.11	23.53	YES	98.35	0.401	98.35	96.98	66.34	20.64	
N-32	1585	36.95	8.19	3.16	YES	95.86	0.376	95.86	89.51	59.34	30.17	
N-33	1512	35.58	9.37	23.23	YES	93.43	0.299	93.43	95.93	61.29	34.64	

*Old Cen Coal

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Figure 39

percent heat released in the bed is plotted against L_f/U_o and L_f/U_b in Figures 41 and 42, respectively. The bubble rise velocity, U_b , was calculated using Equation 6 and assuming that the characteristic bubble diameter is that corresponding to be bubble situated at $(L_f/2)$. From Figure 42, it is evident that percent heat released in the bed is dependent on the bubble residence time, L_f/U_b . This means that the performance is quite dependent of the ability of oxygen interchange rates between bubble and emulsion phases and oxygen transfer rate from emulsion phase to char particle surface.

As has been suggested in the section on bed dynamics, SSU bed combustor is operating in a fast bubble regime. Under this condition the mode of gas flow through the emulsion phase and the rate of inter-phase gas exchange together with the bubble size distribution should determine the char combustion efficiency in the bed. This means that in the present situation bed performance is almost entirely dependent on the bed fluid mechanics (bubble hydrodynamics and mass transfer rates).

In order to obtain an estimate of the extent of oxygen bypassing occurring in the SSU fluid bed, value of the inter-phase gas exchange represented by equation 43 as a dimensionless number X_b was calculated. The calculated values of X_b for L-series runs are presented in Table VI. As can be seen from Table VI, the values of X_b varies from 0.4 to 1.8 implying the inter-phase gas exchange rate is very low, hence the high level of oxygen bypassing of the bed. It is believed that high level of oxygen bypassing from the bed is one of the principal causes of low level of heat release in the bed.

Figure 43 shows the variation in percent air consumed in the bed as a function of total air flow rate. As can be seen from Figure 43, the percent air consumed in the bed remains constant over a

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Figure 4

TABLE VI

Summary of Oxygen Bypassing Results

Run	% $Q_{R, Bed}$	X_b	Run	% $Q_{R, Bed}$	X_b
L-1	66.69	0.65	L-26	59.75	0.45
L-2	63.85	0.69	L-27	58.91	0.45
L-3	68.57	0.91	L-28	61.61	0.63
L-4	71.46	1.06	L-29	76.08	1.65
L-5	64.92	0.81	L-30	64.87	1.25
L-6	60.34	0.39	L-31	64.86	1.17
L-7	63.84	0.51	L-32	68.01	1.24
L-8	62.73	0.55	L-33	67.67	1.82
L-9	60.68	0.89	L-34	70.28	1.48
L-10	62.03	1.09	L-35	67.17	1.0
L-11	68.78	1.01	L-36	63.62	1.34
L-12	63.81	1.13	L-37	69.10	1.00
L-13	65.43	1.5	L-38	62.86	0.88
L-14	63.48	1.70	L-39	61.75	0.76
L-15	64.06	0.88	L-40	61.55	0.67
L-16	61.95	1.57	L-41	63.00	0.58
L-17	67.53	1.40	L-42	63.45	0.52
L-18	61.09	0.98	L-43	61.73	0.50
L-19	57.27	0.77	L-44	59.48	0.61
L-20	52.64	0.51	L-45	63.52	0.62
L-21	56.73	0.88	L-46	72.24	0.70
L-22	70.87	0.84	L-47	68.39	0.90
L-23	77.01	--	L-48	76.16	1.08
L-24	76.72	1.05	L-49	68.71	1.14
L-25	64.82	0.49	L-50	67.90	0.57

range of bed temperature, bed height, and total air flow rates. Approximately 58% of the input air is being utilized in the bed regardless of recycle or no recycle operation. This implies that combustion in the bed is being limited by mass transfer considerations.

2.3.5 Combustion in the Splash Zone

The splash zone is defined in this study as the zone between the surface of active bed and the 40" refractory level. The splash zone height decreased as the active bed height increases. The zonal heat balance results presented in Appendix II shows that approximately 1 to 9% of the coal input energy is being released in the splash zone.

2.3.6 Freeboard Combustion

The SSU freeboard zone is defined as the region between the top of the 40" refractory level and the superheater. The primary function of this zone is to allow the larger particles ejected from the bed to decelerate and fall back into the bed. Another important function of freeboard section is to provide additional gas/gas and/or gas/solid contact so that combustion of CO, volatiles, and/or char particles could take place. In order to determine the extent of combustion taking place in the freeboard and to gain some insight into the combustion mechanism, it is usually desirable to measure the variations in CO₂, CO, and O₂ concentrations with height in the freeboard.

It is commonly observed that the main reaction taking place in the freeboard is the oxidation of CO⁽³⁸⁻⁴¹⁾. The extent of this oxidation is dependent on both the freeboard temperature and excess air. The proportion of combustion in the freeboard decreases with excess air and is small in comparison to that in the bed,

being approximately 7% of the total combustion⁽³⁸⁻⁴¹⁾. It is also possible that splashing of the large particles which are too large to be elutriated and which eventually fall back in the bed may provide good solids mixing and hence help promote some char combustion in the freeboard. Moreover with low levels of combustion in the bed, considerable oxygen is available in the freeboard section to support combustion.

The experimental measurements of this investigation show that the average flue gas temperature in the freeboard is about the same as the bed temperature. This result clearly indicates that freeboard reactions, in addition to CO oxidation, involves char and volatile combustion and to a lesser extent SO₂ absorption and NO_x reduction reactions. Figures 44 and 45 shows the percent heat released for the combined above bed regions as a function of average temperature and the input excess air level. However, the individual reaction heat release rates may still be dependent on these two parameters. Table V shows that heat release in the combined freeboard region is approximately between 25 to 30% of the coal input energy. As can be seen from Figures 44 and 45, somewhat higher heat release is occurring above the bed for recycle runs indicating that additional heat release is due to combustion of some of the fine recycle char.

The extent of char combustion in the freeboard depends on the solids hold up in the freeboard. Solids hold up in turn depends on the amount of solids thrown up into the freeboard by the bursting bubbles at the bed surface. The rate of entrainment of solids from the bed surface, W_o , may be given by⁽²⁾:

$$W_o = A_B (U_o - U_{mf})(1 - \epsilon_{mf}) P_s \alpha \alpha_w \quad (50)$$

Figure 44

where

α = Wake fraction of solids

α_w = Fraction of wake solids thrown into the freeboard

For a given set of operating conditions, increasing the value of α_w increases the solids splashing rate at the bed surface. This means that if a large portion of char leaves the bed, char elutriation rate from the SSU combustor will also be large. However, as can be seen from Figure 32, the char elutriation rate from the SSU furnace is independent of superficial gas velocity implying that the extent of char combustion in the freeboard may be very nearly constant.

In order to estimate the contribution of volatiles/CO and recycle char combustion to the overall heat release in the combined freeboard region, it was assumed that coal feed carbon elutriated during recycle operation was same as that for no recycle operation. Under steady state operating conditions, the recycle and its carbon content will remain constant only if an amount of carbon, equivalent to the feed carbon being elutriated is burnt out from the recycle stream. With this assumption approximately 30-35% of the input volatiles and 20-25% of the recycle char must be combusted in the combined freeboard region in order to achieve the indicated above bed heat release.

2.3.7 Superheater and Economizer Sections

Volatiles, char particles, and CO will continue to burn, out of the bed, until the flue gas has been cooled sufficiently to inhibit combustion (less than 1000°F). The data from Test Series J through N indicate that temperature of the flue gas leaving the superheater section is approximately 1000°F. Therefore, it was assumed that some combustion was occurring in the superheater and economizer sections of the unit. The steam production rate in

the superheater section of the waterwall tubes should equal the difference between the loss in sensible heat of flue gas/solid stream and the superheater duty plus some radiation heat absorption. This means that heat transfer mechanism to superheater section of the waterwall tubes is made up of two modes; namely, convection/conduction and radiation.

Similarly, it was assumed that steam production rate in the economizer section of the waterwall tubes should equal the difference between the loss in sensible heat of flue gas/solid steam and the economizer heat duty plus some radiation heat absorption. Heat transfer to the economizer section of the waterwall tubes is also due to two modes: namely, convection/conduction and radiation.

These data from series L, M & N were used to determine the natural circulation performance of the Sub-Scale Unit in-bed steaming circuits. Figures 46, 47, and 48 are graphs of the circulation ratio (CR) and circuit exit velocities as determined from these test series. Steam generation in each bed circuit varied from 390 to 710 lbs/hr. Over this range of steam generation, circulation ratios varied within a range of 45 to 81.

The observed natural circulation is consistent with conservative design practice. The minimum circulation ratio of 45 corresponds to a percent steam by volume (% SBV) at the exit of 64%. For the operating drum pressure conditions of the SSU, 315 psia, the maximum permissible % SBV is 77%. Since % SBV decreases with increasing CR, the SSU operated well within conservative limits. The circuit entrance velocity was 5 ft/sec and remained constant over the range of test conditions. Also plotted in graphs 46, 47 and 48, the circuit exit velocities varied over the range of 8 to 11.5 ft/sec. These velocities, in conjunction with the acceptable circulation ratios, are adequate to ensure proper wetting of the internal tube surface and sweeping away of forming steam bubbles.

A flow meter was used to determine the waterflow to the header supplying and in-bed steaming circuits. Heat transfer measurements, determined by test probes in the bed, were used to estimate the rates of steam generation in the immersed sections of the circuits. It was necessary to assume that data from the heat transfer probes were valid for calculating the steam generation for the sections immersed in the bed. Data taken from the zoned heat balance analysis of the SSU were used to establish the steam generation in those sections above the bed level.

2-107

Figure 46

2-107-A

Figure 47

SSU: BED CIRCULATION STUDY
 SERIES N
 CIRCULATION RATIO AND EXIT VELOCITY
 vs STEAM PRODUCTION IN BED CIRCUIT

2-108

Figure 48

2.5 Environmental Considerations

2.5.1 Sulfur Dioxide Emissions

The effect of temperature on sulfur removal is shown in Figure 49. The sulfur removal passes through a maximum around 1550°F. The optimum temperature for sulfur removal is between 1500-1600°F. Removal falls off rapidly as the temperature departs from this optimum range. The rate of sulfur retention decrease is uniform on both sides of the optimum. The rate of fall-off is also more severe as the Ca/S mole ratio is decreased.

The optimum temperature range can be explained by examining the kinetics of the sulfation reaction. At atmospheric pressure and temperatures above 1400°F, the CO₂ equilibrium partial pressure is high enough for the limestone to calcine to lime by the following reaction:

At normal bed temperatures of 1500-1600°F, SO₃, SO₂, CO₂, and H₂O gases are available for absorption on the lime. These will tend to react in order of their thermal stability. The reactions and decomposition temperatures of the products are listed in Table 7.

Table 7

<u>Reaction</u>	<u>Decomposition Temperature of Product</u>
$\text{CaO}_{(s)} + \text{CO}_{2(3)} \rightarrow \text{CaCO}_{3(s)}$	1500°F
$\text{CaO}_{(s)} + \text{SO}_{2(3)} \rightarrow \text{CaSO}_{3(s)}$	1100°F
$\text{CaO}_{(s)} + \text{H}_2\text{O}_{(3)} \rightarrow \text{Ca(OH)}_{2(s)}$	1100°F
$\text{CaO}_{(s)} + \text{SO}_{3(3)} \rightarrow \text{CaSO}_{4(s)}$	2600°F

SULFUR REMOVAL vs BED TEMPERATURE

2-110

Figure 49

CaSO_4 has the highest decomposition temperature, hence SO_3 will tend to react first at the surface. The SO_2 is rapidly oxidized by the reaction.

As long as the bed temperature remains sufficiently high, the SO_3 is absorbed by the reaction:

This reaction offsets the equilibrium of Reaction 2 so that additional SO_2 is oxidized. The process continues until the rate of SO_2 oxidation is too slow or the lime particle surface and pores become filled with CaSO_4 .

Below 1400°F , the rate of SO_3 absorption becomes so slow that little reaction can take place. If the bed temperature is raised above 1600°F , the equilibrium partial pressure of SO_2 over CaO is so high that little SO_3 is formed. The SO_3 absorption rate again becomes too slow and little sulfur retention is achieved.

Thus, sulfur retention outside of the optimum temperature range is limited by the low concentration of SO_3 above 1600°F and by the low SO_3 absorption rate below 1400°F .

The Ca/S mole ratio is the most influential parameter in sulfur removal in fluidized beds. SO_3 absorption is the rate limiting step in SO_2 retention. Since higher Ca/S mole ratios increase the total surface area of sorbent available for SO_3 absorption, an increase in sulfur removal can be expected as the Ca/S ratio is increased. The effect of Ca/S mole ratio on sulfur removal is shown in figure 50. At the optimum temperature of 1500°F , 90% SO_2 retention can be achieved with a Ca/S ratio of 2.2. At 1650°F

SULFUR REMOVAL vs Ca/S

and the same Ca/S ratio, SO_2 retention decreases to 75%. The sulfur removal asymptotically approaches 100% as the Ca/S mole ratio is increased.

Calcium utilization is defined as the moles CaO converted to CaSO_4 divided by the total moles CaO available for sulfation. It is equal to the sulfur retention divided by the Ca/S mole ratio. Calcium utilization is limited by a CaSO_4 shell that forms during the sulfation reaction. This shell forms because the molar volume of CaSO_4 is greater than CaO ($52.2 \text{ cm}^3/\text{mole}$ vs. $36.9 \text{ cm}^3/\text{mole}$, respectively). As the reaction proceeds, the pores fill up with CaSO_4 , preventing any further reaction. Unless this skin is worn away by attrition, eventually the sulfation reaction will stop although the core of the lime particle is still unreacted. The calcium utilization varied between 28-46% for tests conducted at 1500°F . The utilization dropped to 25-35% at 1450°F .

Since sulfur removal is very dependent on bed temperature and Ca/S mole ratio, the effects of other variables such as residence time and recycle will be discussed after a model has been presented to account for these effects.

SO_2 Model - The Westinghouse model⁽⁴⁵⁾ was chosen to fit the SSU data because it has been shown to be an accurate predictor of SO_2 emissions using TGA kinetic data. TGA data at 1650°F for Western Materials limestone was available from tests run at Argonne National Laboratory. As shown previously in figure 50, the actual sulfur removal from the SSU closely paralleled the TGA predicted sulfur removal. Although the Westinghouse model was developed for a single pass system, it was found to predict equally well for the present SSU recycle system. A derivation of the model is presented below.

The fluidized bed is treated as a series of plug-flow reactors. The reaction in each differential element is assumed to be first order with respect to SO_2 . This assumption implies that SO_2 removal is independent of the sulfur content of the coal. Thus, the fractional conversion of SO_2 in each element is given by

$$\tau = \int_0^{X_{AF}} \frac{dX_A}{K(1-X_A)}$$

or

$$X_A = 1 - e^{-K\tau}$$

where τ = reaction time

K = mean reaction rate constant

X = fractional conversion

The rate constant K is a mean reaction rate constant for SO_2 absorption throughout the bed. The SO_2 is assumed to be released uniformly throughout the bed. Referring to Figure 51, the fraction of sulfur released in each differential element is given by $\frac{dh}{H}$. The fractional conversion of SO_2 released in each differential element is then given by

$$X_A \frac{dh}{H} = (1 - e^{-K\tau}) \frac{dh}{H}$$

The residence time of the gas above element dh is given by

$$\tau = \frac{(H-h)\epsilon}{v}$$

where $\frac{v}{\epsilon}$ is the superficial velocity through the bed corrected for bed solids. Thus, the sulfur retention associated with each element of bed is

$$\frac{1}{H} \left[1 - e^{-\frac{K(H-h)\epsilon}{v}} \right] dh$$

WESTINGHOUSE MODEL

This term can now be integrated over the bed height to obtain the total sulfur retention in the bed.

$$\frac{1}{H} \int_0^H \left[1 - e^{\frac{-K(H-h)\epsilon}{v}} \right] dh = 1 - \frac{v}{KH\epsilon} \left(1 - e^{\frac{-KH\epsilon}{v}} \right)$$

$$\text{Sulfur retention } R = 100 \left[1 - \frac{v}{KH\epsilon} \left(1 - e^{\frac{-KH\epsilon}{v}} \right) \right]$$

Calcium utilization u is then determined by $\frac{R}{Ca/S}$ or

$$u = \frac{100}{Ca/S} \left[1 - \frac{v}{KH\epsilon} \left(1 - e^{\frac{-KH\epsilon}{v}} \right) \right]$$

TGA data gives the rate constant K' at a fixed calcium utilization. The mean rate constant can be obtained from K' at unit gas concentration. Thus,

$$K = \frac{K' \rho}{C_i}$$

where ρ is the molar density of the lime and C_i is the SO_2 concentration.

The first order rate constant at 1650°F was determined as follows. The TGA rate constant was first plotted as a function of calcium utilization. K was determined by substituting into the calcium utilization equation derived above and solving for K at fixed u . The utilization equation was then plotted at fixed Ca/S mole ratio. The intersection of the two curves gives the first order rate constant K at fixed temperature and Ca/S . The rate constant at 1650°F is plotted in Figure 52.

RATE CONSTANT vs CALCIUM UTILIZATION

Since the sulfur retention predicted by the TGA data agreed reasonable well with the SSU data, it was decided to use the SSU data to predict the first order rate constant at 1500°F. The results are plotted in Figure 53. Using an Arrhenius expression for the rate constant, the temperature dependence of K can be determined by plotting $\ln K$ vs $\frac{1}{T_{\text{abs}}}$.

Since the sulfur removal curve was essentially identical on either side of 1550°F, it was assumed the first order rate constant curve below 1550°F was the mirror image of the curve above 1550. The rate constant K is plotted as a function of temperature and Ca/S mole ratio in Figure 54.

Using the above kinetic data, the Westinghouse model was used to predict sulfur removal for each test point. An average bed voidage of 0.55 was assumed. SO₂ removal is plotted versus predicted SO₂ removal in Figures 55 and 56. For the no recycle cases, the predictions on the average were within 5.9% of the actual sulfur removal. The average deviation for the recycle cases was 6.7%. The accuracy of these predictions is remarkable considering that the first order rate constant was predicted only from actual TGA data and from fluidized bed sulfur removal data. Furthermore, no constants had to be fitted for the model predictions.

With the use of the Westinghouse model, the effects of bed height and superficial velocity can now be examined. An increase in fluidizing velocity will result in a decrease in sulfur removal. This is primarily due to two effects. First, the gas residence time in the bed is reduced, thus limiting the contact time for SO₂ to react with the sorbent. Second, there will be increased bypassing of the gas through the bed as bubbles. Assuming the bed is in the fast bubble regime, any gas entering the bubbles will circulate between the bubble and its cloud. Since there are essentially no solids associated with bubbles, any SO₂ trapped

RATE CONSTANT vs CALCIUM UTILIZATION

RATE CONSTANT vs BED TEMPERATURE

WESTINGHOUSE MODEL - NO RECYCLE

1417 - 1716°F

$1.7 < Ca/S < 4.7$

$\epsilon = 0.55$

2-121

Figure 55

WESTINGHOUSE MODEL - WITH RECYCLE

1440 - 1682°F

$1.7 < Ca/S < 4.3$

$\epsilon = 0.55$

2-122

Figure 56

within a bubble will have little contact with the sorbent particles and will essentially pass unreacted through the bed. Figures 57 and 58 show the effect of velocity on SO_2 removal at 1550°F and 1650°F. The Westinghouse model predictions were made at the two extreme velocities of 5 and 10 ft/sec. The SSU data plotted had velocities within these two extremes.

An increase in bed height will result in an increase in sulfur removal. This is primarily due to increased gas residence time in the bed. However, the increase in SO_2 retention may not be as large as expected because larger bubbles tend to form through coalescence. These large bubbles tend to decrease gas/solid contact, thus limiting SO_2 absorption on the sorbent. Figures 59 and 60 illustrate the effect of bed height on sulfur removal. Again, the Westinghouse model predictions were made at the extremes of 20 and 40 inches. The SSU data was plotted for bed heights between 30-35".

The significance of gas residence time can further be shown if the data is limited to temperatures that correspond to high first order rate constants. Between 1500°F-1600°F, K is high enough to drive the exponential term in the Westinghouse model to zero. Thus, sulfur retention is given by

$$R = 100 \left[1 - \frac{V}{KH\epsilon} \right]$$

The residence time can now be solved for as a function of sulfur removal and rate constant.

$$\tau = \frac{H}{V} = \left(\frac{1}{1 - .01R} \right) \frac{1}{K\epsilon}$$

Figure 61 is a plot of this function. Despite the scatter, the sulfur removal does increase with gas residence time in the bed. Figure 62 is the same plot for gas residence time in the bed and freeboard. Again, the sulfur removal increases with residence time. However, the better fit of the data indicates that some

SULFUR REMOVAL vs CALCIUM/SULFUR
MOLE RATIO

SULFUR REMOVAL vs CALCIUM/SULFUR MOLE RATIO

* FIRST ORDER RATE CONSTANT OBTAINED FROM
TGA DATA TAKEN AT ARGONNE NATIONAL
LABORATORY

SULFUR REMOVAL vs CALCIUM/SULFUR
MOLE RATIO

SULFUR REMOVAL vs CALCIUM/SULFUR
MOLE RATIO

EFFECT OF RESIDENCE TIME ON SULFUR REMOVAL
 $1500^{\circ}\text{F} < T < 1600^{\circ}\text{F}$

EFFECT OF RESIDENCE TIME ON SULFUR REMOVAL

1500 < T° F < 1600

SO_2 removal is taking place in the freeboard. At present, the model accounts only for SO_2 retention that takes place in bed. The remarkable fit of data with the model is largely a result of a very shallow freeboard, such that most sulfur removal occurs within the bed. However, this data indicates that the model may tend to predict low for units with large freeboards, especially for recycle tests with high dust loadings in the freeboard.

As previously shown in Figures 55 and 56, the Westinghouse model predicts sulfur removal equally well for tests run with or without recycle. The model was developed for a once through system and no modifications had to be made to correlate the recycle tests. The implication of these results is that little or no additional sulfur removal is gained by recycling fines back to the bed with the present recycle system. However, the problem is not that recycle doesn't enhance sulfur removal. Rather, the recycle tends to elutriate straight up the side of the furnace where it is injected. Any sulfur removal associated with the higher dust loading along this wall was not detected as the gas sampling probe was located in the middle of the furnace. During the extended steady state run N-14, the furnace was extensively probed for gas and temperature profiles. Figure 63 shows the SO_2 concentration profile at a cross-section 12 ft. above the grid plate. The SO_2 concentration is considerably lower along the wall where the recycle was injected.

The measured sulfur removal for this run was 69%. However, the sulfur retention along the wall was 96%. The high sulfur retention is due to the high level of lime in the recycle. Ash analysis of recycle test M-14 shows that 25% by weight of the recycle is calcium oxide. Assuming the recycle fines rise with no lateral mixing, the effective Ca/S mole ratio along the wall is increased from 2.2 to 7.6. Using these numbers, the Westinghouse model predicts a sulfur removal of 70% at the middle of the

SO₂ CONCENTRATION PROFILE

furnace outlet where the sampling probe is. However, it also predicts a sulfur removal of 92% along the wall. These predictions are close enough to the actual sulfur removals to support the hypothesis that the recycle is not mixing within the bed.

Several tests were run with the recycle material injected through the transport line at the center of the bed. The sulfur removal is plotted against the Westinghouse model predictions in Figure 64. In all cases, the actual sulfur removals were well over 90% and were much higher than previous runs at similar conditions. The enhanced sulfur removal could be due to two effects. First, the recycle fines may be rising straight up with no mixing. In this case, the gas sampling probe is located in the region of extremely high dust loading and measures low SO_2 concentration as a result of SO_2 reaction with recycled lime. The sulfur removal would thus tend to be higher at the center of the bed than along the walls. Second, the recycle material may be more mixed throughout the bed when it is injected in the bed center rather than in a corner of the bed. The mixed recycle would then reduce the SO_2 more uniformly across the bed. In either case, it appears that there is definitely some benefit in sulfur removal when recycling material. What needs to be developed is a better means to inject the recycle so that it will be more distributed throughout the bed.

The following conclusions can be made about sulfur removal:

1. The optimum temperature range for sulfur removal is 1500-1600°F.
2. 90% sulfur removal at 1500°F can be attained with a Ca/S mole ratio of 2.2.
3. Calcium utilization varied between 25-46%.

WESTINGHOUSE MODEL - WITH RECYCLE

1440 - 1682°F

1.7 < Ca/S < 4.3

$\epsilon \approx 0.55$

4. The Westinghouse model predicted sulfur removal within 7% for both with or without recycle. Only TGA or SSU kinetic data is required to use the model. No constants need be fitted.
5. The reaction is 1st order with respect to SO_2 . Thus, it should be independent of the fraction of sulfur in the coal.
6. Sulfur removal increases with gas residence time.
7. The model does not account for sulfur retention in the freeboard. The model's high degree of correlation is partially a result of low sulfur removal in the freeboard because of its very short length.
8. Recycle material tends to flow straight up the side wall. Thus, any SO_2 retention associated with recycled lime cannot be detected by the gas sampling probe.
9. Recycle material injected through the transport line greatly enhanced sulfur removal. This is either because of item 8 above, or because the recycle fines are more distributed throughout the bed.

2.5.2 Nitrogen Oxide Emissions

NO_x is measured by two independent gas analysis systems (GAS). The first system monitors NO_x by sampling gas through a fixed probe just above the economizer. The second system is used for sampling the freeboard with a movable probe.

GAS samples are drawn through a sintered stainless steel filter into a water-cooled stainless steel probe. The sintered filter removes all particulates greater than 20 microns. The gas sample then passes through a heated line to a conditioning unit in the GAS. The heated line maintains the sample temperature above its dew point. The conditioner removes condensate and particulates greater than 0.6 microns from the sample stream. The dry dust-free sample is then delivered to a Thermo Electron's chemiluminescent NO/NO_x Analyzer. In the NO_x mode, the sample is passed through a converter which thermally breaks down any NO₂ to NO. The NO is then combined with ozone in a reaction chamber. The light emissions emitted from the reaction are monitored through an optical filter by a photomultiplier tube. The intensity of chemiluminescence is calibrated to read NO concentration. In the NO mode, the gas sample bypasses the converter and proceeds directly to the reaction chamber.

All NO_x measurements are taken with the analyzer in the NO_x mode. Although the readings include the sum of both NO and NO_x emissions, the NO_x reading is almost completely comprised on NO. In most cases no significant deviations could be detected when switching from NO_x to NO mode.

NO_x emissions were found to increase with bed temperature. With recycle, emissions increased $\frac{0.6\text{ppm}}{^\circ\text{F}}$ or $\frac{.11\text{ lb NO}_x/10^6\text{ BTU}}{100^\circ\text{F}}$. Without recycle, NO_x also appeared to increase with temperature, although scatter was much greater. Figures 65 and 66 show NO_x as a function of temperature. This temperature dependence could be due to several effects. First, higher bed temperatures could result in greater fuel nitrogen conversion as nitrogenous volatile compounds are driven off at a greater rate and to a greater extent. However, for most cases, fuel N conversion was greater than 90%. Thus, temperature effect on conversion is very limited.

NO_x vs BED TEMPERATURE WITH RECYCLE

NO_x vs BED TEMPERATURE NO RECYCLE

The second effect of higher temperatures would be to increase the oxide rate of CO formed in the combustion of coal particles. The reduced concentration of CO would result in higher NO_x concentrations as CO is an important reducing species. The effect of temperature on CO is shown in Figure 67. An increase in temperature is not expected to increase the rate of NO_x reduction reactions significantly as the activation energies for the primary destruction reactions with CO and char are low.

NO_x emissions tended to decrease with increasing gas residence time. The residence time is defined as the time for a unit volume of gas to pass from the grid plate past the superheater. Since the gas temperature drops to about 1200°F after the superheater, it was assumed that little NO_x formation or destruction would take place at these low temperatures. The residence time is calculated from the following equation:

$$t_{\text{res}} = \left(\frac{1}{12}\right) \left(\frac{1}{v_{\text{sup}}}\right) [108 - 0.5H]$$

where

$$\begin{aligned} t_{\text{res}} &= \text{gas residence time, sec.} \\ v_{\text{sup}} &= \text{bed superficial velocity, ft/sec.} \\ H &= \text{bed height, inches} \end{aligned}$$

With recycle, NO_x decreased roughly 20 ppm or 0.039 lb NO_x/MMBTU per 0.1 seconds residence time. The rate of decrease of NO_x lessens as NO_x approaches 150 ppm or 0.2 lb/MMBTU. No dependence on residence time was observed for NO_x without recycle primarily because of excessive scatter. Figures 68 and 69 show NO_x plotted as a function of gas residence time. The decrease in NO_x with increasing residence time is attributed to two effects.

In Figure 70, CO emission can be seen to increase linearly with residence time at a rate of 900 ppm per 0.1 seconds. However, the residence time is inversely proportional to the superficial velocity, which is a function of the bed temperature. Thus, a decrease in bed

CO vs EXCESS AIR

NO_x vs GAS RESIDENCE TIME NO RECYCLE

CO vs GAS RESIDENCE TIME WITH RECYCLE

temperature will result in an increase in CO emissions and a corresponding decrease in NO_x emissions. The second effect of an increase in gas residence time is to allow more time for the NO_x destruction reactions to proceed.

NO_x appears to increase slightly with theoretical excess air. Theoretical excess air is defined as the percent air above that required for complete combustion of the coal. With recycle, it increases by 0.007 lb NO_x /MMBTU per % excess air. The rate of increase is approximately the same without recycle. Figures 71 and 72 present NO_x as a fraction of excess air. An increase in excess air tends to increase the oxygen concentration in the emulsion phase.

Assume the following reaction mechanism for a burning char particle. Oxygen diffuses to the particle surface and reacts with carbon to form CO. The CO then reacts further with O_2 to form CO_2 . With a higher O_2 concentration in the emulsion phase, more O_2 will diffuse through the boundary layer to the particle surface. More CO will combust to form CO_2 and the resulting lower CO concentrations will increase NO_x emissions. As shown earlier in Figure 70, CO emissions tend to increase with residence time.

Increased oxygen diffusion to the particle surface will also result in a larger fraction of fuel nitrogen being converted to NO_x . When a nitrogenous compound is evolved from the coal particle, it will either oxidize to NO_2 or break down, with the fragments then being oxidized to NO_2 or reduced to N_2 . The tendency for the compounds to oxidize to NO_x is enhanced by higher O_2 concentrations around the particles.

NO_x emissions tended to increase with decreasing coal size. Tests were run with recycle for three coal feed sizes. The coal was screened to 1/8" x 0, 1/4" x 0, and 1/4" x 1/8". The results are

NO_x vs EXCESS AIR WITH RECYCLE

NO_x vs EXCESS AIR NO RECYCLE

presented in Figure 73. This effect can again be explained if the combustion reaction is assumed to be diffusion controlled. For large particles, the rate of diffusion of oxygen from the emulsion phase to the particle surface is inversely proportional to the square root of the particle diameter. Thus, as the coal particle size decreases, the rate of O_2 diffusion through the boundary layer to the particle surface will increase. This, in turn, results in greater conversion of evolved fuel N to NO_x rather than N_2 , and in more complete combustion of CO to CO_2 . Decreasing CO results in less NO_x reduction.

NO_x emissions tended to decrease strongly with CO emissions. With recycle, the NO_x seemed to level out around 0.2 lb NO_x /MMBTU or 150 ppm for CO concentrations greater than 4000 ppm. NO_x decreased at a greater rate with CO for the no recycle case. Figures 74 and 75 present NO_x plotted versus CO. The NO_x dependence in CO is due to the NO_x -CO destruction reaction. The high levels of CO encountered in the SSU tend to enhance this reaction. Much higher CO concentrations are obtained while recycling elutriants. This could be due to two effects. First, the recycled carbon could be reacting with CO_2 to form CO. Second, the recycle solids are injected into a corner of the fluidized. The higher carbon concentration in this corner combined with poor lateral mixing will result in the bed and freeboard being locally substoichiometric directly above the recycle injection part. Low O_2 concentrations will then result in incomplete combustion and CO formation.

The accelerated rate of decrease of NO_x with CO for the no-recycle case is probably due to smoother CO concentration gradients throughout the bed. In the recycle case, a sharp CO gradient exists towards the recycle injection point. With limited lateral mixing in the bed, the overall contact between CO and NO_x will be less efficient. NO_x emissions tended to be higher with recycle compared to no-recycle at the same CO levels. Again, this would be due to less efficient gas mixing in the recycle case. It could also be

NO_x vs CO

NO_x vs CO WITH RECYCLE

NO_x vs CO NO RECYCLE

due to greater fuel nitrogen conversion. Fuel conversion is defined as the fraction of the input coal nitrogen being converted to either NO_x or N_2 . In the no-recycle case, N conversion varied between 86 to 95%. However, close to 90% of the residual nitrogen contained in char elutriants was collected in the cyclone underflow. This fraction is reinjected in the bed when recycling. Further combustion of the recycled char particles will release the "fixed" nitrogen in the char. The fuel nitrogen conversion with recycle varied between 91 to 99%. No significant differences in fuel nitrogen conversion with coal feed size were noticed. This is probably because fuel nitrogen not released on the first pass may be released on subsequent passes as the char particles are recycled. Figure 76 shows Fuel N conversion as a function of coal feed size and bed temperature.

NO_x emissions tended to decrease with elutriated carbon for the recycle case. No effect was noticed with recycle. These results are presented in Figure 77. Literature indicates that the carbon- NO_x and CO-NO_x destruction reactions are the primary mechanisms for reducing NO_x . The carbon loading in the bed is estimated about 0.5%. However, the carbon loading is small when compared to the elutriated carbon in both the recycle and no-recycle cases. The lack of NO_x dependence on elutriated carbon in the no-recycle case indicates that the CO-NO_x reduction may be more important in the SSU combustor than the C-NO_x reaction. A good fit of CO as a function of excess air and elutriated carbon indicates that any NO_x dependence on elutriated carbon may be largely due to its effect on CO. The high elutriated carbon in the recycle case is concentrated around the recycle feed point whereas the NO_x would tend to be highest around the feed nozzle. The strongest effect of the recycled carbon for the present SSU configuration seems to be to generate CO rather than directly reduce NO_x . The data is presented in Figure 78. Because of the uncertainty in measuring both the recycle rate and the bed carbon loading, and the uncertainty in

NO_x vs ELUTRIATED CARBON

CO vs ELUTRIATED CARBON

whether the NO_x dependence on elutriated carbon is a result of CO generation or carbon- NO_x reduction, it was decided to model the NO_x emissions on the basis of CO- NO_x reduction only.

A model was derived for predicting NO_x emission by making the following assumptions:

1. NO_x emissions are completely comprised of fuel nitrogen.
2. The fuel nitrogen is completely converted to NO_x initially.
- 2a. The mixing in the bed and freeboard is plug flow.
3. NO_x is reduced by the CO destruction reaction. The reaction is first order with respect to NO_x .
4. The char- NO_x reaction does not contribute significantly to NO_x reduction.

The fuel nitrogen conversion was greater than 90% in most cases. The fraction of nitrogenous compound directly converted to NO_2 is unknown. The NO_x -CO reaction has been reported in literature to be first order with respect to NO_x . The char- NO_x reaction is also reported to be one of the important NO_x reduction reactions. However, the published literature does not deal with recycling fluidized beds. Also, the present SSU configuration with recycle injected into a corner of the bed may inhibit the reaction. The recycle solids most likely elutriated straight up along the wall with little mixing in the bed or freeboard. The model derivation is presented below:

NO_x Model Derivation

$$-\frac{d[\text{NO}]}{dt} = (-r_{\text{NO}}) = k [\text{NO}][\text{CO}]^a \quad (1)$$

For a plug flow reactor, the material balance on NO is given as

Input = Output + Disappearance by Reaction + Accumulation

$$F_{\text{NO}_0} = F_{\text{NO}} + (-r_A) dV \quad (2)$$

dV

F_{NO}

$F_{\text{NO}} + dF_{\text{NO}}$

Where F_{NO_0} = molar rate of NO in, mole/sec

dV = differential volume element

but $F_{\text{NO}} = F_{\text{NO}_0} (1 - X_{100})$

so $dF_{\text{NO}} = -F_{\text{NO}_0} dX_{100} \quad (3)$

Where X_{NO} = fractional conversion of NO

Substituting 1 and 3 into the material balance 2,

$$F_{\text{NO}_0} dX_A = K[\text{NO}][\text{CO}]^a dV$$

but $X_{\text{NO}} = 1 - \frac{[\text{NO}]}{[\text{NO}_0]}$

so $dX_{\text{NO}} = -\frac{1}{[\text{NO}_0]} d[\text{NO}]$

Resubstituting, rearranging, and integrating over reactor volume,

$$- \frac{F_{NO_o}}{[NO_o]} d[NO] = k [NO] [CO]^a dV$$

$$\frac{NO}{NO_o} \frac{d[NO]}{[NO]} = - k [NO]^a \left(\frac{NO_o}{F_{NO_o}} \right)^a dV$$

$$\ln \frac{[NO]}{[NO_o]} = - k [CO]^a \left(\frac{NO_o}{F_{NO_o}} \right)^a V$$

Substituting $F_{NO_o} = \frac{NO_o V}{\tau}$

where τ = gas residence time

$$\ln \frac{[NO]}{[NO_o]} = - k [CO]^a \tau$$

But, $\frac{[NO]}{[NO_o]} = (\text{constant}) \left(\frac{1b NO}{MMBTU} \right)^2 = \frac{(\text{Fraction N})(46)}{HHV} \left(\frac{1b NO}{MMBTU} \right)^2$

$$\ln \frac{16NO}{MMBTU} = -k, [CO]^a \tau$$

or $\frac{16NO}{MMBTU} = \exp [- k, [CO]^a \tau]$

where k = reactor rate constant

k_1 = reaction rate constant

a = order of reaction with respect to CO

The term k_1 is a function of both temperature and coal feed size. As discussed earlier, the coal feed size affects the rate of oxygen diffusion to the coal surface. This in turn affects the fraction of fuel N initially converted to NO_x and also affects the burnout of CO to CO_2 . The gas residence time is inversely proportional to temperature. A log-log plot of NO_x vs. CO yielded a straight line with a slope of -0.5. Thus, the NO_x reduction reaction was assumed to be 1/2 order with respect to CO. The exponential term in the model $k_1 (\text{CO})^a$ was then statistically fit to a function of d_p , T, and $\text{CO}^{1/2}$. The model is presented below with and without recycle. The results are plotted in Figures 79 and 80.

$$\text{With Recycle: } \frac{1\text{b NO}_2}{\text{MMBTU}^2} = \exp \left(-3.03 + \frac{507}{d_{\text{pcoal}}} + 0.0016 T_{\text{Bed}} - 0.02 \text{CO}^{.5} \right)$$

$$\text{Without Recycle: } \frac{1\text{b NO}_2}{\text{MMBTU}^2} = \exp \left(-1.8 + 0.001 T_{\text{Bed}} - 0.026 \text{CO}^{.5} \right)$$

where

T_{Bed} = bed temperature, °F

CO = carbon monoxide concentration, ppm

d_{pcoal} = weight mean coal size, nm

With recycle, the correlation predicts the data with an average deviation of 13.8% for 99 data points. Without recycle, the correlation has an average deviation of 15.2% for 29 points. There was no data on coal size effects for the recycle bypass tests.

A series of tests was run to determine the effect of recycle injection location. In matrix runs M28-M32, the recycle material was injected through the transport line. The NO_x emissions were

NO_x MEASURED vs NO_x CALCULATED
NO RECYCLE

NO_x MEASURED vs NO_x CALCULATED WITH RECYCLE

in each case lower than would be expected under similar conditions with recycle injection in the north east corner of the furnace. The NO_x emissions averaged 14% lower than the recycle correlation predicts. On the other hand, the emissions averaged only 6% higher than the no recycle correlation prediction. These results indicate that significantly lower NO_x emissions may be obtained if the recycle is more mixed throughout the bed. Although CO emissions were lower for these tests, the CO gradient would tend to be more uniform across the bed, rather than peaking near a corner at the furnace. As with the no recycle case, more efficient CO- NO_x contacting would result in lower NO_x .

The C-E demonstration plant at Great Lakes Naval Base should easily meet NO_x emission requirements. The maximum allowable NO_x emission for the demonstration plant is 0.5 lb NO_2 /MMBTU. Assuming typical operating conditions of 1550°F, 1/4" x 0" coal (1400 μm), and 2000 ppm CO, the NO_x emission can be expected to be between 0.29 and 0.39 lb NO_2 /MMBTU. CO emissions would have to drop to roughly 650 ppm to raise NO_x emissions to the maximum allowable limit. However, since the demonstration plant is recycling material through the transport line, it is likely that NO_x emission levels will be lower than those predicted by the presented correlation.

The following conclusions may be made with respect to NO_x emissions:

1. NO_x emissions increase with bed temperature
2. NO_x emissions decrease with gas residence time.
3. NO_x emissions increase slightly with excess air.
4. NO_x emissions decrease with coal feed size.
5. NO_x emissions decrease strongly with CO.
6. NO_x emissions decrease faster with CO for no recycle cases. This is because of more efficient gas contacting.
7. Recycle tends to rise straight up with little mixing in bed.
8. NO_x emissions may be reduced by recycle injection through the transport line.

9. NO_x emissions decrease with elutriated carbon.
10. NO_x emissions are 1st order with respect to NO_x , and 1/2 order with respect to CO.
11. Fuel nitrogen conversion is greater than 90%.
12. The demonstration plants can readily limit NO_x emissions below the maximum allowable 0.5 lb NO_2 /MMBTU.

2.5.3 Particulate Emissions

One of the primary environmental concerns associated with coal combustion is particulate emissions. Current Federal EPA regulations require particulate emissions for steam generators with a net heat input greater than 250 million BTU/hr to be limited to 0.01 lb/mm BTU. Particulate emission limits for smaller units, such as the Great Lakes Demonstration Plant, are set by individual states. Present Illinois state regulations limit particulates to 0.1 lb/mm BTU. More stringent regulations are expected to be passed in the near future. The conditions the demonstration plant has agreed to meet are a maximum particulate emission of 0.03 grains/dry standard cubic foot.

In order to meet these emission requirements, some form of hot-gas clean-up is necessary. On the SSU, a bag filter collector was chosen for final particulate removal. The Air Preheater bag filter contains 48 5" diameter fiberglass bags in 4 compartments. Each bag has over 19 square feet of filtering surface. The bags are cleaned by reversing the air flow through each compartment. This collapses the filters and breaks off the collected dust cake.

Three sets of dust loadings were taken during an extended steady state run. Isokinetic samples were taken in accordance with methods discussed in "Determination of Particulate Emissions from Stationary Sources", by the Federal EPA. The particulate sampling train was a Scientific Glass Blowing Co. STAC-0-LATOR. Six points were sampled in two perpendicular traverses across the 8" diameter stack.

Results from the tests are given in Table 8. Average particulate emissions were 6.1×10^{-3} grain/scf or 1.2×10^{-2} lb/MMBTU. This corresponds to a bag filter efficiency of 99.96%. Based on present regulations, particulate emission limits can be met on the demonstration unit with a dust collector similar to the SSU Air Preheater bag filter. More stringent regulations may require a second stage of clean-up after the bag filter.

Table 8

<u>Test #</u>	<u>Grains/SCF</u>	<u>lb/hr</u>	<u>Particulate Emission</u>	
			<u>lb/10⁶ BTU</u>	<u>Bag Filter Efficiency %</u>
1	0.0057	0.034	0.011	99.97
2	0.0064	0.039	0.013	99.96
3	<u>0.0062</u>	<u>0.038</u>	<u>0.013</u>	<u>99.98</u>
Average	0.0061	0.037	0.012	99.96

2.6 Grid Plate Performance

2.6.1 Mechanical Performance

The original three piece gridplate was removed in July, 1978 because of excessive leakage caused by inadequate seals and severe warpage. It was replaced with a single 1/4" S.S. plate to eliminate any seam leakage. The plenum wall was extended upwards so that the plate could be fastened down before putting on the removeable wall. The plate was held in place by S.S. strips clamped down along its outer edges. This configuration allowed the plate to freely deform such that stresses may be relieved.

Inspections of the plate underside were subsequently made after each two week operating period. The first inspection showed that the entire plate had shifted approximately 1/16" toward the front wall. Minor warpage of the plate had occurred between two coupled stiffeners. The final inspection, nine months after installation, found that the deformation between the two stiffeners welded to the gridplate tended to lean toward the waterwalls, with some deformation between the stiffener support brackets. The plate appeared to still be sealed adequately and was judged capable of sustaining further operation. Figure 81 shows the gridplate deformation observed in the final inspection.

2.6.2 Gridplate Pressure Drop

The pressure drop across the SSU gridplate was measured as a function of plenum velocity, shortly after the new plate was installed. The data is presented in Figure 82. The gridplate pressure drop was correlated to the following equation:

$$\ln \frac{P_{\text{grid}}}{p} = 2.03 \ln U_{\text{in}} + 4.98 \quad (1)$$

where P_{grid} = gridplate pressure drop, lb/ft³
 p = plenum air density, lb/ft³
 U_{in} = plenum air velocity, ft/sec.

SSU GRID PLATE INSPECTION

△ INDICATES A 0.006" FEELER GAUGE COULD PASS
UNOBSTRUCTED BETWEEN THE PLATE AND THE LOWER FLANGE

SSU GRID PLATE DATA PRESSURE DROP vs PLENUM VELOCITY

Calculations based on the above correlation show that the gridplate pressure drop was less than the desired pressure drop at the 70% MCR condition. The initial design criteria for the SSU grid plate was that 30% of the bed pressure drop be maintained across the gridplate at 70% MCR to provide adequate flow distribution. At full active bed depths the design bed pressure drop is approximately 36" H₂O. Thus, at 70% load the grid plate pressure drop must be at least 11" H₂O to satisfy this criteria.

The approximate MCR conditions for combustion air flow and temperature on the SSU are respectively 1960 lb/hr and 485°F. However, the grid plate pressure drop calculated from Equation 1 at 70% load predicts a pressure drop on only 4.3" H₂O. This is below the 11" H₂O pressure drop required at 70% load conditions.

It was decided to increase the grid plate pressure drop by blocking off a number of holes from the grid plate. Calculations indicated that 478 holes out of a total of 1152 holes need to be blocked to obtain the desired pressure drop at 70% MCR. This modification would decrease the grid plate open area from 1.45% to 0.92%. The holes were welded shut in the pattern shown in Figure 83. 196 holes were blocked in a 14 hole x 14 hole square directly below the fuel transport line. The purpose of this blocked area was to eliminate channelling caused by transport air and fluidizing air venting through the center of the bed. This phenomenon was observed both in the two-dimensional fluidized bed and in the SSU cold flow test, where a plexiglass wall was substituted for the removable waterwall. The pressure drop through the modified grid plate is presented in Figure 82. It is correlated to the equation:

$$\ln \frac{P_{grid}}{P} = 2.03 \ln U_{in} + 6.21 \quad (2)$$

SSU GRID PLATE HOLE BLOCKAGE

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Figure 83

The grid plate discharge coefficient C_D is calculated from the relation:

$$U_o = C_D \left[\frac{2g\Delta P_{\text{grid}}}{\rho_{\text{in}}} \right]^{1/2} \quad (3)$$

where,

- C_D = discharge coefficient
- U_o = gas velocity through grid plate holes, ft/sec
- g = gravitational constant, 32.2 ft/sec²

Figure 84 shows the calculated discharge coefficients as a function of the approach Reynolds number in the plenum for both the unblocked and blocked grid plate. The approach Reynolds number range was between about 20,000 and 50,000. There appears to be a tendency for the discharge coefficient to decrease with increasing Reynolds number. This is consistent with studies on grid plate pressure drop that show that plate discharge coefficients tend to decrease as Reynolds number increase to 30,000. Above 30,000, the coefficients are essentially constant. The increase in C_D as the number of holes is decreased is because the gas velocity through the grid plate holes increases as the plate open area decreases.

The average calculated discharge coefficient over the experimental Reynolds number range is approximately 0.70 for the unblocked plate. The discharge coefficient increases to 0.75 for the blocked grid plate. The theoretical discharge coefficient for a square-edged orifice is about 0.62. The higher discharge coefficient for the grid plate implies that there is some rounding of the upstream edges of the holes in the plate. This has decreased the overall pressure drop of the grid plate by about 30% relative to a plate with square-edged holes and a corresponding discharge coefficient of 0.62.

DISCHARGE COEFFICIENT vs REYNOLDS NUMBER

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Figure 84

2.7 Transient Response

2.7.1 Background

The dynamic temperature response of fluid beds determines stability limits of operation and control systems requirements. This analysis has been performed to evaluate temperature transients initiated from steady state conditions in the SSU fluid bed. The transients are started by introducing step changes in the coal flow.

The following assumptions were made:

1. A step change in coal flow corresponds to an equivalent step change in heat generation.
2. Total evaporative heat absorption changes only with bed temperature (steam temperature and overall UA are constant).
3. Bed height is constant.
4. The bed is homogeneous.
5. Radiation from the bed is approximately constant within the range of temperature transients.

2.7.2 Theory

A schematic of the SSU fluid bed steady state heat balance is presented in Figure 85.

The transient energy balance around the bed is written:

$$(Q_{ng} + \sum Q_{in}) - (\sum Q_{out} + Q_s) = \frac{ds}{dt} \quad (1)$$

SSU STEADY STATE FLUID BED HEAT BALANCE

* DOES NOT INCLUDE STEAM/WATER SENSIBLE HEAT

The individual terms are defined below:

The net heat generation in the bed is the heat released in the bed by coal combustion plus the net heat released in the sulfation and calcination reactions.

$$Q_{ng} = \eta_c (1 - \eta_p) M_c (\text{HHV}) + (Q_{sf} - Q_{ca}) \quad (2)$$

The sensible energy of the reactants entering the fluidized bed, assuming a 77°F reference temperature, is given as:

$$\Sigma Q_{in} = \dot{M} C_{p_a} (\theta_a - 77) + \dot{M} C_{p_c} (\theta_c - 77) + \dot{M} C_{p_l} (\theta_l - 77) + \dot{M} C_{p_r} (\theta_r - 77) \quad (3)$$

ΣQ_{out} is the total sensible energy of the combustion gas, elutriated solids, and drain solids leaving the bed, plus any radiation from the bed surface to the superheater or waterwalls in the freeboard.

$$\Sigma Q_{out} = \Sigma \dot{M} C_p (\theta_b - 77) - Q_{rad} \quad (4)$$

The heat absorbed in the bed, Q_s , is the sum of heat absorption in the bed tubes, test loops, and refractory.

$$Q_s = Q_{bt} + Q_{tl} + Q_{ref} \quad (5)$$

$$+ (UA)_{bt} (\theta_b - \theta_s) + (UA)_{tl} (\theta_b - \theta_w) + (UA)_{ref} (\theta_b - \theta_s) \quad (6)$$

The final term $\frac{ds}{dt}$, represents the time rate of change of thermal energy of the bed material and refractory. The thermal content of the bed is $M_b C_{p_b} (\theta_b - 77)$. Assuming a constant thermal conductivity through the refractory wall, the average refractory temperature is

roughly $\frac{\theta_b + \theta_w}{2}$. Thus, the refractory thermal storage is $M_{ref} C_{p_{ref}} (\frac{\theta_b + \theta_w}{2} - 77)$. Assuming constant steam side temperature, the change in bed and refractory heat content is then given as:

$$\frac{ds}{dt} = \left[\frac{M_b C_{p_b} + M_{ref} C_{p_{ref}}}{2} \right] \frac{d\theta_b}{dt} \quad (7)$$

Substituting equations 2, 3, 4, 6, and 7 into 1 and rearranging,

$$\frac{M_b C_{p_b} + M_{ref} C_{p_{ref}}/2}{\Sigma \dot{M} C_{out} + UA_{bt} + UA_{tl} + UA_{ref}} \frac{d\theta_b}{dt} + \theta_b =$$

$$\frac{Q_{ng} - Q_{rad} + \Sigma Q_{in} + \Sigma \dot{M} C_{out} (77) + [UA_{bt} + UA_{ref}] \theta_s + UA_{tl} \theta_w}{\Sigma \dot{M} C_{out} + UA_{bt} + UA_{tl} + UA_{ref}} \quad (8)$$

The time constant τ and forcing function θ_f are defined as:

$$\tau = \frac{M_b C_{p_b} + M_{ref} C_{p_{ref}}/2}{\Sigma \dot{M} C_{out} + UA_{bt} + UA_{tl} + UA_{ref}} \quad (9)$$

and

$$\theta_f = \frac{Q_{ng} - Q_{rad} + \Sigma Q_{in} + \Sigma \dot{M} C_{out} (77) + [UA_{bt} + UA_{ref}] \theta_s + UA_{tl} \theta_w}{\Sigma \dot{M} C_{out} + UA_{bt} + UA_{tl} + UA_{ref}} \quad (10)$$

Equation 8 now becomes

$$\tau \frac{d\theta_b}{dt} + \theta_b = \theta_f \quad (11)$$

This equation is a first order, linear differential equation with constant coefficients. The initial conditions are $\theta_b = \theta_i$ and $t=0$. The general solution is

$$\theta_b(t) = \theta_i + [\theta_f - \theta_i] [1 - e^{-t/\tau}]$$

The forcing function and time constant must be calculated for each step change in coal flow. Since it is assumed that a step change in coal flow corresponds to an equivalent step change in heat generation, the air, recycle, bed drain, and combustion product flows are based on the new rate of coal flow. It should be noted that the forcing function θ_f is the new steady state bed temperature.

2.7.3 Results

Two types of transient temperature response tests were run during the M Series. In the first case, step changes of 10, 20 and 30% of the coal feed were initiated at constant air/fuel ratios. Temperature response curves and model predictions are plotted in Figures 86 and 87 for the 10 and 20% load changes. The 30% load change was invalid because the coal feeder could not meter the coal rate accurately or consistently at such low flow rates. The bed temperature initially fell at a rate of 2.0°F/min and 3.3°F/min for the 10 and 20% step changes, respectively. The overall temperature drops were 41°F and 79°F, and the theoretical time constants τ were 18.5 and 21.0 for 10 and 20% load changes, respectively.

In the second type of transient response test, the coal flow was completely shut off while maintaining constant fluidizing air flow. The bed temperature initially fell at a maximum rate of 58°F/min. The theoretical time constant τ was 18.5 minutes.

The steady state energy balances after the 10 and 20% load changes are presented in Tables 9 and 10. The energy balance before the fuel shut-off transient test is given in Table 11. In each case,

BED TEMPERATURE RESPONSE AT 10% LOAD CHANGE

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BED TEMPERATURE RESPONSE AT 20% LOAD CHANGE

Table 9
Steady State M-33 - Bed Heat Balance
(After 10% load change)
 $T_{BED} = 1393^{\circ}F$

<u>Inputs</u>	<u>BTU/HR $\times 10^{-6}$</u>
Q_{ng}	1.5618
Q_{sf}	.0228
Q_a	.1563
Q_c	.0015
Q_l	.0005
Q_r	<u>.0158</u>
Total In	1.7587
<u>Outputs</u>	
Q_{cg}	.6715
	.0962
Q_e	.0962
Q_{bd}	.0244
Q_{rad}	.0650
Q_{bt}	.4004
Q_{tl}	.1191
Q_{ref}	.0377
Q_{ca}	<u>.0562</u>
Total Out	1.6897
Unacc For	69,000 BTU/HR
Unacc For	3.92 %

Table 10
Steady State M-35 - Bed Heat Balance
(After 20% load change)
 $T_{BED} = 1361^{\circ}F$

<u>Inputs</u>	<u>BTU/HR $\times 10^{-6}$</u>
Q_{ng}	1.3917
Q_{sf}	.0183
Q_a	.0954
Q_c	.0013
Q_l	.0004
Q_r	<u>.0067</u>
Total In	1.5134

Q_{cg}	.5287
	.1884
Q_e	.0633
Q_{bd}	.0216
Q_{rad}	.0650
Q_{bt}	.3807
Q_{tl}	.1129
Q_{ref}	.0365
Q_{ca}	<u>.0436</u>
Total Out	1.4407

Unacc For 72,700 BTU/HR

Unacc For 4.80 %

Table 11
Steady State M-37 - Bed Heat Balance
(Before Shut-down)
 $T_{BED} = 1551^{\circ}F$

<u>Inputs</u>	<u>BTU/HR $\times 10^{-6}$</u>
Q_{ng}	1.8791
Q_{sf}	.0589
Q_a	.2282
Q_c	.0017
Q_l	.0006
Q_r	<u>.0170</u>
Total In	2.1855

<u>Outputs</u>	
Q_{cg}	.9242
	.2736
Q_e	.1235
Q_{hd}	.0252
Q_{rad}	.0650
Q_{bt}	.4488
Q_{tl}	.1302
Q_{ref}	.0438
Q_{ca}	<u>.0666</u>
Total Out	2.1009

Unacc For 84,600 BTU/HR

Unacc For 3.87 %

roughly 75,000 Btu/hr or 4.2% of the total energy input is unaccounted for. The final steady state temperature corresponds to the forcing function when there is closure on the bed heat balance closure was attained by assuming a lower in-bed heat generation rate.

2.7.4 Discussion

The dynamic response of fluidized beds is dependent on the new steady state bed temperature (forcing function) and the bed time constant. The time constant increases with bed mass and decreases with the total rate of energy leaving the bed. For constant air/fuel ratios, the time constant increases as the size of the step change increases. This is because less sensible energy is leaving the bed in the combustion gas. Three time constants are required for the fluidized bed to approach within 95% of its final steady state temperature following a load change.

The new steady state bed temperature is mainly dependent on the size of the step change in coal flow, and to a lesser extent, dependent on the net sensible energy leaving the bed. The steady state temperature decreases as the size of the step change increases. This decrease is primarily because there is less heat generation within the bed itself. The theoretical model plotted in Figures 86 and 87 follows the actual temperature transients extremely well.

The steepest temperature transient occurred when the coal flow was shut off at constant air flow. Without any internal heat generation, the rate of bed temperature drop was determined solely by how fast heat could be conveyed away from the bed. The theoretical model plotted in Figure 88 agrees well with the actual temperature transient over the time span data was collected. However, as the bed temperature falls to below 400°F, the model is expected to break down. This is due to a drop-off in bed heat transfer coefficients as well as a decrease in steam temperature.

SHUTDOWN AT CONSTANT AIR FLOW

Both of these are assumed constants. These effects tend to decrease the rate of heat removal from the bed and increase the time constant.

One effect the model does not account for is the effect of residual char burn-out in the bed. When the fuel is initially stopped, no change in bed temperature is detected until the residual carbon burns up. After the fuel interruption following steady state M-37, the bed temperature remained constant for roughly 90 seconds before dropping. The burnout of the carbon inventory in the bed added a heat input equivalent to that produced by the coal feed for 1-1/2 minutes. The actual carbon inventory in the bed can then be determined to be about 0.5% by weight of the total bed mass. The time lag observed before the transients began was ignored when compared model predictions to actual temperature responses.

2.8 Bed Drain Reinjection

2.8.1 Introduction

Calcium utilization is limited because of the mechanism of sulfation on the limestone particle. The sulfation reaction takes place first at the surface of the particle. The reaction then progresses inward as SO_2 diffuses through available pores in the surface. Particle reactivity falls off rapidly as a hard shell of calcium sulfate forms on the surface covering the pores. Limestone utilization is limited since the inward diffusion of SO_2 becomes inhibited.

The following test was intended to provide an increase in overall calcium utilization. This was accomplished by promoting additional calcium utilization through re-injection of pulverized bed drain material into the bed. Additional limestone utilization would allow a reduction in Ca/S ratio (i.e., limestone feed rate), while maintaining sulfur removal efficiency.

2.8.2 Test Description

Coarse bed drain material containing unreacted limestone was removed from the bed. The bed drain material was pulverized to reduce the average particle size to 50% passing through 200 mesh. The particle size reduction exposed fresh limestone surface thereby increasing particle reactivity. The pulverized bed drain solids were then mixed with fresh coal feed and re-injected into the bed through the fuel transport line. After mixing with coarse bed material, the re-injected bed drain fines were ultimately elutriated from the bed, passed through the primary cyclone, and collected in the baghouse filter.

The primary advantage of reinjecting pulverized bed drain material is the increased particle surface area, making the particles more reactive with SO_2 . The unreacted limestone in the drain solids

is thereby provided with a second pass through the bed to allow additional SO₂ removal.

2.8.3 Results

The bed drain reinjection test was run during matrix steady state point N-16. The test conditions were as follows:

Coal Feed	286 lb/hr
Limestone Feed	66 lb/hr
Bed Drain Rate	69 lb/hr
Pulverized Drain Rate Injection	96 lb/hr
Ca/S Mole Ratio	2.1
Bed Temperature	1450 °F
Superficial Velocity	7.5 ft/sec
Bed Height	31.4 inches
Excess Air	14.4 %

Prior to reinjecting the pulverized bed drain material, SO₂ levels in the stack gas were measured at 990 ppm. Once the reinjection process was started SO₂ levels dropped rapidly. After about 30 minutes SO₂ leveled off to 550 ppm, a reduction of 44%. This reduction in stack SO₂ level is equivalent to an increase in overall sulfur retention from 72% to 85%. The overall calcium utilization increased with reinjection from 36% to 43%.

It should be noted, however, that the apparent reduction in stack SO₂ may have been somewhat affected by the location of the gas sampling probe. The gas sampling probe was located in the center of the cross-section above the economizer. Since gas concentration profiles indicated potential gradients near this probe, it is conceivable that the pulverized material elutriated straight up the middle of the unit. This could possibly result in a decreased SO₂ concentration in the vicinity of the probe.

The results of test N-16 is plotted in Figure 89. For comparison SO_2 removal at 1450°F without pulverized bed drain reinjection is included.

2.8.4 Conclusions

The above test showed that significant reduction in SO_2 emissions can be obtained by reinjection of pulverized bed drain solids. This additional SO_2 removal can be attributed to two effects. First, pulverizing the material exposes fresh sorbent and increases particle reactivity. Second, the drain solids are provided with a second pass through the bed.

From Figure 77, the Ca/S mole ratio may be reduced from 2.7 to 2.0 by reinjection of bed material and still maintain a sulfur retention of 85%. This translates into a 26% reduction of limestone feed with no loss of sulfur removal, for this particular set of test conditions.

SO₂ RETENTION vs Ca/S MOLE RATIO

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Figure 89

2.9 Coal Properties

The purpose of the test program was to evaluate several aspects of the SSU using a single coal (Old Ben Mine 24). No coal-related studies were planned except for a few matrix points operated with 1/4 x 1/8" and 1/8" x 0 coal instead of the standard 1/4" x 0 coal. However, in the course of the extended shakedown, the following operations related closely to the properties of the coal operation with considerably less than stoichiometric air, operation with subbituminous (Jim Bridger coal), test of Old Ben coal on PER PDU, Bench-scale tests of SSU coals, over-bed feeding of coal, changeover to Freeman (Crown II Mine) coal.

2.9.1 Sub-Stoichiometric Tests

As explained previously, during the initial shakedown, the desired 1500-1600 °F bed temperature could only be achieved with a thin bed. Increasing bed height increased the amount of immersed tube area and reduced the bed temperature. The 1550°F beds were too shallow to provide realistic heat transfer data and the operation was unstable at the lower bed temperatures resulting from operation with a deeper bed. Stable operation, with deeper beds, was achieved by increasing the coal feed without correspondingly increasing the combustion air flow rate. Under these conditions the bed drain fraction contained 20 to 25% carbon. High carbon losses in the bed drain were not a problem during any tests with Old Ben, Jim Bridger or Freeman coal when the bed temperatures were <1400°F and more than stoichiometric air was used for combustion.

2.9.2 Tests with Jim Bridger Coal

An alternate approach for obtaining heat transfer data with a thicker, over-cooled bed was to operate with a more reactive coal. It was anticipated that such a coal would continue to burn

efficiently at lower temperatures. A brief test was conducted with some surplus subbituminous coal as used in the "Jim Bridger" generating station.

As explained previously there were several complications during the test - the coal metering, bed drain, gas analysis systems were less than fully effective. However, despite the above problems, operation with this coal was significantly better than with Old Ben. In all tests conducted with Old Ben up to that time the bag filter fractions were black, whereas the bag filter fractions with Jim Bridger were the grey color of acceptable central utility flyash. The carbon contents of these fractions were approximately 30% with Old Ben and 10% with Jim Bridger.

At this time, the operation of the SSU was not adequate with Old Ben but appeared promising with Jim Bridger.

2.9.3 Tests on PER PDU

At the time the natural circulation demonstration plant was proposed coal type was not considered a significant factor in fluidized bed operation. Coal selection was based on competitive bidding with Old Ben being clearly the least expensive.

Some of the early shake-down problems suggested that Old Ben Mine 24 coal might be unduly hard to burn. Some of the same Old Ben coal was shipped from the SSU to the Pope Evans and Robbins, Inc. (PER) site in Alexandria, Virginia. A one day test was conducted with this coal by PER personnel. The test on the PDU, witnessed by C-E, showed that Old Ben coal burned with comparable efficiency and produced comparable bag filter carbon losses as the standard PDU Sewickly seam coal.

2.9.4 Bench-Scale Tests For Reactivity

As explained previously a representative sample of the pile was taken from the shipments of Old Ben and Freeman coal. These samples were extensively analyzed. During shake-down a coal sample was accumulated during each steady state point. Each sample was briefly analyzed. During matrix testing, coal samples were similarly accumulated and "ultimate" analyses were conducted on each sample.

In addition to the above, a series of analyses were conducted with samples of Old Ben, Freeman and Jim Bridger coal to determine how the properties of these coals related to their performance in the SSU.

Tests were conducted on the coals, on the volatile matter buttons produced from these coals, and on low density fractions separated from recycle samples. The tests included flammability, TGA-DTA, surface area (nitrogen) and density (helium). The density and surface area measurements conducted on volatile matter buttons gave the best indication of the amount of carbon that is likely to be found in the bag filter fraction when the coal is burned in the SSU.

2.9.5 Over-Bed Feeding Tests

The over-bed feeding tests were conducted as a means of eliminating the spouting of the bed above the pneumatic feed point. Over-bed feeding with 1/4" x 0 Old Bed coal was tried but no steady state point was taken because the amount of freeboard burning appeared excessive. Over-bed feeding with 1/4 x 1/8" Old Ben coal was encouraging and steady state data indicated the most efficient operation obtained with Old Ben coal. The improved combustion efficiency was undoubtedly due to the absence of fines that tend to be elutriated before combustion is complete.

During matrix testing reasonable operation was obtained with the over-bed feeding of 1/4" x 0 Freeman coal. The combustion efficiency was less than obtained during comparable in-bed feeding tests of the same coal size. The reduced efficiency was undoubtedly because the fines were exposed to a mechanism that reduced the chance of burnout.

2.9.6 Change-Over to Freeman Coal

The change-over to Freeman coal seemed to be advisable for several reasons:

- a. Sulfur content of Old Ben coal was a little low to meet 3% required in contract.
- b. Old Ben coal probably represents some of the least reactive coals mined in Illinois.
- c. Additional coal was required because most of the Old Ben three-car shipment had been exhausted during the extended shake-down.

The change-over to Freeman coal was made at a time when combustion efficiencies were being progressively increased to acceptable levels. However, the trends from all the data indicate that carbon losses varied between 15% and 20% of the coal carbon with Old Ben and 10% and 15% with Freeman. The improved operation with Freeman coal appears to be due principally to the improved burn-out of the fines. Presumably, the same proportions of fines are produced regardless of coal and the retention/elutriation mechanics are the same for only coal. The operation with Freeman coal is presumably more efficient because the same fines burned more completely in the same residence time.

With Freeman coal the least efficient tests were run with 1/8" x 0 and over-the-bed feeding of 1/4" x 0 coal. The most efficient tests were those run with 1/4 x 1/8" coal fed into the bed.

TABLE 12
Standard Analyses of Old Ben, Freeman and
Jim Bridger Coals

Proximate Analysis	Old Ben #24			Freeman Crown 2			Jim Bridger		
	1	2	3	1	2	3	1	2	3
Moisture	12.5	-	-	18.1	-	-	19.2	-	-
Vol Mattle	31.5	36.0	43.8	36.4	44.4	-	28.5	35.3	41.7
Fixed Carbon	40.5	46.3	56.2	36.6	44.7	-	39.9	49.4	58.3
Ash	15.5	17.7	-	8.9	10.9	-	12.4	15.5	-
Ultimate Analysis									
Moisture	12.5	-	-	18.1	-	-	19.2	-	-
Hydrogen	3.9	4.5	5.4	4.0	4.9	5.5	3.6	4.5	5.3
Carbon	57.3	65.5	79.6	55.1	67.3	75.5	52.5	65.5	76.7
Sulfur	2.7	3.1	3.8	3.2	3.9	4.4	0.4	0.5	0.6
Nitrogen	1.1	1.2	1.5	1.0	1.2	1.3	0.9	1.0	1.3
Oxygen	7.0	8.0	9.7	9.7	11.8	13.3	11.0	13.6	16.1
Ash	15.5	17.7	-	8.9	10.9	-	12.4	15.3	-
Heating Value BTU/lb	10,400	12,310	14,000	10,230	12,490	14,013	8,794	10,884	12,850
Mineral Oxides in Ash %									
SiO ₂		53.1			48.6			47.0	
Al ₂ O ₃		18.6			16.7			16.6	
Fe ₂ O ₃		16.0			20.7			17.1	
Ca ₂ O ₃		3.0			5.4			6.6	
Mg O		1.0			1.1			1.1	
Na ₂ O		1.0			1.4			1.4	
K ₂ O		1.0			1.6			2.2	
TiO ₂		1.6			1.0			0.5	
SO ₃		4.7			3.5			7.5	

1 % As Receiver
2 % Moisture-Free
3 % Moisture and Ash Free

2.9.7 Evaluation of Coal Analyses

The standard analyses of the three coals discussed are given in Table 12. Such data can generally be made available on a given coal by the Supplier. With such data some prediction of the operation of a given coal in the SSU can be made. The likely ranks of the coal are Old Ben, hv, Bd, Freeman hv Cb, Jim Bridger subA. These ranks are apparent from the (dry, ash free) carbon, oxygen, and heating values, based on a rationalization of the Classification of Coal by Rank by R. Hensel ⁽⁴⁶⁾.

Rank (moist, mineral matter, free calorific value) is in fact a measure of coal porosity. When this porosity is directly measured, it is found that small increases in area represent large changes in rank. It would seem that the actual porosity of the coal particle has little direct significance to the combustion process since the coal volatiles are readily released after the coal is fed into the fluidized bed. The porosity of principal interest is that which is developed as a result of the particle plasticizing, devolatilizing, and resolidifying. This is the pore structure that is available for reaction with the oxygen in the fluidizing "air". This pore structure is especially important to the fine coal particles which, the consensus believes, reacts according to the continuous reaction model.

For the purposes of making a laboratory determination of char properties the initial residual char can be simulated by preparing a sample in the volatile matter apparatus (ASTM D-3175). The physical properties of the button so produced can be studied nondestructively or by reaction in a small scale apparatus. The properties of three such buttons, crushed and then pulverized to a 100%-200 mesh are listed in Table 13. Reactivity tests in the Flammability and TGA/TA apparatus gave little indication of relative reactivity of the chars. However, as shown in Table 13 the surface area data (using the simplified, single-point determination in nitrogen) show a promising relationship to the degree of burnout of the fines noted in the tests with different coals. It is also likely that the volatile matter button process tends to reduce the differences in surface area between the different coals. The higher rank coals tend to form stronger buttons that require more energy (hence the creation of more surface) to reduce them to -200 mesh.

2.9.8 Conclusions from Coal-Related Tests

Based on the assumption that Freeman coal is somewhat more reactive than Old Ben coal, and the Jim Bridger coal is significantly more reactive than Freeman coal, the coal-related tests suggest the following:

- a. The SSU was generally operated more efficiently when a more reactive coal was fired.
- b. Lower coal reactivity resulted in poorer burnout of the elutriated char and hence in reduced combustion efficiency.
- c. Lower coal reactivity did not appear to result in any significant increase in unreacted carbon in the bed drain material. Hence the burning rate of the coarser particles did not appear to influence combustion efficiency.
- d. Present indications of reactivity - lower rank, improved operation in other reactivity sensitive processes, seem to be applicable to predicting performance in a fluidized bed.

TABLE 13
Analysis of ASTM Volatile Matter
Buttons from Old Ben, Freeman and Jim Bridger Coals

	Old Ben	Freeman	Jim Bridger
Heating Value	10,660	12,210	
% Dry H.V. in F.C.	47.3	46.6	
Ultimate Analysis			
H	0.6	0.6	0.6
L	68.7	68.2	71.9
S	3.0	3.6	0.4
N	1.0	1.2	1.0
O	1.2	6.7	1.7
Ash	25.5	19.7	24.4
Specific Surface M ² /gr			
Dry	42.5	50.6	62.2
Moisture, Ash-Free(1)	57.3	63.0	82.3
Density GR/CC			
Dry	2.0	2.1	1.9
Moisture, Ash-Free(2)	1.9	1.9	1.8

(1) $S_c = S_m / 1 - A$

(2) $1/GS = A/2.7 + 1 - A/G_c$

where A = Fraction Ash
 S = Surface Area
 G = Density
 M = Measured
 C = Correction

- e. Density and surface area measurements of char produced in laboratory appear to be good indications of relative burnout of the fines in the SSU.
- f. At this early stage in the development of this technology historical information on the coals and small scale tests should only be used to guide the initial selection of coals for testing in a fluidized bed unit.

As explained previously, the SSU was operated, as initially assembled, for the boilout and for the A&B test series. The West wall was then removed to allow modifications to the grid plate and to the bed refractory. The SSU was then tested for approximately 750 hours for the C thru G test series. The West wall was then removed again to enable many modifications to be made.

After the wall was removed for the second time, the bed tubes were examined for deposits and erosion. There had been no water flow in the two sets of test tubes during the end of the F and for all of the G series. Therefore, not surprisingly, these tubes were already badly deteriorated and losing surface in the form of flaky iron-rich deposits. Observations of the three remaining bed tubes, those tied into the wall, indicated the following conditions:

1. Approximately 20% of the tube surface was bare and polished. This portion was located in the "windward" faces of the tubes.
2. Approximately 20% of the tube surface was covered with a glazed, yellowish, deposit. These deposits were also on the windward faces of the tubes adjacent to the polished area.
3. Approximately 60%, the balance of the tube surface, was covered with a soft, blank, deposit.

The bare portion of the tubes was distinctly polished but there was no measureable metal loss. The yellowish deposits had a uniform thickness of approximately 1/32-inch and were hard and tightly bonded. A sharp scraper was required to remove a sample.

The black deposits varied in thickness between 1/8 and 1/32-inch and were loosely bonded. The sample was readily collected with aid of a laboratory spatula. The samples of both deposits were analyzed by X-ray defraction to determine the major components. The deposits were also analyzed quantitatively for determining their carbon, hydrogen and nitrogen contents and semi-quantitatively by X-ray spectroscopy to determine the other significant elements. The compounds found were as follows:

Yellowish deposits - Major components: calcium hydroxide, calcium oxide; minor components: calcium sulfate, calcium carbonate.

Black deposits: Major components calcium sulfate, silica dioxide, calcium carbonate; minor components: calcium oxide, iron oxide, sodium chloride.

The details of the elemental analyses are given in Table 14.

No attempt was made to determine the mechanism of the position. The mechanism doubtless varies with load, etc., as an indication black deposits were found on the top of yellow deposits. These deposits certainly reduced somewhat the heat transfer from the bed that would be expected with bare tubes. Neither the polishing of the bent tubes nor the build up on the bed tubes as neted through the end of the G series are considered potentially serious problems nor the Demonstration plant.

TABLE 14

Details of Semi-Quantitative Analysis of
SSU Tube Deposits by X-Ray Spectroscopy and
C-H-N Analyzer

Light Element (By x RF)	Deposits		Oxide Form %	Deposits	
	Yellow	Black		Yellow	Black
	(KDL 5202) %	(KDL 5203) %		%	%
Aluminum	0.36	3.1	Al ₂ O ₃	0.68	5.8
Silicon	0.61	2.7	SiO ₂	1.3	5.7
Sulfur	2.4	5.3	SO ₃	5.9	14.7
Chlorine	1.3	2.8	Cl ₃	1.3	2.8
Potassium	0.041	2.4	K ₂ O	0.049	2.9
Calcium	45.0	10.9	CaO	64.0	15.2
Total	<u>49.7</u>	<u>27.2</u>	TOTAL	<u>73.2</u>	<u>47.1</u>

Heavy Element (By XRF)			Oxide Form %		
	%	%		%	%
Titanium		0.36	TiO ₂		0.61
Chromium		0.013	MnO		0.0080
Iron	0.29	11.2	Fe ₂ O ₃	0.41	6.2
Copper		0.037	CuO		0.046
Zinc	0.0078	0.32	ZnO	0.0097	0.40
Strontium	0.061	0.072	SrO	0.072	0.085
Cadmium		0.16	CdO		0.18
Lead		0.28	PdO		0.30
Bromine		0.13	BR		0.13
Total	<u>0.4</u>	<u>5.6</u>	Total	<u>0.5</u>	<u>8.0</u>

Elements (By CHN)			Oxide Form %		
	%	%		%	%
Hydrogen	1.3	1.1	H ₂ O	11.6	9.8
Carbon	2.0	1.3	CO ₂	7.3	4.8
Carbon		25.8	C		25.8
Nitrogen	0.01		NO ₂	0.033	
Nitrogen		2.9	N ₂		2.9
Total	<u>3.3</u>	<u>31.1</u>	Total	<u>18.9</u>	<u>43.3</u>

		Deposit	
Summary		Yellow	Black
Light Element Oxides	%	73.2	47.1
Heavy Element Oxides	%	0.5	8.0
C-H-N Elements or Oxides	%	<u>18.9</u>	<u>43.3</u>
Total		92.6	98.4

SECTION 3: DISCUSSION OF OPERATIONS

3.1 Shakedown Operations

The shakedown operations portion of test program was started May 1, 1978. This phase of the program proved to be much more difficult and time consuming than initially anticipated. This phase of the program was started May 1, 1978 and not satisfactorily completed until January 22, 1979. The initial assumption was that the unit, built according to standards developed by others, would work acceptably from the time of initial light-off. Shakedown was viewed as the period required to adjust state-of-the-art systems and to make necessary improvements to two experimental items:

1. Natural circulation cooling of the fluidized bed.
2. Recycle reinjection.

The shakedown operation was to be simplified by operating with a single coal (Illinois #6 Seam from the Old Ben Mine 24) and a single limestone (1/8" x 30 mesh stone from the Columbia Quarry in Illinois). In the course of the shakedown the standard coal and limestone were changed and major modifications had to be made to the following systems that are essentially unique to a fluidized bed boiler:

<u>Coal</u>	-	High fines content of normal 1/4" x 0 product. Sulfur marginally low to meet contract specifications and coal reactivity not representative of the range of available Illinois coals.
<u>Limestone</u>	-	Initial limestone met specifications but another limestone found to be more reactive to SO ₂ .

- Gridplate - Was poorly fitted initially, had poor cold flow characteristics. Deteriorated after initial hot operation.
- Bed Tubes - Promoted excessive cooling of the bed.
- Bed Drain - Would only operate with excessive manual attention. Flow was hard to regulate to the needs of steady state operation.
- Recycle Injection - Material hard to feed and meter under all conditions of operation.
- Bed Level Measurement - Subject to frequent blockage. Any blockage made total system inoperative.

In addition, the following systems had unforeseen problems.

- Feed Water Pump - Internal breakdown.
- Turbine Meters - Susceptible to blockage. Bearing service life short.
- Bag Filter - Valves for backflow not fully effective. Bag filter plugged during initial operation.
- Automatic Draft Control - Not suitable for all weather outdoor service.

The SSU was progressively upgraded in several stages. Each stage consisted of a period (1 or 2 weeks) of sustained, around-the-clock operation. The unit was then shut down (1 to 4 weeks) for maintenance and the implementation of improvements. Nine (9)

periods of operation (A through I) were required to achieve the level of operation desired for effective "matrix" testing. The highlight of the periods of operation and the changes made during the later shutdowns are explained below:

3.1.1 Test Series "A"

The "A" Test series was first attempted May 1, 1978 but was delayed (failures of FW pump, turbine meters, and ruptured discs). Sustained operation by 3 shifts of 3 operators was finally started May 14-19 and May 21-26. Some "quick fixes" were required and some additional manpower was made available during day shift and during the May 19-20 weekend. During this period steam was generated most of the time while the unit was operated on gas alone, on gas and coal, and, for increasing periods on coal along. Operating experience was gained and future operating, sampling, and data collection techniques were evaluated. The principal accomplishments during this period were:

1. The operation of a recycle feed and metering system using the counterflow fluidization of a 4" column.
2. The operation of a dilute-phase recycle feed system.
3. Twelve hours of sustained operation with reasonable efficiency using a thinbed.

During these periods most of the problems reported above were encountered.

3.1.2 "A-B" Shutdown

The following items received attention:

<u>Bag Filter</u>	-	Inspected by Representative. Modifications made to recirculation valves.
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- Bed Drain - Reduced flow rate. Modified air-purge system.
- Recycle Hopper - Installed vibrator and purge air system.
- Recycle Valve - Removed, dismantled and checked. Added purge air.
- Solid Sampling - Improved access for bed probes. Prepared dust sampling equipment.
- Crushing System - Increased screener to crusher pipe size and added magnetic separator.

Most of the control systems were recalibrated and an inventory of dry coal was accumulated for the "B" test series. In addition, a number of housekeeping tasks were accomplished - ordering of spare parts, providing storage for parts inventory, providing better lighting and work space.

3.1.3 Test Series "B"

The unit was started up June 12 and operated essentially around the clock until June 23. This test series was plagued with a great many breakdowns and failures. There were a few periods of satisfactory operation but these could not be sustained or restored. Several periods of stable operation were obtained with generally too high coal feed rates and too low air rates. The following subsystems operated more effectively - bag filter, recycle, bed drain.

3.1.4 "B-C" Shutdown

A shallow bed was fluidized with cold air. Observations with a "Boroscope" revealed that many areas of the bed were slumped and

much of the air was channeling through points above the joints . A long shutdown was scheduled for making changes to the grid plate.

A new, single piece, grid plate with sandwich type joints at the margin was designed, constructed and installed. Refractory was also added to the removable wall. Prior to reinstalling the removable wall, a plexiglass window was made to allow the bed to be observed as it was during earlier cold-flow model tests. The objectives of this model type test were to:

1. Compare operation with new plate to operation in earlier model test.
2. Observe the effects of three factors that apparently disrupt the hot fluidized bed:
 - a. Bed burner
 - b. Pneumatic coal and limestone feed
 - c. Pneumatic recycle feed

The cold flow test with the new grid plate indicated effective and uniform fluidization. As expected the disturbance from the bed burner was excessive. However, the disturbance by the pneumatic air feed for the coal was also excessive.

In addition to the work described above the following improvements were also made.

1. The recycle line was redesigned to accommodate a water jacket for solids metering.
2. Further changes were made to the bag filter reverse air valves.

3.1.5 Test Series "C"

This series was run from August 14-18 and August 21-24. The following inter-related improvements were expected as a result of the modifications.

1. Higher combustion efficiencies.
2. Operation with deeper beds.

Operation during the "C" test series was much improved but neither the mechanical operation nor the combustion met the standards desired for "matrix" testing.

The unit now operated consistently and the following conclusions were made:

1. The Old Ben coal was probably hard to burn.
2. The bed was still over-cooled.
3. Recycling of fines back into the bed reduced bed temperature.
4. Bed drain operation was improved by a deeper bed level.
5. Bed level was hard to measure.

Near the end of this series a brief set of tests were made with a small amount of more reactive subbituminous A coal. Unfortunately, several systems - coal feeder, bed drain, recycle, gas sampling, were not fully effective. The results with the new coal were less definitive than desired. However, the following improvements were noted:

1. Less carbon in the bag filter.
2. Less carbon in the recycle and recycle bypass.
3. High burning rate at lower bed temperatures.

It was therefore apparent that the efficient operation of a fluidized bed combustor is made easier by changing to a more reactive coal.

3.1.6 "C-D" Shutdown

The C-D Shutdown occurred over the period August 25 to September 13. Preparatory work was extended into the normal startup period of the D Test Series. During this shut-down the data reduction program was first tried out with the eleven (11) C Series Steady State Runs. The results indicated high unaccounted for losses and a need for further refinement in both data collection and processing. The preparatory work for the D Series included:

1. Complete an alternate 1" pipe coal and limestone feedline.
2. Screen the feed coal into two size splits - 1/4" x 1/8" and 1/8" x 0.
3. Install additional nipples for sampling dust at the top of the freeboard.

In addition to the normal maintenance and calibration, the bag filter was further reworked. All the bags were examined and several were replaced due to holes.

3.1.7 Test Series "D"

The "D" Test Series was started September 3 and continued until September 22. The objectives of this series was to evaluate the operation with both 1" and 1-1/2" feedlines; to feed recycle directly to the bed and into the feedline; and to prepare for the later Matrix Tests with 1/8" x 0 and 1/4" x 1/8" coal. Many additional tests were also planned including clean air velocity traverses of the freeboard, dust sampling in the air heater to bag filter duct, closer characterization of: cyclone, bed drain, recycle rate measurement, bag filter, and rapid coal and limestone moisture measurement.

The unit operated well except for frequent blockages of the 1" coal and limestone feedline and the almost complete breakdown of the bed drain water cooled screw conveyor. During this test period, data were obtained on seventeen (17) steady state points. During these tests, the operation of the recycle cyclone was characterized with 1, 2, and 3 chambers opened. The reworked bag filter consistently produced a clean stack and reproducible hourly rates of discharge.

The conclusions from this test series were:

1. There was no great changes in operation due to any of the following:
 - A. Operation with double screen 1/4 x 1/8" coal.
 - B. Lower transport air rate with a 1" transport line.
 - C. Feeding the recycle into the transport line instead of directly to the bed.

2. A. 1" feedline should only be used for -1/4" coal in an experimental installation.
- B. Effective steady state operation with prescribed calcium/sulfur ratios is not possible without a consistent and adjustable bed drain system.

3.1.8 "D-E" Shutdown

The results from the "D" Series Test were analyzed with essentially the same program used for the C Series. The results again indicated high unaccounted for losses.

The preparatory work required for the E Series included:

- A. The complete rebuilding and modification of the water cooled bed drain screw.
- B. Development of a pneumatic over-bed feed system for coal and limestone.

In addition to normal maintenance, work was also done on the recycle rotary valve and on tuning the crusher to minimize the production of fines. The data retrieval system was also completed by the installation of the preprogrammed microprocessor.

3.1.9 Test Series "E"

The one week "E" Series was started October 9 and completed October 13. The objectives of this series were to evaluate the following factors that might lead to improved operation at higher bed levels.

- a. Operate with high and low transport air flow rates.
- b. Operate with coarser 1/4 x 3/16" limestone.
- c. Operate with overbed feed (hence without transport air).
- d. Shut-off one bed tube.
- e. Shut-off second bed tube.

The test series proceeded adequately except for high bag filter back pressures and frequent microprocessor malfunctions. The conclusions from the test series were:

- a. Doubling the transport air flow rate had negligible effects on the operation.
- b. Increasing limestone size tended to decrease bed expansion and therefore did not produce the desired bed operation.
- c. Feeding 1/4" x 0 coal and limestone onto the top of a deep bed was not successful.
- d. Removing bed surface allowed a significant deepening of the bed but the remaining cooling still higher than optimum.

3.1.10 "E-F" Shutdown

A brief three-day shutdown was scheduled to be followed by a two-day "F" test series during the week of October 16-20. Normal maintenance was carried out and adjustments were made to the bag filter.

3.1.11 Test Series "F"

For the "F" Series four tests were run at a single coal feed rate, with and without recycle, and with no water flow in the two test tubes. Recycle reinjection produced 100 F lower bed temperatures.

3.1.12 "F-G" Shutdown

The four F Series tests showed that major modifications would be required to reduce the cooling of the bed and provide controlled surfaces for measuring heat transfer rates. The G Series of tests was designed to evaluate several other aspects of the operation, with the existing bed cooling.

The "2D" cold-flow apparatus was run to determine how it might be possible to block part of the underlying grid-plate to reduce the "spouting" caused by the transport air. As a result of this study a 14" x 14" area of the grid-plate was blocked-off. A set of smaller, more moveable, probes for gas and temperature sampling were made. Repairs and improvements were made to the bed drain and recycle systems. An extensive amount of calibration was done on the instrumentation, on the gravimetric feeders and on the draft control damper.

In addition to normal maintenance and inventory of 1/4" x 1/8" was prepared.

3.1.13 Test Series "G"

The "G" series was run from November 6-10. The purposes of this series were:

1. To run a steady-state point using gas alone as a means of calibrating the unit for mass and energy balances.
2. To evaluate the operation with the following changes to reduce "spouting" at the fuel discharge:
 - a. 1" Fuel feed line.
 - b. 14" x 14" of grid-plate blocked off
3. Run unit under steady conditions for at least 24 hours to allow extensive probing for gas and temperature distribution in the freeboard.
4. Repeat over-the-bed feeding test with 1/4" x 1/8" coal.
5. Improve the over-all effectiveness of the operation and measurements.

Since the operation was now close to the standards for "Matrix" Testing, greater emphasis was placed on obtaining completely acceptable performance from the following sub-systems - gas analysis system, bed drain, recycle, bed level taps, variations in steam flow rate, microprocessor. Many corrections were made to obtain, to a general tightening of the performance of these systems.

As a result of the additional attention given to the subsystems, all of a planned set of tests were completed. Eight (8) steady-state points were completed and the 24 hour test conditions were maintained except for one excursion, of the feedwater control valve.

The blocking of the grid-plate appeared to boost the efficiency of the unit. An improved insight into the flows and reactions in

the freeboard was obtained from the probe studies. The test with 1/4" x 1/8" coal fed over-the-bed were the most efficient obtained up to that time.

At the conclusion of this series it was evident that most of the serious operating problems were either solved or were reduced to manageable levels. There seemed to be no major obstacle to running the planned Matrix Tests except the modifications required to the bed cooling system, and the ability to control the flow of bed drain material more accurately.

3.1.14 "G-H" Shutdown

The unit was shut down from November 11-December 20 for major modifications to the internals. The frontwall was removed and the following changes were made:

1. The center bend from the three bed tubes was removed.
2. Penetrations for eight water-cooled probes, were made.
3. Two large diameter viewports were added.
4. A series of 1/2" diameter ports included at 45° for bed pressure measurement, were added.
5. The high temperature refractory in the vicinity of the bed burner, was repaired.
6. Castable refractories was extended on all sides 20" further up the freeboard.
7. Fifty percent of the superheater surface was removed.

8. Sixty percent of the economizer surface was removed.
9. Penetration for screw feeder was made.

The grid-plate showed minor distortion but otherwise was in good condition. The seals were still leakproof, but the metallic packing had crept out at one corner and was replaced. Loosely bonded black deposits and tightly bonded beige deposits were removed from the bed tubes and analyzed.

The major changes described above also required a considerable amount of repiping, rewiring, change in instrumentation, etc. The extended shutdown also made time available for many other desirable improvements and maintenance tasks. In addition to normal maintenance and recalibration and the usual work on the bed drain and recycle, the following items received attention:

- Recycle lines - Redo piping.
- Crusher - Open up and inspect.
- Rotary fuel feed valve - Open up, check and replace gear drive seals.
- Gas analysis steam - Strip down and clean all lines.
- I.D. fan damper - Install new valve.
- Silos - Drain and clean out.

In addition, a considerable number of improvements were made to the data reduction computer program. These changes enabled better closing of the mass and energy balances.

3.1.15 Test Series "H"

The "H" series was run from December 20-22. The purpose of the test was to cure the refractory and scope out the effectiveness of the recent improvements. The cooling probes were not installed

for this series and the unit was run with much of the insulation not in place. The unit was run effectively at lower feed rates and higher bed levels than generally obtained during the earlier test series.

It was noted that it was difficult to maintain the bed drain flow at a desired low rate. It was noted, however, that high rate unrestrained flow could be readily stopped and started by cycling the bed drain mechanism.

3.1.16 "H-I" Shutdown

During this shut-down four sets of bed cooling probes were made and installed. The insulation was replaced on the boiler and the new piping was insulated. Normal maintenance was carried out and an adjustable timer was installed to operate the bed drain system.

3.1.17 Test Series "I"

The "I" series was run from January 8 to 12. The unit ran well until the initial bin load of dry coal was used up. Unusual January weather - hard rains followed by 0°F weather - produced unmanageable wet and frozen coal and limestone feeding problems. All the feed lines, the screener and the crusher were blocked at various times. When frozen coal was forced into the bunker, it thawed and it caused characteristic wet coal problems - erratic feed and pluggage of feed lines.

The unit was started up with the 1" feed line. There were several blockages of this line. The original 1-1/2" line was then replaced. Due to all the coal problems, only four steady-state points were completed during this one week test period.

3.1.18 "I-J" Shutdown

During this shutdown, all of the wet coal and limestone was removed from the bins. Additional cleanouts were provided in the coal and limestone feed lines. As well as normal maintenance, the turbine meters and coal and limestone feeders were recalibrated. A switch in test coal was made at this point to coal from Freeman Coal Company, Crown II Mine, also an Illinois No. 6 Seam coal.

Performance of the SSU during the "I" Test Series was analysed and judged to be suitable for the planned matrix testing. This shutdown therefore marked the end of shakedown operations.

Following the equipment shakedown and unit verification phases as well as unit design modifications (covered under shakedown operations) matrix testing began in late January 1979.

The purpose of this section is to document the operation of the subscale unit for that period of time wherein matrix testing was performed. That period bracketed between January 22nd, 1979 and May 2nd, 1979. During this time period five separate sets of runs were accomplished. Each set was given a letter designation. Therefore we began with "J" series wherein 11 matrix test points were taken. Followed by series "K" with 24 points in February of 1979. A test run labelled "L" series was done in March and encompassed to matrix test points. In April and May series "M" and "N" were completed. The former consisted of 37 points, the latter 33 points.

Formal matrix testing ended with the last steady state point taken on May 2nd as point N33. Coincidentally all operational activities ceased on the fluidized bed subscale unit at that time as well. A total of 15 matrix test points were run between January and May of 1979.

The test matrix covered under this phase represents in basic philosophy the guidelines set forth in the test program issued by KDL in February 2nd, 1978 entitled "Test program for Subscale Unit fluidized bed. Changes in the original test program were made as a result of our findings during the shakedown phase. Prior to January 1979 the SSU had undergone several design modifications. The test matrix was altered somewhat in concert with those modifications.

Briefly matrix testing was initiated after

- a. Much in-bed surface had been removed.
- b. Half the superheater and 60% of the economizer were cut out as well.
- c. The freeboard was refractoried an additional 50%.
- d. The center portion of the perforated plate beneath the projection of the feed needle was plugged off.
- e. A commitment was made to maintain the penumatic transport line at 1" diameter and no bigger.

The new test program which is being discussed under the various headings in this report did not change with respect to its objectives or with respect to the primary variables. These objectives being:

1. Thermal characteristics: demonstrations of natural capabilities; determination of heat transfer coefficients; fluidized level explosion and temperatures; transient response to primary input step changes.
2. Combustion characteristics: determination of combustion characterizing porometers, such as volatile matter release; freeboard combustion; sustained operation at load using different limestone sizes, level depths; determining equation temperatures; combustion efficiency.
3. Environmental characteristics: Considerations of gaseous pollutants (No_x So_x) and particulate matter emission. Both

the performance of the fluid bed as well as auxilliary gas handling equipment are to be evaluated during the matrix testing phases.

4. Hardware characteristics: the performance of the major subsystems will be studied, evaluated, documented and catalogued with respect to wear, breakdown frequency, etc. The purpose here will be to obtain technological first hand information with mechanical components as well as control type instrumentation.

As can be seen from the above no changes were made to the objectives of the test program. The changes implemented were:

- a. Fuel feedrates of 150/HR were called for with corresponding additive fuels.
- b. A second stipulation was to call any operation where bed temperatures were at 1400°F or below as not valid and therefore not worth pursuing.
- c. The use of transport air would have to be monitored very closely, using only minimal quantities of air necessary for material transport. Since that is non-preheated air which taxes the bed thermally as well as dynamically.
- d. Test conditions yielding thermal equilibrium at bed levels below 24" would not be considered.
- e. Recycle rates would not be modulated. Recycled char would either be on or off. When "on" a constant pressure drop would be held across the primary separator (collector).

Using the above mentioned 5 constraints the matrix test program was executed in ammended form yielding 155 steady state points without violating any of the parameters originally set down. Figure 90 follows this section and graphically demonstrates the sequence of matrix tests from January thru May 1979. The gaps between those test periods represents downtime drain time wherein operators resumed normal 8 hour work days; maintenance was performed on the equipment; inspections performed; first order calculations were run once on all accumulated data.

Based upon the information it yielded a test plan was formulated for the next period of operation as well as equipment calibration and the preparation of special probes, etc.

It should be noted that a test period consisted of a full run from beginning to end. In other words once series "J" was begun on March 12, 1979 the fluid bed was fixed up at midnight immediately following the end of March 11, 1979 and was not shut down until all the 50 steady state points were taken, on March 23rd 1979. Operation was accomplished by means of 3 shifts. Each shift consisted of 2 operators and one shift engineer. An analyst, an instrument technician assisted as well on a need basis during non-normal working hours and were always there from 8:00 until 4:00 p.m.

During periods of downtime the entire fluidized bed operating crew, engineers and technicians reverted to their normal assigned duties which included, maintenance of equipment, modifications, calibrations for the technicians. Data analysis, test planning and system component redesign if necessary for the engineers.

Prior to commencing the discussion on the individual points two more important items have to be mentioned, as it will help the reader understand the data gathering and operations process better.

CALENDAR SEQUENCE OF PARAMETRIC TESTWORK
CE/DOE SUB-SCALE UNIT FLUID BED BOILER

Figure 30

These items are 2 concepts actually that were used throughout the testing period to insure a high degree of accuracy and repeatability. The first concept was operational steady state. As was mentioned earlier a test series would encompass an entire block of time (5 to 10 days) wherein the SSU would be on line 24 hours a day. Except for interruption beyond our control. Such as power outages, plugged feed lines, frozen fuel or additive. These forced outages were short in duration however (1-2 hours each). It therefore follows that overall thermal and mass equilibrium were achieved early in the run and not disturbed grossly throughout the run. The only disturbances were those imposed upon the system by the test conditions. The individual tests were sequenced in such a manner as to create the least upsets in the system. In other words, if in a given series fuel feed introduction into the bed was made from more than one point or in more than one maneuver (mechanical vs. pneumatic) at various bed levels and air flow rates, all tests in a given mode were done first, thus minimizing the upsets. A steady state point was taken during one hour period wherein conditions were established for that one hour 2 hours prior and held for 1 hour after, unless the change for the next steady state point was so minimal so as not to warrant the one hour past the steady state hour, an example of that would be step changes in excess air value only while all other conditions remained constant.

The second concept was that of baseline tests. A single set of test conditions were picked (conditions approximating normal operation at design condition for the SSU) which subsequently became a standard by which the unit's calibration was checked. These baseline tests were incorporated into the matrix test program, and run at the beginning of each series; at the end of each series and in between as well whenever test conditions called for equipment changes that were different from the last set of points run within a series. In order to illustrate this

point consider series "L" a baseline test is called for at the beginning, prior to the start of test #25 and one at the end after point #54 was taken. Now consider series "M" where a baseline was run prior to the start of point #95 then one after point #75 then one more at the end of point #124. A baseline test was run after point #75 because transient response tests began which represented a drastic departure from the previous 29 points in series "M". Therefore a baseline test having the following parameters:

1. Total Air Flow
2. Limestone Feedrate
3. Bed Height
4. Combustion Air Temperature
5. Coal Feedrate or Steam Flow

The point will be evaluated by looking at variations from baseline of:

1. Steam Flow or Coal Feedrate
2. Superheater Outlet Temperature
3. Gas Temperature Out of Economizer
4. Flue Gas Analysis

Became a gauge and a continuous running check on the

- a. Operation of the system
- b. Calibration of instruments
- c. A time flow for steady state operation.

The next five segments of this section report on the operation only during matrix testing. No conclusions or analysis will be drawn here since this is done in other sections of the overall report.

3.2.1 Test Series "J"

Series J tests which represents the first block of tests in the official matrix series represents test points 1 thru 15 in of the original program.

A baseline test was run without recycle at 30" bed depth and coal feed rate of 200 lb/hr. The matrix points were run at full expanded bed depths varying coal feed rates from 60 lb/hr to 250 lb/hr with concurrent step changes in excess air from 10 to 45%.

Operation at 160 lb/hr feedrates resulted in unstable bed temperatures. Even at 200 lb/hr feedrate the average bed temperature stabilized at 1340°F lower than the minimum acceptable limit of 1400°F.

The operation at the higher feedrates (above 200 lb/hr to 240 lb/hr) yielded a thermally stable bed with temperatures ranging from 1450°F to 1510°F; bed depths of 30" to 34" expanded. Combustion efficiencies ranged from 90 to 94% based on chemical heat balance. Nominal gas velocity in the bed ranged from 7.5 to 9.0 ft/sec depending on the excess air to the unit.

The Ca/S mole ratio was maintained between 2.5 to 3.5. The resultant sulfur capture although acceptable at time 65% to 92% deviated excessively to be able to see any significant operational trends at this point.

3.2.2 Test Series "K"

The second block of matrix tests began with the light off on February 5 of the SSU. This block was designated series "K" which roughly encompassed original matrix points 16 thru 45.

This block of tests was highlighted by several events worth noting. After baseline tests, steady state criteria tests were run to determine the time required for bed composition to stabilize after or step change in Ca/S mole ratio of 4 to 2. Bed drain samples were taken once an hour and outlet SO₂ monitored until the levels steadied out.

Extensive temperature and gas probing of the freeboard and convection passes was attempted as well. The gas probing was for CO, CO₂, O₂, SO₂, NO_x.

In order to minimize localized in bed turbulence suspected to occur at the center of the bed, which resulted in excessive splashing of material to the freeboard an initial attempt was made at feeding the bed (coal and limestone) mechanically via a screw conveyor; thus eliminating the need for transport air. This initial attempt at screw feeding proved inconclusive as the feed system was improperly designed for this application. The idea was not abandoned however, the vendor was encouraged to design a screw feed system specifically tailored for this application.

The K series tests which encompassed 24 separate steady state points were run in groups with the major input variable being Ca/S mole ratio, therefore:

K1 and K2 and K29 were run at Ca/S ratio of 4.7 to 4.2

K3 to K15 were run at Ca/S ratios of 2.1 to 1.9

K16 to K23 were run at Ca/S ratios of 2.9 to 3.0.

Bed depths were controlled throughout this series between 30 and 45" as coal and limestone rates were changed to obtain the proper mole ratios and hold bed temperatures up. These temperatures fluctuated between 1470° and 1660°F. Combustion efficiencies in the low to mid 90% were obtained.

3.2.3 Test Series "L"

Series "L" began on March 12 and lasted 11 days of consecutive operation. The purpose of this particular block of tests which represents roughly the mid point of the original matrix program was essentially to gather large quantities of steady state data which running the subscale unit for an extended period of time.

Input variables where: the coal feedrate set at increments from 180 lb/hr, 230 lb/hr to 300 lb/hr in addition to two baseline tests at 250 lb/hr. Mole ratios of Ca/S of 3.5, 2.5 and 2.0. Bed levels to be maintained at a high expanded level, a medium level and or low level. These would correspond to roughly 40", 32" and 25". Excess air values for these tests would be bracketed between 10% and 30% with the use of recycle throughout. The recycle would tend to load up the unit with char thereby theoretically decreasing the excess air.

Operational information relative to running those tests showed variations in bed temperatures of 200°F between 1450°F and 1600°F. Sulfur removal as well seemed to have an exceptional wide bond ranging from as low as 50% to a high of 99%. Seemingly such primary variables as excess air, Ca/S mole ratio, bed depth have a significant effect on unit operational efficiencies. The effect no doubt is felt in gas residence time, optimum additive temperature for sulfur capture, char burn-out after devolatilization and elutriation rates.

This series of tests indeed demonstrated predicted performance response to a given set of input variables.

3.2.4 Test Series "M"

Test block M encompassed 37 steady state points. In this series several novel things were attempted. These were (a) overload feeding of coal and limestone; (b) recycle through the main transport line; and (c) transient response tests.

The overbed feed tests, matrix points 101, 102, 103, 104 coal and limestone was fed to the unit above the expanded bed by gravity through an observation port located in the center of the unit's right wall in the freeboard. The reasoning in this case was similar to that tried unsuccessfully in the "K" series with a mechanical feeder. Namely, the elimination of transport air in the bed as an additional source of air and therefore a potential zone of geyzering or excessive disturbance. The results from these tests were not conclusive. It is felt that more refinement of the actual hardware to accomplish this end would have yielded more favorable results.

The use of the transport line for means of conveying recycle char back to the bed was tried in three reported points #91, #92, #93. The reasoning here was to try and enhance carbon burnout. When char recycle was brought back to the unit through the standard mode, a line was provided in the bed at a place where short circuiting of the char could occur. Thus resulting in poor carbon burnup. It was therefore argued that if the char were to be fed back through the coal feedline and reintroduced into the unit with fresh fuel and additive, thereby acting in a sense as the binder in a coarse/fine heterogeneous mixture which subsequently discharged into the bed could only result in beneficiation for char burnout. Increased residence time being the overriding parameter in this case.

The transient response tests scheduled as tests 117 thru 124 in the matrix were designed to obtain thermal data on the unit. Namely, the heat capacity of the bed in terms of a relight on loss of fuel. These data would be significant for the operational constraints on the 50,000 lb/hr demonstration plant at The Great Lakes Naval Station.

3.2.5 Test Series "N"

The last block of tests run on the subscale unit which completed the laboratory phase of C-E's development of an atmospheric fluidized bed boiler was designated series "N" tests.

In this series 33 points were attempted with the subscale unit operating in several different modes. Halfway through the test the unit was shutdown for a weekend and the plenum opened up to unblock the center of the perforated plate. It had been blocked up during the modifications of December 1978 described in the shakedown section. The mechanical feed screw was re-installed at that time as well. The interruption to accomplish this work was after point #20 or at the end of matrix point #16.

Therefore, the test conditions for "N" were as follows:

- a. Run tests 105 thru 110 with coal sized at $1/8"$ x $0"$.
- b. Run tests 111 thru 116 with coal topsized at $1/4"$ x $1/8"$.
- c. Run extended steady state tests while performing extensive gas and temperature probing along the entire combustion chamber column, (freeboard, superheater, economizer) perform outlet dust traversing.

- d. Run or test by reinjecting pulverized spent bed drain material with the fresh coal.
- e. Run a test with the original contract coal (Old Ben).
- f. Perform a hot slump test from a 1550°F bed.

The rationale behind these variations to the standard primary parameter variables test was as follows:

In the case of firing "fine" (1/8" x 0") coal and "coarser" (1/4" x 1/8") coal we would want to obtain data of changes in combustion rates in the one case and elutriation rates in the second case. For similar Ca/S mole ratios (3.0) bed temperatures averaged in both cases 1450° to 1575°F for depths averaging 35". However SO_2 removal seemed to be somewhat consistently higher for the finer coal (99% vs. 90%).

With respect to the extended steady state run which lasted 36 hours, complete sets of data including gaseous components were obtained. Three complete sets of dust loadings were taken as well demonstrating particulate emission levels (non-combustibles) well below the EPA limit. The dust train was an EPA approved apparatus as described in Volume 17 of the Federal Register Clean Air Act of 1973.

A test was set aside wherein spent bed drain material (99% CaSO_4 , CaCO_3 , $\text{Ca}(\text{OH})_2$) was pulverized to 60% - 200 mesh and then fed to the bed with the row fuel. It has been postulated by several investigators that sulfur capture occurred on the surface of the lime particle therefore, if more surface was exposed during a grinding operation the rate of fresh additive makeup could be reduced by recycling spent bed material. This was done in Test #16 and #26, the former with pneumatic feed, the latter with

mechanical feed. The data gathered in either case with respect to sulfur capture, Ca/S mole ratio does not show dramatic changes. However a 50 lb/hr addition of recycled bed material does show a potential saving in fresh additive usage.

Two tests with the Old Ben coal #17 and 18 which originally showed to be a highly unreactive agglomerating coal demonstrated again the burning difficulty with which its been attributed. Bed temperatures of 1440° and 1480°F were experienced with relatively high feedrates of coal 290 lb/hr and a medium level bed height of 28" and 32".

At this point the mechanical feed screw was returned to operation after it was successfully modified. Results were ambiguous. They proved no better/no worse than penumatic feed. However due to a scheduling problem and time limitations the perforated plate was unplugged in the center prior to the feed screw test. Therefore a true comparison against the pneumatic feed system with a plugged plate was never made. It can be argued that this particular test was somewhat biased. Operational parameters such as bed levels, bed temperatures, Ca/S ratio of fuel feedrates yielded results comparable to ones performed with pneumatic feed under similar conditions.

SECTION 4 SYSTEM DESCRIPTION

4.1 Introduction

The fluidized bed boiler is designed to burn coal in a fluidized mixture with limestone to provide 200 lb/hr, maximum continuous rating steam at a superheater outlet temperature of 635°F and pressure of 365 psig. The unit has a 516 sq. ft. bed, 925 psig design pressure, 216 sq. ft. water wall surface, 525 sq. ft. bed tube surface, 45 sq. ft. superheater surface, and 137 sq. ft. economizer surface. The boiler has a 36" diameter drum, with safety valves capable of relieving the total steam capacity. Water walls are membrane welded 2" O.D. tubes.

The superheater tubes are 1-1/2" O.D. Economizer tubes are 1" O.D. and the bed loop tubes are 1-7/8" O.D.

A bed plate is used to distribute fluidizing air to the bed material. The bed plate has a 1.6% open area providing a 7 ft/sec nominal velocity at full load.

A bed burner is used to heat the bed material to achieve the desired ignition temperature. A duct burner is used to heat the fluidizing air so as to maintain the temperatures.

Bed loop tubes are provided to remove heat from the bed material. These tubes may be run as natural circulation or as a controlled circulation loop by utilizing the circulating pump. Bed test tubes are provided to determine the rate of heat transfer in the bed.

TUBULAR AIR HEATER - This is the tubular air heater system. It is designed to provide heater air to the furnace bed for fluidizing and remove the heat from the exit gas stream. This is a tubular type air heater manufacture by C-E Air Preheater.

DUST COLLECTOR - This is the ZURN cyclone flyash collector. Most of the flyash leaving the unit is separated from the gas stream by spinning the gas in the flyash collector. The ash is either re-injected or sent to a collection system.

BAG FILTER - The C-E Air Preheater bag filter system is intended to remove the small particles of dust or flyash from the exit gas stream.

I.D. FAN - The induced draft fan is manufactured by Spencer Turbine. It is a multi-stage centrifugal blower. A 25 HP 3500 RPM electric motor is used to drive the blower.

LIMESTONE STORAGE BIN - The vibra-screw live bottom limestone storage bin will hold properly sized limestone until it is needed. The bin bottom has a vibrator to assure continuous flow.

COAL BIN - The vibra-screw live bottom storage bin will hold properly sized coal until it is needed. The bin bottom has a vibrator to assure continuous flow.

SCREENER - The prater vibrating screener is used to determine the proper sizing of the coal and limestone being loaded in the storage bins. Rejects are sent to waste or to the Pennsylvania crusher.

BUCKET ELEVATOR - This is the #2 Universal bucket elevator, which transports coal or limestone material from the screw conveyor to the vibrating screener.

VIBRASCREW FEEDERS - The vibrascrew gravimetric feeds control the rate of coal and limestone material fed to the bed.

SCREW CONVEYOR - #2 screw conveyor, manufactured by Taunton Engineering Co., provides a variable speed drive screw type conveyor for transporting material from the Pennsylvania crusher or the crusher by-pass line to the No. 2 bucket elevator.

CRUSHER - Pennsylvania rotary reversable crusher. The crusher will be used to crush any coal and limestone which is rejected from the screener. From the crusher, material is fed back to the screener and rechecked for size. The crusher is driven by a 15 hp Allis Chalmers motor.

BUCKET ELEVATOR - The No. 1 Universal bucket elevator transports material from the No. 1 screw conveyor to the divertor valve where it may be routed to the crusher or to the No. 2 screw conveyor.

RECEIVING HOPPER - Coal or limestone is loaded in the receiving hopper to be transported to the material handling system.

CONVEYOR - The No. 1 screw conveyor, manufactured by Taunton Engineering Company, provides a variable speed drive screw type conveyor for transporting material from the receiving hopper to the No. 1 Universal bucket elevator.

F.D. FAN - This is a Lamson Diebold multi-stage centrifugal fan. A 30 hp Lincoln motor drives the fan.

EXPLOSION DOOR - In the event of a tube rupture or shear failure, the explosion door will relieve the pressure in the furnace.

BOILERFEED PUMP - Goulds multi-stage centrifugal boiler fuel pump driven by 50 hp Mariobhen motor.

The coal & limestone storage for SSU use is being handled in a staged maneuver.

Originally Old Ben Illinois #6 coal & later on Illinois #6 Freeman coal along with Medesa Aggregate limestone in a top size distribution of 1/8" x 0" was transported by rail and/or truck from their respective sources to a scaling in Hartford. The local siding selected is a storage facility where space is available on a rental basis.

At this point the coal and limestone is off, loaded, weighed and stored where cold. Periodically these piles are topped to replenish the active storage facilities which are located at the laboratory site. These active storage piles are also covered. From that point on any material drawn off is destined for the fuel preparation area of the SSU located in the auxiliary equipment enclosure (AEE).

The fuel/sorbent preparation system is made up of several pieces of equipment which are located in the AEE and are part of the auxiliary equipment complex which includes material preparation as well as the metered feeding system.

The preparation system consists of an outdoor receiving hopper of approximately 1 to 3" in capacity. The hopper whose 60° sloped scales feed a screw which conveys the makeup indoors to a bucket elevator. This bucket elevator then has the flexibility of feeding a Penn Crusher or directly conveying the material to screens located on the top of live bottom bins. In the case of coal handling, the bucket elevator will direct the coal to the crusher. When reduced down to size, (top size 1/4" x 0") the coal is conveyed through a second bucket elevator to moving screens located in top of the coal silo. Oversized coal runs off the screener and is returned to the crusher for a second time, the process is repeated till all the coal is satisfactorily crushed down to size.

In the case of limestone, which is purchased of size (1/8" x 0" top size) the crushing operation is bypassed. The limestone is diverted from the bucket elevator directly to the limestone silo.

A Griffin bag filter located within the makeup handling complex is also part of the preparation process and insures a safe environment in the enclosure. Figure 91 shows a perspective view of the equipment layout for makeup preparation. The systems capacity for loading the live bottom lines using the crusher is approximately 1 ton/hour depending on the condition of the coal and limestone. Each line holds 150 ft³ of makeup or 8000 of crushed coal and 10,000 of crushed limestone. Discussion of this bottom line will follow in the next section on feeding since it is considered part of the process demands' conditions.

PERSPECTIVE VIEW OF EQUIPMENT FOR FEEDING COAL AND LIMESTONE
FROM BINS INTO THE FLUIDIZED BED

1. COAL BIN
2. LIMESTONE BIN
3. COAL FEEDER
4. LIMESTONE FEEDER
5. FEEDERS TO AIR-LOCK TRANSITION
6. ROTARY AIR-LOCK VALVE AND PICK-UP ADAPTOR
7. FEED PIPE AND LANCE
8. ANNUBAR
9. CONVEYING AIR SUPPLIES

The raw coal and limestone feed system incorporate several pieces of hardware beginning with the live bottom bins and ending at the discharge point in the bed. After leaving the coal and limestone bins the material discharges onto a gravimetric belt feeder whose speed is remotely controlled. There is one feeder under each bin. Therefore limestone feed is metered separately from coal feed. They are joined into a single stream of material further downstream.

The point at which both materials join occurs in a duct level in the shape of Y. Each leg of the Y is located under the discharge chute of each gravimetric feeder belt. Both legs join and then the materials flow in a common stream although still segregated by gravity into a rotary air lock valve referred to as the fuel RAV, see figures 92 and 93. The rotational speed of that valve is fixed and set at a rate above the maximum feedrates of the combined streams.

At the discharge of the RAV the material drops into a small cylindrical lock hopper positioned laterally and flared out at the top to meet the RAV's flanged connections.

At one end of the drop hopper a 3/8" air line is connected where transport air is supplied to convey the material through a 1-1/2" diameter line to the needle located in the bed. As mentioned the feedline is a 1-1/2" diameter although at times, for test purposes it can be changed out to a 1" line. It is of schedule 40 wall thickness and runs from the rotary valve to the flanged connection at the units' point of penetration, see figure 82. The line has a total length of 35'. It runs uphill at 3% grade with 3 very long radius bends. This is due to structural interferences. As shown in figure 94 the discharge end of the pipe consists of an X

DETAILS OF VIBRASCREW
COAL AND LIMESTONE BINS

SIDE VIEW OF COAL/LIMESTONE FEED LINE

DETAILS OF X FITTING

arrangement, a 2-1/2 standard flange and a 2-1/2" lance. The lance penetrates the unit. The feedline or needle is located within this pipe is 1-1/2" diameter. The cross design provides for cleanout with the lance and a pocket to provide for autogenous wearing surface at the normal point of wear.

Dust Collector (Griffin Unit)

The Griff Jet-Vent pulse air dust collector and filter is capable of exhausting 900 cfm of air at a 6" negative pressure. The Unit is a system of 2" suction lines attached to the following critical points in the preparation plant.

- a. Bucket Elevator No. 1
- b. Bucket Elevator No. 2
- c. Crusher Inlet
- d. Limestone Bin
- e. Coal Bin
- f. Duct to Rotary Air Lock Valve

The purposes of the Griffin dust collector are as follows:

- 1. Maintain a negative pressure throughout the system to prevent the escape of explosive and respirable dust.
- 2. Prevent the accumulation of such dust within the system.

The line is fitted with a valve and two tees. The valve will be set before start-up to maintain the required balance of draft throughout the system. The valves must not be adjusted during normal operation. The air from all the suction lines is fed through a series of filter bages and exhausted as clean air into the room. The bags are sequentially shaken down by the automatically times pulse air system into a 55 gallon drum at the base of the unit.

Screw Feeder No. 1

This adjustable rate feeder moves the coal or limestone from the receiving hopper and meters it into the preparation system. The screw feeder No. 1 is an inclined Taunton Engineering 9" helical screw driven by a Reeves-Reliance Vari Speed Drive. This 5-1/2 ft. long feed screw is a Taunton design made for this application.

The Reeves drive consists of a Reliance 1016 3/4 hp parallel variable speed drive coupled to a 1040 CG 16A 60-1 right angle gear reducer. The screw speed is further reduced by the final sprocket drive. The maximum speed of the screw is 37 and the minimum 9 rpm. The torque for the final drive is applied to the chain through a Browning sprocket. If the screw is jammed, the sprocket will turn on the hub and shear the tube 1/4" screws. Before restoring the drive the feed screw should be freed and any obstructions removed.

Bucket Elevator No. 1

The Bucket Elevator No. 1 is a vertical conveyor for lifting the coal or limestone from the outlet of feed screw No. 1 to the main diverter valve. This valve provides gravity flow to either the crusher or directly to feed screw No. 2. The bucket elevator is a 24" Universal Model U2 800 Elevator driven by a 1 electric motor through a 1107 J-20 Reducer. The elevator is 24" wide, 16" deep and 27' high with a 24' discharge height. Vertical conveying is provided by the movement of a series of belt-mounted 7" x 4-1/2" buckets.

The elevator is rated at 16 TPH for coal and 30 TPH for limestone. The actual feed rate will be regulated by the adjustment of Screw Feeder No. 1. The feed rate will be governed by the effective capacity of the gravity feed lines, screener, or crusher - whichever is lowest. The construction of the elevator and the inlet and outlet are dust tight. Any leakage is drawn inward by the dust collector connection at the top of the elevator. Inclined slide gates are provided for access at the bottom of the inlet and outlet sides of the unit. These are kept closed during operation. These gates are used for cleaning out the base of the unit when changing over from coal to limestone and back. There is also a bolted access panel at the rear of the unit.

Prater Screener

The Prater screener consists of an enclosure, drive, inlet, outlets, and adjustments devices to enable a single steel screen to effectively separate the feed material into coarse and fine fractions. The operation of this equipment is described in detail in the attached booklet. This unit, Model VS-1-25-IE is a one-cut screener set with a 15° pitch of the screen cloth. The efficiency of screening may be the factor that determines the rate of coal processing.

In summary, the unit has a top inlet at the rear and outlets for coarse and fine fractions at the lower front end. The inlet and outlets are fitted with flexible "sicks" that will not restrain motion of the screener. The outlets consist of transitions from the top and bottom rectangular outlets of the screener to direct the material to appropriate round outlets.

The lower, fine fraction, outlet is connected to the screener product diverter valve. This valve directs the coal or limestone to the appropriate bin. The upper, coarse fraction, outlet is connected to the oversized return line and then to the oversized discharge valve and to the crusher.

Crusher

The Pennsylvania Crusher AS 00-116 Unit consists of the necessary enclosure, drive, inlet, outlet and adjustments to allow coarse feed coal to be reduced to $-1/4"$ by the action of the rotating swing hammers against breaker plates. The operation of the equipment is described in detail in the attached instructions and drawing. The crusher has a variable speed between 1400 and 1700 RPM. The capacity of the crusher at the crushing speed selected determines the rate of coal processing.

In summary, the unit has a top inlet and a bottom outlet. The breaker plates consist of two hinged half sections that can be moved closer to the hammers to compensate for hammer and plate wear. The plate adjustments are spring-loaded to relieve the impact of any unusually hard particles. The crusher is reversible so as to equalize the wear on both the breaker plates.

Screw Feeder No. 2

Screw Feeder No. 2 has two possible modes of operation:

- a. It takes all the feed from the crusher, or
- b. It takes all the feed directly from the bucket elevator and the crushed oversize from the crusher.

In either case, all the material is fed to the inlet of bucket elevator No. 2.

The feed screw is a Taunton Engineering screw similar to No. 1. Twin inlets have been provided for this 5' long unit. For details of the drive see feed screw No. 1 (Item 5).

Bucket Elevator No. 2

Bucket Elevator No. 2 is a vertical conveyor similar to No. 1. This unit takes coal or limestone from the outlet of feed screw and discharges it into the Prater Screener. The dimensions, capacity, drive, operation, cleanout, and catalog details are the same as for Bucket Elevator No. 1.

4.6 System Air Requirements

4.6.1 Forced Draft Air System

The forced draft (FD) air supply system, is composed of the following sub-systems, each having the stated specific purpose.

<u>Sub-System</u>	<u>Purpose</u>
FD Fan	Supply air at pressures up to 120 in. std. water.
Fluidizing Air	Supply fluidizing (and combustion) air to the fluidized bed.
Material Transport	Supply material transport air.
Ignitor Air	Supply air to the in-bed ignitor.

The primary purpose of the FD fan is to supply air to the other sub-systems at pressures up to above atmospheric pressure. The forced draft fan is a 30 Lamson Blower. The fan characteristic curve shows a maximum flow rate of 1000 ACFM at 95 in. std. water. A constant speed (3510 rpm), 30 hp Lincoln electric motor drives the fan.

The fluidizing air sub-system is provided to supply heated fluidizing air (up to 750°F) to the fluidized bed plenum. The heating of the air is accomplished by either the tubular air heater, the in-duct ignitor (described in the ignitor system description), or by a combination of tubular air heater and in-duct ignitor. The flow rate of fluidizing air is controlled by the 4" actuator control damper. Flow control is achieved from the control panel through the use of the Combustion Air Flow controller.

The operator is responsible for computing the required combustion air flow rate from the furnace superficial velocity, percent excess air, and coal feed rate requirements of the data point. Heated air flow rate is monitored by the 6" venturi preceding the furnace plenum. Air temperature is controlled from the control panel by the combustion air temperature controller which actuates a 2" damper

located in the tubular air heater by-pass line. After passing through the air flow venturi, the combustion air passes through the in-duct ignitor. The purpose of the in-duct ignitor is to supply further heat if the tubular air heater is unable to raise the fluidizing air temperature to the set temperature called for by the combustion temperature air controller. The heated air then passes through the furnace plenum, through the grid plate, and into the fluidized bed where fluidization and combustion occur. The surface of a fluidized bed is referred to as the balance plane since the SSU was designed to experience atmospheric pressure at that plane. Therefore, it is considered the terminal plane of the FD fan air supply system. The region from that plane to the furnace stack is considered to be within the induced draft fan system and is described under that system description.

The material transport sub-system is shown schematically on Figure 91. It is used to supply sufficient transport air for injecting the limestone and coal into the SSU furnace. Transport air versus material feed has been experimentally determined and the plots of transport air versus material feed rate are on file in SSU log. The flow rate of transport air is measured with an annubar. The procedure for providing transport air to the material transport sub-system is:

1. The FD fan (or shop air system) must be providing transport
2. Adjust the air flow rate to that required for material feeding to the SSU using the 2" damper annubar, and manometer mounted in the auxiliary equipment room.
3. Proceed to start the coal and limestone feed system as described in the description of that system.

The material transport sub-system may also be run with the shop air supply. The usage of the shop air system with the material transport system is described in the description of the coal and limestone system.

The purpose of the ignitor air sub-system is to supply combustion air to the bed burner heat to the bed prior to fuel injection. The in-bed ignitor air is supplied from a tap-off preceding the transport air and fluidizing air sub-systems. Air flow to the bed burner is controlled by the bed burner temperature controller in the control room no-flow or temperature control. A description of the bed burner system is contained in the description of the ignitor system.

4.6.2 Induced Draft Air System

The primary purpose of the Induced Draft (ID) Fan System is to remove the gaseous products of combustion (flue gas) from the SSU furnace. The ID fan draws flue gas through the cyclone, tubular air heater, and the bag filter; these systems are described in their respective system descriptions. The ID fan information is summarised as follows:

Spencer four bearing outboard turbo-compressor

Model: 1215-H

Horsepower: 26

RPM: 3500

CFM: 1500

Inlet Pressure: 13.5 PSIA (=33 in. std. water vacuum)

Inlet Temperature: 350°F (maximum)

The ID fan is driven by a constant speed GE electric motor having the following characteristics:

GE horizontal, ball bearing, polyphase induction motor

Model: 5K284BL115

Horsepower: 25

RPM: 3540

Voltage: 230/460

Since the ID fan is being used to remove flue gas from the furnace, the entire system is under a vacuum which varies from approximately 1/2 in. std. water of (vacuum) at the balance plane (immediately above the bed surface) to the maximum vacuum (33 in. std. water) at the FD fan inlet. This increasing vacuum profile through the system is consistent with the principal of pressure decreasing in the direction of flow; the ID fan is the device which brings the system pressure from a vacuum at the fan inlet to a pressure slightly above atmosphere at the damper outlet. The pressure decreases slightly as it passes through the stack and at the exit of the stack it is discharged at atmospheric pressure.

Both the ID fan and motor are mounted on a common base located on the roof of the KDL Laboratory above the SSU floor area. Also located on the roof are the 8" furnace pressure control damper, the flue gas stack, and an opacity meter. The furnace pressure is maintained at approximately negative -1/2 in. std. water at the balance plane by the controller labelled furnace pressure which controls the 8" pressure control damper. Furnace pressure is monitored by three pressure taps in the furnace area. The furnace pressure controller located in the control room, selects the medium signal (rejects the high and rejects the low value) as the furnace pressure.

The air preheat system on the fluidized bed boiler provides heated air to the plenum for fluidizing and preheating the bed. The air preheat system plays an important part in the overall cycle efficiency by removing heat from the gas stream and lowering the exit gas temperature. The system consists of a C-E tubular air heater, inlet and outlet ducts for air and gas, an airheater by-pass duct with actuator driven damper, and a F.D. fan control damper.

Instrumentation and control consists of temperature and pressure measurements for the airheater inlet and outlet on both the gas and air sides. Two control dampers with Foxboro automatic controllers for control of the air heater by-pass damper, and FD fan control actuator damper, which are regulated to control outlet air temperature.

The C-E tubular air heater is a four pass system with gas traveling through the inside of the tubes and air from the F.D. fan passes around the tubes. Baffles are used to assure a positive flow path of air streams to prevent any short circuiting and maintain counter flow between the air and the gas sides. Tubes and baffles run vertically through the air heater. Furnace gas, at approximately 660°F, exiting from the Zurn ash collector, enters the right side of the air heater via the air heater inlet duct, travels down through the air heater tubes to the bottom, is forced to reverse direction and travel up the right side outlet tubes. When the gas reaches the top it reverses direction and travels down the left side inlet tubes, up through the left side outlet tubes and out the air heater gas outlet duct. From the outlet duct about 295°F gas goes to the bag filter fly ash collector. Ash which settles out in the air heater double bottom hoppers is collected thru the fly ash collection pipe in a fly ash collection barrel. During controlled tests the fly ash barrel will be removed and weighed to determine the amount of fly ash collected.

Air at approximately 109°F from the F.D. fan goes through the 8" F.D. fan actuator damper and air inlet duct to the air heater. Air travels outside the tubes, making two passes on the left side and two passes on the right side. The air is kept in a flow path opposite the gas by baffles. This system maintains the highest differential temperature between the gas and air for efficient heat transfer. Air exiting the air heater can be controlled at 575°F. Upstream of the 8" F.D. fan control actuator damper is a 6" air duct which supplies unheated air to the bed burner and to the 4" air heater by-pass duct and actuator damper. The air heater by-pass damper is used to regulate the amount of unheated air which mixes with the air from the air heater, to control air temperature to the bed. The operator can set the temperature desired from the control room and the Foxboro controller will regulate the air mixture to give a constant air temperature.

During the startup of the fluidized bed boiler, auxiliary heat is required to raise the temperature of the bed material and the fluidizing air to a level which will induce and support combustion of the primary fuel. Auxiliary heat is supplied by the combustion of natural gas in two burners. A "duct burner" is located in the air supply duct to the bed plenum. This heats the fluidizing air to help support combustion of the coal. A "bed burner" is mounted in the left sidewall and directs heated air and natural gas into the bed to raise the temperature of the bed material and to initiate the combustion process. These are shown in Figure 95.

The duct burner is manufactured by North American Manufacturing Co. The burner is designed to operate in very high excess air and has a firing rate controllable between 50,000 Btu/hr and 800,000 Btu/hr.

The duct burner is intended to control automatically the fluidizing air temperature at a temperature set by the controller, located in the central control room with a control range of 100 to 1200°F. The burner firing rate is controlled by an adjustable port gas valve positioned by a Honeywell proportioning controller. A gas orifice metering unit is included to measure gas flow to the burner. The burner uses spark ignition of the fuel and flame is monitored by an ultra-violet scanner. The gas train includes supervisory valves and interlock switches to meet flame safety requirements.

The flame safety system is discussed in the control manual. Included in the flame safety system are purge controls, low fire start, ignition trial and monitoring systems. A high temperature limit control is used to shut down the burner if the temperature should exceed 1000°F.

DIAGRAMMATIC VIEW OF DUCT AND BED BURNERS
AS USED IN SSU

The bed burner is also manufactured by North American Manufacturing Co. and designed to operate in a high excess air environment. The bed burner is capable of being fired to a maximum input of 2,500,000 Btu/hr by the combustion of natural gas, with a minimum controllable rate of 150,000 Btu/hr. The design of the bed burner provides stoichiometric operation in the internal primary combustion chamber, with the excess air going around the primary combustion chamber. This burner is designed to be capable of raising the bed temperature to 1000°F for preheating the bed material. Temperature may be pre-set from the controller in the central control room.

Primary control of firing rate is accomplished by controlling the air flow rate to the burner with a North American adjustable port control valve positioned by a Honeywell controller. Percent of excess air is regulated by a limiting orifice air valve located in the air line between the primary air control valve and burner. A calibrated air orifice plate is located in the air duct for determining air flow rate to burner. The fuel flow rate is controlled by a cross connected pressure metering governor, which receives its signal from the air supply line to the burner.

A gas orifice metering station is used for measuring gas flow to the burner. The gas train controls include supervisory gas shut-off valves, and vent valves for flame safety requirements. Air and fuel interlock switches are provided to prove availability of fuel and air for operation. The flame is ignited by a high energy spark ignitor. Main burner flame is monitored by a ultra-violet scanner. A high temperature limit control is provided to shut down the burner in the event of excessive temperature. The high temperature limit is set for 2000°F. The flame safety system, covered in the control section, provides for prepurge low fire start, trial for ignition and main flame supervision.

This is a system description of the SSU Fluidized Bed Steam Generator which is shown schematically on Fig. 91. Reference to Figures 96 & 97 will be made throughout this discussion. The system consists of the following water side subsystems each having the specific purposes as indicated below:

<u>Sub-system</u>	<u>Purpose</u>
Economizer	Supply feedwater to the steam generator and test tubes, Heat Feedwater.
Drum	Steam water separation. Establish a circulation covering pressure drop. Mix feedwater with circulating water. Supply water to respective headers.
Waterwalls, In-bed Tubes	Absorb heat which generates steam.
Superheater and Pressure Control Valve	Superheat - saturated steam maintains SSU waterside pressure
In-bed test tubes	Absorbs heat, used for experimental purposes.

Before describing the specifics of each subsystem, it is worthwhile to point out the following characteristics of the steam generator sub-systems. All sub-systems are at approximately drum pressure (385 PSIG) which is primarily controlled by the pressure control valve; the small differences in pressure from component to component are flow-causing pressure drops which decrease in the direction of flow. This is a natural circulation steam generator; no pump is intended to be used to circulate the water.

The in-bed tubes circulation pump (Afton pump) is a backup to be used should the natural circulation concept not prove feasible; the feedwater pump raises feedwater to system pressure. All water and steam streams (with the exception of the in-bed test tubes)

SSU - FLUIDIZED BED
WATER & STEAM VALVE ARRANGEMENT
LEFT SIDE VIEW

Figure 96

TABLE II Continued

Run	T _B °F	Uo, ft/sec	RECY	T.L.I. BTU Ho, m, ft ² , °F	T.L.II Ho	T.L.I. Ho, conv	T.L.I. Ho, Rad	T.L.II. Ho, conv	T.L.II. Ho, Rad
N-1	1410	4.73	YES	56.76	40.69	41.92	14.84	25.59	14.79
N-2	1417	5.21	YES	55.14	39.42	40.18	14.96	24.51	14.91
N-3	1581	6.32	YES	55.64	43.29	37.13	18.51	24.82	18.46
N-4	1566	7.36	YES	55.51	44.61	37.34	18.16	26.49	18.12
N-5	1625	8.58	YES	57.42	48.00	37.87	19.56	28.49	19.52
N-6	1594	9.51	YES	58.33	48.65	39.51	18.82	29.87	18.75
N-7	1434	4.68	YES	58.52	50.34	43.21	15.31	25.05	15.28
N-8	1411	5.46	YES	57.17	49.14	42.33	14.54	34.32	14.81
N-9	1483	6.22	YES	56.48	48.86	40.56	16.33	32.57	16.30
N-10	1505	7.16	YES	55.78	49.30	38.98	16.80	32.53	16.77
N-11	1559	8.36	YES	57.87	51.63	39.16	18.70	32.96	18.67
N-12	1516	9.43	YES	57.87	51.85	40.81	17.06	34.82	17.03
N-13	1464	7.82	yes	59.54	51.28	43.65	15.94	35.35	15.92
N-14	1457	7.82	YES	57	48.30	41.23	15.78	32.54	15.76
N-15	1446	7.75	YES	56.22	41.38	40.99	15.22	31.18	15.20
N-16	1420	7.51	YES	56.27	46.28	41.60	14.67	31.62	14.65
N-17	1440	8.91	YES	51.54	43.03	36.50	15.03	28.02	15.01
N-18	1489	7.88	YES	54.69	45.92	38.59	16.10	29.85	16.07
N-19	-	-	YES	-	-	-	-	-	-
N-20	-	-	YES	-	-	-	-	-	-
N-21	1437	6.21	YES	57.70	51.47	42.52	15.18	36.31	15.16
N-22	1426	7.02	YES	56.56	46.12	41.62	14.94	31.19	14.93
N-23	1566	8.13	YES	57.62	50.91	39.73	17.90	33.03	17.88
N-24	1485	8.94	YES	57.85	51.35	41.72	16.13	35.24	16.11
N-25	1348	8.04	YES	62.48	54	49.03	13.46	40.56	13.44
N-26	1415	5.93	YES	59.88	50.64	45.14	14.73	35.93	14.72
N-27	1461	7.09	-	54.42	46.37	38.73	15.69	30.69	15.68
N-28	1581	6.39	YES	61.42	58.81	43.13	18.29	40.52	18.29
N-29	1541	6.32	YES	62.19	59.34	44.30	17.39	41.94	17.39
N-30	1539	6.34	YES	62.93	60.37	45.57	17.36	43.01	17.36
N-31	1497	7.11	YES	62.03	59.80	45.19	16.44	43.36	16.44
N-32	1585	8.19	YES	63.14	60.75	44.81	18.33	42.42	18.33
N-33	1512	9.37	YES	60.84	58.07	44.18	18.66	41.41	16.66

pass through the drum. Feedwater flow rate into the drum equals the steam flow rate out of the drum (with increased flow for adjustments to drum water level changes being the exception). The water in the economizer, water walls, in-bed tubes and test tubes is not completely boiled dry to steam (i.e., from water entering to steam only exiting). The superheater flow stream is composed of steam only; the steam temperature is increased by the hot flue gases. The source of all heat absorbed by the heat absorbing surfaces (economizer, waterwalls, in-bed and test tubes) is the combustion process in, the bed and freeboard.

4.9.1 Deaerated Feedwater System:

The feedwater Sub-System is designed to supply deaerated water from the storage tank to the steam drum and in-bed test tubes, at a sufficient pressure and quantity to provide water for steam production and adequate flow through the in-bed test tubes. Provisions are made for chemical addition to the feedwater to protect the boiler metal from corrosion.

The feedwater sub-system consists of a Goulds model 3935 feed pump, deaerator isolating valve, feedwater control valve, feedwater check valve, feedwater isolation valve, chemical feed valve, test loop isolation valves, water column, feedwater line and test tubes line. Orifices are provided in the feedwater line and turbine meters in the test loop line for flow indication.

Deaerated water from the laboratory deaerator storage tank at a temperature of 220°F is supplied to the Goulds Co. feedwater pump through a 4" feed pump supply line and deaerator isolating valve. The Goulds pump is a model 3935 multi-stage centrifugal pump supplying a 1600 ft. head at 20 GPM flow. Power to the pump is supplied by a 50 hp 3540 RPM Marathon Corporation electric motor. The pump shaft is provided with mechanical seals. Feedwater discharges from the pump through a 3" discharge flange to a 1" economizer feed pipe and a 1" test loop supply pipe. The feed-

water line is equipped with a drain valve for complete system draining, feedwater chemical addition line, and a Foxboro automatic control valve and drum level controller which control feedwater flow steam flow based on drum level and steam flow.

The bed test tubes form an independent circuit which receives water from the feed pump discharge, circulates the water through the test tubes submerged in the coal and limestone bed, into the test loop cooler, and back to the feed water pump inlet. The bed test tubes are designed for a 5GPM flow rate through each loop with a low flow alarm at 3 GPM. Flow Technology turbine meters in each loop are provided a flow measurement and low flow alarm. Temperature elements (RTD's) are used to measure temperature rise in each test tube. The test tubes cooler is designed to remove heat picked up in the bed before the water is returned to the feed pump suction header.

The feedwater line supplies water to the economizer for preheating to 390°F before entering the boiler. From the economizer water enters the steam drum by means of the 1-1/2" feedwater distribution heater, which distributes the water uniformly the length of the steam drum.

The steam drum is equipped with a Clark Reliance electrode type water column. Electrodes which activate a relay are provided for high water alarm, low water alarm and low water fuel cut-out. The water column also has two lighted sight glasses for visual indication of drum level. Remote drum level indication sight glasses for visual indication of drum level. Remote drum level indication is by a Foxboro drum level transmitter to the control room. If drum level falls below the low water fuel cut level, the fuel to the unit will be automatically shut off by the flame safety system.

4.9.2 Economizer Sub-system

The economizer is located near the top of the furnace generator as shown on Fig. 96. Deaerated water is boosted to system pressure by the feedwater pump, passes through a 1-1/2" Sch. 80 pipe and experiences a flow split. Approximately 10 GPM goes to the test tube circuit through the 1-1/2" sch. 80 horizontal pipe while from 0-6 GPM goes into the 1-1/2" Sch. 80 vertical pipe and through the 1" flow control valve, the stop valve and the feedwater flow measuring orifice. The feedwater flow control valve is a 1" Fisher valve actuated by a 3-element controller. The controller receives 3 signals (steam flow, feedwater flow rate, and drum level) and processes these 3 signals through a logic network to set a feedwater flow control valve position, that is, a flow rate. Feedwater next passes through a 1" check valve and a 1" stop valve. Feedwater next enters the economizer through the 1-1/2" Sch. 80 inlet. Temperature is measured at the economizer inlet and outlet. The purpose of the economizer is to heat feedwater from the 200-240°F range to approximately 400°F at the system pressure of 385 PSIA. Feedwater heating is accomplished by the flue gas which is cooled approximately 400°F after passing over the outside economizer surface area of 137 sq. ft.

4.9.3 Drum Sub-system

Feedwater exits the economizer and passes through a 1-1/2" Sch. 80 pipe and enters the drum at the connection shown on Fig. 96. The drum is a 36" inside diameter by 7 ft. long fabricated cylinder with attached 10" Clark Reliance water column, sight glass, and associated connections. The functions of the drum are:

Separation of steam and water entering the drum from waterwalls and in-bed tubes.

Mixing of feedwater with recirculated water.

Establishment of a water level which promotes natural water circulation and control.

Once in the drum, the approximately 400°F feedwater mixes with and slightly cools the saturated water which has entered the drum from the steam generating surfaces of the waterwalls and in-bed tubes. The normal drum water level is 4" below the horizontal drum center line with a high level alarm at 2" above the centerline. Above the water level steam is being drawn off to the superheater and below the water level water is being fed through the downcomer and headers to the waterwalls and in-bed tubes. The drum, downcomers & waterwalls are shown schematically on Figs. 84 and 85.

4.9.4 Waterwalls and In-bed Tubes Sub-system

Water which enters the waterwalls and in-bed tubes, absorbs fluidized bed heat, and then boils. Therefore, a water and steam mixture (referred to as a 2 phase mixture) flows vertically upward through the waterwalls. This boiling or steam generation results in the 2 phase fluid in the waterwalls having a lower bulk density than the water in the drum, downcomers and lower headers. This density difference between essentially two vertical columns, the downcomers and the waterwalls, results in a pumping action from the drum, through the downcomers, and up through the waterwalls and in-bed tubes. This pumping action is referred to as natural circulation. It is a major objective of the SSU program to demonstrate that the natural circulation capabilities of a fluidized bed combustor and it is expected that most of the boiler will occur in the in-bed tubes; since those tubes will experience the greatest exposure to the fluidized and consequently the most heat absorption. Should the in-bed tubes not be able to provide a sufficiently high water flow rate to keep the in-bed tubes from overheating, it will be necessary to provide pump-assisted circulation through the in-bed tubes with

the to GPM Afton pump. Initially, the in-bed tubes will be operated with pump assisted-circulation based on the results of the bedside heat transfer coefficients, determined by the in-bed test tubes (described in the in-bed test tube subsystem), a decision to reduce pump flow and eventually attempt natural circulation will be made. All waterwall of the boiler are refractory covered and not in direct contact with the hot fluidized bed. These 4 sides receive heat from the hot flue gases (approximately 1600°F) above the bed. The right side is not refractory covered and receives heat by direct contact with the fluidized bed and with the hot flue gases above the bed. The 2 phase steam and water mixture generated flows vertically into the headers at the top of the furnace or directly into the steam water separation apparatus located in the drum. After passing through the separation apparatus, the saturated water contributes to the water inventory (i.e., level) in the drum and will mix with feedwater and recirculate through the downcomers, headers, and waterwalls and in-bed test tubes. The saturated steam produced passes through the top half of the drum and through the 2" Sch. 80 steam outlet located on the top half of the drum and through the 2" Sch. 80 Sch. 80 steam outlet located on the top of the drum. Located immediately downstream of the steam outlet is the isokinetic steam sampler which extracts a 5% (of total steam flow) sample. The sample passes through a Hagen cooler.

4.9.5 Superheater Sub-System

The purpose of the superheater sub-system, is to heat saturated steam to a superheated condition. The superheater is basically a heat exchanger located below the economizer as shown schematically on Figure 90. The superheater is composed

of 45 ft² of 1-1/2" diameter 0.135" wall thick tubing arranged on a 3" triangular pitch. The superheater heats the inlet saturated steam from 440°F to 560°F with heat absorbed from the hot flue gas which experiences a temperature drop of approximately 500°F. After passing through the superheater, the steam next flow through the 2" horizontal 2" Sch. 40 pipe and then vertically upward through a 1" Fisher steam pressure control valve. The steam pressure control valve allows control of the pressure within the steam generator and in doing so experiences a 385 pressure drop across it. The 2" Sch. 80 horizontal pipe supports the superheater vent valves, the 1" Fisher pressure control valve, a 1" Consolidated safety valve, and a Hancock isolation valve. The superheater vent is normally blocked closed by the two 1" Hancock valves placed in series. These two valves are hand operated. The 1" Fisher steam pressure control valve is operated from the control room with the controller label Steam Pressure. The valve is closed at 0% output or setpoint on the controller and fully open at 100%. The steam exists through a 8" diameter silencer located on top of the roof. The safety valve is a 1" Consolidated safety valve which discharges through a 2" stack passing through the roof. The bypass valve is a manually operated Hancock valve, and is normally closed in order to force the saturated steam exiting the drum to pass through the superheater. Opening this valve will allow saturated steam to bypass the superheater and pass through the 2" Sch. 80 pipe to the 1" Fisher steam pressure control valve. After passing through this valve, the steam then exits through a silencer located on the roof.

4.9.6 In-Bed Test Tubes Sub-System

The two in-bed test tubes will be used to determine the bed-side heat transfer coefficient within the fluidized bed combustor. This will be accomplished by determining the temperature rise experienced by the water flowing through these test tubes. The

heat pick-up information thus obtained from the in-bed test tubes will also be used to determine if the in-bed tubes can be operated in a natural circulation mode. Water is supplied to the in-bed test tubes at rates up to 8 GPM per test tube by the feedwater pump through the horizontal 1-1/2" Sch. 80 pipe. The 1-1/2" pipe runs vertically upward along a structural steel column through a 1-1/2" manually operated flow control valve. The 1-1/2" pipe then turns horizontally and runs west at about a 7 ft. elevation for approximately 8 ft. and is then split into two flowstreams, each going into a separate in-bed test tube. The test tube's geometry is similar to the in-bed tube's geometry with the exception that the in-bed test tubes middle bend is made outside of the furnace. The test tubes then re-enter the furnace for another pass through the fluidized bed where more heat pick-up occurs. After exiting the furnace, the heated water passes through the test tube cooler located in front of the furnace right (removable) wall where it is cooled to approximately 220°F. The cooled test tube water flow rates are measured downstream of the cooler by two 1/2" Flow Technology turbine meters. Individual readouts for the turbine meters are located in the control room. Immediately downstream of the turbine meters are 1" flow control valves which are manually set prior to test start to provide 3 to 8 GPM through the test tubes. The test tube water next returns to the feedwater pump inlet through a 1" Sch. 80 pipe running east at about a 7 ft. elevation. This pipe immediately drops through the feedwater pump inlet and thru it in-bed test tube water is returned for mixing with pump inlet deaerated water. This water then begins its journey through the feedwater pump and is discharged into a 1-1/2" Sch. 80 pipe from which a flow split between the drum feedwater or the in-bed test tubes water occurs. In addition to the two turbine meters located in the in-bed test tubes, other instrumentation includes RTDs at each of the test tubes inlet, middle bend, and outlet from the fluidized bed. Pressure transmitters located in the test tubes inlet. The control room BETALARM

contains a high and low flow rate alarm for each of the two in-bed test tubes. In the event of an alarm sounding, the corrective action to be taken requires opening each 1" manual flow control valve located downstream of the water cooler. Equal flow should be maintained through both test tubes.

The particulate collection system is designed to collect the fly ash carried out of the furnace and not picked up in the cyclone duct collector. Light fly ash of about 10 microns or less will pass through the cyclone ash collector and the air heater to the C-E Ray-Jet dust collector, where it can be removed from the gas stream and collected for disposing. The air cleaning system consists of a fly ash collection barrel, reverse air fan, and bag filter dust collector with a solid state sequential timer, solenoid valves, pneumatic cylinders, and cylinder operated dampers.

Fly ash carried in the gas stream enters the lower part of dust collector and up through the 4 compartments which each have 12 fiberglass/teflon filter tubes. The filter tubes are designed to collect the fine dust particles, thus removing them from the exit gas stream.

The dust collector is equipped with an automatic sequential timer which controls the cleaning operation. The controller signals the timer to actuate the solenoid valves causing the outlet damper, in the compartment being cleaned, to close and the reverse air damper to open, thus reversing the flow through the compartment, collapsing the filter tubes and breaking off the collected dust cake. After a preset interval the reverse air damper is closed, stopping flow, and allowing the dust to settle out. After the dust is settled out, the timer sends a signal to the solenoid valve controlling the outlet damper, causing it to open and putting the compartment back on stream. The timer then advances to the next compartment and repeats the cleaning process. Dust which settles out in the hopper is collected in a barrel and weighted to determine the rate of collection.

Pressure gauges are provided in the control room to monitor the dust collector pressure drop.

During normal operation the pressure drop, from the inlet to the outlet, should be about 4 to 6 inches. Should the pressure drop become excessive the operator will need to check the frequency of cleaning and the dust loading which the system is expected to remove. A low pressure drop will indicate a worn or torn collector bag.

Provisions for checking the dust collection efficiency by sampling the fly ash in and out, have been provided. During initial operation the collection efficiency will be low until some dust has been collected by the bags.

The particulate collection system is completely automatic and should require little attention by the operator, except for checking the pressure drop and the fly ash removal system. If any problems occur with the control panel an instrument electrician should be contacted for troubleshooting.

SECTION 5: DATA ACQUISITION AND PROCESSING

5.1 Data Collection

5.1.1 Introduction

The primary objectives of the Sub-Scale Unit (SSU) Test Program are to demonstrate that an atmospheric fluidized bed boiler can be operated in a natural circulation mode and to provide the necessary experimental data to finalize the design of the demonstration unit. To meet these objectives, numerous data points must be acquired at each point in the test matrix. This data must then be reduced to give an indication of the SSU's performance and to determine if the unit is operating at steady state.

The data Acquisition System of the Sub-Scale Unit (SSU) collects frequently required data and transmits it to a microprocessor. In steady state mode, the microprocessor inputs five minute averages of the raw data and stores it on magnetic diskettes. Data is rerecorded on cassette tape and transferred to a data file on the main computer. The data file is then edited to include manually recorded data such as solid flow rates and analyses.

The data is finally converted by the main computer into engineering units and the SSU's performance is evaluated. The complete data analysis system consists of the data acquisition system, a microprocessor, a magnetic cassette recorder, a remote time-sharing terminal, the computer and the computer programs. A flow diagram for this system is shown in Figure 98.

Flowsheet of Data Acquisition System and Computer Program

Figure 98

5.1.2 Data Acquisition System

The acquisition and recording of SSU data at a steady state point is facilitated by the use of a computerized data acquisition and logging system known as the Doric Digitrend 220 Data Acquisition System. This system is capable of monitoring up to 100 data points or any sequence of data points continuously. Reference junction compensation allows for direct temperature readout. The Digitrend also performs analog to digital signal conversion for digital output display. The data collected by the acquisition system includes 47 temperatures, 11 pressures, 6 differential pressures, 4 flow rates, and 10 gas composition measurements. Complete data scans are taken once a minute. This information is then reduced for operator assistance or stored on magnetic diskettes by a Data General microNOVA microprocessor.

The microNOVA is a general purpose, user-programmable processor. It has a word length of 16 bits and has the capacity to address up to 32K words of memory. Data storage takes place in the diskette subsystem, which as a direct access, moving head disc memory. Each diskette is capable of storing up to 157K 16-bit words. A CRT is used to display data scans and reduced data locally.

The function of the microprocessor is two-fold. One use is for the storage and retrieval of raw data collected by the data acquisition system. Data may be stored in any one of 6 operating modes. These modes store data every 1, 5, 15, 30, or 60 minutes. The last 4 modes store an average of the last five 1 minute data scans in the selected time interval. The final storage mode is the steady state mode. This mode stores 5 minute averaged data and performs a 3% deviation check on the data. Selected data can

later be retrieved and recorded on cassette tape. The data is then read into the main computer from the tape for further data reduction.

The second function of the microprocessor is to display reduced data on the CRT for operator assistance. These displays include summaries of all pertinent parameters in the fluidized bed, combustion air system, the flue gas system, and the water/steam system. Two additional displays provide summaries of all important operating and performance parameters.

5.1.3 Manually Recorded Data

A sufficient amount of manually recorded data is taken throughout the SSU facility for the following reasons:

1. To evaluate the performance of subsystems (such as the air heater, bag filter, forced draft fan, and induced draft fan).
2. To provide a backup for the primary instruments that will be recorded by the data acquisition system.
3. As part of the surveillance of SSU equipment.

The manual data includes all local readouts (gauges, strip charts, and manometers) in the control room and surrounding the boiler and auxiliary equipment. This data is taken once per hour.

5.1.4 Sampling and Analysis

Sampling and analysis of gaseous and solid streams on the SSU provide extremely important information for characterizing the fluidized bed's behavior. A lot of time and effort are required to acquire reliable data. Sampling and analysis will be discussed under subheadings of: (1) Gas Sampling and Analysis; (2) Solids Sampling and Analysis; and (3) Water/Steam Sampling and Analysis.

1. Gas Sampling and Analysis

Gas sampling and analysis provides key information relative to combustion performance and emission levels of gaseous pollutants of the SSU.

The complete Gas Analysis Systems (GAS) are utilized. Each of these systems is capable of monitoring gaseous streams for CO, O₂, CO₂, NO_x, and SO₂ concentrations. A manual SO₂ station is also available for checking the SO₂ concentration leaving the boiler. The SO₂ is found by bubbling stack gas through a scrubber, adding an indicator, and then filtrating the solution with hydrogen peroxide. One of the GAS continuously monitors the gas concentrations leaving the boiler by sampling through a fixed probe just above the economizer. Constant gas concentrations, especially for SO₂, are required for a good steady state point. The second GAS is attached to a movable probe for gas sampling throughout the freeboard. Nine probe locations are used ranging from just above the bed surface to just above the economizer. Penetration depths across the furnace are varied in increments of seven inches.

2. Solids Sampling and Analysis

Solids sampling and analysis provides key information relative to combustion performance, emissions control, and the performance of ancillary equipment such as cyclone collectors and bag filters. In addition to the collection of representative solids samples for qualitative analysis, the sampling of solids in many instances provides a measure of mass flow rates. These rates are necessary for the calculation of mass and energy balances.

Solids samples are taken from any of nine locations, as shown in Figure 99. These samples correspond to the following streams:

1. Coal feed.
2. Limestone feed.
3. Griffin Collection.
4. Bed drain.
5. Recycle solids.
6. Recycle bypass solids.
7. Air heater dropout.
8. Bag filter dropout.
9. Stack solids effluent.

Coal samples are taken off the feeder belt 4 times during the course of a steady state point. These fractions are then riffled together to form a representative sample. The coal is routinely analyzed for proximate analysis, moisture, carbon, hydrogen, sulfur, and nitrogen. Less frequent analysis includes size distribution, higher heating value, and ash composition. These tests are all conducted in accordance with ASTM standards. Coal feeder calibrations are conducted routinely at the start and end of each weeks operation.

SCHEMATIC ARRANGMENT OF SSU SHOWING
SOLIDS METERING AND SAMPLING POINTS

- 1&2 COAL AND LIMESTONE METERING AND SAMPLING
- 3 DUST COLLECTOR WEIGHING AND SAMPLING
- 4 BED DRAIN WEIGHING AND SAMPLING
- 5 RECYCLE SOLIDS FLOWRATE AND SAMPLING
- 6 RECYCLE BYPASS WEIGHING AND SAMPLING
- 7 AIR HEATER DROPOUT WEIGHING AND SAMPLING
- 8 BAG FILTER WEIGHING AND SAMPLING
- 9 STACK METERING AND SAMPLING

Figure 99

Limestone samples are also taken off the feeder belts several times during a steady state point and mixed together to form one composite sample. The limestone is routinely analyzed only for moisture content. As with the coal, the limestone feeder is calibrated both at the start and end of each test week.

The griffin collection rate is a measure of coal and limestone fines which are entrained by air leakage through the rotary feed valve and carried away to the dust collection unit. The rate is computed by measuring the rate of material collected in the dropout drum. This sample is routinely sampled only for carbon.

Bed material is sampled by collecting four increments of material coming off the drain screw and riffing them together. Samples from the fluidized bed itself were taken for comparison with bed drain samples. Results show that the two samples are very closely sized, with hardly any attrition noticeable in the bed drain screw. Due to the similarity of the two samples, it was decided to sample bed material from the drain screw because it is simpler and safer. A comparison of the two sampling techniques is shown in figure 100. Bed drain samples are routinely analyzed for C, H, N, and S. Size analysis is also performed on each sample.

Recycle and recycle bypass material are both sampled by withdrawing a small increment from the bypass line. These samples are routinely sampled for C, H, N, and S. Size analysis is performed on an infrequent basis.

Air heater dropout is collected in a drum below the air heater. Dropout rates typically run about 5 lbs. per week. This sample is collected only infrequently.

BED DRAIN SAMPLE SIZING

LOCATION	OPERATION	SYMBOL	MEAN SIZE (MM)
BED DRAIN	K-24	○	0.64
LOW PORT	SLUMPED	●	0.56
HIGH PORT	SLUMPED	×	0.72

The bag filter dropout rate is calculated by measuring the rate of material collected in the drum below the bag filter. The samples are routinely analyzed for C, H, N, and S. Size analysis is performed on an infrequent basis.

Stack solids effluent rate is computed from a measured dust loading. Dust loadings are taken infrequently because of the long sampling time involved. During a six hour sampling period, sufficient material could not be collected to perform a size analysis due to the low solids rate.

5.1.5 Water/Steam Sampling and Analysis

Water and steam samples are taken to determine quality. Water samples are taken from the steam drum on an infrequent basis during operation of the SSU. These samples are subjected to a pH and conductivity analysis to monitor water quality. Steam samples are taken from the steam drum to monitor the quantity of water carryover. This is determined by using the ASTM sodium-tracer method for measuring steam purity. The measured steam quality is 99.999 + %.

Two computer programs, SSU77 and COMPAR, are available for data conversion, mass and energy balance calculations, data reduction, and retrieval of results from previous runs. The two programs are available on the Network Operating System (NOS), which is a multimode job processing system developed by Control Data Corporation (CDC). These programs are executed through a remote timesharing terminal on the CDC Cyber 172 computer.

A data file is created primarily of data recorded on magnetic diskettes by the Data General MicroNOVA microprocessor. This data is rerecorded on magnetic cassette tapes for transferal to the CDC computer. Data is read into the computer by tape and by manual input at the terminal. The tape data consists of 47 temperatures, 11 pressures, 6 differential pressures, 4 flow rates, and 10 gas composition measurements. The manual input data includes test number and test date, recycle water flow rate, drain rate, bag filter and recycle bypass rate, and griffin collection rate. It also includes partial solid analysis for all streams and the humidity.

Program SSU77 may be executed once a data file is established. The data is first converted from volts into engineering units. Calibration curves for the feedwater and steam flow orifices, the bed tubes and combustion air flow venturis, and the transport air flow annubar are generated by standard ASME procedures, as outlined in Principles and Practice of Flowmeter Engineering, by L. K. Spink. Orifice coefficients for the feedwater and steam orifices are obtained from calibration tests conducted by Alden Research Laboratories at Worcester Polytechnic Institute. Manufacturer's calibration curves for the two test loop turbine meters are curve fitted by line segments over several regions. The fluidized bed temperature is computed as the weighted average

of temperature measurements taken from those thermocouples fully immersed in the bed. The recycle solids flow rate is computed from an energy balance around the waterjacket.

After data conversion is complete, program SSU77 performs an overall mass and energy balance around the SSU. The balance envelope includes only the fluidized bed boiler. Since the cyclone remains outside of the envelope, the recycle stream is treated as an input. The balance envelope is shown in Figure 101. The objective of this portion of the program is to determine if the fluidized bed is actually at steady state conditions. Closure of the mass and energy balance within reasonable limits indicate the validity of a test point. Closure limits are determined by operating experience.

The inlet streams for the overall mass balance are the coal and limestone feed rates, the recycle rate, the feedwater and test loop water flow rates, and the combustion air and transport air flow rates. The outlet streams for the mass balance include the drain and elutriated solid flow rates, the test loop water flow rates, the steam flow rate, and the stack gas flow rate.

The heat inputs for the overall energy balance include the sensible heat of the limestone and coal feed, the recycle stream, the feedwater and test loop water streams, and the combustion air and transport air flow streams. Heat inputs also include the chemical heat from combustibles in the coal and the recycle stream, and the heat of sulfation. The higher heating value (HHV) of the coal is calculated from a fixed value of HHV on a moisture-free ash-free basis. HHV's from numerous tests were found to correlate well with this value. The heat of sulfation is released in the exothermic reaction of SO_2 with calcium oxide to form calcium sulfate. The steady state rate of formation of calcium sulfate is determined by a sulfur balance.

Mass and Energy Balance Envelope

The heat outputs for the overall energy balance include the sensible heat of the bed drain and elutriated solid streams, the test loop water streams, the superheated steam, and the stack gas. Heat outputs also include the chemical heat from combustibles of the bed drain, elutriated solids, stack gas, and the heat of calcination and heat losses. The heat of calcination reflects the endothermic decomposition of limestone to calcium oxide. The steady state rate of calcium oxide formation is equal to the molar rate of calcium carbonate feed. This rate assumes complete calcination of the limestone. The heat loss is determined from a temperature probing of the boiler exterior. The furnace exterior is divided up into zones with heat losses determined for each region. Enthalpy and density values for water and steam are determined by an in-house steam table subroutine called SRSORT. This routine is based on the ASME steam tables, and requires only water or steam temperature and pressure as inputs. Enthalpy values from the steam tables are then corrected from a 32°F basis to a 77°F basis to be consistent with the rest of the program.

Following the mass and energy balance, the program performs a carbon balance as another check on data validity. Inputs include both carbon and carbonate carbon in the coal, limestone, and recycle solids. Outputs include carbon in the recycle or recycle bypass, bag filter, and bed drain. It also includes carbon in the stack gas as SO and CO₂. A large percentage of unaccounted for carbon is a good indicator that the stack gas analysis or coal calibration is suspect.

The program next proceeds to data reduction. Combustion efficiency is calculated by three methods. The reported efficiency is computed on a chemical heat basis. It is defined as 1 minus the total chemical heat out divided by the heat input from the coal. The chemical heat includes CO and carbon heat content in the stack gas, bag filter, recycle bypass, and the bed drain. A

second combustion efficiency is computed on a sensible heat basis. It is defined as the difference between the total sensible heat output and input, divided by the heat input from the coal. The third efficiency is based on the stack gas oxygen concentration. It compares the stack O_2 concentration with the O_2 concentration expected for complete combustion. The difference is then related to unburnt carbon. The efficiency is defined as 1 minus the heat content of the unburnt carbon and CO divided by the heat input of the coal. The last two efficiencies are performed as a check on the reported chemical heat combustion efficiency.

Bed height is obtained by measuring the bed static pressure at 4 different levels. A straight line is fitted through these points on a pressure versus height plot and extrapolated to furnace pressure. The height at this point corresponds to the bed level. The data is fitted by a least squares curve fitting. The resulting simultaneous equations are solved by Cramer's rule. The grid plate pressure drop is calculated from a correlation determined from grid plate pressure drop versus velocity data. The equation predicts that pressure drop divided by the gas density is directly proportional to the velocity thru the plate raised to the 2.03 power. A check on the calculated pressure drop can be made by extrapolating the pressure versus height line to zero bed height. The difference between the plenum pressure and pressure at zero bed height is the grid plate pressure drop. The bed pressure drop is the sum of the pressure at zero height and the furnace pressure.

Excess air is defined as the percent air flow above the theoretical air requirement. The theoretical air is determined from the oxygen required for complete combustion of the coal, based on the test coal analysis. It also accounts for the oxygen required for the reaction between CaO and SO_2 to form $CaSO_4$. The bed superficial velocity is defined as the stack gas volumetric flow rate corrected

to bed temperature and pressure divided by the bed cross-sectional area. The freeboard superficial velocity is computed from the stack gas rate corrected to the freeboard temperature and pressure divided by the freeboard cross-sectional area. A check on the bed velocity is made by substituting a calculated gas flow through the bed for the corrected stack gas flow. The calculated flow rate assumes complete combustion of the coal volatiles. The stack gas analysis, the nitrogen flow rate in the inlet air and the computed stack gas moisture from combustion products and feed moisture. It assumes complete combustion of the coal volatiles. The computed stack gas flow rate agrees reasonably well with the measured rate when explosion damper and air heater leakage is accounted for.

The calcium to sulfur mole ratio is defined as the moles of calcium in the limestone feed divided by the moles of sulfur in the coal feed. The sulfur removal is determined by the difference between the sulfur feed rate in the coal and the sulfur rate out the stack as SO_2 , divided by the sulfur feed rate in the coal. Calcium utilization is defined as the moles of calcium sulfate formed divided by the moles of calcium available for reaction with SO_2 . It is computed by dividing the sulfur removal by the calcium to sulfur mole ratio. NO_x and SO_2 emissions are both reported on a lb per million Btu basis. The NO_x is treated as if it were completely NO_2 for this calculation. The NO_x and SO_2 emissions are also corrected to a 3% O_2 concentration in the stack gas.

Heat transfer coefficients for both test loops, the economizer, and the superheater are calculated. The overall heat transfer coefficient is defined as the heat absorbed by each stream as it passes thru the tubes, divided by the tube outside surface area and the log mean temperature. The inside heat transfer coefficient is calculated by the Dittus-Boelter correlation. This correlation

is based on the bulk mean temperature in the tube and assumes turbulent flow through the tubes. Thermodynamic properties for the correlation are curve fitted by Stirlings Interpolation Formula. The outside heat transfer coefficient is back-calculated from the overall and inside heat transfer coefficients. The outside heat transfer coefficient is then broken down into its two components, a convective term and a radiative term. The radiative term assumes an emissivity of 0.85. Freeboard heat release is determined by a heat balance on the gas side between the bed surface and the furnace outlet. It is defined as the sum of the above bed heat pickup in the water circuit and the sensible heat of the gas and elutriated solids leaving the furnace minus the sensible heat of the gas and solids leaving the bed. It neglects any heat transfer to the freeboard by solids splashing from the bed surface. It also neglects any radiation from the bed surface to the freeboard.

Following execution of program SSU77, key results are automatically stored in a data base. These results may later be compared with other runs by executing program COMPAR. This program will tabulate up to eight runs in columns for direct comparison of results. Among the values retained are stream flow rates, gas compositions, heat transfer coefficients, and the reduced data discussed above.

In the appendix are tables containing a listing of the two computer programs, sample inputs and sample outputs. A section is also devoted to derivation of important SSU77 equations. Another section includes a list of nomenclature for SSU77.

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APPENDIX I

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I. Combustion Efficiencies

A. Overall Combustion Efficiency on Sensible Heat Basis

$$\eta_c H_{\text{COMB}} + Q_{\text{sen}_{\text{in}}} = Q_{\text{sen}_{\text{out}}}$$

$$\eta_c = \frac{Q_{\text{sen}_{\text{out}}} - Q_{\text{sen}_{\text{in}}}}{H_{\text{COMB}}}$$

where η_c = overall combustion efficiency

H_{COMB} = heat content in coal feed, Btu/hr.

$Q_{\text{sen}_{\text{in}}}$ = sensible heat in, Btu/hr.

$Q_{\text{sen}_{\text{out}}}$ = sensible heat out, Btu/hr.

B. Overall Combustion Efficiency on Chemical Heat Basis

$$H_{\text{COMB}} + Q_{\text{sen}_{\text{in}}} = Q_{\text{sen}_{\text{out}}} + Q_{\text{chem}}$$

$$H_{\text{COMB}} = Q_{\text{sen}_{\text{out}}} - Q_{\text{sen}_{\text{in}}} + Q_{\text{chem}}$$

$$1 = \frac{Q_{\text{sen}_{\text{out}}} - Q_{\text{sen}_{\text{in}}}}{H_{\text{COMB}}} + \frac{Q_{\text{chem}}}{H_{\text{COMB}}}$$

$$1 = \frac{Q_{\text{chem}}}{H_{\text{COMB}}}$$

$$\eta_c = 1 - \frac{Q_{\text{chem}}}{H_{\text{COMB}}}$$

where Q_{chem} = chemical heat out, Btu/hr. (includes CO and carbon heat content in stack gas, bag filter, recycle bypass, and bed drain.)

C. Overall Combustion Efficiency based on Stack Gas O_2 Concentration

$$HCHAR = (O_2 - O_{2TH}) \left(\frac{12 \text{ lb C}}{\text{mole}} \right) (14093 \frac{\text{Btu}}{\text{lb C}})$$

$$\eta = \frac{HCOMB - HCHAR - HCO}{HCOMB}$$

where HCHAR = chemical heat retained in unburnt carbon, Btu/hr.

HCO = chemical heat retained in stack CO, Btu/hr.

O_2 = molar flow rate of O_2 in stack, moles/hr.

O_{2TH} = molar flow rate of O_2 in stack assuming complete combustion, moles/hr.

II. Fluidized Bed Parameters

A. Bed Height

Bed height is obtained by measuring the bed static pressure at four bed levels. A straight line is fitted through these points on a pressure vs height plot and extrapolated to furnace pressure. The height at this point corresponds to the bed level.

A least squares curve fitting is used to fit a straight line. Cramer's rule is used to solve the resulting simultaneous equations.

The two simultaneous linear equations derived from least squares fitting are given below. Their derivation is given in section VI.

$$\begin{array}{ccc} \begin{bmatrix} N & \Sigma x_i \\ \Sigma x_i & \Sigma x_i^2 \end{bmatrix} & \begin{bmatrix} a_0 \\ a_1 \end{bmatrix} & = & \begin{bmatrix} \Sigma P(x_i) \\ \Sigma x_i P(x_i) \end{bmatrix} \\ \longleftarrow C \longrightarrow & \longleftarrow a_i \longrightarrow & & \longleftarrow R \longrightarrow \end{array}$$

The coefficients a_0 and a_1 can be solved for by employing Cramer's rule. The unknown a_k is given by

$$a_i = \frac{\det C_k}{\det C}$$

where C_k is the matrix C with its k^{th} column replaced by the right hand side column R . Thus, the two coefficients are given by

$$\begin{aligned} a_0 &= \frac{\det \begin{bmatrix} \Sigma P(x_i) & \Sigma x_i \\ \Sigma x_i P(x_i) & \Sigma x_i^2 \end{bmatrix}}{\det C} \\ &= \frac{\Sigma P(x_i) \Sigma x_i^2 - \Sigma x_i \Sigma x_i P(x_i)}{N \Sigma x_i^2 - \Sigma x_i \Sigma x_i} \end{aligned}$$

$$a_1 = \frac{\det \begin{bmatrix} N & \sum P(x_i) \\ \sum x_i & \sum x_i P(x_i) \end{bmatrix}}{\det C}$$

$$= \frac{N \sum x_i P(x_i) - \sum P(x_i) \sum x_i}{N \sum x_i^2 - \sum x_i \sum x_i}$$

where N = number of pressure taps

x_i = pressure tap level

P_i = pressure

det = determinant

The straight line approximation can now be written as

$$P_i = a_0 + a_1 x_i$$

Referring to Figure , it can be seen that a_0 equals the sum of the furnace pressure and bed pressure drop. The bed pressure gradient is given by a_1 . Bed height can now be solved for setting the pressure equal to the furnace pressure.

$$P_{\text{Furn}} = a_0 + a_1 H_{\text{Bed}}$$

$$H_{\text{Bed}} = \frac{P_{\text{Furn}} - a_0}{a_1}$$

where P_{Furn} = furnace pressure

H_{Bed} = bed level

The grid plate pressure drop can be found by

$$\Delta P_{\text{grid}} = \text{plenum pressure} - \text{bed pressure drop} - \text{furnace pressure}$$

$$= P_{\text{plenum}} - (\Delta P_{\text{bed}} + P_{\text{furn}})$$

$$= P_{\text{plenum}} - a_o$$

B. Grid Plate Pressure Drop

The grid plate pressure drop equation was determined by measuring the pressure drop across the grid plate at various superficial velocities. The data was correlated by

$$\ln \frac{\Delta P_{\text{grid}}}{f} = 2.03 \ln u_{\text{in}} + 5.71$$

$$\frac{\Delta P_{\text{grid}}}{f} = \exp (2.03 \ln u_{\text{in}} + 5.71)$$

But,

$$u_{\text{in}} = \frac{\text{ACFM}}{(\text{AREA})(60)}$$

$$\frac{\Delta P_{\text{grid}}}{f} = \exp \left(2.03 \ln \frac{\text{ACFM}}{(4.78)(60)} + 5.71 \right)$$

$$= \exp (2.03 \ln \text{ACFM} - 5.76)$$

where ΔP_{grid} = grid plate pressure drop, psf

ρ = gas density in plenum, lb/ft³

ACFM = gas flow rate through grid plate, ACFM

C. Bed Weight

When a bed is fluidized, the particles are suspended by an upward drag force from the passing gases and the downward pull of gravity. This balance of forces can be expressed by

$$\begin{aligned} & (\text{total weight of particles}) - (\text{pressure drop across bed}) \times \\ & \quad (\text{cross-sectional area of bed}) \end{aligned}$$

$$W = \Delta P_{\text{bed}} A_{\text{bed}}$$

Data from a two-dimensional fluidized bed indicated that the bed weight was somewhat less than $\Delta P_{\text{bed}} A_{\text{bed}}$. These tests were run prior to the sub-scale unit testing. The data was correlated to

$$W = \frac{\Delta P_{\text{bed}} A_{\text{bed}}}{0.9}$$

The bed weight is now given as

$$\begin{aligned} W &= \frac{(\Delta P_{\text{bed}})(4.78\text{ft}^2)}{0.9} \left(\frac{\text{Psi}}{27.7\text{H}_2\text{O}} \right) \left(\frac{144\text{ in}^2}{\text{ft}^2} \right) \\ &= 27.61 \Delta P_{\text{bed}} \end{aligned}$$

where ΔP_{bed} = bed pressure drop, inches H_2O

W = bed weight, lb

D. Static Bed Height

$$\text{static bed height} = \frac{\text{bed weight}}{(\text{bed area})(\text{bed density})}$$

$$\begin{aligned} \text{LSTAT} &= \frac{W}{A_{\text{bed}} \rho_{\text{bed}}} \\ &= \left(\frac{\Delta P_{\text{bed}} A_{\text{bed}}}{0.9} \right) \frac{1}{A_{\text{bed}} \rho_{\text{bed}}} \end{aligned}$$

$$\text{Let } \rho_{\text{bed}} = 86 \text{ lb/ft}^3$$

$$= \left(\frac{\Delta P_{\text{bed}}}{0.9} \right) \left(\frac{1}{86 \text{ lb/ft}^3} \right) \left(\frac{\text{Psi}}{27.7 \text{ "H}_2\text{O}} \right) \left(\frac{144 \text{ in}^2}{\text{ft}^2} \right) \left(\frac{12 \text{ in}}{\text{ft}} \right)$$

$$= 0.818 \Delta P_{\text{bed}}$$

where LSTAT = static bed height, inches

E. Bed Expansion

$$\text{Expansion} = 100 \left[\left(\frac{\text{Expanded Bed Height}}{\text{Static Bed Height}} \right) - 1 \right]$$

$$\text{EXPAN} = 100 \left(\frac{H_{\text{bed}}}{\text{LSTAT}} - 1 \right)$$

where EXPAN = bed expansion, %

F. Bed Voidage

$$\frac{\text{HBED}}{\text{LSTAT}} = \frac{1 - \epsilon_{mf}}{1 - \epsilon}$$

where ϵ_{mf} = void fraction at minimum fluidization

ϵ = bed void fraction

Rearranging,

$$\epsilon_f = \left[1 - \frac{\text{LSTAT}}{\text{HBED}} (1 - \epsilon_{mf}) \right] 100$$

where ϵ_f = bed voidage, %

Assume $\epsilon_{mf} = 0.4$

$$\epsilon_f = 100 \left[1 - 0.6 \frac{\text{LSTAT}}{\text{HBED}} \right]$$

III. Gas Flow Relationships

A. Excess Air

<u>component burned</u>	<u>moles O₂ required/lb coal</u>	<u>reaction</u>
C	$\frac{CCOAL}{12}$	$C + O_2 \rightarrow CO_2$
H	$\frac{HYDRO}{4}$	$2H + \frac{1}{2}O_2 \rightarrow H_2O$
S	$\frac{SCOAL}{32} \left[1 + 0.5 \frac{SREMP}{100} \right]$	$S + O_2 \rightarrow SO_2$ and $CaO + SO_2 + \frac{1}{2}O_2 \rightarrow CaSO_4$
O	$-\frac{OCOAL}{32}$	

$$RAIR = (\text{theoretical air to burn 1 lb coal})(\text{coal flow rate})$$

$$= \left[\frac{CCOAL}{12} + \frac{HYDRO}{4} + \frac{SCOAL}{32} \left(1 + 0.5 \frac{SREMP}{100} \right) - \frac{OCOAL}{32} \right] \left(\frac{\text{mole air}}{0.21 \text{ mole } O_2} \right) \left(\frac{28.84 \text{ lb air}}{\text{mole air}} \right) FCOAL$$

$$EXAIR = 100 \left[\frac{AIR}{RAIR} - 1 \right]$$

where CCOAL = weight fraction C in coal

HYDRO = weight fraction H in coal

SCOAL = weight fraction S in coal

OCOAL = weight fraction O in coal

SREMP = % sulfur removal, %

RAIR = theoretical air flow required for complete combustion, lb/hr

FCOAL = coal feed rate, lb/hr

EXAIR = excess air, %

AIR = actual air flow, lb/hr

B. Superficial Velocity

- 1) Calculated bed superficial velocity (assumes complete combustion of coal volatiles)

$$\begin{aligned}
 VSUP &= \frac{\text{combustion gas flow rate}}{\text{bed cross-sectional area}} \\
 &= (GAS) \left(\frac{\text{hr}}{3600 \text{sec}} \right) \left(\frac{359 \text{ft}^3}{\text{mole}} \right) \left(\frac{14.7 \text{ psia}}{P + 14.7} \right) \left(\frac{1}{4.78 \text{ft}^2} \right) \left(\frac{TBED + 460}{492^\circ \text{R}} \right) \\
 &= 6.23 \times 10^{-4} (GAS) \frac{(TBED + 460)}{(P + 14.7)}
 \end{aligned}$$

where VSUP = bed superficial velocity, ft/sec

GAS = combustion gas flow rate, moles/hr

TBED = bed temperature, °F

P = furnace pressure, psig

- 2) Measured bed superficial velocity

$$\begin{aligned}
 VSUP &= \frac{\text{stack gas flow rate corrected to bed temperature}}{\text{bed cross-sectional area}} \\
 &= 6.23 \times 10^{-4} (SGTOT) \frac{(TBED + 460)}{(P + 14.7)}
 \end{aligned}$$

where SGTOT = stack gas flow rate, moles/hr.

- 3) Measured freeboard superficial velocity

$$\begin{aligned}
 VFREE &= (\text{bed velocity}) \left(\frac{\text{area bed}}{\text{area freeboard}} \right) \left(\frac{\text{freeboard temperature}}{\text{bed temperature}} \right) \\
 &= VSUP \left(\frac{AB}{AF} \right) \left(\frac{TF+460}{TBED+460} \right)
 \end{aligned}$$

where VFREE = freeboard superficial velocity, ft/sec

AB = bed cross-sectional area, ft²

AF = freeboard cross-sectional area, ft²

TF = freeboard temperature, °F

C. Stack Gas Flow Rate

stack gas flow rate = nitrogen + moisture + CO + CO₂ + O₂ + NO_x + SO₂

$$\text{SGTOT} = \text{N}_2 + \text{H}_2\text{O} + \frac{\text{SGTOT}}{100} (\text{CO}_{\text{wet}} + \text{CO}_{2\text{wet}} + \text{O}_{2\text{wet}})$$

$$\text{CO}_{\text{wet}} = \text{CO} \left(1 - \frac{\text{H}_2\text{O}}{\text{SGTOT}}\right)$$

$$\text{SGTOT} = \text{N}_2 + \text{H}_2\text{O} + \left(\frac{\text{SGTOT}}{100} - \frac{\text{H}_2\text{O}}{100}\right)(\text{CO} + \text{CO}_2 + \text{O}_2)$$

$$= \frac{\text{N}_2 + \text{H}_2\text{O} \left[1 - \frac{\text{CO} + \text{CO}_2 + \text{O}_2}{100}\right]}{\left[1 - \frac{\text{CO} + \text{CO}_2 + \text{O}_2}{100}\right]}$$

$$= \text{N}_2 / \left(1 - \frac{\text{CO} + \text{CO}_2 + \text{O}_2}{100}\right) + \text{H}_2\text{O}$$

where SGTOT = stack gas flow rate, moles/hr.

N₂ = nitrogen flow rate, moles/hr.

H₂O = stack moisture

CO_{wet} = CO stack analysis, wet basis

CO_{2wet} = CO₂ stack analysis, wet basis

O_{2wet} = O₂ stack analysis, wet basis

CO = CO stack analysis, dry basis

CO₂ = CO₂ stack analysis, dry basis

O₂ = O₂ stack analysis, dry basis

IV. Stack Emission Relationships

A. Calcium/Sulfur Mole Ratio

$$\text{CASMR} = \frac{\text{moles Ca in limestone}}{\text{moles S in coal}}$$

$$= \frac{(\text{FCCO}_3)(\text{CaCO}_3)/100.1}{(\text{FCOAL})(\text{SCOAL})/32.1}$$

where FCCO_3 = limestone feed rate, lb/hr.

CaCO_3 = weight fraction CaCO_3 in limestone

FCOAL = coal feed rate, lb/hr.

SCOAL = weight fraction S in coal

CASMR = calcium/sulfur mole ratio

B. Sulfur Removal

$$\text{SREMP} = \frac{(\text{sulfur in coal feed}) - (\text{sulfur in stack gases})}{(\text{sulfur in coal feed})}$$

$$= 100 \frac{(\text{SCOAL})(\text{FCOAL}) - 32.1(\text{FSO}_2)}{(\text{SCOAL})(\text{FCOAL})}$$

where FSO_2 = sulfur in stack gases, moles/hr

SREMP = sulfur removal, %

C. Calcium Utilization

$$\text{CAUTL} = \frac{\text{moles calcium sulfate formed}}{\text{moles calcium available}}$$

$$= 100 \left[\frac{(\text{SCOAL})(\text{FCOAL}) - 32.1 (\text{FSO}_2)}{32.1} \right] \left[\frac{100.1}{(\text{FCCO}_3)(\text{CACO}_3)} \right]$$

$$= 100 \left[\frac{(\text{SCOAL})(\text{FCOAL}) - 32.1 (\text{FSO}_2)}{(\text{SCOAL})(\text{FCOAL})} \right] \left[\frac{(\text{SCOAL})(\text{FCOAL})/32.1}{(\text{CACO}_3)(\text{FCCO}_3)/100.1} \right]$$

$$= \frac{\text{SREMP}}{\text{CASMR}}$$

where CAUTL = calcium utilization, %

D. NO_x Emissions

$$\text{NO}_x \text{ BTU} = \frac{1 \text{b NO}_2}{\text{heat content in coal feed}}$$

$$= \frac{(\text{NO}_x) \left(\frac{461 \text{b}}{\text{mole}} \right)}{\text{HCOMB}}$$

where $\text{NO}_x \text{ BTU} = 1 \text{b NO}_x / 10^6 \text{ Btu}$

$\text{NO}_x = \text{NO}_x \text{ flow rate, mole/hr.}$

$\text{HCOMB} = \text{heat content in coal feed}$

E. SO₂ Emissions

$$SO_2 \text{Btu} = \frac{1b \text{ SO}_2}{\text{heat content in coal feed}}$$

$$= \frac{(S) \left(\frac{2 \text{ lb SO}_2}{1b \text{ S}} \right)}{HCOMB}$$

where $SO_2 \text{Btu} = 1b \text{ SO}_2 / 10^6 \text{Btu}$

S = sulfur flow rate in stack gas, lb/hr.

V. Thermal Effects

A. Heat Transfer Coefficients

Overall heat transfer coefficient given by

$$U_o = \frac{q}{A_o \Delta T_{lm}}$$

but

$$\frac{1}{U_o} = \frac{D_o}{D_i h_i} + \frac{\Delta r D_o}{k D_{lm}} + \frac{1}{h_o}$$

Solving for the outside heat transfer coefficient,

$$h_o = \frac{1}{\frac{1}{U_o} - \frac{D_o}{D_i h_i} - \frac{\Delta r D_o}{k D_{lm}}}$$

The outside heat transfer coefficient can be broken down into two components, a convective term and a radiation term. The radiation term is given by

$$h_{rad} = \frac{0.1714 \times 10^{-8} \epsilon (T_{Bed}^4 - T_{Surf}^4)}{(T_{Bed} - T_{Surf})}$$

The emissivity was assumed to be 0.85. The convective term is given by

$$h_{conv} = h_o - h_{rad}$$

The inside heat transfer coefficient is calculated by the Dittus - Boelter correlation. The functions for c_p , k and μ contained in the Reynolds and Prandtl numbers were curve fit using Stirling's Interpolation Formula. A derivation is given in section .

$$h_i = 0.023 \frac{k}{D} (Re)^{0.8} (Pr)^{0.4}$$

- where
- U_o = overall heat transfer coefficient, Btu/hr ft²°F
 - q = heat transfer, Btu/hr.
 - A_o = outside tube area, ft²
 - ΔT_{lm} = log mean temperature, °F
 - D_o = outside diameter, ft.
 - D_i = inside diameter, ft.
 - D_{lm} = log mean diameter, ft.
 - Δr = tube thickness, ft.
 - k = thermal conductivity, Btu/hr. ft. °F
 - h_i = inside heat transfer coefficient, Btu/hr ft²°F
 - h_o = outside heat transfer coefficient, Btu/hr ft²°F
 - ϵ = emissivity
 - h_{rad} = radiative heat transfer coefficient, Btu/hr ft²°F
 - T_{Bed} = bed temperature, °R
 - T_{Surf} = tube surface temperature, °R
 - h_{conv} = convective heat transfer coefficient, Btu/hr ft²°F
 - Re = Reynolds number
 - Pr = Prandtl number

B. Freeboard Combustion

A heat balance on the gas side between the bed surface and the furnace outlet may be written as

$$Q_{\text{GEN}} + \sum m C_p T_{\text{Bed}} = Q_{\text{AB}} + \sum m C_p T_{\text{B}}$$

where Q_{GEN} = freeboard heat release, Btu/hr

$\sum m C_p$ = summation of mass flow x specific heat of
elutriated solids and gas, Btu/hr °F

T_{Bed} = bed temperature, °F

T_{B} = gas temperature at furnace outlet, °F

Q_{AB} = above bed heat pickup in water circuit, Btu/hr

Q_{AB} may be found by

$$Q_{\text{AB}} + Q_{\text{BT}} = m_s (h_{\text{SH}} - h_{\text{ECON}})$$

where Q_{BT} = bed tube heat pickup, Btu/hr

m_s = steam flow rate, lb/hr

h_{SH} = superheated steam enthalpy, Btu/lb

h_{ECON} = feedwater enthalpy, Btu/lb

Q_{BT} may be determined by

$$Q_{\text{BT}} = U_o A_o (T_{\text{Bed}} - T_s)$$

where T_s = saturated steam temperature, °F

An inside heat transfer coefficient of 5000 Btu/hr ft² °F was assumed. Substituting for Q_{AB} and Q_{BT} , freeboard heat release can be solved for

$$Q_{GEN} = m_s (h_{SH} - h_{ECON}) - \sum m C_p (T_{Bed} - T_G) - UA (T_{Bed} - T_s)$$

It should be noted that heat transfer by solids splashing into the freeboard is not accounted for. Radiation from the bed surface into the freeboard was assumed to be negligible.

C. Heat Loss

Boiler:

Depth - 45.5 inches
Width - 45.5 inches
Height - 225 inches

Height is taken from 2.5 ft. above ground level to the start of turn for mechanical collector.

Area of Drum & Downcomers	-	197.94 ft ²
Avg. Cold Face Temperature	-	120°F
Heat Loss	-	9,600 Btu/hr
Area of Boiler Casing	-	369.82 ft ²
Avg. Cold Face Temperature	-	170°F
Heat Loss	-	90,600 Btu/hr

Area of Probes	-	17.89 ft ²
Avg. Cold Face Temperature	-	180°F
Heat Loss	-	3,900 Btu/hr
Area of View Ports	-	2.66 ft ²
Avg. Cold Face Temperature	-	180°F
Heat Loss	-	2,900 Btu/hr
Area of Drum Connections	-	4 ft ²
Avg. Cold Face Temperature	-	200°F
Heat Loss	-	<u>1500 Btu/hr</u>
Total Heat Loss		108,500 Btu/hr

D. Heat of Calcination

Heat of Formation @ 77°F

CaO	- 151.9 Kcal/gmole
CaCO ₃	- 288.5 Kcal/gmole
CO ₂	- 94.1 Kcal/gmole

$$\Delta h_f = h_p - h_R$$

$$= (-151.9 - 94.1) - (-288.5)$$

$$= +42.5 \text{ Kcal/gmole CaO}$$

where Δh_f = heat of calcination at 77°F, Kcal/gmole

h_p = heat of formation of products, Kcal/gmole

h_r = heat of formation of reactants, Kcal/gmole

$$\Delta h_f = (42.5 \frac{\text{Kcal}}{\text{gmole}}) (\frac{\text{mole}}{56 \text{ gm CaO}}) (\frac{454 \text{ gm}}{\text{lb}}) (\frac{3.969 \text{ Btu}}{\text{Kcal}})$$

$$= 1368 \text{ Btu/lb. CaO @ 77°F}$$

The increase in Δh_f with temperature was considered negligible.

E. Heat of Sulfation

Heat of Formation @ 77°F

CaO - 151.9 Kcal/gmole

CaSO₄ - 342.4 Kcal/gmole

SO₂ - 71.0 Kcal/gmole

O₂ -----

$$\Delta h_f = h_f - h_R$$

$$= -342.4 - (-151.9 - 71.0)$$

$$= 119.5 \text{ Kcal/gmole CaSO}_4$$

where Δh_f = heat of sulfation @ 77°F, Kcal/gmole

$$\Delta h_f = (-119.5 \frac{\text{Kcal}}{\text{gmole CaSO}_4}) (\frac{\text{mole}}{136 \text{ lb}}) (\frac{454 \text{ gm}}{1 \text{ lb}}) (\frac{3.969 \text{ Btu}}{\text{Kcal}})$$

$$= 1590 \text{ Btu/lb CaSO}_4 @ 77^\circ\text{F}$$

The increase in Δh_f with temperature was considered negligible.

VI. Curve Fitting Approximations

A. Least Squares Curve Fitting

Assume a function $F(x)$ is to be approximated. In a least squares fit, an approximate function $g(x)$ is chosen such that the square of the differences ϵ between the two functions is minimized.

$$\epsilon = \sum_{i=1}^n \left[g(x_i) - f(x_i) \right]^2$$

can be minimized by setting its partial derivatives equal to 0.

$$\frac{\partial \epsilon}{\partial a_0} = 0 = 2 \sum_{i=1}^n \left[a_0 + a_1 x_i - f(x_i) \right]$$

$$n a_0 + a_1 \sum_{i=1}^n x_i = \sum_{i=1}^n f(x_i)$$

$$\frac{\partial \epsilon}{\partial a_1} = 0 = 2 \sum_{i=1}^n \left[a_0 + a_1 x_i - f(x_i) \right] x_i$$

$$a_0 \sum_{i=1}^n x_i + a_1 \sum_{i=1}^n x_i^2 = \sum_{i=1}^n x_i f(x_i)$$

These two equations can now be solved simultaneously for the two coefficients. They may be set up in matrix form as below.

$$\begin{bmatrix} n & \sum x_i \\ \sum x_i & \sum x_i^2 \end{bmatrix} \begin{bmatrix} a_0 \\ a_1 \end{bmatrix} = \begin{bmatrix} \sum f(x_i) \\ \sum x_i f(x_i) \end{bmatrix}$$

B. Stirling Interpolation Formula

Central differences are an approximate means of representing derivatives. Central differences are obtained by a Taylor series expansion in each direction of point x . The difference formula is:

$$\delta^n = \frac{d^n F}{dx^n} = \frac{\nabla F_{j+n/2} + \nabla F_{j-n/2}}{2h^n} \quad n = \text{even}$$

$$\delta^n = \frac{d^n F}{dx^n} = \frac{\nabla F_{j+(n-1)/2} + \nabla F_{j-(n-1)/2}}{2h^n} \quad n = \text{odd}$$

$$\text{where } \nabla F_j = F_j - F_{j-1}$$

The value $F(x)$ at any point on an analytical function can be determined by using a Taylor series expansion of $F(x)$ about another point on the function. The Taylor series expansion about $x=0$ is given by

$$F(x) = F(0) + x F'(0) + \frac{x^2}{2!} F''(0) + \frac{x^3}{3!} F'''(0) + \dots$$

The Stirling interpolation formula is derived by substituting central differences for the derivatives in the Taylor series expansion. The Sterling interpolation formula is given by

$$F(x) = F(0) + x (\delta y_0) + \frac{x^2}{2!} (\delta^2 y_0) + \frac{x (x^2-1)}{3!} (\delta^3 y_0) + \frac{x^2(x^2-1)}{4!} (\delta^4 y_0) + \dots$$

This formula should be used with an odd number of evenly spaced points. The points should be scaled such that each independent variable x is one unit away from each other, with the central value of x having a value of 0.

VII. Derivation of Equation 17

Let us assume that an average bubble diameter of the bed is given by:

$$\bar{d}_B = \frac{\int_0^{L_f} n_i d_{B_i} dh}{\int_0^{L_f} n_i dh} \quad (A-1)$$

where

$$n_i = \frac{f_{pt.}}{\frac{\pi}{4} d_B^2} = \frac{16.7 R^{-0.72}}{\frac{\pi}{4} d_B^2} \quad (A-2)$$

Assuming that for bubbling bed the bubble growth equation is given by:

$$d_{B_i} = 0.136 (U_0 - U_{mf})^{0.4} (\bar{R})^{0.8} \quad (A-3)$$

Substituting N_i into Equation A-1, we have:

$$\begin{aligned} \bar{d}_B &= \frac{\int_0^{L_f} \frac{16.7 R^{-0.72}}{\frac{\pi}{4} d_{B_i}^2} d_{B_i} dh}{\int_0^{L_f} \frac{16.7 R^{-0.72}}{\frac{\pi}{4} d_{B_i}^2} dh} \\ \bar{d}_B &= \frac{\int_0^{L_f} \frac{R^{-0.72}}{d_{B_i}} dh}{\int_0^{L_f} \frac{R^{-0.72}}{d_{B_i}^2} dh} \end{aligned} \quad (A-4)$$

Again substituting for d_{B_i} into Equation A-4, we have:

$$\begin{aligned} \bar{d}_B &= \frac{\int_0^{L_f} \frac{R^{-0.72} dh}{[0.136 (U_0 - U_{mf})^{0.4} R^{0.8}]}}{\int_0^{L_f} \frac{R^{-0.72} dh}{[0.136 (U_0 - U_{mf})^{0.4} R^{0.8}]^2}} \\ \bar{d}_B &= 0.136 (U_0 - U_{mf})^{0.4} \frac{\int_0^{L_f} R^{-1.52} dh}{\int_0^{L_f} R^{-2.94} dh} \\ \bar{d}_B &= 0.3505 (U_0 - U_{mf})^{0.4} L_f^{0.82} = \frac{0.3505 (U_0 - U_{mf})^{0.4} L_{mf}^{0.82}}{(1-\delta)^{0.82}} \end{aligned} \quad (A-5)$$

Assuming $L_{mf} \approx 2 \text{ ft.}$, we have

$$\bar{d}_B = 10.195 \frac{(U_0 - U_{mf})^{0.4}}{(1-\delta)^{0.82}}$$

(A-5 Cont'd.)

Substituting $\bar{d}_B$ into following equation:

$$U_{br} = 0.711 \sqrt{g \bar{d}_B}$$

$$\therefore U_{br} = \frac{71.12 (U_0 - U_{mf})^{0.2}}{(1-\delta)^{0.41}} \quad \text{cm/sec}$$

Therefore, the average rise velocity of an isolated bubble in a stationary medium is given by:

$$U_{br} = \frac{71.12 (U_0 - U_{mf})^{0.2}}{(1-\delta)^{0.41}} \quad \text{cm/sec}$$

Therefore, Equation 11 can be used to predict δ , i.e., R.

$$\delta = \frac{(U_0 - U_{mf})}{U_0 (1 + \alpha \epsilon_{mf} + \beta \epsilon_{mf}) + U_{br} (1 + \alpha \epsilon_{mf} + \beta \epsilon_{mf}) - U_{mf} (2 + \alpha \epsilon_{mf} + \beta \epsilon_{mf} + \alpha \beta)}$$

Assuming $\alpha = 0.2$ and $\epsilon_{mf} = 0.42$, we have:

$$\delta = \frac{(U_0 - U_{mf})}{U_0 (1.084 + 0.42\beta) + U_{br} (1.084 + 0.42\beta) - U_{mf} (2.284 + 1.42\beta)}$$

where

$$\beta = \frac{3.6}{\left(\frac{0.42 U_{br}}{U_{mf}} - 1 \right)}$$

$$U_{br} = \frac{71.12 (U_0 - U_{mf})^{0.2}}{(1-\delta)^{0.41}} \quad \text{cm/sec}$$

VIII. Zonal Heat Balance

The primary function of zonal heat balance is to determine the heat release in various sections of the fluid bed furnace. The approach is to divide the SSU furnace into six well-defined zones and estimate the performance of each individual zone by performing mass and energy balance calculations. The present approach considers heat balance calculations for the active bed section, the splash zone, the superheater section, the economizer section, the furnace exit zone, and the freeboard section in that order.

Immediately after release of heat by combustion there is a transfer of heat by conduction/convection and radiation modes to various heat absorbing surfaces. In the various zones of the SSU furnace, all of the principal mechanisms of heat transfer take place simultaneously. Although the calculational procedure involving heat transfer is quite complex, an attempt has been made to estimate the extent of combustion and steaming or heat absorption rates occurring in the individual zone.

Ideally, it is desirable to achieve most of the combustion of coal in the active bed section. However, not all of the BTU in the coal are converted to heat and absorbed by immersed surfaces in the bed. Some of the unburned carbon leaves the bed in the elutriated stream and/or overflow stream. In addition, some CO and unburned volatiles leave the bed and reduce the level of combustion in the bed. Subsequently, some of the elutriated carbon and CO and all of the unburned volatiles are combusted in the freeboard and possibly in the superheater section of the SSU furnace.

By far the greatest heat loss is the loss up the stack. Since the heat in the fuel is determined from a base of ambient temperature, all of the products of combustion must be cooled to the same temperature if all of the heat is to be

utilized. This loss is made up of: 1) the sensible heat in dry flue gas, and 2) the sensible and latent heat in the moisture in the flue gas.

In the active bed section, the heat generated by combustion of coal is transferred to various immersed surfaces. In addition, there are certain associated heat losses. For the active bed section, these losses include radiation to plenum and freeboard, through refractory walls and to the atmosphere, splashing of solids from the bed surface, in the combustion gases and elutriated solids stream leaving the bed, etc. Heat losses due to radiation from the bed to the plenum and freeboard were calculated in the usual manner by assuming 0.85 for the emissivity.

The general equation for heat flow rate can be used to calculate the heat content of individual flow stream:

$$q = U \cdot S \cdot \Delta t = W^* \cdot C \cdot (t_1 - t_2) = W' C' (t_2' - t_1') = W (h_2 - h_1) \quad (\text{A-II-1})$$

where

q = Rate of heat flow, BTU/hr

U = Overall heat transfer coefficient, BTU/hr ft² °F

Δt = Temperature difference causing heat flow, °F

W^* = Weight of hotter fluid, lbs/hr

C = Temperature dependent specific heat of hotter fluid, BTU/lb °F

t_1, t_2 = Temperature of hotter fluid entering and leaving heat transfer surface, °F

W' = Weight of colder fluid, lbs/hr

C' = Specific heat of colder fluid, BTU/lb °F

t_1', t_2' = Temperature of colder fluid entering and leaving heat transfer area, °F

h_1, h_2 = Enthalpy of colder fluid entering and leaving, BTU/lb

Throughout zonal heat balance calculations, the relation expressed by Equation A-II-1 was used to estimate the heat content of various streams entering and leaving the section. It should be noted that throughout this investigation (Test Series J through N) the active bed height was always greater than 25" implying that both test loops were submerged in the bed at all times.

In order to estimate in-bed steaming rate, the variation in in-bed steam tube area with active bed height must be known accurately. The following relations were used to estimate the variation in in-bed steam tube surface area with bed height:

$$\begin{aligned}
 S &= 0.454 L_f - 5.631 & 12 < L_f < 19 \\
 &= 0.132 L_f - 0.302 & 19 \leq L_f < 38 \\
 &= 0.453 L_f - 11.121 & 38 \leq L_f \leq 42 \\
 &= 7.75 & 42 < L_f \leq 62
 \end{aligned}
 \tag{A-II-2}$$

where L_f is the active bed height in inches and S is the total in-bed steam tube surface area in sq ft.

After performing an overall heat balance for the active bed section, the percent heat released in the bed is equal to:

$$Q_{R,Bed} = \frac{(Q_{out} - Q_{in})}{W_c \cdot HHV} \times 100
 \tag{A-II-3}$$

where

$Q_{R,Bed}$ = Percent heat released in bed

Q_{out} = Total heat out from bed, BTU/hr

Q_{in} = Total heat in from bed, BTU/hr

W_c = Coal feed rate, lbs/hr

HHV = Higher heating value, BTU/lb

Throughout this study a base temperature of 80°F was used. The composition of flue gas leaving the bed was assumed to be the same as that leaving the SSU furnace. The average temperature of flue gas/elutriated solids stream leaving and entering a given section is measured experimentally.

The combustion gases and elutriated solids stream leaving the bed pass successively through a splash zone, freeboard, superheater, economizer, and the furnace exit zone (containing screen tubes). The heat balance calculations for the splash zone (if the active bed height is less than 40"), the superheater and economizer sections are quite similar in nature to those for standard boiler calculations except that convection passes in the SSU furnace experience a significantly higher solids carryover rate. In addition, some char, volatiles, and CO combustion may occur. The overall heat transfer coefficient value for the water wall tubes in the superheater section was taken as 65% of the experimentally measured value of the superheater bundle. Similarly, the overall heat transfer coefficient values for the water wall tubes in the economizer and furnace exit zone sections was taken to be equal to 85% of the experimentally measured value of the economizer bundle. In addition, the emissivity of the nonluminous gas stream exchanging radiation heat between superheater and economizer tube assembly was taken to be equal to 0.4.

It should be realized that whenever the temperature of the feedwater stream leaving the economizer is below its saturation temperature, additional steam (above that actually measured) must be generated in the system to bring economizer feedwater exit stream to its saturation temperature. Therefore, in the zonal heat balance calculations, total steaming rate should be equal to the steaming rate actually being measured plus that required for the economizer feedwater stream. The steaming rate in the freeboard water wall tubes was calculated by subtracting total steaming rate required from the

steaming rate in the bed splash zone, superheater and economizer wall tubes, bed/freeboard/splash zone refractory wall tubes, bare wall tubes ($3\frac{1}{2}$ " level above the distributor plate) and the furnace exit zone wall tubes. The percent heat released in the freeboard was than calculated using Equation A-II-3. The final results of the heat balance calculations of Test Series J through N are summarized on the following pages.

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APPENDIX II

I. PRIMARY DATA REDUCTION PROGRAM: SSU77

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C      PROGRAM SSU IS A FIRST ORDER DATA REDUCTION PROGRAM
C      FOR THE C-C/DCE INDUSTRIAL FLUIDIZED BED SUB-SCALE UNIT
C      THIS CURRENT PROGRAM VERSION IS: SSU77
C
      PROGRAM SSU (INPUT, OUTPUT, OUT, SSU1, CONS, BASE, TAPE3=SSU1, TAPE10=
10OUT, TAPE1=BASE, TAPE4=CONS, TAPE5=INPUT, TAPE6=OUTPUT, DEBUG=OUTPUT)
      DIMENSION TITLE(20), TIME(10), X(150), DATE(20), LPROR(20)
      DIMENSION UO(4), TV(4), F(4), HI(4), HO(4), RE(4), HTM(2), GC(150)
      DIMENSION COM1(10), COM2(10), COM3(10), COM4(10)
      DIMENSION HT(10), PRESS(5), DATA(43), INDEX(201)
      INTEGER ERROR, TEST
      REAL MASSI, MASSO, MOIST, N2, NOX, K1, K2, NOXBO, LHV1, MOISTL, MGCOSL,
1INERTL, LSTAT, MV, LSTATM
C
C      READ SSU INPUT DATA
      M=88
      REWIND 3
      REWIND 4
      REWIND 10
      READ(3,30) (DATE(I), I=1,2), (TIME(I), I=1,2)
      READ(3,35) DUMMY
      READ(3,40) (X(I), I=1,M)
30  FORMAT(25X, A4, A4, 12X, A4, A4)
35  FORMAT(A10)
40  FORMAT(5(6X, F10.4))
      READ(3,53) (COM1(I), I=1,10)
      READ(3,53) (COM2(I), I=1,10)
      READ(3,53) (COM3(I), I=1,10)
      READ(3,53) (COM4(I), I=1,10)
53  FORMAT(10A4)
C
C      READ CONSTANTS
      READ(3,39) TEST
39  FORMAT(I3)
      READ(3,41) VAR, REH2O, DRAIN, C1, C2, C3, C4, C5, FBAG, CBAG,
1CRECY, CDRAN1, CDRAN2, FLSOL, CFLSOL, FORUM, CDRUM, MOIST, COCAL,
2HYDRO, SCAL, COALN, COALA, MOISTL, CLOSS, H2OAMB
41  FORMAT(3F10.3)
      READ(4,42) COALVM, COALFC, CACOSL, MGCOSL, INERTL,
1CACOS, CAO, CASO4, SIO2, CCO3R, CAOR, CSO4R, SIO2R,
2TWALL, TAMB, PAMB, AREA, HCOEF, SQUAL, HT(1), HT(2),
3HT(3), HT(4), CC1, CC2, CL1, CL2
42  FORMAT(4F10.3)
C
C      CONVERT DATA TO ENGINEERING UNITS
      HA=(X(1)*0.5)**2
      HR=X(2)**2
      PA=X(5)*10./27.7
      PRESS(1)=X(6)*7.5
      PRESS(2)=X(7)*4.0
      PRESS(3)=X(8)*4.2
      PRESS(4)=X(9)*2.5
      TA2=X(11)
      TA=X(11)
      TR=(X(14)+70.)/2.
      TFD=X(14)
      PR=X(3)*15./27.7
      PFURN=(X(18)-5.0)*0.0722
      PSG=0.
      HSG=(X(19)*.447)**2
      TFREEB=X(20)
      TCOND=X(22)
      TSC=X(54)
      TSC=X(54)

```

C GAS STREAM CONCENTRATIONS

O2SG=X(27)*C1
CO2G=X(28)*C2
CO2SG=X(29)*C3
NOX=X(30)*C4
FLS02=X(31)*C5
O2BD=X(32)*C1
CO2BD=X(33)*C2
CO2BD=X(34)*C3
NOXBD=X(35)*C4
SO2BD=X(36)*C5

C
C
C

HF=(X(40)*.707)**2
HS=(X(41)*1.225)**2
TF=X(43)*1000.
TECND=X(44)*1000.
TS=X(45)*1000.
TSUP=X(46)*1000.
TG=X(47)
HC=(X(48)*.5)**2
PS=X(51)*100.
PSUPD=X(49)*10.0
PSUP=PS-PSUPD
PF=X(50)*100.
PC=X(51)*100.
TSOL=(X(14)+70.)/2.
TREC=X(16)

C
C

TTL1=X(74)*1000.
TTL1M=X(76)*1000.
TTL10=X(76)*1000.
TTL2=X(77)*1000.
TTL2M=X(79)*1000.
TTL20=X(79)*1000.
PTL1=X(80)*100.
PTL2=X(81)*100.
VOLT1=X(83)
VOLT2=X(82)
VCCOG=X(84)
VCOAL=X(85)
VDRAN=X(86)

C
C
C
C
C
C
C

INSTRUMENT CALIBRATION CURVES

FEEDWATER FLOW ORIFICE

SUBROUTINE SRSORT IS A STEAM TABLES SUBROUTINE THAT IS USED
TO CALCULATE DENSITY AND ENTHALPY OF STEAM AND WATER

P=PF+14.7
T=TF
CALL SRSORT(P, T, V, H, S, ISAT, VG, HG, SG)
RHOF=1./V
HFEED=H-45.09
FA1=1.80E-5*TF+0.9988
YA1=1.0-0.0114*HF/(PF+14.7)
CF1=0.6175
FLOWF=99.96*CF1*YA1*FA1*SQRT(HF*RHOF)
RED1=FLOWF*(0.124*TF-5.08)
CF1=0.623-RED1*.01757E-5
FLOWF=99.96*CF1*YA1*FA1*SQRT(HF*RHOF)
GPMF=FLOWF*0.1247/RHOF

C
C

$PHID=70.492-0.0421*TC$
 $FLOWD=1462.5*FA2*YA2*SQRT(HD*RHOD)$
 $GPMO=0.1247*FLOWD/RHOD$

STEAM FLOW ORIFICE

$FA3=1.8E-5*TS+.9989$
 $YA3=1.0-0.013*HS/(PS+14.7)$
 $T=TS$
 $K=2$
 CALL SATUR(P, T, VF, HF, SF, VG, HG, SG, K)
 $RHOS=1./VG$
 $CF3=0.6158$
 $FLWS=416.36*CF3*FA3*YA3*SQRT(RHOS*HS)$
 $RED3=FLWS*(916.6-1.54*TS)$

MOLECULAR WEIGHT OF MOIST AIR
 $MW=(1.+H2OAMB)/(1./28.84+H2OAMB/18.)$

TRANSPORT AIR FLOW ANNULAR

$CNSTR=(TFD+460.)/(PR+14.7)$
 $RHOR=MW/(10.73*CNSTR)$
 $FLOWR=RHOR*60.*2.483*SQRT(HR*CNSTR)$
 $H2OTR=FLOWR*H2OAMB/(H2OAMB+1.)$
 $FLOWRD=FLOWR-H2OTR$
 $CFMRD=FLOWRD*CNSTR/(2.688*60.)$

COMBUSTION AIR VENTURI

$CNSTR1=(TA2+460.)/(PA+HA/27.7+14.7)$
 $FA4=1.9E-5*TA2+0.9986$
 $YA4=1.0-0.0245*HA/(PA+14.7+HA/27.7)$
 $RHOD=2.688*(PA+HA/27.7+14.7)/(TA2+460.)$
 $RHOA=MW/(10.73*CNSTR1)$
 $FLOWA=3923.4*YA4*FA4*SQRT(HA*RHOA)$
 $H2OCA=FLOWA*H2OAMB/(H2OAMB+1.0)$
 $FLOWAD=FLOWA-H2OCA$
 $CFMAD=FLOWAD/RHOD/60.0$
 $CFMAD=CFMAD*(PA+14.7+HA/27.7)/(PA+14.7)*(TA+460.)/(TA2+460.)$
 $CFMA=FLOWA/RHOA/60.$
 $TOTAIRD=FLOWAD+FLOWRD$
 $CFMA=CFMA*(PA+14.7+HA/27.7)/(PA+14.7)*(TA+460.)/(TA2+460.)$

STACK GAS FLOW VENTURI

$FA5=1.9E-5*TSG1+.9986$
 $YA5=1.0-0.0235*HSG/(PSG+14.7)$
 $S=.23045$
 $RHOSG=2.798*(PSG+14.7)/(TSG1+460.0)$
 $FLWSG=22862.5*S*FA5*YA5*SQRT(RHOSG*HSG)$
 $CFMSG=(FLWSG/RHOSG/60.0)*(TSG+460.)/(TSG1+460.0)$

COAL FEEDER (1VOLT=40LB/HR)

$FCCAL=(40.*VCCAL*CC1+CC2)-CLOSS$
 $FCCALD=FCCAL*(1.-MOIST)$

LIMESTONE FEEDER (1VOLT=25LB/HR)

$FCCO3=25.*VCCO3*CL1+CL2$
 $FCCO3D=FCCO3*(1.-MOISTL)$

ASH STREAM

DRSQL=DRAIN

RECYCLE SOLIDS FLOW

IF (VAR.EQ.1.0) GO TO 680
 $RECYC=27.6*REH20*(X(87)-X(83))/(X(15)-X(16))$
 GO TO 690
 RECYC=0.0

```

C      COALAD=COAL*(1.-MOISTL+HYDRO+SUQAL+COALN+COALA+MOIST))
      COALID=COAL/(1.-MOIST)
      COALAD=COALA/(1.-MOIST)
      CONST=COALVM/COALFC
      COALFC=(1.-COALAD)/(1.+CONST)*(1.-MOIST)
      COALVM=1.-MOIST-COALA-COALFC
C
      CACOSL=CACOSL
      CACOSL=CACOSL*(1.-MOISTL)
      MGCOSL=MGCOSL*(1.-MOISTL)
      INERTL=INERTL*(1.-MOISTL)
C
C
C      PERCENT EXCESS AIR
      GO TO 710
700  RECC=CCOAL/12.
      REQH=HYDRO/4.
      REQS=SCCOAL/32.1*(1.+5*SREMP/100.)
      REED=CCOAL/32.
      RAIR=137.3*(RECC+REQH+REQS-REED)*FCOAL
      EXAIR=100.*((FLOWAD+FLOWRD)/RAIR-1.0)
C
C      CALCULATED SUPERFICIAL VELOCITY
      TMOIST=MTIST*FCOAL+H2OCTR+H2OCCA+FCOQ3*MOISTL
      TMOLES=(FLOWAD+FLOWRD)/28.84+TMOIST/18.+HYDRO/2.*FCOAL
      VSUPC=6.264E-4*TMOLES*(TBED+460.)/(PFURN+14.7)
      GO TO 140
C
C
C      BED HEIGHT CALCULATED FROM BED PRESSURE MEASUREMENTS
710  NHT=4
      SUMHT=0.
      SUMHT2=0.
      SUMP=0.
      SUMHP=0.
      DO 1200 I=1,NHT
      SUMHT=SUMHT + HT(I)
      SUMHT2=SUMHT2 + HT(I)**2
      SUMP=SUMP + PRESS(I)
      SUMHP=SUMHP + HT(I)*PRESS(I)
1200  CONTINUE
      DET=NHT*SUMHT2-SUMHT**2
      A0=(SUMP*SUMHT2-SUMHT*SUMHP)/DET
      A1=(NHT*SUMHP-SUMHT*SUMP)/DET
      BEDHTM=(27.7*PFURN-A0)/A1
      BEDDEL=A0
      PLATE=PA*27.7-A0
      GRID=100.*PLATE/A0
C
C      FLUIDIZED BED CONDITIONS CALCULATED FROM
C      BED STATIC PRESSURE MEASUREMENTS
      LSTATM=.8134*BEDDEL
      EXPAN=100.*(BEDHTM/LSTATM-1.0)
      BEDWT=28.26*BEDDEL
      BEDVOID=(1.-.60*LSTATM/BEDHTM)*100.
      BEDGRAD=BEDDEL/BEDHTM
C
C
C      CALCULATED GRID PLATE DELTA P
      RHOP=RHQA*(PA+14.7)/(PA+HA/27.7+14.7)*(TA2+460.)/(TA+460.)
      DPGRID=RHOP*EXP(2.03*ALOG(CFMA)-5.76)
      DPGRID=DPGRID/144.*27.6
C
      IF(BEDHTM.GT.24.) GO TO 650
      TBED=(X(57)+X(58)+X(62)+X(63))/4.0
      GO TO 670

```



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      TBCD=(X(57)+X(58)+X(59)+X(60)+X(61)+X(62)+X(63)+X(64)+X(65))/8.0
      GO TO 670
651  IF(PEDNTM.GT.43.) GO TO 652
      TDCD=(X(57)+X(58)+X(59)+X(60)+X(62)+X(63)+X(64)+X(65))/8.0
      GO TO 670
652  TBED=(X(57)+X(58)+X(59)+X(60)+X(61)+X(62)+X(63)+X(64)+X(65)
1+X(66))/10.0
C
C
C
C      TEST TUBE 1      (TURBINE METER 80647)
670  IF(VOLT1.GT.1.423) GO TO 800
      T1GPM=.978*VOLT1-.0023
      GO TO 820
800  IF(VOLT1.GT.6.995) GO TO 810
      T1GPM=.9985*VOLT1-.0318
      GO TO 820
810  T1GPM=1.0475*VOLT1-.3949
C
C      TEST TUBE 2      (TURBINE METER 80648)
820  IF(VOLT2.GT.1.861) GO TO 830
      T2GPM=.9833*VOLT2-.0052
      GO TO 850
830  IF(VOLT2.GT.6.996) GO TO 840
      T2GPM=.9999*VOLT2-.03175
      GO TO 850
840  T2GPM=1.0294*VOLT2-.2492
850  P=PTL1+14.7
      T=TTL1
      CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
      RHOT1=1./V
      HEAT1=H-45.09
      FTL1=8.02*RHOT1*T1GPM
      P=PTL2+14.7
      T=TTL2
      CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
      RHOT2=1./V
      HEAT2=H-45.09
      FTL2=8.02*RHOT2*T2GPM
C
C
C      CALCIUM/SULFUR MOLE RATIO
      CASMR=0.321*FCD03*CAD03L/(FCDAL*SCD0AL)
C
C      ELUTRIATED SOLIDS FLOW
      ELUTR=RECYC+BA9+FLSOL+FD0RM
      GO TO 870
C
C      OVERALL ENERGY BALANCE
C
C      HEAT INPUTS
C
C      HEAT OF SULFATION
860  TBED1=(TBED+460.)/1.8
      FCAS04=SREM*136./32.
      HUSU4=FCAS04*1590.
      GO TO 700
C
C      ENTHALPY OF COMBUSTION AIR
870  TA1=(TA+460.)/1.8
      AIRM=FLOWAD/20.84
      HAIR=AIRM*(11.59*TA1+1.48E-3*TA1**2-1.16E-7*TA1**3-3582.)
C
C      ENTHALPY OF FEEDWATER
      HFLW0=FLW0*HFFWD

```

```

HTL P1=HEAT1*FTL1
HTL P2=HEAT2*FTL2
P=PTL1+14.7
T=TTL10
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
H=H+45.09
HLP10=H*FTL1
P=PTL2+14.7
T=TTL20
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
H=H+45.09
HLP20=H*FTL2

C
C      ENTHALPY OF TRANSPORT AIR
TR1=(TR+460.)/1.8
TRANM=FLOWRD/29.84
HTRAN=TRANM*(11.59*TR1+1.43E-3*TR1**2-1.14E-7*TR1**3-3582.)

C
C      ENTHALPY OF LIMESTONE FEED
TL=(TSOL+460.)/1.8
HCCO3=FCCO3D*(19.63+.01189*TL-307600./TL**2)*(TSOL-77.)/100.1

C
C      ENTHALPY (SENSIBLE) OF COAL FEED
HCCAL1=FCCALD*.24*(TSOL-77.)

C
C      ENTHALPY OF FEED MOISTURE
HMOIST=(MOIST*FCCAL+MOISTL*FCCO3)*(TSOL-77.)
HH2OTR=H2OTR*(.46*TR+1015.)
HH2OCA=H2OCA*(.46*TA+1015.)
HH2OIN=HH2OTR+HH2OCA+HMOIST

C
C      HEAT OF COMBUSTION - AVERAGED ASH AND MOISTURE FREE HIGHER
C      HEATING VALUE FOR FREEMAN COAL
HHV=15005.*(1.-MOIST-COALA)
HCOMB=FCCAL*HHV
HCCAL=HCOMB+HCCAL1
PERMCR=HCOMB/2570000.*100.

C
C      HEAT OUTPUTS
C
C
C      ENTHALPY OF DRAIN SOLIDS
HSOL=.29*(TBED-77.)
HDRAN1=HSOL*DRSOL
HDRAN2=DRSOL*CDRAN1*14093.
HDRAN=HDRAN1+HDRAN2
HELUTB=HSOL*ELUTR

C
C      ENTHALPY OF ELutriATED SOLIDS
HSOL=.24*(TSO-77.)
HELUT1=ELUTR*HSOL
HELUT2=14093.*(FGAQ*CBAG+RECYC*CRECY+FLSOL*CFLSOL+FCDRM*CDRM)
HELUT=HELUT1+HELUT2

C
C      ENTHALPY OF RECYCLE SOLIDS
HSOL=.22*(TREC-77.)
HREC1=RECYC*HSOL
HREC2=RECYC*CRECY*14093.
HREC=HREC1+HREC2

C
C      HEAT OF CALCIINATION
TBED1=(TBED+460.)/1.8
FCAQ=FCCO3*CAO3L*56./100.
HCAID=FCAQ*1349.

```

```

C      STACK GAS COMPOSITION IN MOLES
TMOIST=MOIST*(FOAL+H2O)+H2O*(H2O+FOAL)*FOAL*MOISTL
H2O=(TMOIST+H2O*(9.*FOAL)/18.0
H2OLB=H2O*18.
N2=(AIRM+TRASH)*0.79
CONV2=(1.-(CO2G+CO2SG+O2SG+NOX/10000.+FLSO2/10000.)/100.)
SGTOT=N2/CONV2
CONV1=SGTOT/100.
O2=O2SG*CONV1
CO=CO2SG*CONV1
CO2=CO2SG*CONV1
FNOX=NOX/10000.*CONV1
FSO2=FLSO2/10000.*CONV1
W1=N2/SGTOT
W2=O2/SGTOT
W3=CO/SGTOT
W4=CO2/SGTOT
W5=FNOX/SGTOT
W6=FSO2/SGTOT
SGMW=W1*28.+W2*32.+W3*28.+W4*44.+W5*18.+W6*44.+W6*64.
FLWSGD=SGTOT*SGMW
CFMSGD=0.179*SGTOT*(TSG+460.)/(PFURN+14.7)
PRCTGAS=O2SG+CO2SG+CO2SG+(NOX+FLSO2)/10000.

C
C      MEASURED SUPERFICIAL VELOCITY
VSUPM=6.149E-4*(SGTOT+H2O)*(TBED+460.)/(PFURN+14.7)
VFREEM=VSUPM*.553*(TFREEB+460.)/(TBED+460.)

C
C      ENTHALPY OF DRY STACK GAS
TB1=(TSG+460.)/1.8
XO=0.0
DO 150 I=1,2
  HGAS=((6.52*W1+6.15*W2+6.42*W3+6.21*W4)*(TB1-298.)
  1+(.00125*W1+.0031*W2+.00167*W3+.0104*W4)*
  2(TB1**2-88804.)/2.0 -(1.0E-7*W1+9.23E-7*W2+1.96E-7*W3+3.545E-6
  36*W4)*(TB1**3-2.65E7)/3.0)*1.8
  IF (XO.GT.0.0) GO TO 150
  HFLUE1=HGAS*SGTOT
  HFLUE2=CO*121000.
  HFLUE= HFLUE1+ HFLUE2
  XO=1.0
  TB1=(TBED+460.)/1.8
150 CONTINUE
  HFLUEB=HGAS*SGTOT

C
C      ENTHALPY OF STACK MOISTURE
HSTH2O=H2OLB*(.46*TSG+1016.)

C
C      HEAT LOSS - DETERMINED FROM BOILER CASING INSPECTION
HLOSS=110000.

C
C      ENTHALPY OF SUPERHEATED STEAM
P=PSUP+14.7
T=TSUP
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
H=H-45.09
HSUP=H *FLOWS

C
C      NOX AND SO2 EMISSIONS
CNOX=18.*NOX/(21.-O2SG)
CSO2=18.*FLSO2/(21.-O2SG)
BTUN2=46.*FNOX*1.55/HCOMB
GRFH=SGAL*HCOMB+PRCTGAS*SGAL

```

```

C
C      TOTAL HEAT IN
140  CHEMI=HCOMB+HREC2+HCSO4
      SENSIN=HCOAL1+HCOO3+HAIR+HTRAN+HTLP1+HTLP2+HREC1+HFH2O
      1+HH2O1N
      HEATI=CHEMI+SENSIN
C
C      TOTAL HEAT OUT
      CHEMO=HORAN2+HELUT2+HCALC+HFLUE2
      SENOUT=HELUT1+HFLUE1+HLP10+HLP20+HSUP+HORAN1+HLOSS+HSTH2O
      HEATO=CHEMO+SENOUT
      UNACH=HEATI-HEATO
      PERCH=100.*UNACH/(HEATI-HREC1-HREC2)
C
C      OVERALL MASS BALANCE
      MASSI=FCCALD+FCCO3D+FLOWAG+FLOWF+FTL1+FTL2+FLOWRD+RECYC+TMOIST
      MASSO=FLWSOD+DRSOL+FLWS+FTL1+FTL2+ELUTR+H2OUB
      UNACM=MASSI-MASSO
      PERCM=100.*UNACM/(MASSI-RECYC)
C
C
C      CALCIUM UTILIZATION
      CAUTL=SREMP/CASMR
C
C      OVERALL CARBON BALANCE
C
C      CARBON IN
      FCCALC=FCCAL*CCOAL
      CLIME=CACCO3L*.120
      CLIMEO=CACCO3D*.120
      FLIMEC=FCCO3*CLIME
      FRECYC=RECYC*CRECY
      CARBI=FCCALC+FLIMEC+FRECYC
C
C      CARBON OUT
      FBAGC=FBAG*CBAG
      FDRUMC=FDRUM*CDRUM
      FLSOLC=FLSOL*CFLSOL
      CELUT=(FBAGC+FLSOLC+FRECYC+FDRUMC)/ELUTR
      FELUTC=ELUTR*CELUT
      CDRAN=CDRAN1+CDRAN2
      FDRANC=DRSOL*CDRAN
      SGCO=C0*12.
      SGCO2=C02*12.
      SGCARB=SGCO+SGCO2
      CSTGAS=SGCARB/FLWSGD
      CARBO=FELUTC+FDRANC+SGCARB
      UNACC=CARBI-CARBO
      PERCC=(UNACC/(CARBI-FRECYC))*100.
C
C
C      OVERALL COMBUSTION EFFICIENCY CALCULATED
C      FROM SENSIBLE AND CHEMICAL HEAT BALANCES
      CEFF1=100.*(SENOUT-SENSIN+HCSO4-HCALC)/HCOMB
      CEFF2=100.*(HCOMB-14093.*(FBAG*CBAG+FLSOL*CFLSOL
      1+FDRUM*CDRUM)-HFLUE2-HBRAN2)/HCOMB
C
C
C      COMBUSTION EFFICIENCY CALCULATED
C      FROM STACK GAS O2 CONCENTRATIONS
      OETH=(OYATP*RAIR/20.9)*.21/100.
      HCHAR=(O2-O2TH)*12.*14093.
      CEFF3=100.*(HCOMB-HCHAR-HFLUE2)/HCOMB

```

HFO AND ECONOMIZER HEAT TRANSFER COEFFICIENTS

ENTHALPY OF TEST LOOP STREAMS AT LOOP MID-POINT

```

DTLP=0.1563
DTLM=.1436
THLP=.01447
ARTL=1.064
CORN=33.0
AECOR=58.70
DECOR=0.0833
P=PTL1+14.7
T=TTL1M
DO 401 I=1,2
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
HTM(I)=H-45.09
P=PTL2+14.7
401 T=TTL2M
HTM1=HTM(1)*FTL1
HTM2=HTM(2)*FTL2

```

ENTHALPY OF ECONOMIZER OUTLET

```

P=PE+14.7
T=TECNO
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
H=H-45.09
HECON=H*FLOWF

```

LOG MEAN TEMPERATURES

```

TLG1M=(TTL1M-TTL1)/ALOG((TBED-TTL1)/(TBED-TTL1M))
TLG2M=(TTL2M-TTL2)/ALOG((TBED-TTL2)/(TBED-TTL2M))
TENLM=(X(21)-TF-TECNO+TECNO)/ALOG((X(21)-TF)/(TECNO-TECNO))

```

OVERALL HEAT TRANSFER COEFFICIENT

```

UO(1)=(HTM1-HTLP1)/ARTL/TLG1M
UO(2)=(HTM2-HTLP2)/ARTL/TLG2M
UO(3)=(HECON-HFH20)/TENLM/AECOR

```

APPROXIMATE RADIATION COMPONENT TO OUTSIDE

HEAT TRANSFER COEFFICIENT ON TEST LOOPS

```

RAD1=.1714E-8*.85*((TBED+460.))**4.-((TBED-TLG1M+460.))**4.)/TLG1M
RAD2=.1714E-8*.85*((TBED+460.))**4.-((TBED-TLG2M+460.))**4.)/TLG2M

```

BULK MEAN TEMPERATURE

```

TV(1)=(TTL1M+TTL1)/2.
TV(2)=(TTL2M+TTL2)/2.
TV(3)=(TECNO+TF)/2.
F(1)=FTL1
F(2)=FTL2
F(3)=FLOWF

```

HEAT CAPACITY, THERMAL CONDUCTIVITY AND VISCOSITY VALUES

CURVE FITTING BY STIRLING INTERPOLATION FORMULA

```

DO 402 I=1,3
XX=(TV(I)-250.)/100.
CP=1.015+.024*XX+.007*XX**2+.0014*XX**3+.00063*XX**4
TCOR=0.3174+5.83E-4*TV(I)-1.034E-6*TV(I)**2
Y=(TV(I)-302.)/90.
VIS=(1810.-602.5*Y+217.6*Y**2-122.5*Y**3+35.4*Y**4)*2.419E-4
PRAN=CP*VIS/TCOR
IF(1.E2.3) GO TO 405.
RE(I)=6.050*F(I)/VIS

```

INSIDE HEAT TRANSFER COEFFICIENT - DITTUS-BOELTER CORRELATION

```

HI(I)=.6210*TCOR*RE(I)**0.3*PRAN**0.4 45

```

```

C
C      REYNOLDS NUMBER
405 RE(3)=20.972*F(3)/VIS
    HI(3)=.3731*TCOVR*RE(3)**0.8*PRAN**0.4
    HQ(3)=1.0/(1.0/HQ(3)-1.370/HI(3)-.0113*DECON/(CONW*.0714))
402 CONTINUE
C
C      APPROXIMATE CONVECTION COMPONENT TO OUTSIDE
C      HEAT TRANSFER COEFFICIENT ON TEST LOOPS
CONV1=HQ(1)-RAD1
CONV2=HQ(2)-RAD2
C
C      SUPERHEATER HEAT TRANSFER COEFFICIENT
DSUPR=0.125
THSUP=0.0113
ASUPR=23.45
TS1=X(45)*1000.
TS2=X(46)*1000.
TG1=X(20)
TG2=X(21)
TSPLM=(TG1-TS1-TG2+TS2)/ALOG((TS1-TG1)/(TS2-TG2))
TVSUP=(TS1+TS2)/2.
XX=(TVSUP-500.)/50.
CPD=0.55-0.025*XX-.0046*XX**2+.0004*XX**4
SL=(6.67-1.429*XX-.169*XX**2-1.026*XX**3+.524*XX**4)*1.0E-4
UP=CPD+SL*(PSUP-200.)
Y=(TVSUP-392.)/90.
THCND=(33.2+4.783*Y+.258*Y**2-.033*Y**3-.008*Y**4)*.5778/1000.
THCN=THCND*(1.0+PSUP/4000.)
VIS=(160.2+19.358*Y+1.038*Y**2+.192*Y**3-.188*Y**4)*2.419E-4
RE(4)=12.43*FLOWS/VIS
PRAN=CP*VIS/THCN
T=TS
K=2
C
C      ENTHALPY OF SUPERHEATER INLET STEAM
CALL SATUR(P,T,VF,HF,SF,VB,HG,SG,K)
HG=HG-45.09
DELH=HSUP-FLOWS*HG
UQ(4)=DELH/TSPLM/ASUPR
HI(4)=.2244*THCN*RE(4)**.8*PRAN**.4
HQ(4)=1.0/(1.0/UQ(4)-1.2195/HI(4)-.0124/CONW)
C
C
C      FREEBOARD HEAT RELEASE - ESTIMATED
C
C      IMMERSIED BED TUBE SURFACE AREA
IF(BEDHTM.GT.11.) GO TO 396
ABT1=0.
GO TO 400
396 IF(BEDHTM.GT.16.5) GO TO 397
ABT1=.041*(12.36*BEDHTM-136.)
GO TO 400
397 IF(BEDHTM.GT.33.) GO TO 398
ABT1=.041*(2.97*BEDHTM+19.)
GO TO 400
398 IF(BEDHTM.GT.38.4) GO TO 399
ABT1=.041*(13.33*BEDHTM-323.)
GO TO 400
399 ABT1=7.75
C
C      BED TUBE HEAT PICK-UP (IN BED)
400 HIBT=5000.
    THOUT1=TBED-TS

```

```

C
C
C      FREEDBOARD HEAT RELEASE
HBT=HBT1
HDELTA=HSHR-HFEED*FLOWS-HBT-HLOSS
HFREE=HDELTA-((HFLUE2+HFLUT6)-(HFLUT1+HFLUE1))
PROTFD=(HFREE/(CEFF2*HCOMB))*10000.

C
C
C      CYCLONE EFFICIENCY
CYCEFF=(1.-FBAG/ELUTR)*100.

C
C
C      REDUCED DATA CONVERTED TO MATRIX FORMAT
DATA(1)=FCCAL
DATA(2)=FCCO3
DATA(3)=TOTAIRD
DATA(4)=PERMCR
DATA(5)=EXAIR
DATA(6)=TBED
DATA(7)=CASMR
DATA(8)=VSUPM
DATA(9)=VFREEM
DATA(10)=BEDHTM
DATA(11)=BEDGRAD
DATA(12)=PLATE
DATA(13)=BEDDEL
DATA(14)=FLOWC
DATA(15)=FLOWS
DATA(16)=RECYC
DATA(17)=FORUM
DATA(18)=FBAG
DATA(19)=DRSOL
DATA(20)=CYCEFF
DATA(21)=HO(1)
DATA(22)=HO(2)
DATA(23)=HO(3)
DATA(24)=HO(4)
DATA(25)=O2SG
DATA(26)=CO2SG
DATA(27)=CO2SG
DATA(28)=NOX
DATA(29)=FLSO2
DATA(30)=BTUN2
DATA(31)=SMBT
DATA(32)=CNOX
DATA(33)=CSO2
DATA(34)=SREM
DATA(35)=SREMP
DATA(36)=CAUTL
DATA(37)=CEFF2
DATA(38)=CRECY
DATA(39)=CIRUM
DATA(40)=CBAG
DATA(41)=HFREE
DATA(42)=PROTFD
DATA(43)=TEST

C
C
C      WRITE REDUCED DATA INTO DATA BASE
WRITE(6,6000) TEST
6000 FORMAT(I3)
CALL OPENMS(1,INDEX,201,0)
CALL WRITMS(1,DATA,43,TEST,-1) 47
CALL CLOSMS(1)

```

```

C      DATA OUTPUT FORMATING
      WRITE(6,70)
70  FORMAT(20X, 'PRELIMINARY SSU DATA ANALYSIS', 11X,
1  'PROGRAM VERSION: SSU77')
      WRITE(6,71) (DATE(I), I=1,2)
71  FORMAT(60X, 'DATA RECORDED ON ', 2A4)
      WRITE(6,72) ((CON1(I), I=1,8), (TIME(I), I=1,2))
72  FORMAT(20X, 8A4, 8X, 'DATA RECORDED AT ', 2A4)
      WRITE(6,74) (COM2(I), I=1,10)
      WRITE(6,74) (COM3(I), I=1,10)
      WRITE(6,74) (COM4(I), I=1,10)
74  FORMAT(20X, 10A4)
      WRITE(6,75) FCCAL, FCCOS, TOTAIRO, PERMCR, EXAIR, TBED, CASMR, VSUPM,
1  VFREE, BEDHTM
75  FORMAT(/ /40X, '- SUMMARY OF OPERATING CONDITIONS -' / /30X,
1  'COAL FEED RATE (AS FIRED)', F11.2, ' LB/HR', /30X,
1  'LIME FEED RATE (AS FIRED)', F11.2, ' LB/HR', /30X,
1  'TOTAL DRY AIR FLOW', F11.2, ' LB/HR', /30X,
2  'PERCENT OF MCR LOAD', F11.2, ' PRCT', /30X,
2  'THEOR EXCESS AIR', F11.2, ' PRCT', /30X,
3  'BED TEMPERATURE', F11.2, ' DEG F', /40X,
4  'COA/S MOLE RATIO', F11.2, /30X,
5  'SUPERFICIAL VELOCITY', F11.2, ' FT/SEC', /30X,
6  'FREEBOARD VELOCITY', F11.2, ' FT/SEC', /30X,
7  'ACTIVE BED HEIGHT', F11.2, ' IN',
      WRITE(6,76) BEDGRAD, PLATE, BEDDEL, FLOWC, FLOWS, RECYC,
1  FDRUM, FBAG, DRCSL, CYCEFF
76  FORMAT(30X, 'BED PRESSURE GRADIENT', F11.2, ' IN H2O/IN BED', /30X,
3  'GRID PRESSURE DROP', F11.2, ' IN H2O', /30X,
4  'BED PRESSURE DROP', F11.2, ' IN H2O', /30X,
5  'BED TUBES FLOW RATE', F11.2, ' LB/HR', /30X,
5  'STEAMING RATE', F11.2, ' LB/HR', /30X,
6  'RECYCLE SOLIDS', F11.2, ' LB/HR', /30X,
7  'RECYCLE BYPASS SOLIDS', F11.2, ' LB/HR', /30X,
8  'BAGHOUSE SOLIDS', F11.2, ' LB/HR', /30X,
8  'BED DRAIN SOLIDS', F11.2, ' LB/HR', /30X,
9  'CYCLONE EFFICIENCY', F11.2, ' PRCT')

C
      WRITE(6,77) UO(1), HI(1), HO(1), CONV1, RAD1, UO(2), HI(2), HO(2),
1  CONV2, RAD2, UO(3), HI(3), HO(3), UO(4), HI(4), HO(4)
77  FORMAT(/ /36X, '- SUMMARY OF STEADY STATE TEST RESULTS -' / /
134X, 'HEAT TRANSFER COEFFICIENTS - BTU/HR FT2 DEG F',
241X, 'UO', HI, HO, HO(CONV), HO(RAD)',
320X, 'TEST LOOP NO. 1', 5(F10.2)/20X, 'TEST LOOP NO. 2', 5(F10.2)/
420X, 'ECONOMIZER', 3(F10.2)/20X, 'SUPERHEATER', 3(F10.2))
      WRITE(6,78) Q2SG, COSG, Q2SG, NOX, FLSO2
78  FORMAT(/38X, ' DRY STACK GAS STREAM COMPOSITIONS ' /
141X, 'O2', CO, CO2, NOX, SO2',
240X, 'PRCT', PRCT, PRCT, PPM, PPM',
320X, 'FURNACE OUTLET', 4X, F6.2, F10.3, F11.2, F10.0, F10.0)
      WRITE(6,79) BTUN2, SMO2, CNOX, CSO2, SREM, SREMP, CAUTL, CEFF2
79  FORMAT(/30X, 'NOX EMISSIONS (AS FIRED)', F8.2, ' LB NOX/MMBTU', /30X,
1  'SO2 EMISSIONS (AS FIRED)', F8.2, ' LB SO2/MMBTU', /30X,
2  'NOX EMISSIONS (AT 3PRCT O2)', F8.2, ' PPM', /30X,
3  'SO2 EMISSIONS (AT 3PRCT O2)', F8.2, ' PPM', /30X,
4  'SULFUR RETENTION', F8.2, ' LB S/HR', /30X,
5  'SULFUR RETENTION', F8.2, ' PRCT', /30X,
6  'CALCIUM UTILIZATION', F8.2, ' PRCT', /30X,
8  'COMBUSTION EFFICIENCY', F8.2, ' PRCT' //////////////)

C
      WRITE(6,74) (COM1(I), I=1,10)
      WRITE(6,210)
210  FORMAT(/ /48X, 'SSU MASS AND ENERGY BALANCES' /
1  / /100X, 'SENSIBLE', 6X, 'CHEMICAL'
1  /48X, 'TEMPERATURE', 8X, 'PRESSURE', 9X, 'MASS FLOW', 10X, 'HEAT FLOW

```



```

230LD/HR*, 10X, *STU/HR*, 7X, *STU/HR*/10X, * IN4UT5*)
WRITE(6, 220) TSOL, FCCALD, HCCAL1, HCCMB, TSOL, FCCO3D, HCCO3, TREC, RECYC
1, HREC1, HREC2, TF, PF, FLOWF, HFH2O, TTL1, PTL1, FTL1, HTLP1, TTL2, PTL2,
2FTL2, HTLP2, TMOIST, HH2OIN, TA, PA, FLOWAD, HAIR, TR, PR, FLOWRD, HTRAN,
3HCO3D4, MASSI, SENGIN, CHEMI
220 FORMAT(20X, *COAL (DRY)*, 19X, F5.0, 31X, F5.0, 13X, F9.0, 4X, F9.0/
120X, *LIMESTONE (DRY)*, 14X, F5.0, 31X, F5.0, 13X, F9.0/
120X, *RECYCLE SOLIDS*, 15X, F5.0, 31X, F5.0, 13X, F9.0, 4X, F9.0/
220X, *FEEDWATER*, 20X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0/
320X, *WATER TEST LOOP 1*, 12X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0/
420X, *WATER TEST LOOP 2*, 12X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0/
420X, *FEED MOISTURE*, 52X, F5.0, 13X, F9.0/
520X, *COMBUSTION AIR (DRY)*, 9X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0/
620X, *TRANSPORT AIR (DRY)*, 10X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0/
720X, *SULFUR REMOVAL*, 82X, F9.0//
820X, *TOTAL*, 58X, F7.0, 13X, F9.0, 4X, F9.0////)

C
C
WRITE(6, 230) TBED, DRSQL, HDRAN1, HDRAN2, TSG, ELUTR,
1HELUT1, HELUT2, TTL10, PTL1, FTL1,
1HLP10, TTL20, PTL2, FTL2, HLP20, TSUP, PSUP, FLOWS, HSUP, TSG, PFURN,
2FLWSGD, HFLUE1, HFLUE2, H2OLB, HSTH2O, HCALC, HLOSS, MASSO, SENDUT,
1CHEMD
230 FORMAT(10X, * OUTPUTS*/
120X, *DRAIN SOLIDS*, 17X, F5.0, 31X, F5.0, 13X, F9.0, 4X, F9.0/
220X, *ELUTRIATED SOLIDS*, 12X, F5.0, 31X, F5.0, 13X, F9.0, 4X, F9.0/
320X, *WATER TEST LOOP 1*, 12X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0/
420X, *WATER TEST LOOP 2*, 12X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0/
520X, *STEAM*, 24X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0/
620X, *STACK GAS (DRY)*, 14X, F5.0, 13X, F7.2, 11X, F5.0, 13X, F9.0, 4X, F9.0/
620X, *MOISTURE*, 57X, F5.0, 13X, F9.0/
720X, *CALCINATION*, 85X, F9.0/
820X, *HEAT LOSS*, 74X, F9.0//
920X, *TOTAL*, 58X, F7.0, 13X, F9.0, 4X, F9.0////)

C
C
WRITE(6, 240) MASSI, MASSO, UNACM, PERCM, HEATI, HEATO, UNACH, PERCH
240 FORMAT(20X, *TOTAL MASS FLOW IN, LB/HR *, 3X, F9.0/20X,
1*TOTAL MASS FLOW OUT, LB/HR*, 4X, F9.0/20X, *UNACCOUNTED *
2*FOR, LB/HR*, 6X, F11.0/20X, *UNACCOUNTED FOR, PRCT*,
35X, 4X, F9.1//20X, *TOTAL HEAT FLOW IN, BTU/HR *
41X, F9.0/20X, *TOTAL HEAT FLOW OUT, BTU/HR *, 2X, F9.0/20X,
5*UNACCOUNTED FOR, BTU/HR*, 7X, F9.0, /20X, *UNACCOUNTED FOR, PRCT*,
68X, F9.1, 23(/))

C
C
WRITE(6, 74) (COM1(I), I=1, 10)
WRITE(6, 250)
250 FORMAT(//44X, *SSU CARBON MASS BALANCE*, ///42X, *STREAM FLOW*, 7X,
1*CARBON FRACTION*, 8X, *CARBON FLOW*, /20X, *STREAM*, 19X, *LB/HR*, 13X
2*IN STREAM*, 14X, *LB/HR*, /12X, *INPUTS*)
WRITE(6, 260) FCCALD, CCOALD, FCCALC, FCCO3D, CLIMED, FLIMEC, RECYC, CRECY,
1FRECYC, CARBI
260 FORMAT(20X, *COAL*, 20X, F6.0, 14X, F5.3, 16X, F6.0, /20X,
2*LIMESTONE*, 15X, F6.0, 14X, F5.3, 16X, F6.0, /20X,
3*RECYCLE SOLIDS*, 10X, F6.0, 14X, F5.3, 16X, F6.0, //20X,
4*TOTAL*, 60X, F6.0/////11X, *OUTPUTS*)
WRITE(6, 270) RECYC, CRECY, FRECYC, FORUM, CDUM, FDRUMC, FBAG, CBAG, FBAGC,
1DRSQL, CDAN, FORANC, FLWSGD, DSTGAS, SDCARB, CARBO
270 FORMAT(20X, *ELUTRIATED SOLIDS:*/25X,
2*RECYCLE*, 12X, F6.0, 14X, F5.3, 16X, F6.0/25X,
2*RECYCLE BYPASS*, 5X, F6.0, 14X, F5.3, 16X, F6.0/25X,
2*BAGHOUSE*, 11X, F6.0, 14X, F5.3, 16X, F6.0/20X,
2*DRAIN SOLIDS*, 12X, F6.0, 14X, F5.3, 16X, F6.0, /20X,
4*STACK GAS (CO+CO2)*, 6X, F6.0, 14X, F5.3, 16X, F6.0//20X,

```

1*UNACCOUNTED FOR CARBON, PRCT*,F6.1,25(/))

SSU COAL AND LIMESTONE ANALYSIS(AS FIRED).
CONVERTED TO PERCENT BY TOTAL WEIGHT

COALVM=COALVM*100.
COALFC=COALFC*100.
COALA=COALA*100.
MOIST=MOIST*100.
HYDRO=HYDRO*100.
CCOAL=CCOAL*100.
SCCOAL=SCCOAL*100.
COALN=COALN*100.
CCOAL=CCOAL*100.

MOISTL=MOISTL*100.
CACO3L=CACO3L*100.
MGC03L=MGC03L*100.
INERTL=INERTL*100.
TOTAL=100.0

WRITE(6,74) (COM1(I), I=1,10)
WRITE(6,320)MOIST, COALVM, COALFC, COALA, TOTAL, MOIST, HYDRO, CCOAL,
1SCCOAL, COALN, CCOAL, COALA, TOTAL, HHV, MOISTL, CACO3L, MGC03L,
2INERTL, TOTAL
320 FORMAT(/40X, *SSU COAL AND LIMESTONE ANALYSIS - AS FIRED*///
120X, *PROXIMATE COAL ANALYSIS:*, 8X, *WEIGHT PRCT*//
235X, *MOISTURE*, 11X, F6.1/28X, *VOLATILE MATTER*, 11X, F6.1/
331X, *FIXED CARBON*, 11X, F6.1/40X, *ASH*, 11X, F6.1/38X, *TOTAL*, 11X,
4F5.0///21X, *ULTIMATE COAL ANALYSIS:*/35X, *MOISTURE*, 11X, F6.1/
535X, *HYDROGEN*, 11X, F6.1/37X, *CARBON*, 11X, F6.1/
537X, *SULFUR*, 11X, F6.1/
635X, *NITROGEN*, 11X, F6.1/37X, *OXYGEN*, 11X, F6.1/
740X, *ASH*, 11X, F6.1/38X, *TOTAL*, 11X, F5.0///
923X, *HIGHER HEATING VALUE:*, 5X, F6.0, * BTU/LB*///
925X, *LIMESTONE ANALYSIS:*, 8X, *WEIGHT PRCT*//
135X, *MOISTURE*, 11X, F6.1/38X, *CACO3*, 11X, F6.1/
238X, *MGC03*, 11X, F6.1/37X, *INERTS*, 11X, F6.1/38X, *TOTAL*, 11X, F5.0,
329(/))

WRITE(6,74) (COM1(I), I=1,10)
WRITE(6,330)X(12), X(13), X(20), X(21), X(54), X(22), X(23), X(24), X(25),
1TF

330 FORMAT(/40X, *ADDITIONAL SSU STREAM TEMPERATURES - DEG.F*//
1/20X, *AIR TEMPERATURE BEFORE AIR HEATER*, 10X, F6.0
2/20X, *AIR TEMPERATURE AFTER AIR HEATER*, 11X, F6.0/
3/20X, *GAS TEMPERATURE BEFORE SUPERHEATER*, 9X, F6.0
4/20X, *GAS TEMPERATURE AFTER SUPERHEATER*, 10X, F6.0/
5/20X, *GAS TEMPERATURE AFTER ECONOMIZER*, 11X, F6.0
5/20X, *GAS TEMPERATURE AFTER SCREEN TUBES*, 9X, F6.0
6/20X, *GAS TEMPERATURE BEFORE AIR HEATER*, 10X, F6.0
7/20X, *GAS TEMPERATURE AFTER AIR HEATER*, 11X, F6.0
7/20X, *GAS TEMPERATURE AT STACK*, 19X, F6.0/
8/20X, *FEEDWATER TEMPERATURE AT ECON. INLET*, 7X, F6.0)

WRITE(6,340)TECNO, TS, TSUP, TC, X(57), X(58), X(59), X(60), X(61)

340 FORMAT(20X, *FEEDWATER TEMPERATURE AT ECON. OUTLET*, 6X, F6.0/
1/20X, *STEAM TEMPERATURE BEFORE SUPERHEATER*, 7X, F6.0
2/20X, *STEAM TEMPERATURE AFTER SUPERHEATER*, 8X, F6.0
3/20X, *DOWNCOMER WATER TEMPERATURE*, 16X, F6.0/
4/20X, *BED TEMP. AT 6 INCH LEVEL - LEFT WALL*, 5X, F6.0
5/20X, *BED TEMP. AT 12 INCH LEVEL - LEFT WALL*, 5X, F6.0
6/20X, *BED TEMP. AT 24 INCH LEVEL - LEFT WALL*, 5X, F6.0
7/20X, *BED TEMP. AT 36 INCH LEVEL - LEFT WALL*, 5X, F6.0

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WRITE(6,350)X(62),X(63),X(64),X(65),X(66)
350 FORMAT(/20X,*BED TEMP. AT 6 INCH LEVEL - REAR WALL*,5X,F6.0
1/20X,*BED TEMP. AT 12 INCH LEVEL - REAR WALL*,5X,F6.0
2/20X,*BED TEMP. AT 24 INCH LEVEL - REAR WALL*,5X,F6.0
3/20X,*BED TEMP. AT 36 INCH LEVEL - REAR WALL*,5X,F6.0
4/20X,*BED TEMP. AT 48 INCH LEVEL - REAR WALL*,5X,F6.0,29(/))
C
C
WRITE(6,80) (COM1(I),I=1,10)
80 FORMAT(10A4)
WRITE(6,360)
360 FORMAT(/25X;*SSU STEADY STATE RAW DATA FILES*//)
C
C
WRITE(10,365) (COM1(I),I=1,10)
365 FORMAT(20(/),20X,10A4,10X,/)
WRITE(10,370)RED3,RED1,CF1,PLATE,DPGRID,LSTATM,EXPAN,BEDWT,BEDVOID
370 FORMAT(20X,*STEAM FLOW REYNOLDS NUMBER*,11X,F10.3/20X,
1*FEEDWATER FLOW REYNOLDS NUMBER *,F10.3/20X,
2*FEEDWATER ORFICE DIS. COEFF. *,F10.3//20X,
3*MEASURED GRID PLATE DELTA P *,F10.3,* IN H2O*/20X,
4*CALCULATED GRID PLATE DELTA P *,F10.3,* IN H2O*///20X,
5*FLUID BED CONDITIONS BASED ON THE*/20X,
6*MEASURED BED PRESSURE GRADIENT:*/20X,
7*STATIC BED HEIGHT *,F10.3,* IN*/20X,
8*BED EXPANSION *,F10.3,* PRCT*/20X,
9*BED MASS *,F10.3,* LB*/20X,
9*BED VOIDAGE *,F10.3,* PRCT*//)
WRITE(10,375) H2O,N2,O2,CO2,CO,FNOX,FSO2,SGTOT,SGMW ,PRCTGAS
375 FORMAT(20X,*CALCULATED STACK GAS CONSTITUENTS BASED*/20X,
1*ON THE MEASURED GAS CONCENTRATIONS:*/20X,
2*MOISTURE *,F10.3,* MOLES/HR*/20X,
3*NITROGEN *,F10.3,* MOLES/HR*/20X,
4*OXYGEN *,F10.3,* MOLES/HR*/20X,
5*CARBON DIOXIDE *,F10.3,* MOLES/HR*/20X,
6*CARBON MONOXIDE *,F10.3,* MOLES/HR*/20X,
7*NITROGEN OXIDE *,F10.3,* MOLES/HR*/20X,
8*SULFUR DIOXIDE *,F10.3,* MOLES/HR*/20X,
9*TOTAL (DRY) *,F10.3,* MOLES/HR*///20X,
9*DRY STACK GAS MOLECULAR WEIGHT *,F10.3,* LB/LB MOLE*/20X,
9*SUMMATION OF CONSTITUENT GASES *,F10.3,* PRCT*//)
WRITE(10,380) HFREE,PRCTFB,CEFF1,CEFF3,VSUPM,VSUPC
380 FORMAT(/20X,*ESTIMATED ABOVE BED COMBUSTION *,F11.3,
1* BTU/HR*/20X,
2*ESTIMATED ABOVE BED COMBUSTION *,F10.3,* PRCT*///20X,
3*COMBUSTION EFFICIENCY (SENSIBLE HEAT)*,F10.3,* PRCT*/20X,
4*COMBUSTION EFFICIENCY (STACK GAS O2) *,F10.3,* PRCT*///20X,
5*MEASURED SUPERFICIAL VELOCITY *,F10.3,* FT/SEC*/20X,
6*CALCULATED SUPERFICIAL VELOCITY *,F10.3,* FT/SEC*//)
325 STOP
END

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PROGRAM COMPARE (INPUT, OUTPUT, BASE, TAPES=INPUT, TAPES=OUTPUT,
1TAPES=BASE)
  DIMENSION DATA(43), INDEX(201), RUN(8), ARRAY(8, 43)
  DIMENSION HEAD1(3, 17), HEAD2(3, 16), HEAD3(3, 9)
  INTEGER TEST, NUMB, RUN
  DATA HEAD1/10HCOAL FEED , 10H(LB/HR - A, 10HS FIRED)
110HLINE FEED , 10H(LB/HR - A, 10HS FIRED) ,
210HTOTAL DRY , 10HAIR FLOW (, 10HLD/HR) ,
310HPERCENT OF, 10H MCR LOAD , 10H(PRCT) ,
410HTHEOR EXCE, 10HSS AIR (PR, 10HCT) ,
510HBED TEMPER, 10HATURE (DEG, 10H F) ,
610HCA/S MOLE , 10HRATIO , 10H ,
710HSUPERFICIA, 10HL VELOCITY, 10H (FPS) ,
810HFREEDBOARD , 10HVELOCITY (, 10HFPS) ,
910HACTIVE BED, 10H HEIGHT (I, 10HN) ,
110HBED PRES G, 10HRAD (IN H2, 10HQ/IN BED) ,
210HGRID PRESS, 10HURE DROP (, 10HIN H2O) ,
310HBED PRESSU, 10HRE DROP (I, 10HN H2O) ,
410HBED TUBES , 10HFLOW RATE , 10H(LB/HR) ,
510HSTEAMING R, 10HATE (LB/HR, 10H) ,
610HRECYCLE SO, 10HLIDS (LB/H, 10HR) ,
710HRECYCLE BY, 10HPASS SOLID, 10HS (LB/HR) /
  DATA HEAD2/10HBAGHOUSE S, 10HOLIDS (LB/, 10HHR)
110HBED DRAIN , 10HSOLIDS (LB, 10H/HR) ,
210HCYCLONE EF, 10HFICIENCY (, 10HPRCT) ,
310HOUTSIDE HQ, 10H - LOOP NO, 10H. 1 ,
410HOUTSIDE HQ, 10H - LOOP NO, 10H. 2 ,
510HOUTSIDE HQ, 10H - ECONOMI, 10HZER ,
610HOUTSIDE HQ, 10H - SUPERHE, 10HATER ,
710HSTACK O2 (, 10HPRCT) , 10H ,
810HSTACK CO (, 10HPRCT) , 10H ,
910HSTACK CO2 , 10H(PRCT) , 10H ,
110HSTACK NOX , 10H(PPM) , 10H ,
210HSTACK SO2 , 10H(PPM) , 10H ,
310HNOX EMISSI, 10HON (LB NOX, 10H/MMBTU) ,
410HSO2 EMISSI, 10HON (LB SO2, 10H/MMBTU) ,
510HNOX EMISSI, 10HON (PPM AT, 10H 3PRCT O2),
610HSO2 EMISSI, 10HON (PPM AT, 10H 3PRCT O2)/
  DATA HEAD3/10HSULFUR RET, 10HENTION (LB, 10H S/HR)
110HSULFUR RET, 10HENTION (PR, 10HCT) ,
210HCALCIUM UT, 10HILIZATION , 10H(PRCT) ,
310HCOMBUSTION, 10H EFFICIENC, 10HY (PRCT) ,
410HCARBON FRC, 10HTN IN RECY, 10HCLE ,
510HCARBON FRC, 10HTN IN RECY, 10HC BYPASS ,
610HCARBON FRC, 10HTN IN BAGH, 10HOUSE ,
710HABOVE BED , 10HCOMBUSTION, 10H (BTU/HR) ,
810HABOVE BED , 10HCOMBUSTION, 10H (PRCT) /
  REWIND 3
  REWIND 5
  REWIND 6
  WRITE(6, 10)
10 FORMAT(5X, *TYPE IN NUMBER OF RUNS TO BE COMPARED (8 RUNS MAX):*)
  READ(5, 20) NUMB
20 FORMAT(I1)
  WRITE(6, 30)
30 FORMAT(5X, *TYPE IN THE RUN INDEX NUMBERS WHEN PROMPTED:*)
  READ(5, 40) (RUN(I), I=1, NUMB)
40 FORMAT(I3)
  DO 60 I=1, NUMB
    TEST=RUN(I)
    CALL OPENMS(3, INDEX, 201, 0)
    CALL READMS(3, DATA, 43, TEST)
    CALL CLOSMS(3)
    DO 50 J=1, 43

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50 CONTINUE
60 CONTINUE
   WRITE(6,70) (RUN(I), I=1, NUMB)
70 FORMAT(/////5X, *SSU MATRIX TEST PROGRAM*, /5X, *REDUCED *,
1 *DATA COMPARISON*///5X, *RUN INDEX NUMBER*, 14X, 8(6X, 13, 3X)/)
   DO 90 J=1, 17
     WRITE(6,80) (HEAD1(K, J), K=1, 3), (ARRAY(I, J), I=1, NUMB)
80 FORMAT(5X, 3A10, 8F12.2)
90 CONTINUE
   DO 110 J=1, 16
     WRITE(6,100) (HEAD2(K, J), K=1, 3), (ARRAY(I, J+17), I=1, NUMB)
100 FORMAT(5X, 3A10, 8F12.2)
110 CONTINUE
   DO 130 J=1, 9
     WRITE(6,120) (HEAD3(K, J), K=1, 3), (ARRAY(I, J+33), I=1, NUMB)
120 FORMAT(5X, 3A10, 8F12.2)
130 CONTINUE
140 STOP
    END

```

Definitions of SSU77 Variables

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
Ao	Bed pressure drop	inches H ₂ O
A1	Bed pressure gradient	inch H ₂ O/inch bed
ABT1	Immersed bed tubes area	ft ²
AECON	Outside surface area of economizer tubes	ft ²
AIRM	Combustion air flow rate (dry basis)	moles/hr.
AREA	Furnace outside surface area	ft ²
ARTL	Test loop outside surface area	ft ²
ASUPR	Superheater outside surface area	ft ²
BEDDEL	Bed pressure drop	
BEDGRAD	Bed pressure gradient	inch H ₂ O/inch bed
BEDHTM	Measured bed height	inches
BEDVOID	Bed voidage	%
BEDWT	Bed mass	lb
BTUN2	NO _x emissions	lb NO ₂ /MMBTU
C1	O ₂ analyzer scale	%
C2	CO analyzer scale	%
C3	CO ₂ analyzer scale	%
C4	NO _x analyzer scale	PPM
C5	SO ₂ analyzer scale	PPM
CAC03	Weight fraction CaCO ₃ in bed	
CAC03D	Limestone feed rate (dry basis)	lb/hr
CAC03L	Weight fraction CaCO ₃ in limestone	
CAO	Weight fraction CaO in bed	
CAOR	Weight fraction CaO in recycle stream	

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
CARBI	Total carbon input	lb/hr
CARBO	Total carbon output	lb/hr
CASMR	Calcium/Sulfur mole ratio	
CASO4	Weight fraction CaSO_4 in bed	
CAUTL	Calcium utilization	%
CBAG	Weight fraction C in bag filter	
CC1	Coal feeder calibration factor	
CC2	Coal feeder calibration factor	
CCO3R	Weight fraction CaCO_3 in recycle stream	
CCOAL	Weight fraction C in coal	
CCOALD	Weight fraction C in coal (dry basis)	
CDRAN	Total weight fraction C in bed drain	
CDRAN1	Weight fraction C in bed drain	
CDRAN2	Weight fraction carbonate C in bed drain	
CDRUM	Weight fraction C in recycle bypass	
CEFF1	Combustion efficiency based on sensible heat balance	%
CEFF2	Combustion efficiency based on chemical heat balance	%
CEFF3	Combustion efficiency based on stack gas O_2 concentration	%
CELUT	Weight fraction C of elutriated solids	
CF1	Orifice coefficient	
CF3	Orifice coefficient	
CFLSOL	Weight fraction C in stack solids	
CFMA	Combustion air flow rate	ACFM
CFMAD	Combustion air flow rate (dry basis)	ACFM
CFMRD	Transport air flow rate (dry basis)	ACFM
CFMSG	Stack gas flow rate	ACFM

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
CFMSGD	Stack gas flow rate (dry basis)	ACFM
CHEMI	Total chemical heat input	Btu/hr
CHEMO	Total chemical heat output	Btu/hr
CL1	Limestone feeder calibration factor	
CL2	Limestone feeder calibration factor	
CLIME	Weight fraction carbonate C in limestone	
CLIMED	Weight fraction carbonate C in limestone (dry basis)	
CLOSS	Carbon loss from cool feeder belt	lb/hr
CNOX	NO _x emissions	ppm NO _x @ 3% O ₂
CNSTR	Constant	°R/psia
CNSTR1	Constant	°R/psia
CO	CO flow rate in stack gas	moles/hr
CO ₂	CO ₂ flow rate in stack gas	moles/hr
CO2BD	CO ₂ above bed	%
CO2SG	CO ₂ at top of boiler	%
COALA	Weight fraction ash in coal	
COALAD	Weight fraction ash in coal (dry basis)	
COALFC	Weight fraction fixed carbon in coal	
COALN	Weight fraction nitrogen in coal	
COALVM	Weight fraction volatile matter in coal	
COBD	CO above bed	
CONST	Constant	
CONV1	Conversion factor	
CONV2	Conversion factor	
CONW	Thermal conductivity of tubes	Btu/hr ft °F
COSG	CO at top of boiler	%
CP	Specific heat	Btu/lb°F

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
CPO	Specific heat of steam @ 200 Psi	Btu/lb °F
CRECY	Weight fraction C in recycle stream	
CSO2	SO ₂ emissions	PPM SO ₂ @ 3% O ₂
CSO4R	Weight fraction CaSO ₄ in recycle stream	
CSTGAS	Weight fraction C in stack gas (dry basis)	
CTCEFF	Cyclone efficiency	%
DATA (I)	Reduced data converted to matrix format	
DATE (I)	Test date	
DECON	Economizer outside diameter	ft
DELH	Heat pickup across superheater	Btu/hr
DET	Determinant	
DPGRID	Grid plate pressure drop	inches H ₂ O
DRAIN	Bed drain rate	lb/hr
DRSOL	Bed drain rate	lb/hr
DSUPR	Superheater outside diameter	ft
DTLM	Test loop tube log mean diameter	ft
DTLP	Test loop tube outside diameter	ft
ELUTR	Elutriation rate	lb/hr
EXAIR	Excess air	%
EXPAN	Bed expansion	%
F(I)	Flow rates	lb/hr
FA1	Thermal expansion factor	
FA2	Thermal expansion factor	
FA3	Thermal expansion factor	
FA4	Thermal expansion factor	
FA5	Thermal expansion factor	
FBAG	Bag filter collection rate	lb/hr

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
FBAGC	Weight fraction C in bag filter	
FCAO	Rate of CaO formation	1b/hr
FCASO4	Rate of CaSO ₄ formation	1b/hr
FCCO3	Limestone feed rate	1b/hr
FCCO3D	Limestone feed rate (dry basis)	1b/hr
FCOAL	Coal feed rate	1b/hr
FCOALC	Flow rate of coal C	1b/hr
FCOALD	Coal feed rate (dry basis)	1b/hr
FDRANC	Flow rate of bed drain C	1b/hr
FDRUM	Recycle bypass rate	1b/hr
FDRUMC	Flow rate of recycle bypass C	1b/hr
FELUTC	Flow rate of elutriated C	1b/hr
FLIMEC	Flow rate of limestone carbonate C	1b/hr
FLOWA	Combustion air flow rate	1b/hr
FLOWAD	Combustion air flow rate (dry basis)	1b/hr
FLOWC	Bed tube water flow rate	1b/hr
FLOWF	Feedwater flow rate	1b/hr
FLOWR	Transport air flow rate	1b/hr
FLOWRD	Transport air flow rate (dry basis)	1b/hr
FLWS	Steam flow rate	1b/hr
FLSO2	SO ₂ at top of boiler	PPM
FLSOL	Stack solids rate	1b/hr
FLSOLC	Weight fraction C in stack solids	
FLWSG	Stack gas flow rate	1b/hr
FLWSGD	Stack gas flow rate (dry basis)	1b/hr
FNOX	NO _x flow rate in stack gas	moles/hr

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
FRECTC	Flow rate of recycle C	lb/hr
FSO2	SO ₂ flow rate in stack gas	moles/hr
FTL1	Test loop 1 water flow rate	lb/hr
FTL2	Test loop 2 water flow rate	lb/hr
GPMC	Bed tube water flow rate	GPM
GPMF	Feedwater flow rate	GPM
GRID	Percent of bed pressure drop through grid plate	%
H	Steam tables parameter	Btu/lb
H2O	Flow rate H ₂ O in stack gas	moles/hr
H2OAMB	Humidity	lb H ₂ O/lb dry air
H2OCA	Flow rate H ₂ O in combustion air	lb/hr
H2OLB	Flow rate H ₂ O in stack gas	lb/hr
H2OTR	Flow rate H ₂ O in transport air	lb/hr
HA	Combustion air venturi pressure drop	inches H ₂ O
HAIR	Sensible heat in combustion air (dry basis)	Btu/hr
HBT	Bed tubes heat pickup	Btu/hr
HBTI	Bed tubes heat pickup	Btu/hr
HC	Bed tubes venturi pressure drop	inches H ₂ O
HCALC	Heat required for calcination reaction	Btu/hr
HCCO3	Sensible heat in limestone feed (dry basis)	Btu/hr
HCHAR	Chemical heat in clutriated C	Btu/hr
HCOAL	Sensible heat in coal feed	Btu/hr
HCOAL1	Sensible heat of coal feed (dry basis)	Btu/hr
HCOEF	Convective heat transfer coefficient	Btu/hr ft ² °F
HCOMB	Heat of combustion	Btu/hr
HCSO4	Heat released in sulfation reaction	Btu/hr
HDELTA	Wall tubes heat pickup	Btu/hr
HDRAN	Total heat in bed drain	Btu/hr

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
HDRAN1	Sensible heat in bed drain	Btu/hr
HDRAN2	Chemical heat in bed drain	Btu/hr
HEAT1	Enthalpy of test loop 1 water in	Btu/hr
HEAT2	Enthalpy of test loop 2 water in	Btu/hr
HEATI	Total heat input	Btu/hr
HEATO	Total heat output	Btu/hr
HELUT	Total heat in elutriated solids	Btu/hr
HELUT1	Sensible heat in elutriated solids	Btu/hr
HELUT2	Chemical heat in elutriated solids	Btu/hr
HELUTB	Sensible heat in elutriated solids	Btu/hr
HF	Feedwater orifice pressure drop	inches H ₂ O
HFEED	Enthalpy of feedwater	Btu/lb
HFH2O	Sensible heat in feedwater	Btu/hr
HFLUE	Total heat in stack gas (dry basis)	Btu/hr
HFLUE1	Sensible heat in stack gas (dry basis)	Btu/hr
HFLUE2	Chemical heat in stack gas	Btu/hr
HFLUEB	Sensible heat in stack gas (dry basis)	Btu/hr
HFREE	Freeboard heat release	Btu/hr
HG	Steam tables parameter	Btu/lb
HGAS	Specific heat of stack gas	Btu/mole °F
HH2OCA	Sensible heat in combustion air moisture	Btu/hr
HH2OIN	Sensible heat in total feed moisture	Btu/hr
HH2OTR	Sensible heat in transport air moisture	Btu/hr
HHV	Higher heating value	Btu/lb
HI(I)	Inside heat transfer coefficient	Btu/hr ft ² °F
HIBT	Inside heat transfer coefficient	Btu/hr ft ² °F
HLOSS	Heat loss	Btu/hr

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
HLP10	Sensible heat in test loop 1 water out	Btu/hr
HLP20	Sensible heat in test loop 2 water out	Btu/hr
HMOIST	Sensible heat in solid feed moisture	Btu/hr
HO(I)	Outside heat transfer coefficient	Btu/hr ft ² °F
HR	Transport air annubar pressure drop	inches H ₂ O
HREC1	Sensible heat in recycle stream	Btu/hr
HREC2	Chemical heat in recycle stream	Btu/hr
HRECY	Total heat in recycle stream	Btu/hr
HS	Steam flow orifice pressure drop	inches H ₂ O
HSG	Stack gas venturi pressure drop	inches H ₂ O
HSOL	Specific heat of solids	Btu/lb °F
HSTH20	Sensible heat in stack gas moisture	Btu/hr
HSUP	Sensible heat in superheated steam	Btu/hr
HT(I)	Pressure tap height above grid plate	inches
HTLP1	Sensible heat in test loop 1 water in	Btu/hr
HTLP2	Sensible heat in test loop 2 water in	Btu/hr
HTM1	Sensible heat in test loop 1 water at midpoint	Btu/hr
HTM2	Sensible heat in test loop 2 water at midpoint	Btu/hr
HTRAN	Sensible heat of transport air (dry basis)	Btu/hr
HYDRO	Weight fraction H in coal	
INERTL	Fraction inerts in limestone	
K	Steam tables parameter	
LSTATM	Static bed height	inches
M	Number of data points	
MASSI	Total mass in	lb/hr
MASSO	Total mass out	lb/hr
MGC03L	Weight fraction MGC03 in limestone	

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
MOIST	Weight fraction H ₂ O in coal	
MOISTL	Weight fraction H ₂ O in limestone	
MW	Molecular weight of moist air	lb/mole
N2	N ₂ flow rate in stack gas	mole/hr
NHT	Number of bed pressure taps	
NOX	NO _x at top of boiler	PPM
NOXBD	NO _x above bed	PPM
O2	O ₂ flow rate in stack gas	moles/hr
O2BD	O ₂ above bed	%
O2SG	O ₂ at top of boiler	%
O2TH	Flow rate O ₂ above theoretical requirement	moles/hr
OCOAL	Weight fraction O in coal	
P	Steam tables parameter	psia
PA	Combustion air pressure	psig
PAMB	Ambient pressure	psig
PC	Bed tube water pressure	psig
PERCC	Carbon unaccounted for	%
PERCH	Heat unaccounted for	%
PERCM	Mass unaccounted for	%
PERMCR	Percent of maximum continuous rating heat input	%
PF	Feedwater pressure	psig
PFURN	Furnace pressure	inches H ₂ O
PLATE	Measured grid plate pressure drop	inches H ₂ O
PR	Transport air pressure	psig
PRAN	Prandtl number	
PRCTFB	Freeboard heat release	%
PRCTGAS	Summation of stack constituent gases	%

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
PRESS(I)	Bed static pressure	inches H ₂ O
PS	Steam pressure	psig
PSG	Stack gas pressure	psig
PSUP	Superheated steam pressure	psig
PSUPD	Pressure drop across superheater	psi
PTL1	Test loop 1 water pressure	psig
PTL2	Test loop 2 water pressure	psig
RAD1	Radiative heat transfer coefficient for test loop 1	Btu/hr ft ² °F
RAD2	Radiative heat transfer coefficient for test loop 2	Btu/hr ft ² °F
RAIR	Theoretical air required for complete combustion	moles/hr
RE(I)	Reynolds number	
RECYC	Recycle rate	lbs/hr
RED1	Reynolds number	
RED3	Reynolds number	
REH ₂ O	Recycle water cooler flow rate	GPH
REQC	O ₂ required to burn C	mole/lb coal
REQH	O ₂ required to burn H	mole/lb coal
REQO	O ₂ contained in coal	mole/lb coal
REQS	O ₂ required to burn S	mole/lb coal
RHOA	Combustion air density	lb/ft ³
RHOAD	Combustion air density (dry basis)	lb/ft ³
RHOC	Bed tube water density	lb/ft ³
RHOF	Feedwater density	lb/ft ³
RHOP	Combustion air density	lb/ft ³
RHOR	Transport air density	lb/ft ³
RHOS	Steam density	lb/ft ³
RHOSG	Stack gas density	lb/ft ³

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
RHOT1	Test loop 1 water density	lb/ft ³
RHOT2	Test loop 2 water density	lb/ft ³
S	Stack gas venturi design constant	
SCOAL	Weight fraction S in coal	
SENIN	Total sensible heat in	Btu/hr
SENOUT	Total sensible heat out	Btu/hr
SGCARB	Flow rate C in stack gas	lb/hr
SGCO	Flow rate C in stack gas CO	lb/hr
SGCO2	Flow rate C in stack gas CO ₂	lb/hr
SGMW	Stack gas molecular weight	lb/mole
SGTOT	Stack gas flow rate (dry basis)	moles/hr
SI02	Weight fraction SI0 ₂ in bed	
SI02R	Weight fraction SI0 ₂ in recycle	
SL	Constant	
SMBT	SO ₂ emissions	lb/SO ₂ /MMBtu
S02BD	SO ₂ above bed	PPM
SQUAL	Steam quality	
SREM	Sulfur removal	lb/hr
SREMP	Sulfur removal	%
SUMHP	Matrix coefficient	
SUMHT	Matrix coefficient	
SUMHT2	Matrix coefficient	
SUMP	Matrix coefficient	
T	Steam tables parameter	°F
TA	Combustion air temperature	°F
TA1	Reduced temperature	°F
TA2	Reduced temperature	°F
TAMB	Ambient temperature	°F

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
TB1	Absolute temperature	°R
TB2	Absolute temperature	°R
TBED	Bed temperature	°F
TBED1	Bed temperature	°R
TC	Bed tube water temperature	°F
TCON	Thermal conductivity of water	Btu/hr ft °F
TECNG	Gas temperature after economizer	°F
TECNO	Economizer water outlet temperature	°F
TENLM	Economizer log mean temperature	°F
TEST	Test number	
TF	Feedwater temperature	°F
TFREEB	Freeboard temperature	°F
TFD	Transport air temperature (before adding solids)	°F
TG1	Gas temperature before superheater	°F
TG2	Gas temperature after superheater	°F
THCN	Thermal conductivity of steam	Btu/hr ft °F
THCNO	Thermal conductivity of steam at 1 atm	Btu/hr ft °F
THLP	Test loop tube thickness	ft
THSUP	Superheater tube thickness	ft
TIME(I)	Test time	
TL	Limestone feed temperature	°R
TLG1M	Test loop 1 log mean temperature at midpoint	°F
TLG2M	Test loop 2 log mean temperature at midpoint	°F
TLGBT1	Temperature drop across bed tube	°F
TMOIST	Total feed moisture	lb/hr
TMOLES	Stack gas flow rate (assuming complete combustion)	moles/hr
TOTAIRD	Total dry air flow	lb/hr

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
TOTAL	Summation of constituent solids	%
TR	Transport air temperature	°F
TR1	Reduced temperature	°F
TRANM	Transport air flow rate	moles/hr
TRECY	Recycle stream temperature	°F
TS	Steam temperature	°F
TS1	Steam temperature before superheater	°F
TS2	Steam temperature after superheater	°F
TSG	Stack gas temperature	°F
TSOL	Feed solids temperature	°F
TSPLM	Superheater log mean temperature	°F
TSUP	Superheated steam temperature	°F
TTL1	Test loop 1 water temperature at inlet	°F
TTL1M	Test loop 1 water temperature at midpoint	°F
TTL1O	Test loop 1 water temperature at outlet	°F
TTL2	Test loop 2 water temperature at inlet	°F
TTL2M	Test loop 2 water temperature at midpoint	°F
TTL2O	Test loop 2 water temperature at outlet	°F
TV(I)	Averaged temperature	°F
TVSUP	Averaged steam temperature in superheater	°F
TWALL	Outside wall temperature	°F
UBT1	Overall heat transfer coefficient	Btu/hr ft ² °F
UNACC	Unaccounted carbon	lb/hr
UNACH	Unaccounted heat	Btu/hr
UNACM	Unaccounted mass	lb/hr
UO(I)	Overall heat transfer coefficient	Btu/hr ft ² °F
VAR	Recycle flow/bypass indicator	
VCCO3	Limestone feed rate	volts

<u>Variable</u>	<u>Definition</u>	<u>Units</u>
VCOAL	Coal feed rate	volts
VDRAN	Bed drain solids collected	volts
VFREEM	Measured freeboard velocity	ft/sec
VIS	Viscosity	lb/hr ft
VOLT1	Test loop 1 flow rate	volts
VOLT2	Test loop 2 flow rate	volts
VSUPC	Calculated superficial velocity	ft/sec
VSUPM	Measured superficial velocity	ft/sec
W1	Mole fraction N_2 in stack gas (dry basis)	
W2	Mole fraction O_2 in stack gas (dry basis)	
W3	Mole fraction CO in stack gas (dry basis)	
W4	Mole fraction CO_2 in stack gas (dry basis)	
W5	Mole fraction NO_x in stack gas (dry basis)	
WG	Mole fraction SO_2 in stack gas (dry basis)	
X0	Constant	
X(I)	Digitrend 220 data input	volts
XX	Reduced temperature	$^{\circ}F$
Y	Reduced temperature	$^{\circ}F$
YA1	Adiabatic expansion factor	
YA2	Adiabatic expansion factor	
YA3	Adiabatic expansion factor	
YA4	Adiabatic expansion factor	
YA5	Adiabatic expansion factor	

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II. ZONAL HEAT BALANCE PROGRAM

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C PROGRAM SSU IS A FIRST ORDER DATA REDUCTION PROGRAM
C FOR THE C-F/DOE INDUSTRIAL FLUIDIZED BED SUB-SCALE UNIT.
C THIS CURRENT PROGRAM VERSION IS: SSU77
C
PROGRAM SSU (INPUT,OUTPUT,OUT,SSU1,CONS,BASE,TAPE3=SSU1,TAPE10=
OUT,TAPE1=BASE,TAPE4=CONS,TAPE5=INPUT,TAPE6=OUTPUT,DEFUG=OUTPLT)
DIMENSION TITLE(20),TIME(10),X(150),DATE(20),LPROB(20)
DIMENSION UN(4),TV(4),F(4),HE(4),HO(4),KE(4),HT(2),CC(150)
DIMENSION COM1(10),COM2(10),COM3(10),COM4(10)
DIMENSION HT(10),PRESS(5),DATA(74),INDEX(201)
INTEGER EPRGR,TEST
REAL MASS1,MASS2,MOIST,H2,NOX,K1,K2,NOYRD,LHV1,MOISTL,MGC03L,
INERTL,LSTAT,M2,LSTATM

```

```

C READ SSU INPUT DATA
C
M=8
REWIND 3
REWIND 4
REWIND 10
READ(3,30) (DATE(I),I=1,2),(TIME(I),I=1,2)
READ(3,35) DUMMY
READ(3,40) (X(I),I=1,M)
30 FORMAT(25X,A4,A4,12X,A4,A4)
35 FORMAT(A10)
40 FORMAT(56Y,F10.4)
READ(3,53) (COM1(I),I=1,10)
READ(3,53) (COM2(I),I=1,10)
READ(3,53) (COM3(I),I=1,10)
READ(3,53) (COM4(I),I=1,10)
53 FORMAT(10A4)

```

```

C READ CONSTANTS
C
READ(3,39) TEST
39 FORMAT(I3)
READ(3,41) VAR,REH2O,DRAIN,C1,C2,C3,C4,C5,FHAG,CHAG,
1CHFCY,CDRAM1,CDRAM2,FLSOL,CFLSOL,FDRUM,CDRUM,MOIST,CCOAL,
CHYDFO,SCOAL,COALN,COALA,MOISTL,CLOSE,H2OAMB
41 FORMAT(3F10.3)
READ(3,42) COALVM,COALFC,CAC03L,MGC03L,INERTL,
1CAC1,CAC0,CAS04,S102,CC03R,CADR,CS04R,S102R,
2TLALL,TAMP,PAHH,AREA,HCOEF,FOHAL,HT(1),HT(2),
3HT(3),HT(4),CC1,CC2,CL1,CL2
42 FORMAT(4F10.3)

```

```

C CONVERT DATA TO ENGINEERING UNITS
C

```

```

MA=(X(1)+0.5)**2
HE=X(2)**2
PA=X(3)**10./27.7
PRESS(1)=X(4)*7.5
PRESS(2)=X(7)*4.0
PRESS(3)=X(8)*4.25
PRESS(4)=X(9)*2.5
TA2=X(11)
TA=X(11)
TR=(X(14)+70.)/2.
TFD=X(14)
PR=X(3)*15./27.7
PEJFH=(X(12)-5.9)*0.0722
PSO=0.
H3O=(X(1)+0.447)**2
IFRTH=X(20)
TEC=X(21)
TSC=X(24)
TSC=X(24)

```

72

BED TUBES FLOW VENTURI

FA2=1.65E-5*TA+0.9988

YA2=1.0-0.024*HC/(C+14.7)

RHOC=70.442-0.0421*IC

FLOWC=1462.5*(FA2+YA2)*SQRT(HC*RHOC)

GMHC=1.1247*FLOWC/RHOC

C

STEAM FLOW ORIFICE

FA3=1.0E-5*TS+0.999

YA3=1.0-0.013*HS/(PS+14.7)

T=TS

K=2

CALL SATUR(T,VF,HF,SF,VG,HG,SG,K)

RHOS=1./VG

CF3=0.615P

FLU3S=916.28*CF3*FA3*YA3*SQRT(RHOS*HS)

RED3=FLU3S*(916.2-1.54*TS)

C

MOLECULAR WEIGHT OF MOIST AIR

MW=(1.*H2OAMR)/(1./28.84+H2OAMR/18.)

C

TRANSPORT AIR FLOW ANNULAR

CNSTR=(TFD+460.)/CPR+14.7)

RHOF=HW/(10.73*CNSTR)

FLOWR=RHOF*EO.*2.483*SQRT(HR*CNSTR)

H2OIR=FLOWR*H2OAMR/(H2OAMR+1.)

FLOWRU=FLOWR-H2OIR

CFMRD=FLOWRU*CNSTR/(2.688*EO.)

C

COMBUSTION AIR VENTURI

CNSTR1=(FA2+460.)/(PA+HA/27.7+14.7)

FA4=1.0E-5*TA2+0.9946

YA4=1.0-0.0245*HA/(PA+14.7*HA/27.7)

RHOAD=2.58P*(PA+HA/27.7+14.7)/(TA2+460.)

RHOA=HW/(10.73*CNSTR1)

FLOWA=3923.4*YA4*FA4*SQRT(HA*RHOA)

H2OCA=FLOWA*H2OAMR/(H2OAMR+1.0)

FLOWAU=FLOWA-H2OCA

CFMAD=FLOWAU/RHOAD/60.0

CFMAD=CFMAD*(PA+14.7*HA/27.7)/(PA+14.7)*(TA+460.)/(TA2+460.)

CFMA=FLOWAU/RHOA/60.

TOTAIRD=FLOWAU+FLOWRU

CFMA=CFMA*(PA+14.7*HA/27.7)/(PA+14.7)*(TA+460.)/(TA2+460.)

C

STACK GAS FLOW VENTURI

FA5=1.0E-5*ISG1+0.9986

YA5=1.0-0.0235*MSG/(PSG+14.7)

S=23145

RHOSG=2.70P*(PSG+14.7)/(ISG1+460.0)

FLMSG=22962.5*S*(FA5+YA5)*SQRT(RHOSG*MSG)

CFMSG=(FLMSG/RHOSG/60.0)*(ISG+460.)/(ISG1+460.0)

C

COAL FEEDER (1VOLT=45LB/HK)

FCCAL=(49.*VCCAL*CC1*CC2)-CLOSS

FCCALU=FCCAL*(1.-MOIST)

C

LIMESTONE FEEDER (1VOLT=25LB/HK)

FCCLS=25.*VCCLS*CL1*CL2

FCCLSU=FCCLS*(1.-MOISTL)

C

ASH STREAM

DROPL=BRATP

C

RECYCLE SOLIDS FLO

IF (VAR1+2.1*P3) TO 600

RECYC=7.0*EPRD*(XCF1+XCF2)/(VCF1+VCF2)

GO TO 690

680 RECYC=0.0

```

C      SSU COAL AND LIMESTONE ANALYSIS (AS FINE)
690  OCOAL=(1.-(CCOAL+HYDRO+SCOAL+COALH+COALA+MOIST))
      CCOALD=CCOAL/(1.-MOIST)
      COALAU=COALA/(1.-MOIST)
      CONST=COALVH/COALFC
      COALFC=(1.-COALAU)/(1.+CONST)*(1.-MOIST)
      COALVH=1.-POIST-COALA-COALFC

```

```

C      CACO3=CACO2L
      CACO2L=CACO2L*(1.-MOISTL)
      H2CO3L=H2CO3L*(1.-MOISTL)
      INFRTL=INFRTL*(1.-MOISTL)

```

```

C      PERCENT EXCESS AIR

```

```

GO TO 710
700  REOC=CCOAL/32.
      REQH=HYDRO/4.
      REQS=SCOAL/32.+(1.+5.*SRFMP/100.)
      REQO=OCOAL/32.
      RAIR=137.3*(REQC+REQH+REQS+HECO)+FCOAL
      EXAIR=100.+(FLOWAD+FLOWRD)/RAIR-1.1)

```

```

C      CALCULATED SUPERFICIAL VELOCITY

```

```

      TMOIST=MOIST*FCOAL+H2OIR+H2OCA+FCO3*MOISTL
      TMOLES=(FLOWAD+FLOWRD)/28.94+TMOIST/18.+HYDRO/2.*FCOAL
      VSUPC=6.24E-4*TMOLES*(TRED+460.)/(FURN*34.7)
      GO TO 140

```

```

C      RED HEIGHT CALCULATED FROM BED PRESSURE MEASUREMENTS

```

```

710  NHT=4
      SUMHT=0.
      SUMHT2=0.
      SUMH=0.
      SUMHP=0.
      DO 1230 I=1,NHT
        SUMHT=SUMHT + HT(I)
        SUMHT2=SUMHT2 + HT(I)**2
        SUMH=SUMH + PRESS(I)
        SUMHP=SUMHP + HT(I)*PRESS(I)
1230 CONTINUE

```

```

      DET=NHT*SUMHT2-SUMHT**2
      AU=(SUMH*SUMHT2-SUMHT*SUMHP)/DET
      A1=(NHT*SUMHP-SUMHT*SUMH)/DET
      BEDHT=(27.7*FURN-A1)/A1
      BEDDEL=A1
      PLATE=PA+27.7-A1
      GRIF=196.*PLATE/AC

```

```

C      FLUIDIZED BED CONDITIONS CALCULATED FROM
      BED STATIC PRESSURE MEASUREMENTS

```

```

      LSTAT=9194*BEDDEL
      EXFAN=100.*(BEDHT/LSTAT-1.0)
      BEDHT=28.26*BEDDEL
      REDVOID=(1.-60.*LSTAT/BEDHT)*100.
      BEDGRAD=3BEDDEL/BEDHT

```

```

C      CALCULATED GRID PLATE DELTA P

```

```

      RHH=(HHA+HA*14.7)/4PA+HA/(27.7+14.7)-11A2*454.)/(TA+454.)
      DPGT=HHH*EXP(2.453*ALOG(EXFAN)-5.7)
      DPGT=DPGT/144.*27.7

```

```

C
IF (EUNTH.GT.24.) GO TO 659
TRCD=(X(57)+X(58)+X(62)+X(63))/4.0
GO TO 679
658 IF (EUNTH.GT.36.) GO TO 651
TRCD=(X(57)+X(58)+X(59)+X(62)+X(63)+X(64))/6.0
GO TO 679
651 IF (EUNTH.GT.43.) GO TO 652
TRCD=(X(57)+X(58)+X(59)+X(60)+X(62)+X(63)+X(64)+X(65))/8.0
GO TO 679
652 TRCD=(X(57)+X(58)+X(59)+X(60)+X(61)+X(62)+X(63)+X(64)+X(65)
+X(66))/10.0

```

```

C
C
C
C
TEST TYPE 1 (TURBINE METER 86647)
670 IF (VOLT1.GT.1.423) GO TO 800
T1GPM=.978+VOLT1-.0623
GO TO 820
800 IF (VOLT1.GT.6.995) GO TO 810
T1GPM=.7905+VOLT1-.0318
GO TO 820
810 T1GPM=1.3475+VOLT1-.3949

```

```

C
TEST TYPE 2 (TURBINE METER 86648)
820 IF (VOLT2.GT.1.861) GO TO 830
T2GPM=.9333+VOLT2-.0152
GO TO 850
830 IF (VOLT2.GT.6.996) GO TO 840
T2GPM=.9949+VOLT2-.63175
GO TO 850
840 T2GPM=1.3254+VOLT2-.2492
850 P=PIL1+14.7
T=TIL1
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
RHO1=1./V
HEAT1=H-45.19
FIL1=.4.92*RHO1+T1GPM
P=PIL2+14.7
T=TIL2
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
RHO2=1./V
HEAT2=H-45.19
FIL2=.4.92*RHO2+T2GPM

```

```

C
C
C
C
CALCIUM/SULFUR MOLE RATIO
CASMR=C.321+FCO3L+CAC03L/(FCOAL+SCOAL)
C
C
ELUTRIATED SOLIDS FLOW
ELUTR=RECYC+FRAG+FLSOL+FDROM
GO TO 870

```

```

C
C
C
C
OVERALL ENERGY BALANCE
HEAT INPUTS

```

```

C
HEAT OF SULFATION
860 TRCD1=(TRCD+460.)/1.8
FCAS04=SRFH+136./32.
HCS04=FCAS04+140.
GO TO 870

```

```

C
C
ENTHALPY OF PRODUCTION H1
870 TAI=(TA+460.)/1.8
AIRPI=FLOWAD/28.84

```

```

C      MAIN=FAIRH*(11.59*TA1+1.48E-3*TA1**2-1.16E-7*TA1**3-3582.)
C
C      ENTHALPY OF FEEDWATER
C      HFM20=FLOWF*HFETD
C
C      ENTHALPY OF TEST LOOP STREAMS
C      HTL1=HEAT1*FTL1
C      HTL2=HEAT2*FTL2
C      P=PTL1+14.7
C      T=TTL10
C      CALL SRSORT(P,T,V,H,S,ISAT,VG,HC,SG)
C      H=H-45.39
C      HLP10=H*FTL1
C      P=PTL2+14.7
C      T=TTL20
C      CALL SRSORT(P,T,V,H,S,ISAT,VG,HC,SG)
C      H=H-45.39
C      HLP20=H*FTL2
C
C      ENTHALPY OF TRANSPORT AIR
C      TR1=(TR+960.)/1.8
C      TRANH=FLOWRD/28.84
C      HTRAN=TRANP*(11.59*TR1+1.48E-3*TR1**2-1.16E-7*TR1**3-3582.)
C
C      ENTHALPY OF LIMESTONE FEED
C      TL=(TSOL+460.)/1.8
C      HCCO3=FCCO3D*(19.68+.01189*TL-307.05/TL**2)*(TSOL-77.)/100.1
C
C      ENTHALPY (SENSIBLE) OF COAL FEED
C      HCOAL1=FCCOALD*.24*(TSOL-77.)
C
C      ENTHALPY OF FEED MOISTURE
C      HHMO1=(MOIST*FCCOAL*MOISTL*FCCO3)*(TSOL-77.)
C      HH20TR=H20TF*(.46*TR+1015.)
C      HH20CA=H20CA*(.46*TA+1015.)
C      HH20TH=HH20TR+HH20CA+HHMO1
C
C      HEAT OF COMBUSTION - AVERAGE ASH AND MOISTURE FREE HIGHER
C      HEATING VALUE FOR FREEMAN COAL
C      HHV=15095.*(1.-MOIST-EOAL)
C      HCOMB=FCCOAL*HHV
C      HCOAL=HCOMB*HCOAL1
C      PERHCR=HCOMB/2570000.*100.
C
C      HEAT OUTPUTS
C
C      ENTHALPY OF DRAIN SOLIDS
C      HSOL=.29*(TFED-77.)
C      HDRAIN1=HSOL*DRSOL
C      HDRAIN2=DRSOL*CDRAIN1+14093.
C      HDRAIN=HDRAIN1+HDRAIN2
C      HELUTR=HSOL*FLUTR
C
C      ENTHALPY OF FLUTRIATED SOLIDS
C      HSOL=.24*(TSG-77.)
C      HELUT1=FLUTE*HSOL
C      HELUT2=14193.*(FDRAG*CFAR+PECYC*CRECY+FLSOL*CLSOL+FORUM*CDRUM)
C      HELUT=HELUT1+HELUT2
C
C      ENTHALPY OF RECYCLE SOLIDS
C      HSOL=.22*(TECY-77.)
C      HREC1=FLCYC*HSOL
C      HREC2=RECYC*CRECY+14093.
C      HRECY=HREC1+HREC2

```

```

C
C      HEAT OF CALCINATION
C      THEO=(THEO+460.)/1.8
C      FCAO=FCCO3*CAO3L*56./100.
C      HCALC=FCAO*1368.
C
C      STACK GAS COMPOSITION IN MOLES
C      IMOIST=MOIST*FCOAL+H2OIR+H2OCA*FCCO3*MOISTL
C      H2O=(IMOIST+HYDRO*9.*FCOAL)/18.0
C      H2OLH=H2O*11.
C      N2=(AIR*TRANH)*0.79
C      CONV2=(1.-(COSG+C2SG+O2SG+NOX/10000.*FLSO2/10000.)/100.)
C      SGTOT=N2/CONV2
C      CONV1=SGTOT/100.
C      D2=O2SG*CONV1
C      CO=COSG*CONV1
C      C2=C2SG*CONV1
C      FNOX=NOX/10000.*CONV1
C      FSO2=FLSO2/10000.*CONV1
C      W1=N2/SGTOT
C      W2=O2/SGTOT
C      W3=CO/SGTOT
C      W4=C2/SGTOT
C      W5=FNOX/SGTOT
C      W6=FSO2/SGTOT
C      SGHVE=W1*28.+W2*32.+W3*28.+W4*44.+W5*18.+W6*44.+W6*64.
C      FLVSGU=SGTOT*SGHV
C      CFMSGU=G.175*SGTOT*(TSG+460.)/(PFURN*14.7)
C      PRTGAS=O2SG+COSG+C2SG+(NOX+FLSO2)/10000.
C
C      MEASURED SUPERFICIAL VELOCITY
C      VSUIM=6.149E-4*(SGTOT*H2O)*(THEO+460.)/(PFURN*14.7)
C      VFREEH=VSUIM*.553*(TFREEB+460.)/(THEO+460.)
C
C      ENTHALPY OF DRY STACK GAS
C      TH1=(TSG+460.)/1.8
C      X0=0.9
C      DO 150 I=1,2
C      HGAS=((6.52*W1+6.15*W2+6.42*W3+6.21*W4)*(TB1-298.)
C      1+(1.60125*W1+.0031*W2+.05167*W3+.6104*W4)*
C      2(TB1**2-APPA4.)/2.0 -(1.0E-9*W1+9.25E-7*W2+1.96E-7*W3+3.545E-6
C      36*W4)*(TB1**3-2.65E7)/3.6)*1.8
C      IF (X0.GT.0.0) GO TO 150
C      HFLUE1=HGAS*SGTOT
C      HFLUE2=CO*121060.
C      HFLUE= HFLUE1+ HFLUE2
C      X0=1.0
C      TB1=(THEO+460.)/1.8
C 150 CONTINUE
C      HFLUER=HGAS*SGTOT
C
C      ENTHALPY OF STACK MOISTURE
C      HSTH2O=H2OLF*(1.46*TSG+1016.)
C
C      HEAT LOSS - DETERMINED FROM BOILER CASING INSPECTION
C      HLGSS=110000.
C
C      ENTHALPY OF SUPERHEATED STEAM
C      P=PSUP*14.7
C      T=TSUP
C      CALL JBOOTIP(F,V,H,G,IPAT,MS,IN,SC)
C      H=H-45.09
C      HSUP=H*FLOWS

```

```

C
C      NOX AND SO2 EMISSIONS
CNOX=16.*NOX/(21.-O2S6)
CSO2=18.*F(SO2)/(21.-O2S6)
BTUN2=46.*FACX+1.*F6/HCOMB
SRFH=SCOAL*FCOAL-FSO2*32.1
SMST=(SCOAL*FCOAL-SMST)*1.E6*2./HCOMB
SRFMP=100.*SRFH/(SCOAL*FCOAL)
GO TO 863

C
C      TOTAL HEAT IN
140 CHEMI=HCOMB*HREC2+HCSO4
SENSIN=HCOAL1*HCCO3+HAIR*HTRAN*HTLP1+HTLP2*HREC1*HFH2O
H*H2OIN
HEATI=CHEMI+SENSIN

C
C      TOTAL HEAT OUT
CHEMO=HDRAN2*HEUT2+HCAIC*HFLUE2
SENOUT=HEUT1*HFLUE1+HLP10+HLP20+HSTP*HURAN1+HLOSS*HSTH2O
HEATO=CHEMO+SENOUT
UNACH=HEATI-HEATO
PERCH=100.*UNACH/(HEATI-HREC1-HREC2)

C
C      OVERALL MASS BALANCE
MASSI=FCOALD*FCCO3D*FLOWD*FTL1*FTL2*FLOWRD*RECYC*THOIST
MASSO=FLWGG*ORSOL*FLOUS*FTL1*FTL2*ELUTR+H2OLB
UNACH=MASSI-MASSO
PERCH=100.*UNACH/(MASSI-RECYC)

C
C      CALCIUM UTILIZATION
CAUTL=SRFMP/CASHR

C
C      OVERALL CARBON BALANCE
C
C      CARBON IN
FCOALC=FCOAL*CCOAL
CLIME=CACO3L*.120
CLTHEU=CACO3O*.120
FLIMEC=FCCO3*CLIME
FRECYC=RECYC*CRECY
CARET=FCOALC+FLIMEC+FRECYC

C
C      CARBON OUT
FBACC=FBAG*CBAG
FDRUMC=FDRUM*CDRUM
FLSOLC=FLSOL*FLSOL
CELUT=(FBACC+FLSOLC+FRECYC+FDRUMC)/ELUTR
FELUTC=FLUTR*CELUT
CDRAN=CDRAN1*CDRAN2
FDRANC=ORSOL*CDRAN
SGCO=CO*12.
SGCO2=CO2*12.
SGCARB=SGCO+SGCO2
CSTGAS=SGCARB/FLWGG
CARMO=FELUTC+FDRANC+SGCARB
UNACC=CARPI-CARMO
PERCC=((UNACC/(CARPI-FRECYC))*100.

C
C      OVERALL COMBUSTION EFFICIENCY CALCULATED
C      FROM SENSIBLE AND CHEMICAL HEAT BALANCES
CEFF1=100.*(HEI-OUT-SENSIN+HCOAL1*HCCO3)/HCOMB
CEFF2=100.*(HCOMB-1493.*CFBAG*CBAG+FLSOL*FLSOL
1+FDRUM*CDRUM)-HEI-OUT-HDRAN2)/HCOMB

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C
C
C COMBUSTION EFFICIENCY CALCULATED
C FROM STACK GAS O2 CONCENTRATIONS
O2TH=(FXAIR*RAIR/28.8)*.21/100.
HCHAR=(O2-O2TH)*12.*14093.
CEFF3=100.*(HCOMP-HCHAR-HFLUE2)/HCOMP
C
C RED. AND ECONOMIZER HEAT TRANSFER COEFFICIENTS
C
C ENTHALPY OF TEST LOOP STREAMS AT LOOP MID-POINT
DTLP=.1563
DTLM=.1436
HTLP=.91667
ARTL=1.054
CONL=33.0
AECON=58.76
DECON=1.0433
P=PTL1*14.7
T=TTL1M
DO 401 I=1,2
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
HTM(I)=H-45.09
P=PTL2*14.7
401 T=TTL2M
HTM1=HTM(1)*FTL1
HTM2=HTM(2)*FTL2
C
C ENTHALPY OF ECONOMIZER OUTLET
P=PF*14.7
T=TECNO
CALL SRSORT(P,T,V,H,S,ISAT,VG,HG,SG)
H=H-45.09
RECON=H*FLOWF
C
C LOG MEAN TEMPERATURES
TLG1M=(TTL1M-TTL1)/ALOG((THED-TTL1)/(TFID-TTL1M))
TLG2M=(TTL2M-TTL2)/ALOG((THED-TTL2)/(TFID-TTL2M))
TECHM=(X(21)-TF-TECNG*TECNO)/ALOG((X(21)-TF)/(TECNG-TECNO))
C
C OVERALL HEAT TRANSFER COEFFICIENT
UO(1)=(HTM1-HTLP1)/ARTL/TLG1M
UO(2)=(HTM2-HTLP2)/ARTL/TLG2M
UO(3)=(RECON-HFHP2)/TECHM/AFCOR
C
C APPROXIMATE RADIATION COMPONENT TO OUTSIDE
C HEAT TRANSFER COEFFICIENT ON TEST LOOPS
RAD1=.1714E-8*.85*(.1714E+460.)*.4*(THED-TLG1M+460.)*.4)/TLG1M
RAD2=.1714E-8*.85*(.1714E+460.)*.4*(TLED-TLG2M+460.)*.4)/TLG2M
C
C BULK MEAN TEMPERATURE
TV(1)=(TTL1M+TTL1)/2.
TV(2)=(TTL2M+TTL2)/2.
TV(3)=(TECNO+TF)/2.
F(1)=FTL1
F(2)=FTL2
F(3)=FLOWF
C
C HEAT CAPACITY, THERMAL CONDUCTIVITY AND VISCOSITY VALUES
C CURVE FITTING BY STIRLING INTERPOLATION FORMULA
DO 402 I=1,3
XX=(TV(I)-200.)/100.
CF=1.0151-.024*XX+.009*XX**2+.0014*XX**3+.00063*XX**4
TC=N*0.3174*5.63E-4*TV(I)-1.034E-6*TV(I)**2
Y=(TV(I)-302.)/90.
VIS=C1810.-602.5*Y+219.6*Y**2-122.5*Y**3+35.4*Y**4)*2.419E-4

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PRAN=CP*VIS/TCOH
IF(1.E0.3) GO TO 405
RE(1)=6.050*F(1)/VIS

C
C      INSIDE HEAT TRANSFER COEFFICIENT - DITTUS-BDELTER CORRELATION
C      HI(1)=.6210*TCOH*RE(1)**0.8*PRAN**0.4
C
C      OUTSIDE HEAT TRANSFER COEFFICIENT
C      HO(1)=1.0/(1.0/UC(1)-1.187/HI(1)-THLP*DTLP/CONV/DTLM)
C      GO TO 402

C
C      REYNOLDS NUHNER
C      .05 RE(3)=20.972*F(3)/VIS
C      HI(3)=.5781*TCOH*RE(3)**0.8*PRAN**0.4
C      HO(3)=1.0/(1.0/UC(3)-1.370/HI(3)-.0113*DECON/(CDNV*.0714))
C      402 CONTINUE

C
C      APPROXIMATE CONVECTION COMPONENT TO OUTSIDE
C      HEAT TRANSFER COEFFICIENT ON TEST LOOPS
C      CONV1=HO(1)-RAD1
C      CONV2=HO(2)-RAD2

C
C      SUPERHEATER HEAT TRANSFER COEFFICIENT
C      DSUPR=0.125
C      THSUP=0.3113
C      ASUPR=23.45
C      TS1=X(45)*100.
C      TS2=X(46)*100.
C      TG1=X(20)
C      TG2=X(21)
C      TSPLN=(TG1-TS1)-TG2+TS2)/ALOG((TS1-TG1)/(TS2-TG2))
C      TVSUP=(TS1+TS2)/2.
C      XX=(TVSUP-50.)/50.
C      CP0=.55-0.025*XX-.0045*XX**2+.0004*XX**4
C      SL=(6.67-1.499*XX-.169*XX**2-1.026*XX**3+.524*XX**4)*1.0E-4
C      CP=CP0+SL*(FSUP-200.)
C      Y=(TVSUP-302.)/90.
C      THCNO=(33.2+4.783*Y+.258*Y**2-.233*Y**3-.008*Y**4)*.577H/1000.
C      THCN=THCNO*(1.0+FSUP/4000.)
C      VIS=(160.2+19.358*Y+1.038*Y**2+.192*Y**3-.189*Y**4)*2.419E-4
C      RE(4)=12.43*FLOWG/VIS
C      PRAN=CP*VIS/THCN
C      T=TL
C      K=2

C
C      ENTHALPY OF SUPERHEATER INLET STEAM
C      CALL SATUR(F,T,VF,HF,SF,VG,HG,SG,K)
C      HVAR=HG-HF
C      HG=HG-45.06
C      DLLH=HSUP-FLOWS*HG
C      UC(4)=DLLH/TSPLN/ASUPR
C      HI(4)=.2244*THCN*RE(4)**0.8*PRAN**0.4
C      HO(4)=1.0/(1.0/UC(4)-1.2125/HI(4)-.1124/CONV)

C
C
C
C      FREEBOARD HEAT RELEASE - ESTIMATED
C
C      IMMERSIED RFD TUBE SURFACE AREA
C      IF(PEDITH.GT.11.) GO TO 396
C      ART1=0.
C      GO TO 403
C      396 IF(PEDITH.GT.15.) GO TO 397
C      ART1=.041*(12.36*(PEDITH-12.))

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ABT1=.041*(2.97*BEDHTM*19.)
 GO TO 400
 398 IF (BEDHTM.GT.38.4) GO TO 399
 ABT1=.041*(13.33*BEDHTM-323.)
 GO TO 400
 399 ABT1=7.75

C
 C BED TUBE HEAT PICK-UP (IN BED)

400 HIRT=5000.
 TLGRT=TBED-TS
 UBT1=1./(1.214/HIRT+1./H011)+.855*.165/CONV1
 HBT1=URT1*ABT1*TLGRT1

C
 C
 C FREEBOARD HEAT RELEASE

HBT=HBT1
 HDELTA=HSUP-HFRED*FLOWS-HET-HLOSS
 HFRFE=HDELTA-(HFLUER*HFLUTR)-(HCLUT1*HFLUE1)
 PRCTFB=(HFRFE/(CEFF2*HCONR))*10000.

C
 C
 C CYCLONE EFFICIENCY
 CYCFFF=(1.-FBAG/CLUTR)*100.

C
 C
 C
 C
 C CALCULATE PERCENT HEAT RELEASED IN BED

TMST=H2OLP
 HHV1=13985.*(1.-MOIST-COALA)
 FLWSG=FLWSGD
 BEDHT=BEDHTM
 TS=X(45)*1000.0
 TG1=X(20)
 TECHG=X(22)
 TLEMO=X(34)*1000.0
 TF=Y(43)*1000.0
 TRFF=40.0
 TREF1=(80.0**2)
 TREF2=(80.0**3)
 TREF3=(80.0**4)
 UCAL=FCOAL*(1.0-MOIST)*0.24*(TSOL-TRIF)
 A=TA-TRFF
 TL1=(TSOL+460.)/1.8
 TL2=(TL1+460.)/1.8
 TL3=(TL1-TL2)
 TL4=(TL1**2)-(TL2**2)
 TL5=(1./TL1)-(1./TL2)
 QLIME=(FCO3D/100.1)*419.68*TL3+5.945(-3*TL4-3*.076E5*TL5)*1.8
 A1=(TA**2)-TREF1
 A2=(TA**3)-TREF2
 QCAIR=(FLOWAD/28.969)*(6.953*A+3.798E-4*A1-2.207E-8*A2)
 A3=TR-TREF
 A4=(TR**2)-TREF1
 A5=(TR**3)-TREF2
 QTAIR=(FLOWAD/28.969)*(6.953*A3+3.798E-4*A4-2.207E-8*A5)
 QCMOIS=(H2OCA/18.)*47.827*A+0.674E-3*A1+2.747E-8*A2
 QTMOIS=(H2OIR/18.)*47.827*A3+1.674E-3*A4+2.747E-8*A5
 TMCL=(FCOAL*MOIST)*(FCO3S*MOISTL)
 AG=TSOL-TR1
 A7=(TSOL**2)-TREF1
 A8=(TSOL**3)-TREF2
 QCLM=(TMCL/10.0)*(1.0**4*AL40.68E-3*A7+0.047E-6*A8)
 QRECY=RECYC*0.2*(TREC1-TRCF)
 QIN=QCOAL+QLIME+QCAIR+QTAIR+QCMOIS+QTMOIS+QCLM+QRECY

A10=((TA+460.0)/100.0)**4
 AA9=((X(67)+X(68)+X(65)+X(70)+X(71))/5.0
 AS=((AA9+460.0)/100.0)**4
 GRADP=4.69*0.85*0.173*(A9-A10)
 A11=((TBED+460.0)/100.0)**4
 A12=TB1
 A13=((A12+460.0)/100.0)**4
 QTRANS=1.0-((0.40/0.85)*((TG1/TBED)**4))
 GRADSH=4.69*0.85*0.173*A11*QTRANS*0.26
 QRF1=9560.0*(BEDHT/12.0)
 GSPL=H00.0*0.3*(TBED-A12)
 QLS=0.333*0.50*(TBED-450.0)
 C
 CALCULATE STEAM PRODUCTION IN HFD
 HISTH=5000.0
 DOD=0.1563
 DID=0.128H
 DRL=0.142
 XWALL=0.31375
 A14=DOD/(DID*XWALL)
 A15=(DOD*XWALL)/(CONW*DRL)
 A16=1.0/H0(1)
 UOS1=1.0/(A14+A15+A16)
 A17=1.0/H0(2)
 UOS2=UOS1
 UOS=(UOS1+UOS2)/2.0
 IF (BEDHT.GE.19.0) GO TO 1
 STOT=1.95392857*BEDHT-5.630928571
 GO TO 2
 1 IF (BEDHT.GE.38.0) GO TO 3
 STOT=0.1321578947*BEDHT+0.3018421653
 GO TO 2
 3 IF (BEDHT.GT.42.0) GO TO 4
 STOT=0.453375*BEDHT-11.12125
 GO TO 2
 4 STOT=1.75
 GO TO 5
 2 CONTINUE
 5 CONTINUE
 TSA1=TS+25.0
 USBF0=(UOS*STOT*(TBED-TSA1))/HVAP
 QSBF0=USBF0*HVAP
 IF (BEDHT.LE.21.0) GO TO 7
 IF (BEDHT.GT.21.0.AND.BEDHT.LE.25.5) GO TO 9
 QTL1=FTL1*1.0*(TTL10-TTL1)
 QTL2=FTL2*1.0*(TTL20-TTL2)
 GO TO 11
 9 QTL1=FTL1*1.0*(TTL1H-TTL1)
 QTL2=FTL2*1.0*(TTL2H-TTL2)
 GO TO 12
 7 QTL1=0.0
 QTL2=0.0
 GO TO 8
 12 CONTINUE
 11 CONTINUE
 8 CONTINUE
 A18=TBED-TRLF
 A19=(TRLO**21-TRIF1
 A20=(TBED**21-TRIF2
 QSPL=ELUTR*0.29*A18
 B1=101.76
 B2=F1**2
 B3=B1**3
 B4=TBED-41
 B5=(TBED**2)-B2
 B6=(CROCDW*3)-B3

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QTH202=(THST*QTH201)/18.0
QTH20=QTH202*(THST+1037.1)+(THST+21.1)
QK2=Q2*(6.855*A18+0.3812E-3*A19-0.2336E-7*A20)
QURN=ORSOL*0.29*(TRED-R0.0)
QK2=Q2*(7.104*A18+0.3926E-3*A19-0.1843E-7*A20)
A21=(TRED**4)-TREF3
QC02=C02*(1.44P*A18+2.8785E-3*A19-7.197E-7*A20+0.7648E-10*A21)
QC0=C0*(6.865*A18+0.4012E-3*A19-0.2456E-7*A20)
QGB=QK2+Q02+QC02+QC0
A1CB=((1.33*95.1)+111.18)*20.8
GLATH=A100*(REDHT/12.0)
C FOLLOWING CALCULATES HEAT OF CALCINATION
WCAC03=FCC03*.9800
WMC07=WCAC03/100.091
WCAOM=WMC03
WCAOM=WCAOM*56.08
TREDK=(TRED+460.1)/1.8
A61=278.9
A62=(TREDK-A61)
A63=(TREDK**2)-(A61**2)
A64=(1./TREDK)-(1./A61)
DHPR0=20.34*A62+3.79E-3*A63+3.035E5*A64
A65=(TSOL+40.0)/1.8
A66=(A65-A61)
A67=(A65**2)-(A61**2)
A68=(1./A65)-(1./A61)
DHREC=19.0P*A66+5.95E-3*A67+3.076E5*A68
DHPR0=20.34*A66+3.79E-3*A67+3.035E5*A68
DHNET=43748.0*DHPR0-DHREC
DHNT=DHNET*1.8
C ASSUME COMPLETE CALCINATION
HTCALC=DHNT*WCAOM
C FOLLOWING CALCULATES HEAT OF SULFATION
C ASSUME 85 PERCENT SO2 RETENTION
A69=TREDK**2
A70=1./TREDK
A71=TREDK**3
HTS04=-115477.7P-3.315*TREDK+5.05E-3*A69-45050.0*A70+27.67E-7*A71
HTS04=HTS04*1.8
SULM=(SREM/32.0)
HEATSU=SULM*HTS04M
HEATSU=ABS(HEATSU)
HTNET=(DHPR0-DHREC)*1.8*WCAOM
QOUT=QRAOP+QRAOSH+QREF1+QSPL+QLS+QSPFD*QTL1+2TL2*QTH20+QSOL+QGG*BL
1ATH=HTNET+QTH20
QREF=(QOUT-QIN)/TFCOAL*HHV1)
C
C
C CALCULATE PERCENT HEAT RELEASED IN SPLASH ZONE
T43=(A61)*Y(66)/2.0
A22=TRED-T43
A23=(TRED**2)-(T43**2)
A24=(TRED**3)-(T43**3)
A25=(TRED**4)-(T43**4)
QK25=Q2*(6.195*A22+0.3812E-3*A23-0.2336E-7*A24)
QC25=Q2*(7.104*A22+0.3926E-3*A23-0.1843E-7*A24)
QC025=C02*(1.44P*A22+2.8785E-3*A23-7.197E-7*A24+0.7648E-10*A25)
QC05=C0*(6.865*A22+0.4012E-3*A23-0.2456E-7*A24)
QGS=QK25+Q025+QC025+QC05
QTH205=(THST/18.0)*(7.827*A22+0.6791-3*A23+2.747E-8*A24)
QSOLS=FLUTH*0.2P*A22
QAVL=QGS+QTH205+QSOLS
C CALCULATE CEMENT PRODUCTION IN SPLASH ZONE
DT1=TRED-(T1+21.0)
DT2=T43-(T5+35.0)
DTLPI=(QDT1-DT2)/(CALOG(QDT1/DT2))

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SPTOT=7.75E-STOT
UOSPL=UOS*(EXP((0.03328*REDHT)-1.3312))-1.3312
WSSPL=(UOSPL*SPTOT*DTLM)/HMAP
QSSPL=UOSPL*HMAP
QTL1S=(FTL1*1.0)*(TTL10-TTL1)-QTL1
QTL2S=(FTL2*1.0)*(TTL20-TTL2)-QTL2
IF(REDHT.GT.40.0) GO TO 5000
QLSPL=A100*(46.0-REDHT)/12.
QRSPL=9560.0*(40.0-REDHT)/12.
GO TO 6000
5000 QLSPL=A100*(61.-REDHT)/12.
QRSPL=9560.0*(61.-REDHT)/12.
GO TO 7000
6000 CONTINUE
7000 CONTINUE
IF(REDHT.GT.40.0) GO TO 8000
QLSFH=A100*3.27
GO TO 9000
8000 QLSFH=A100*(61.-REDHT)/12.
GO TO 9200
9000 CONTINUE
9200 CONTINUE
QOUTSP=QSSPL*QTL1S+QTL2S*QLSPL+GRSPL
QRESP=(QOUTSP-QAVAL)/(FCOAL*HHV1)
C CALCULATE PERCENT HEAT RELEASED IN SH
TSHO=X(21)
A26=TG1-TSHO
A27=(TG1**2)-(TSHO**2)
A24=(TG1**3)-(TSHO**3)
A29=(TG1**4)-(TSHO**4)
Q12SH=N2*(6.875*A26+0.3812E-3*A27-4.2336E-7*A28)
Q02SH=Q2*(7.194*A26+0.3926E-3*A27-0.1143E-7*A28)
Q00SH=C0*(8.444*A26+2.8795E-3*A27-7.197E-7*A28+0.7648E-10*A29)
QCUSH=C0*(6.865*A26+0.4012E-3*A27-0.2456E-7*A28)
QCSH=Q12SH+Q02SH+Q00SH+QCUSH
QH2GSH=(TMS1/16.0)*17.827*A26+0.674E-3*A27+2.747E-8*A28)
QSOLSH=ELUTH*0.24*A26
QRAD1=9.3*6.40*0.173*((TG1+46.0)/100.0)**4
QRAD2=8.8*7.85*0.173*((TS+56.0)/100.0)**4
QRUSH=(QRAD1-QRAD2)*0.85
QAVSH=QUSH+PH20SH+QSOLSH+QRUSH+QRADSH
QSTMSH=DELH
TSHO=X(46)*1E-0.
A30=(TSHO*50.0+460.0)/100.0)**4
A31=((TF+10.0+460.0)/100.0)**4
QRDOUT=0.0*0.85*0.173*(A30-A31)
TSHM=(TS+TSHO)/2.0
A32=((TSHM*50.0+460.0)/100.0)**4
A33=((TS+25.0+460.0)/100.0)**4
ORDUL=12.0*(4.0/12.0)*0.85*0.173*(A32-A33)
A34=((TG1-TS)-(TSHO-TS))
A35=A34/(ALOG((TG1-TS)/(TSHO-TS)))
QCONB=0.65*UO(4)*21.73*A31
QLSAT=A100*1.5
QSHOUT=QSTPSH+QRDOUT+ORDUL+QCONB+QLSAT
QRFH=(QSHOUT-QAVSH)/(FCOAL*HHV1)
C CALCULATE PERCENT HEAT RELEASED IN ECONOMIZER ZONE
A36=TSHO-TECHG
A37=(TSHO**2)-(TECHG**2)
A38=(TSHO**3)-(TECHG**3)
A39=(TSHO**4)-(TECHG**4)
Q12EC=N2*(6.875*A36+0.3812E-3*A37-4.2336E-7*A38)
Q02EC=Q2*(7.194*A36+0.3926E-3*A37-0.1143E-7*A38)
Q00EC=C0*(8.444*A36+2.8795E-3*A37-7.197E-7*A38+0.7648E-10*A39)
QCEC=C0*(6.865*A36+0.4012E-3*A37-0.2456E-7*A38)
QSEC=Q12EC+Q02EC+Q00EC+QCEC

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QH2OFC=(TMSI/18.3)*(7.827*A48+0.674E-3*A47+2.747E-8*A48)
QSOLFC=FLUTH*0.25*A36
A40=((TSHO-100.0+46.0)/100.0)**4
A41=((TF+520.0)/100.0)**4
QRDEC1=9.0*0.40*0.173*A40
QRDEC2=9.0*0.85*0.173*A41
A42=A40-A41
QRDEC=QRDEC1-QRDEC2
QAVFC=QRDEC+QH2GLC*(QSOLFC+QRDEC*(0.1+QRDOUT))*(0.15*(QRAD1-QRAD2))
QFWFC=HECON*HFH2O
A43=(TF+TECI.0)/2.0
A44=((A43+51.0+46.0)/100.0)**4
A45=A44-A33
QRDEC=0.0
A46=((TSHO-TS)-(TECNG-TS))/(ALOG((TSHO-TS)/(TECNG-TS)))
QCONFC=0.00*UO(3)*32.13*A46
QLEFC=A100*2.17
QOUTFC=QFWFC+QRDEC+QCONFC+QLEFC
QDIFC=QOUTFC-QAVFC
QREFC=0.0
C
CALCULATE HEAT RELEASED IN FURNACE EXIT ZONE
A92=(TECNG*(FLWSG*0.24*ELUTR*0.30-0.50*3.0*23.78))+3.0*23.78*TS
A93=0.5*3.0*23.78*FLWSG*0.24*ELUTR*0.30
A94=A92/A93
A48=TECNG-A94
A49=(TECNG**2)-(A47**2)
A50=(TECNG**3)-(A47**3)
A51=(TECNG**4)-(A47**4)
Q2FE=N2*(6.895*A48+0.3812E-3*A49-0.2336E-7*A50)
QO2FE=O2*(7.104*A48+0.3926E-3*A49-0.1443E-7*A50)
QCO2FE=CO2*(8.444*A48+2.875E-3*A49-1.197E-7*A50+0.7648E-10*A51)
QCOFE=CO*(6.865*A48+0.412E-3*A49-0.2456E-7*A50)
QCFE=Q2FE+QO2FE+QCO2FE+QCOFE
QH2OFC=(TMSI/18.3)*(7.827*A48+0.674E-3*A49+2.747E-8*A50)
QSOLFC=ELUTH*0.24*A48
QLFC=A100*7.0
QAVFC=QCFE+QH2OFC+QSOLFC+QLFC
QSTFC=GAVFC-QLFC
C
CALCULATE PERCENT HEAT RELEASED IN FIREBOARD
HF=397.
HLCOH=1.045*TECI-0.43*6364
A52=(FLOWF*HF)-(FLOWF*HECON)
A53=A52/HVAP
TFLOWS=FLOWF+A53
C
C
TO SUMMARIZE STEAM PRODUCTION RATES IN VARIOUS ZONES
FLOWSB=VSREP
FLOWSS=SSPL
FLOWSH=(QFDVL+QRDND+0.45*QRDOUT)/HVAP
FLOWSC=(QRDEC+QCONFC+0.45*QDOUT)/HVAP
FLOWSR=(9560.0*REDHT)/(12.0*HVAP)
IF(REDHT.GT.40.0) GO TO 6
FLOWSP=(9560.0*(40.0-REDHT))/(12.0*HVAP)
GO TO 111
6 FLOWSP=0.0
GO TO 112
111 CONTINUE
112 CONTINUE
FLOWPK=QSTFI/HVAP
FLOWSI=QLS/HVAP
FLOWSF=12.0*P/HVAP
TOTAL=FLOWP+FLOWSP+FLOWH+FLOWSC+FLOWSR+FLOWSP+FLOWSI+FLOWSF
FLOWPK
A54=T43-T41
A55=(T43*A2)-(TG1*A2)

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A56=(T43**3)-(T61**3)
A57=(T43**4)-(T61**4)
Q02FB=N2*(6.875*A54+0.3012E-3*A55-0.2336E-7*A56)
Q02FR=O2*(7.104*A54+0.3926E-3*A55-0.1643E-7*A56)
QCO2FR=C02*(R.448*A54+2.875E-3*A55-7.197E-7*A56+0.7648E-10*A57)
QCOFR=C0*(6.865*A54+0.4012E-3*A55-0.2456E-7*A56)
QGF=Q02FR+Q02FB+QCO2FR+QCOFR
QH2OF0=(THST/10.0)*(7.027*A54+0.674E-3*A55+2.747E-8*A56)
QSOLF0=ELUTR*0.2P*A54
QAVFR=QGF+QH2OF0+QSOLF0
TOTFR=TFLC0S-TOTALS
QSHFR=(TOTFR*HHV1)/12600.0
QLSFR=A101*3.27
QOUTFR=QSHFR+QLSFR
QREFR=(QOUTFR-QAVFR)/(FCOAL*HHV1)
C CALCULATE OVERALL COMBUSTION EFFICIENCY USING ZONAL APPROACH
QLSPM=A100*3.
A101=((1.33*50.)*52.)*200.
QLSD0=A101
COMPEF1=(QLSPM+QLSD0*OLFF)/(FCOAL*HHV1)
COMPEF=QREFR+QREFSP+QRESH+QREEC+QREFB+COMPEF1
COMPEF2=100.0*COMPEF
9700 WRITE(6,9700) TEST
      FORMAT(//13)
      WRITE(6,13)(COM1(I),I=1,10)
      HEAT=HHV1*FCOAL
      HCS04=0.0
      HCALC=0.0
      UNACC1=0.0
      COMPEF2=14.86.*(FBAG*CRAG*FLSOL+CFLSOL*UNACC1+FD0RM*CD0RM)
      COMPEF3=100.*(HEAT-COMBLF2-HFLUE2-HDRAN2-HCS04*HCALC)/HEAT
13  FORMAT(20X,20A4//)
      WRITE(6,14)
14  FORMAT(//47X,*SSU STREAM FLOW RATES*,//20X,*STREAM*,19X,*LB/HR*,//
1//)
      HCS04=HEATSU
      HCALC=78746.4*WCA0H
      WRITE(6,15)FCOAL,FCG03,FL0WA,FL0WR,FLLSF,THST
15  FORMAT(20X,*COAL*,21X,F6.0/20X,*LINES IONE*,16X,F6.0/20X,*COMBUST IO
IN AIR*,11X,F6.0/20X,*TRANSPORT AIR*,12X,F6.0/20X,*STACK GAS*,16X,F
26.0/20X,*TOTAL MOISTURE OUT*,7X,F6.0/)
      WRITE(6,16)
16  FORMAT(//47X,*STEAM FLOW RATES*,//20X,*LOCATION*,19X,*LB/HR*,//)
      WRITE(6,17)HLOW,FL0WSH,FL0WSS,FL0WSN,FL0WSE,FL0WSR,FL0WSP,FL0WST,
1FL0WSF,FL0WSK,TOTFR
17  FORMAT(20X,*TOTAL STEAM*,16X,F6.0/20X,*RED*,24X,F6.0/20X,*SPLASH*,
221X,F6.0/20X,*SH WALL TURES*,14X,F6.0/20X,*ECO WALL TURES*,13X,F6.
30/20X,*BED REFR*,19X,F6.0/20X,*SPLASH REFR*,16X,F6.0/20X,*RARE WAL
4L TUPES*,12X,F6.0/20X,*FR REFR*,20X,F6.0/20X,*FUR FAIT ZONE*,14X,F
56.0/20X,*FR OPEN WALL TUPES*,9X,F6.0)
      WRITE(6,18)
18  FORMAT(//47X,*PERCENT HEAT RELEASED*,//)
      WRITE(6,19)QREFR,QRESP,QRESH,QREEC,QREFB
19  FORMAT(20X,*RED*,30X,F10.5/20X,*FLASH*,27X,F10.5/20X,*SH*,31X,F10
1.5/20X,*E*,30X,F10.5/20X,*FR*,31X,F10.5)
      WRITE(6,22)COMPEF2
22  FORMAT(//20X,*COMB EFF*,16.2)
      WRITE(6,935)HHV1,COMPEF3
935  FORMAT(//20X,*HIGHER HEATING VALUE*,10X,F10.3/20X,*CHEMICAL COMP
LEFF*,15X,F6.2)
      A202=HHV1*FCOAL
      QREF1=QREFR*A202
      QRESP1=QRESP*A202
      QRESH1=QRESH*A202
      QREEC1=QREEC*A202
      QREFB1=QREFB*A202

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QLDE1=CONHFF1*A202
QNOUT=QREH1+QREFS1+QREFSH1+QREFC1+QREFB1+QLDE1
WRITE(6,8P)
88 FORMAT(//4X,*,NET SENSIBLE HEAT OUT*,///20X,*,LOCATION*,19X,*,BTU/
1HR*,///)
WRITE(6,8P)QREH1,QREFS1,QREFSH1,QREFC1,QREFB1,QLDE1,QNOUT
89 FORMAT(20X,*,FID*,24X,F10.0/20X,*,FLASHZONF*,16X,F9.2/20X,*,SH*,
125X,F9.2/20X,*,ECON*,23X,F9.2/20X,*,FF*,25X,F9.2/20X,*,DRUMHEAT
2LOSS*,13X,F9.2/20X,*,TOTAL SENSIBLE HEAT OUT*,9X,F10.0)
WRITE(6,563)001EC
563 FORMAT(20X,*,ECONOMIZER HEAT BALANCE*,F10.2//)
WRITE(6,564) HTCALC,HCS04
554 FORMAT(20X,*,HEAT OF CALCINATION*,10X,F10.2/20X,*,HEAT OF SULFATION*
1,11X,F10.2//)
DATA(1)=FCCAL
DATA(2)=FCC03
DATA(3)=TOTALRO
DATA(4)=PERMCR
DATA(5)=EXAIR
DATA(6)=THFD
DATA(7)=CASHR
DATA(8)=VSUHM
DATA(9)=VPRFEM
DATA(10)=REFGTHM
DATA(11)=REFGRAD
DATA(12)=PLATE
DATA(13)=HFRDFL
DATA(14)=FLOWC
DATA(15)=FLOWS
DATA(16)=RECYC
DATA(17)=FDRUM
DATA(18)=FRAG
DATA(19)=RPSOL
DATA(20)=CYCEFF
DATA(21)=HC(1)
DATA(22)=HC(2)
DATA(23)=HC(3)
DATA(24)=HC(4)
DATA(25)=O2SG
DATA(26)=CO2G
DATA(27)=CO2SG
DATA(28)=NOX
DATA(29)=FLS02
DATA(30)=FTUN2
DATA(31)=SMPT
DATA(32)=CHOX
DATA(33)=CSO2
DATA(34)=SPFM
DATA(35)=SHMP
DATA(36)=CAUTL
DATA(37)=CFF2
DATA(38)=CRCLY
DATA(39)=CDKUM
DATA(40)=CPAG
DATA(41)=HFREE
DATA(42)=FRCTFR
DATA(43)=TGST
DATA(44)=FLOWA
DATA(45)=FLOWR
DATA(46)=FLCWS
DATA(47)=TGST
DATA(48)=FLCTI
DATA(49)=HOC(1)
DATA(50)=HOC(2)
DATA(51)=CONVEF2
DATA(52)=HHV1

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DATA(53)=COMBFF3
DATA(54)=QREB
DATA(55)=QREB
DATA(56)=QREFSH
DATA(57)=QREEC
DATA(58)=QREEC
DATA(59)=QREIFR
DATA(60)=STGT
DATA(61)=TS
DATA(62)=TG1
DATA(63)=T42
DATA(64)=TSHO
DATA(65)=TFING
DATA(66)=TF
DATA(67)=TECNO
DATA(68)=VSPED
DATA(69)=QSIMCH
DATA(70)=TSHO
DATA(71)=QFUEC
DATA(72)=FLOWF
DATA(73)=HFH2D
DATA(74)=HVAP
CALL OPLNMS(1,INDEX,201,0)
CALL WRITMS(1,DATA,74,TEST,-1)
CALL CLOSMS(1)
STOP
END

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