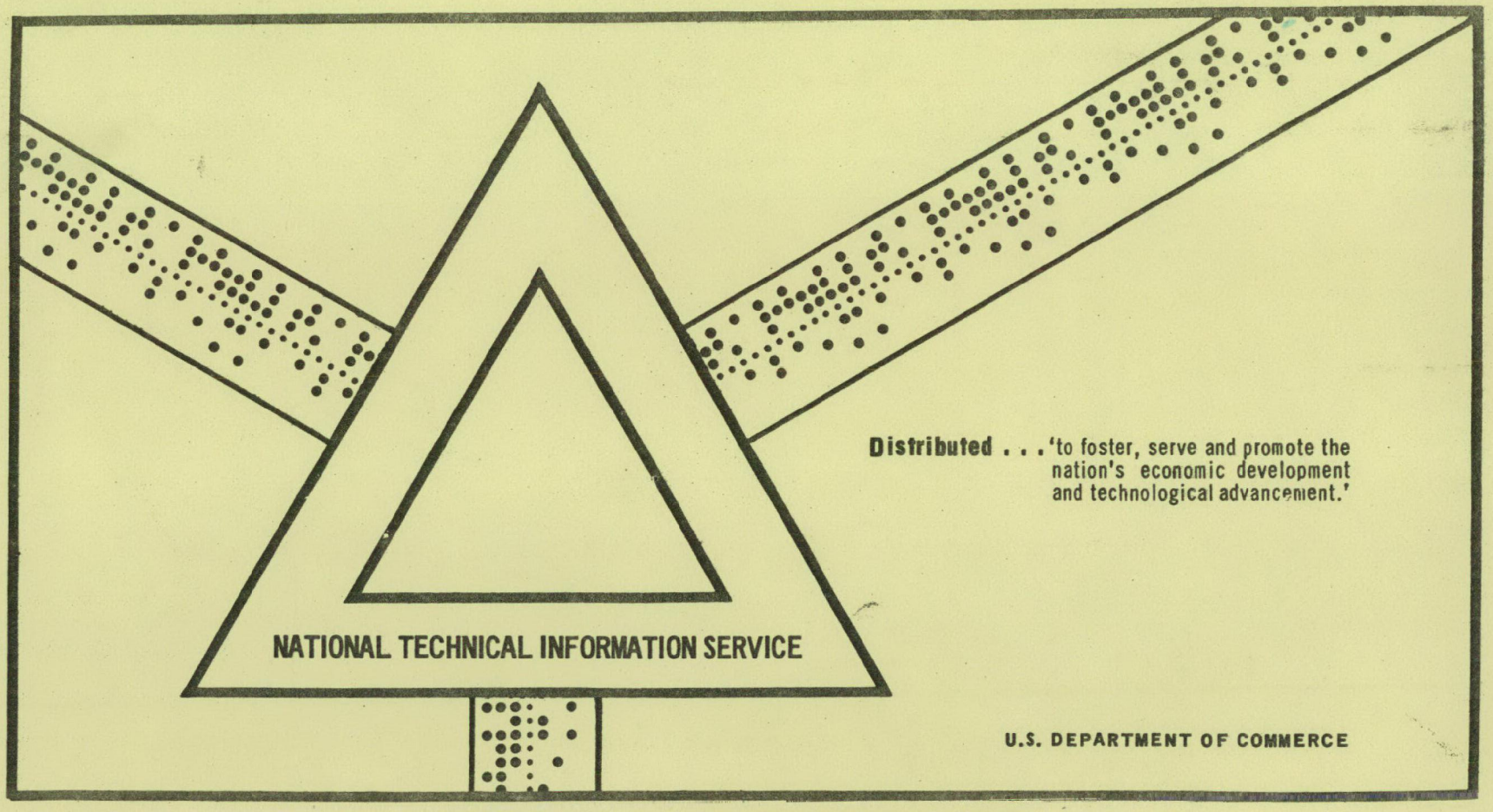
PB 198 413

CHARACTERIZATION AND CONTROL OF GASEOUS EMISSIONS FROM COAL-FIRED FLUIDIZED-BED BOILERS

E. B. Robison, et al

Pope, Evans and Robbins
Alexandria, Virginia

October 1970

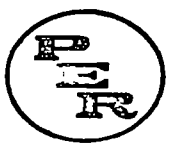


CHARACTERIZATION AND CONTROL OF GASEOUS EMISSIONS FROM COAL-FIRED FLUIDIZED-BED BOILERS

INTERIM REPORT
OCTOBER, 1970

FOR

DIVISION OF PROCESS CONTROL ENGINEERING
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
ENVIRONMENTAL HEALTH SERVICE
PUBLIC HEALTH SERVICE
DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE



POPE, EVANS AND ROBBINS
CONSULTING ENGINEERS
A DIVISION OF PERATHON INCORPORATED

INTERIM REPORT

on

CHARACTERIZATION AND CONTROL OF
GASEOUS EMISSIONS
FROM
COAL-FIRED FLUIDIZED-BED BOILERS

to

DIVISION OF PROCESS CONTROL ENGINEERING
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
ENVIRONMENTAL HEALTH SERVICE
PUBLIC HEALTH SERVICE
DEPARTMENT OF HEALTH, EDUCATION, AND WELFARE

October 1970

by

E. B. Robison, A. H. Bagnulo,
J. W. Bishop, and S. Ehrlich

POPE, EVANS AND ROBBINS
Consulting Engineers
A DIVISION OF PERATHON INCORPORATED
Alexandria, Virginia 22314

TABLE OF CONTENTS

TABLE OF CONTENTS (Continued)

LIST OF FIGURES

LIST OF FIGURES (Continued)

LIST OF FIGURES (Continued)

LIST OF TABLES

LIST OF ENCLOSURES

APPENDIX A

LIST OF ENCLOSURES. (Continued)

LIST OF ENCLOSURES. (Continued)

1. SUMMARY

Pope, Evans and Robbins, in a continuing project* sponsored by the National Air Pollution Control Administration, has monitored air pollutant emissions from the combustion of coal in a fluidized bed under a comparatively large number of different conditions. Efforts were made to reduce emissions of oxides of sulfur by the use of limestone-based sorbents and to determine the conditions most favorable for the reduction. The major test variables and ranges are summarized as follows:

Coal Type: Medium and high sulfur

Bed Temperature: 1500°F to 1900°F

Bed Depth: 6 to 20 inches

Bed Material: Sintered ash and limestone

Flue Gas Oxygen Content: 0.5 to 5%

Superficial Gas Velocity: 6 to 14 fps

Sorbent Type: A dolomite designated 1337 and a limestone designated 1359

Sorbent State: Raw, hydrated and precalcined

Sorbent Particle Size: -7 to -325 mesh

Fly-Ash Recirculation: Full Range (0% to 80%)

Method of Sorbent Feed: Pneumatic feed with the coal, pneumatic feed remote from the coal feed and premixed with the coal

The tests were conducted on both pilot-scale and full-scale test units. The pilot scale fluidized-bed combustor, designated the FBC, contained a rectangular bed 12" x 16". The full scale unit, designated the FBM, contained a rectangular bed ~20" x 72" and constituted one half cell of a full-scale multicell boiler concept. Emissions of sulfur dioxide, nitric oxide and hydrocarbons were monitored continuously with periodic samples taken for measurement of particulates and wet-test determination of SO_x and NO_x. When conditions most favorable for air pollution control were established on a pilot scale, the conditions were reproduced in tests with the Fluidized-Bed Boiler Module (FBM).

* This report describes the results of experiments carried out between November 1967 and August 1969.

1.1 TEST PROCEDURES

The FBC test program was begun with combustion of the high sulfur coal in a sintered ash bed and addition of sorbents in a particle size approximating the bed particle size (-7 +14 mesh). The 1337 dolomite and 1359 limestone*, in the raw and precalcined state, were injected at varying rates and operating conditions, i.e., bed temperature and excess air. The superficial velocity was held in the range of 12 - 14 feet per second for all tests except two designed for this parameter.

The program was continued with the use of finely divided sorbents in the sintered ash bed again with the high sulfur coal. The decision to employ a smaller sorbent particle size was based on increasing evidence that the desulfurization reaction was limited by product shell formation. Reducing the particle size increases the surface-to-mass ratio and, in turn, the sorbent reactivity.

The fine sorbents, both the dolomite and limestone, were injected in the raw and calcined states, ground to a -325 mesh particle size. A third state, the hydrate, was studied because of its natural occurrence in a -325 mesh particle size. The test procedures involved, principally, changes in bed temperature, sorbent feed rate, and ash recirculation. The excess air was held constant at a level which effected a 3% oxygen content in the flue gas, a minimum value found necessary to control hydrocarbons emission. The superficial velocity was held in the 12 - 14 fps range.

The effect of varying the method of sorbent feed was investigated. Three injection methods were studied with lime hydrate fed at rates varying over the 1 - 3 stoichiometric range. The methods are distinguished as follows:

a. Pneumatic injection of the sorbent at a single port with the coal after having been mixed with coal in the coal feed line.

b. Pneumatic injection into the fluidized bed at two

*Designations established by Bituminous Coal Research, Inc., an affiliate of the National Coal Association, as follows: 1337 - 53% calcium carbonate and 46% magnesium carbonate; 1359 - 97% calcium carbonate.

ports remote from the coal feed port.

c. Premixing the sorbent and coal in the coal hopper.

The tests were conducted at the bed temperatures found to be most favorable (1500°F - 1600°F). The excess air and superficial velocity were restricted as before. The two-port feed system was extended subsequently to four-port feed in a study of sorbent distribution in the bed.

A medium sulfur coal was tested with the dolomite and limestone sorbents in the raw state, ground to -325 mesh, and as the hydrate. Bed temperature and feed rates were varied for comparison of the response to that observed with the high sulfur coal. Excess air and superficial velocity were again held constant.

An investigation was conducted using the medium sulfur coal to determine the independent effects of sorbent particle size, bed depth and bed temperature. A cooling coil inserted in the bed provided a variable heat transfer surface for independent temperature control. Raw 1359 limestone with close cut particle sizes in the range of -325 to -12 mesh was injected into beds 10 and 18 inches deep.

The last FBC investigation involving the use of a sintered ash bed concerned the effect of reducing superficial velocity (to 6 feet/sec) and the comparative effect of sorbent injection above the bed.

The feasibility of burning coal in a fluidized bed of limestone was demonstrated. A medium sulfur coal was burned in the FBC containing a bed of 1359 limestone initially in the raw state. Operating conditions, i.e., bed temperature and excess air, were varied for the effect on emissions, sorption of sulfur in the bed, subsequent desorption, calcination and bed loss. Heat transfer measurements were made in the bed for comparison with values determined in the sintered ash bed.

Tests conducted in the full scale unit, FBM, were devoted to the use of fine sorbents with combustion of the medium and high sulfur coals in a sintered ash bed. Emissions were monitored with injection of 1337 dolomite and 1359 limestone in the raw state, ground to -325 mesh particle size, and as the hydrate. The principal variables were the sorbent feed rate and ash recirculation. The temperature was held generally in the range of 1500°F - 1600°F, and the flue gas oxygen

at 3%. The superficial velocity was held in the 12- 14 fps range for all tests. One test was conducted to ascertain a possible correlation between nitrogen content in the coal and nitrogen oxides emission.

1.2 SULFUR DIOXIDE EMISSION

1.2.1 Reduction with Coarse Sorbents

Emission of sulfur, in the form of sulfur dioxide, from the combustion of high sulfur coal in a sintered ash bed was found to vary from 90% to 95% of the input sulfur. When raw sorbents were injected into the bed in a relatively coarse particle size (-7 +14 mesh), the sulfur dioxide emission was reduced more effectively with the 1337 dolomite than with the 1359 limestone at the same Ca/S molar feed ratios. The dolomite produced a reduction of 54% at a ratio of 1.44, for a utilization of 37.4%, whereas the limestone utilization was limited to 20.7%. The tendency of the dolomite to decrepitate in the bed may have contributed to its higher reactivity. Utilization is defined as the percentage of input calcium which combines with sulfur. The magnesium contained in the dolomite was assumed to be inert.

The reduction in sulfur dioxide emission was found to improve somewhat with increase in oxygen content in the flue gas. Near reducing conditions in the bed were found to result in less effective sulfur capture. The reduction in sulfur oxides was found to be more favorable at bed temperatures of 1500°F - 1600°F than at 1800°F when using the dolomite. With the coarse limestone addition, the effect of bed temperature was not well defined although the reduction in sulfur oxide emissions improved somewhat with increase in bed temperature. Both sorbents precalcined by the supplier were found to be less effective than the raw stone under similar test conditions.

1.2.2 Reduction with Finely Divided Sorbents

a) Effect of fine grinding

FBC test results with -325 mesh sorbents indicated an improvement over the coarse sorbent performance in both sulfur dioxide reduction and sorbent utilization. The performance was markedly improved in the case of the 1359 limestone tests with the high sulfur coal; this raw limestone fed at a Ca/S ratio of 1.5 indicated an increase in utilization from 18% to 37% with the reduction in particle size. At the same stoichiometric ratio, the 1337 dolomite utilization increased from 38% to 46% when the particle size was reduced from -7 +14 mesh to -325 mesh. The tests conducted to determine the independent effects of particle size, bed

depth, and bed temperature indicated that desulfurization is strongly dependent on sorbent particle size for a medium sulfur coal. Under similar test conditions using the 1359 limestone, a 78% reduction in sulfur dioxide emission observed with a -325 mesh particle size was decreased to a 48% reduction when the particle size was increased to 100 mesh. Reductions were even less with particle sizes larger than 100 mesh.

b) Effect of hydrating and precalcining

Performance of the fine raw sorbents in terms of sulfur dioxide reduction at various Ca/S ratios was found to be about the same as the corresponding hydrate. When the hydrate of the 1337 dolomite was injected at a Ca/S ratio of 2.0 burning the high sulfur coal, the reduction in sulfur oxide emissions was 80% to 85%. The most favorable single reduction was 88%, observed at a stoichiometric ratio of 1.8 with this hydrate. Injection of the 1359 limestone hydrate produced an 80% reduction at a Ca/S ratio of 2.6. These results were found in the FBC with a 10-inch deep bed operating at 1500°F to 1600°F, 3% oxygen in the flue gas, a -325 mesh particle size and a superficial gas velocity of 12-14 fps.

The precalcined, finely divided, sorbents were found to be considerably less reactive than the raw or the hydrated sorbents.

c) Effect of sorbent type

The results indicate the dolomite to be more effective than the limestone when the stoichiometric ratio is based on the calcium fraction of the dolomite only (51% $CaCO_3$), but was less effective on a total sorbent weight basis. The limestone containing 97% $CaCO_3$ would be the more economical of the two sorbents in terms of sulfur removal per unit weight of sorbent when the cost, per ton of stone, is comparable.

d) Effect of coal S content

Percentage reduction in sulfur dioxide emission, as a function of stoichiometric feed rate, was approximately the same in the FBC for both the high sulfur coal (4.5% S) and the medium sulfur coal (2.6% S). Under the most favorable conditions, burning the 4.5% S coal and injecting the finely divided, raw, 1359 limestone, utilization was found to be 40%, 33% and 20% at Ca/S stoichiometric ratios of 1.0, 2.0 and 3.0 respectively. Comparable utilizations were indicated in tests in the larger FBM.

e) Effect of sorbent feed method

The tests to determine the most advantageous method of sorbent feed failed to point up a significant advantage for any one of the three tested, although in general the best results were observed with a "two point" feeder. This feeder provided pneumatic injection of the sorbent into the bed at two points remote from the coal feed port. Increasing the number of feed ports in the FBC to four did not improve the sulfur capture.

The sorption efficiency was considerably less when a sorbent stream directed into the bed was suddenly diverted to a feed port above the bed. These results might have been anticipated -- a fluidized bed is a good mixer; injection above a fluidized bed is similar to injection into a conventional boiler.

f) Effect of bed temperature

Tests made to determine the effect of bed temperature showed the sorbents to be more effective at the lower end of the operating range (1550°F). Sulfur dioxide reductions of 78% and 24% were observed at respective temperatures of 1550°F and 1800°F.

g) Effect of bed height

At 1500°F a reduction of 73% with a 10-inch deep bed increased to 78% with an 18-inch deep bed.

h) Effect of superficial velocity

Tests conducted with successive lowering of the superficial gas velocity but with injection of fine limestone at a constant Ca/S ratio did not show a significant improvement in sulfur control despite the decrease in velocity.

i) Effect of fly ash recirculation

Recirculation of fly ash with fine sorbent injection improved the sulfur control in some instances but the results were inconsistent.

1.2.3 Reduction with the Use of Limestone Beds

The tests conducted with a medium sulfur coal burning in a bed composed of 1359 limestone indicated that the emission of sulfur dioxide could be controlled almost completely for a period of 2 to 3 hours with the favorable sorption condition, i.e., 1550°F temperature and 3% O_2 in the flue gas. When the breakthrough of sulfur dioxide becomes significant, most of the sulfur may be driven out of the bed by increasing the bed temperature and lowering the oxygen concentration. The bed thus "regenerated" could be reused for sulfur control

by reverting to operation under the sorption conditions. During the regeneration phase, sulfur dioxide concentrations as high as 8.1% were observed, a value some 30 times the untreated gas concentration. A cyclic process for carrying out the sorption and desorption on a continuous basis was devised but remains undeveloped.

The bed remained active after two cycles of sorption and regeneration. Additional work is indicated to establish the reactivity over a number of cycles and for a number of stones. Bed attrition rates were found to be high during calcination (5% to 7% of initial calcium charge lost per hour) but lower during sorption and regeneration (3% and 4% per hour respectively).

Measurement of the overall heat transfer coefficient in the limestone bed indicated the same value (47 Btu/ft^{2}hr°F) observed in the sintered ash bed.

1.3 SULFUR TRIOXIDE EMISSION

Average values of sulfur trioxide concentrations observed in the flue gas from the process were found to be 30 to 50 ppm in a field of 3800 ppm sulfur dioxide. The sulfur trioxide invariably disappeared when a sorbent material was injected. None was observed with the limestone bed tests.

1.4 HYDROCARBONS EMISSION

The fluidized-bed combustor can be operated with as little as 5% excess air without evolution of smoke, but hydrocarbons concentration in the flue gas may be as much as 1500 ppm (methane) at this excess air level. The test data show that hydrocarbons emission is sharply dependent on oxygen content in the flue gas determined by the excess air rate. An excess air rate of 17% was necessary to burn up hydrocarbons in the FBC, while 24% was required for the FBM. These values correspond respectively to 3% and 4% oxygen in the flue gas.

The heat loss incurred by increasing the excess air from 5% to 17% is approximately 0.8% of the input energy based on a flue gas exit temperature of 400°F. The heat recovered from complete combustion of the hydrocarbons is about 0.9% of the input energy. These results indicate that operation with less than 17% excess air would not be advantageous in terms of thermal efficiency, whereas operation at 17% excess air has the obvious advantage of lower hydrocarbons emission.

The 17% excess air rate was considered minimum for the bed operation. During a few tests with reducing conditions in the bed, sufficient air was added overbed to complete hydrocarbons combustion and to result in 3% oxygen in the flue gas.

A measurable concentration of carbon monoxide does not appear in the flue gas at a 3% or higher oxygen concentration.

1.5 OXIDES OF NITROGEN EMISSION

Emission of nitric oxide from the FBC was found to increase with oxygen content in the flue gas. Values of 320 ppm at 1% oxygen increased to 440 ppm at 5% oxygen. A typical value of 380 ppm at 3% oxygen corresponds to ∿0.30 pounds NO per million Btu (MBtu) input.

The nitric oxide concentrations do not correlate with bed temperature. This would be expected since the measured values are well above those predicted by thermodynamic equilibrium. This result suggests the presence of local temperatures higher than the measured bed temperature or that the nitrogen content in the coal plays a role. Infrequently the emission may rise to 0.38 lb per MBtu with no increase in oxygen concentration.

Comparison of infrared determinations for NO and wet tests for NO_x indicate that nitric oxide (NO) is the dominant oxide of nitrogen. Oxides of nitrogen other than NO, determined by difference, were found to vary in the range of 10 to 30 ppm.

Emission of nitric oxide from the FBM was observed to be less than the level found with the FBC at the same flue gas oxygen content (3%) and temperature. The average emission of NO from the last sixteen FBM tests was 0.22 pounds per MBtu, a value equivalent to ∿275 ppm concentration. In general, nitric oxide emission was not affected by addition of sulfur control sorbents.

1.6 PARTICULATE EMISSION

Particulates passing the FBC cyclone collector represented about 10% of the fly ash input without fine sorbent addition. When the fine sorbent was added, the particulate emission rate was increased (from ∿2.0 to ∿4.0 lb/hr), but the percentage of the total input that was emitted remained the same. Most of the fine sorbent was collected in the cyclone.

Somewhat higher collection efficiencies (95%) were found with the FBM collector. One sample of the fly ash discharged to atmosphere during fine sorbent injection was analyzed for particle size. The analysis showed that 90% of the particulate emitted was smaller than 5 microns.

2. CONCLUSIONS

The results obtained from the test program thus far and the economic study led to the following conclusions:

a. Emission of sulfur dioxide from the combustion of coal can be reduced to currently acceptable levels by burning the coal in a fluidized bed and injecting finely divided limestone into the bed. A 4.5% sulfur coal can be converted to an equivalent 1.0% sulfur coal with the injection of -325 mesh 1359 raw limestone at a rate of 27 lb/100 pounds of coal, equivalent to a stoichiometric ratio of 1.9. A 2.6% sulfur coal can be converted to the 1% equivalent with addition of 10 lb/100 pounds of coal, equivalent to a stoichiometric ratio of 1.2.

b. Limestone injection equipment involves a comparatively low capital investment, approximately $220,000 for a 500,000 lb/hr boiler plant containing two 250,000 lb/hr boilers.

c. The cost of reducing sulfur dioxide emission to the equivalent 1% sulfur coal is estimated to be $.54 per ton of coal for the 2.6% sulfur coal and $1.06 per ton for the 4.5% sulfur coal with the use of -325 mesh 1359 raw limestone at the rates indicated above where limestone is available at $2.05 per ton. These are incremental costs based on the assumption that the plant is built with air pollution control in mind. Improvement in costs will depend largely on an economical method of increasing the sorbent utilization. Possibility for improvement exists in the use of limestone beds in a cyclic process or in the processing of partially reacted stone to expose the unreacted core.

d. For a once through process, grinding to a fine particle size (-325 mesh) appears necessary for the 1359 limestone which is very durable in comparison with the dolomite. Fine grinding should be beneficial with other limestones, but perhaps not necessary if the stone tends to decrepitate in the bed. The 1359 lime hydrate, which occurs naturally in a fine size, is as reactive as the finely ground raw stone but at $15.00 to $20.00 per ton is much more costly.

e. Utilization of the finely ground raw limestone for sulfur control varies in the range of 40% - 33% at stoichiometric ratios of 1 to 2. Slightly higher sorbent utilization is indicated for the 1337

dolomite if utilization is based on the calcium fraction alone. On a total weight basis, however, the dolomite removes less SO_2 per pound of stone fed than the limestone. Utilization appears to be limited by product shell formation even with the fine sorbent particles.

f. When a medium sulfur coal is burned in a bed made up entirely of -10 + 20 mesh 1359 limestone, 99% of the sulfur dioxide is captured initially. The sulfur dioxide emission rises with time; after 2 to 3 hours, the capture rate may drop to 90%. The emission would be expected to rise steadily in time until the capture rate becomes negligible.

   To maintain a high capture rate, the stone must be either replaced or regenerated. By raising the bed temperature and decreasing the oxygen concentration, 90% or more of the sulfur may be driven out of the spent stone and the stone thus regenerated. Makeup of the bed to replace attrition losses was indicated to be ~5% per hour of operation.

   During the regeneration phase, sulfur dioxide concentrations as high as 8.1% were observed--some 30 times the untreated gas concentration. The high concentration should facilitate sulfur recovery or scrubbing, if this is desired.

g. Sulfur trioxide emission is completely eliminated by limestone injection or by combustion of the coal in a limestone bed. This would permit coal-fired boilers to be designed with lower flue gas temperatures than is normally permitted when low temperature corrosion is a problem.

h. Emission of oxides of nitrogen from the FBM was found to average 0.22 pounds per MBtu input at 17% excess air. Values reported for conventional coal-fired boilers of similar capacity vary from .31 to 2.2 lb/MBtu. NO_x emission is not affected by limestone injection and is higher than predicted by thermodynamic equilibrium at the measured bed conditions. Emissions from the FBC were somewhat higher (.30 lb/MBtu). NO_x emission increases with increasing excess air and may be decreased by operating with reducing conditions in the bed.

i. Emission of hydrocarbons from the fluidized-bed combustion process can be controlled effectivley with ~24% excess air based on FBM test results. This rate is favorable in comparison to values of 40% to 50% excess air commonly employed in conventional boilers. Carbon monoxide was not detected in the flue gas with 24% excess air.

3. RECOMMENDATIONS

For continued research to improve the air pollution control capability of the fluidized-bed combustion process, the following measures are recommended:

a. Improvement in sulfur emission control without added cost necessarily implies increased sorbent utilization. The present utilization limit of 40% reflects the theoretical potential for improvement. Increasing sorbent utilization would seem to require a method of gaining access to the calcined core of the sorbent or a repeated use of the product shell area of the sorbent particle in a cyclic sorption-regeneration operation.

   The spent sorbent particle can probably be broken down by hydration because of the heat generated in the process and the naturally fine state of the product hydrate. This breakdown was observed during the test program when spent sorbent particles were dropped into water or exposed to humid air. The application to the fluidized-bed boiler would involve wetting the spent sorbent fly-ash mixture with a minimum amount of water at a point downstream from the dust collector and then reinjecting the mixture into the bed. Another technique which might give access to the core is grinding of the spent sorbent before reinjection.

b. The investigation of combustion in a limestone bed should be continued as a means of increasing the effective sorbent utilization for possible application in industrial or utility-size boilers. Optimum concentration of sulfur dioxide in the off-gas during the sorbent regeneration phase should be determined for its bearing on sulfur recovery.

c. While emissions of NO may be somewhat lower than from conventional boilers, they are still present. Therefore, methods for reducing NO emissions should be sought.

   The possibility of finding an inexpensive sorbent which acts as effectively on NO as limestone does on sulfur oxides appears remote. Unlike sulfates and sulfides, most nitrates, nitrites and nitrides are not stable at the bed-operating temperature. The systems study by Esso Research provides a valuable checklist of methods that might find application either as an in-situ control process or

on the stack gas. These include catalytic decomposition, catalytic reduction, adsorption, absorption and modification of operating conditions.

Recommended for evaluation are lowering the oxygen gradient between the bottom and the top of the bed by recirculation of flue gas, operation of the bed under slightly reducing conditions, and reduction of the oxygen partial pressure at the base of the bed by combustion of a premixed hydrocarbon fuel, such as natural gas. Reduction of NO emissions of 50% have been obtained under certain operating conditions indicating a potential for NO control via fluidized-bed combustion.

4. INTRODUCTION

Pope, Evans and Robbins in 1965 undertook a program sponsored by the Office of Coal Research, United States Department of the Interior, to develop low cost, high capacity, coal-fired boilers. In comparison with oil and gas-fired units, the conventional coal-fired boiler suffered a competitive disadvantage in higher capital cost for any given steam capacity. The primary aim of the program was to improve the economic position of coal as a boiler fuel. A report on the boiler development program will be published by the Office of Coal Research.

The problem of increasing steam capacity while reducing the capital cost (furnace size) necessitated an increase in combustion intensity, i.e., heat release per unit volume and also an increase in heat transfer rate to reduce heat transfer surface requirements at the high volumetric heat release rates. These requirements demanded a new approach in coal combustion technology. The concept of fluidized-bed combustion provided the most promising area of investigation as the basis for this new approach.

Early test results indicated that fluidized-bed combustion afforded order-of-magnitude increases in both combustion intensity and heat transfer rates. From these results it was predicted that a coal-fired, railroad transportable, multicell boiler capable of producing 250,000 pounds of steam per hour was feasible. Development and testing of a full-scale, single cell of a multicell boiler concept has been in progress since 1967.

The fluidized-bed combustion principle is illustrated in Figure 1. Crushed coal is injected into a bed of granular, inert material which is fluidized by air flowing upward through the bed. The coal particles are dispersed rapidly in the bed because of its turbulent motion and burned with oxygen supplied by the fluidizing air. Most of the ash residue accompanied by a fraction of carbon is blown out of the bed and entrained in the gas stream.

Rapid oxidation of the coal particles gives rise to combustion intensities (heat release rates) as high as 350,000 Btu per hour per cubic foot of bed volume. The high heat transfer rate permits rapid removal of heat through the walls surrounding the bed. This, in turn, permits control of bed temperature to a comparatively low 1600°F despite the rapid heat release.

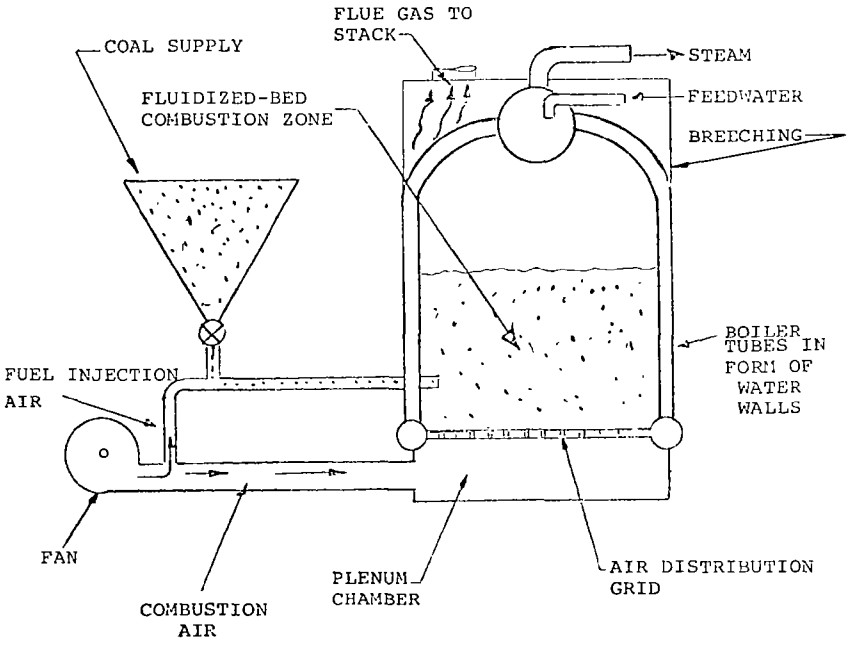


FIGURE 1. SCHEMATIC OF FLUIDIZED-BED BOILER

POPE EVANS AND ROBBINS

Measured overall heat transfer coefficients of 45 Btu/ft^{2}hr°F result in an average heat flux of ~70,000 Btu/ft^{2}hr. This flux is such that a sizable fraction of the heat released in the bed can be converted to steam energy with a relatively small amount of heat transfer surface surrounding the bed. The total heat transfer surface in the boiler can thus be markedly reduced.

The fluidized-bed combustion principle, therefore, makes it possible to decrease the capital cost of coal-fired boilers by increasing the steam capacity per unit volume of furnace. Other advantages include the fact that the coal need not be cleaned. Poor quality fuels, having high ash fractions, can be burned in a fluidized-bed combustor. In addition, the coal need not be pulverized but merely crushed. These factors would reduce both coal and coal preparation costs.

The low bed operating temperature (1600°F) should reduce boiler tube corrosion and fireside ash deposition. The uniform temperature distribution throughout the bed should reduce the possibility of tube distortion from local, high thermal stresses. The low excess air requirement showed promise of increased thermal efficiency over conventional coal-fired boilers.

Principal disadvantages are the higher fan power required because of the pressure drop across the bed and air distributors, and piping and control requirements that possibly may be more complex than those employed with conventional boilers. Operation with some coals would require makeup of bed material. The present state-of-the-art requires that the coal be single-screened to preclude buildup of large inert particles in the bed.

It appears impractical to attempt to burn coal completely in one pass through a fluidized-bed combustor. Special methods must therefore be employed to insure high levels of combustion efficiency. One such method, the Carbon-Burnup Cell, is now under intensive study. The potential disadvantage of high dust loadings and subsequent erosion from ash recirculation may also be overcome through the use of the Carbon-Burnup Cell.

The combustion principle and the performance characteristics pointed up a number of potential advantages for air pollution control. Among these was the fact that the random motion of the bed particles could provide an ideal environment for contacting limestone with

sulfur oxides in the flue gas. Sulfur emission control by injection of limestone into boiler flue passes has been demonstrated by others, but effectiveness of the method was limited apparently by formation of a sulfate shell around the injected particles which prevented further reaction. The possibility existed that the fluidized bed could provide not only the gas-solids contacting and the residence time for the desulfurization reaction but could erode a product shell and continuously expose unreacted surface. Bench scale studies of this reaction by others also indicated that the bed operating temperature range would be favorable for sulfur capture.

The low bed operating temperature was felt to be a characteristic favorable for the control of nitrogen oxides emission. Thermodynamic considerations and experience with other combustion processes indicated that nitrogen oxides emission increases with rise in flame temperature. Operating at a temperature of 1600°F, the fluidized-bed boiler was felt to have a clear advantage over conventional boiler systems which burn coal at temperatures of 2500°F and higher.

A further potential advantage, noted earlier, is the fact that the coal could be burned at near stoichiometric air rates without visible smoke in the flue gas discharged to atmosphere. This meant that smoke emission could be eliminated without loss in thermal efficiency which necessarily follows the use of excess air for smoke control.

In November 1967, the National Air Pollution Control Administration (NAPCA) of the United States Department of Health, Education, and Welfare initiated an air pollution test program through an interagency transfer of funds from NAPCA to OCR. The test program was undertaken to characterize the pollutant emissions from the combustion of coal in a fluidized bed and to assess the potential of fluidized-bed combustion for air pollution control.

The test program entailed initially the investigation of the operating variable effects on emissions and the effect of injecting sorbent materials (limestone and dolomite) into the fluidized bed of inert material. Subsequently the investigation was expanded to include the use of sorbent material as the bed material.

The variables are itemized in the discussion below.

4.1 OPERATING VARIABLES

a. Bed temperature
   1500°to 1900°F

b. Bed depth, static
   6 to 20 inches

c. Bed composition
   sintered ash and limestone

d. Air rate
   superficial velocity 6 to 14 fps

e. Fuel rate
   required to match superficial velocity 6 to 14 fps

f. Ash recirculation
   full range (0% to 80%)

4.2 ADDITIVE (SORBENT) VARIABLES

a. Sorbent type
   limestone, dolomite and a natural mine additive

b. Sorbent state
   raw, calcined and hydrated

c. Sorbent feed rate
   stoichiometric ratio 1 to 3

d. Sorbent particle size
   -7 +14 to -325 mesh

e. Method of sorbent injection

f. Water (or steam) injection

4.3 COAL COMPOSITION

a. Ash content
   7.2 and 10.7 wt. percent

b. Sulfur content
   4.5, 3.0, and 2.6 wt. percent

5. EQUIPMENT AND PROCEDURES

5.1 PILOT SCALE COMBUSTOR, FBC

Initial tests were conducted in a pilot scale combustor, designated the FBC. The FBC consisted of a rectangular combustion space, 12" x 16", enclosed by an air distribution grid at the bottom, and waterwalls around the periphery as shown in Figures 2 and 3. Air was passed into a plenum below the grid, through the grid buttons and into the combustion chamber where it fluidized the bed material and provided the combustion oxygen. Coal, crushed to pass through a 1/4" screen, was injected through a port at the base of the bed.

The air distribution grid contained a matrix of grid buttons mounted in a mild steel plate. The buttons were fabricated in stainless steel and designed to direct the air slightly downward toward the grid plate. This downward flow tended to eliminate stagnant areas around the buttons and provided cooling air for the grid plate. A cross section of a button is shown in Figure 4.

The bed material consisted generally of sintered coal ash crushed and screened to a mesh size of -7 +14. The bed was heated to coal ignition temperatures with a premix gas burner flame directed downward onto the bed as shown in Figure 2. The ignition procedure involved fluidizing the bed material with minimum air flow, raising the bed temperature to 800°F and then injecting coal until the combustion was self sustaining. About 10 minutes is required for ignition. The bed temperature was monitored with a number of thermocouples spaced vertically in the combustor. Kaowool seals were provided to prevent flue gas leakage out of the system. Specifications for the FBC are presented in Appendix A, Enclosure 1. The coal feed rate to the unit was approximately 110 lb/hr for an energy input of 1.35×10^6 Btu/hr.

The FBC test system is shown in a photograph, Figure 5, and schematically in Figure 6. Combustion products from the FBC were passed through a heavy gauge welded seam duct, through an induced draft fan, through a dust collector and on to atmosphere. The slanted configuration of the duct between the FBC and the induced draft (I.D.) fan was intended to provide gas cooling without causing wall surface temperatures to fall below the dew point of sulfur trioxide ∿360°F. This was accomplished since the flue gas temperature declines in this region from ∿1300°F to ∿800°F. The control damper provides a variable back pressure on the system to create a slightly

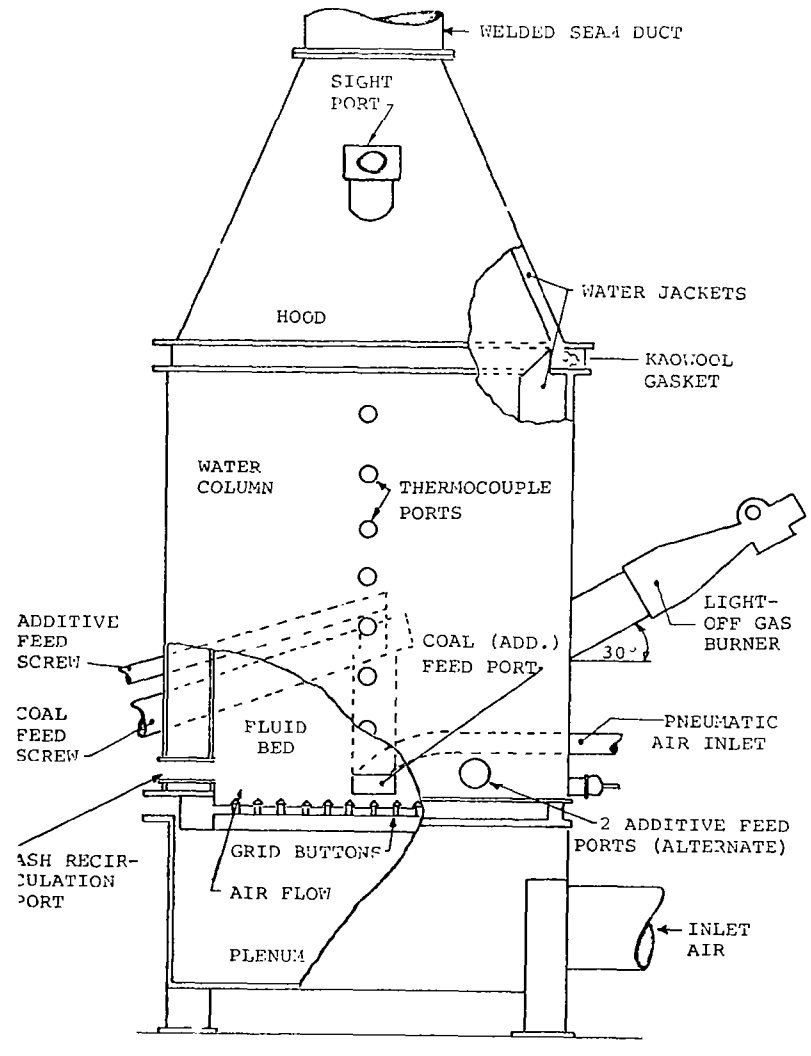


FIGURE 2. FLUIDIZED-BED COLUMN (FBC) CONSTRUCTION DETAIL - FRONT VIEW

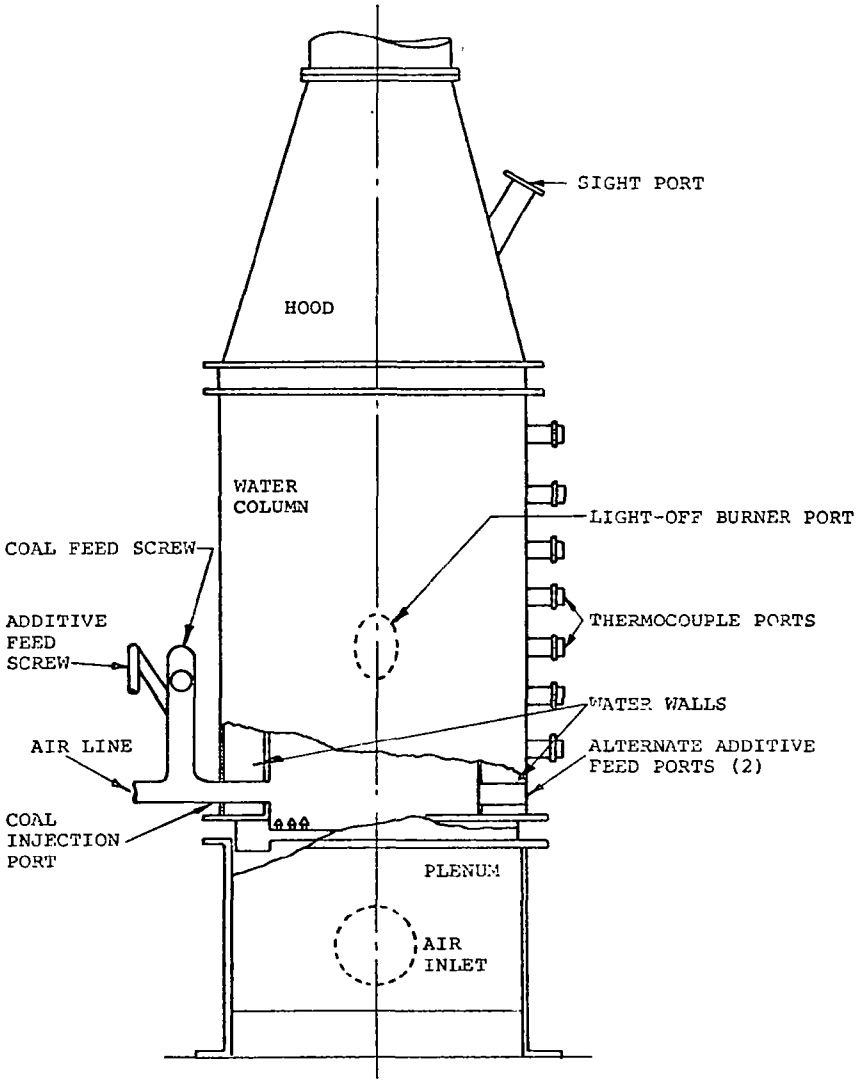


FIGURE 3. FLUIDIZED-BED COLUMN (FBC) CONSTRUCTION DETAIL - SIDE VIEW

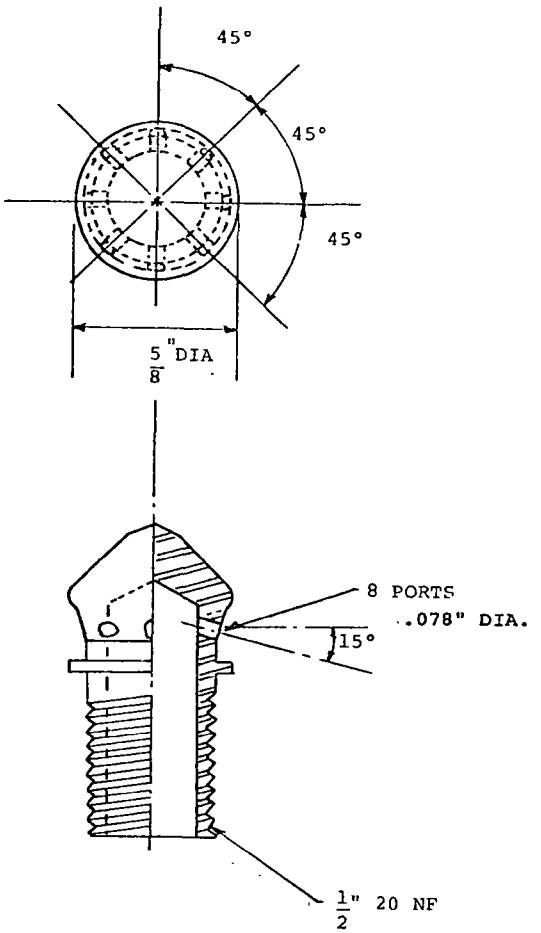


FIGURE 4. AIR DISTRIBUTION GRID BUTTON

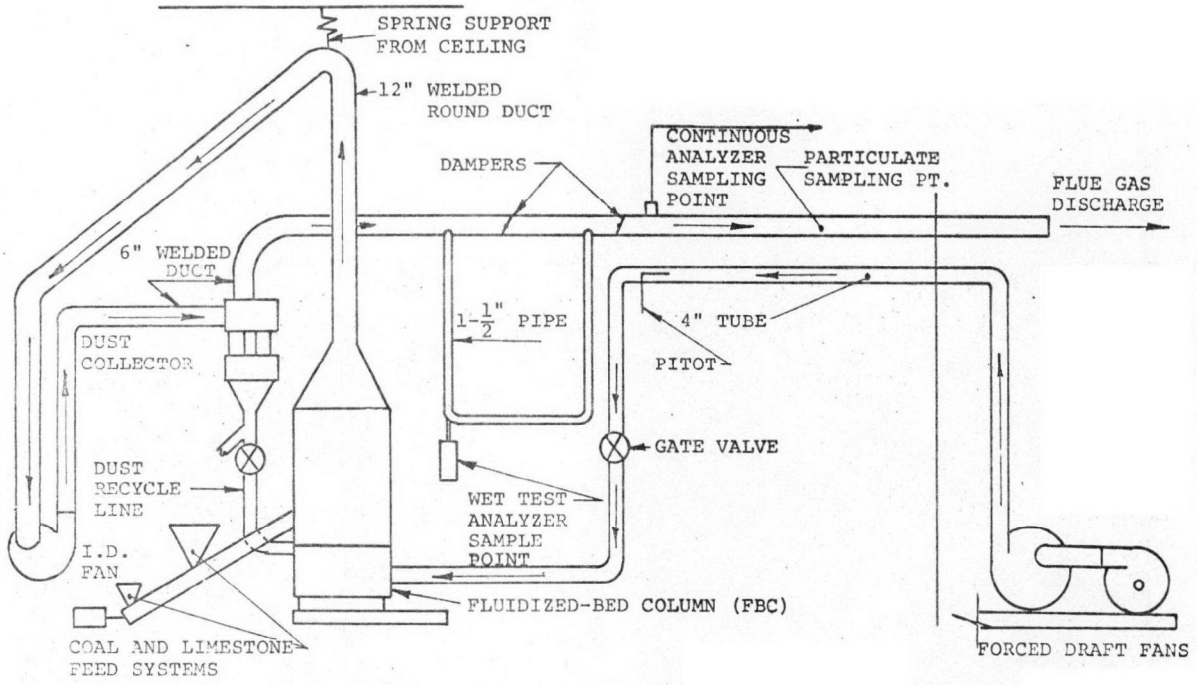


FIGURE 6. FBC AIR AND EXHAUST GAS DUCTING SHOWING SAMPLING POINTS

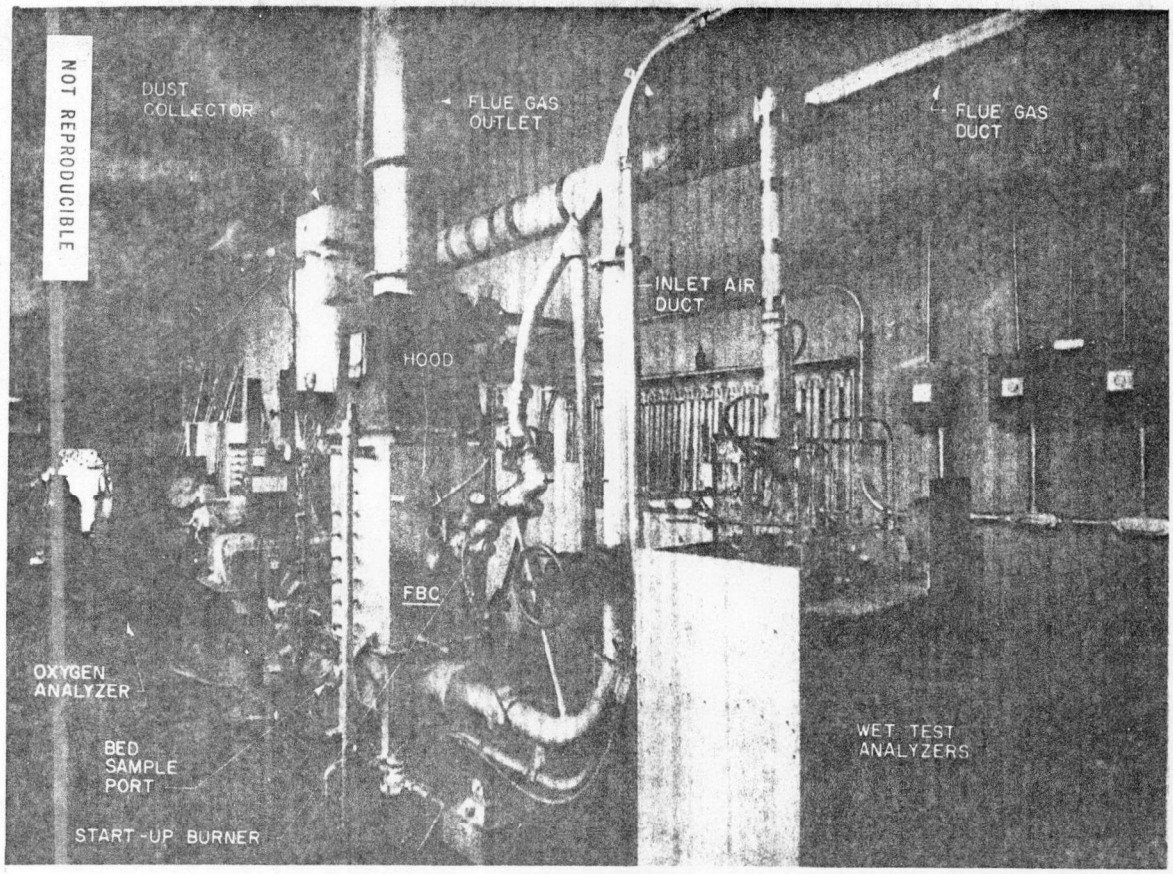


FIGURE 5. FBC TEST SYSTEM

POPE, EVANS AND ROBBINS

positive pressure on the FBC and thus prevent infiltration of air at the hood connection.

Combustion air was provided from an external fan to reduce the noise level in the test area. The air flow rate was monitored by a pitot tube in the long entrance duct and a gate valve in the line provided air flow control to the unit. The coal feed rate was controlled by a variable speed drive on the coal feed screw. Fly ash collected was discharged into bags or recirculated into the FBC as indicated in Figure 6. Locations of thermocouples are described in Section 5.4-Instrumentation.

The temperature of the bed during operation of the FBC (or FBM) depends for the most part on the bed depth which governs the total transfer surface at the water walls. This dependency created a problem in determining the separate effects of temperature and depth. The problem was solved by insulating the periphery of the bed and installing an internal cooling coil as shown in Figure 7. The bed temperature was then adjusted at various depths by raising or lowering the coil. This mode of temperature control was also used in the limestone bed tests. The temperature vs. depth tests are discussed in Section 6.3 and the limestone tests in Section 6.9.

5.2 FULL-SCALE BOILER MODULE, FBM

The full-scale boiler module, designated the FBM, is a boiler unit capable of generating steam under pressure. In this unit the fluidized bed is contained in a rectangular enclosure in which each wall is a row of vertical boiler tubes seal-welded so as to form a gas-tight enclosure. The FBM represents one half cell of the multicell, fluidized-bed packaged boiler concept developed under the OCR project. Two modules placed back to back would comprise one cell. A number of cells placed side by side without intervening insulation would make up the full-scale boiler.

A cutaway sketch of the FBM is provided in Figure 8. The fluidized-bed cross section is ~18 x 72 inches, roughly seven times the FBC cross section. The bed is surrounded by vertical water tubes which extend from the grid plate to the overhead drum. No other tubes are placed in the bed. The water tubes are joined together by a steel webbing and are backed by insulation. Flue gas from the bed passes between the tubes at the top of the unit and around the steam drum.

FIGURE 7. FBC COOLING COIL AND INSULATING SLEEVE VIEWED FROM ABOVE

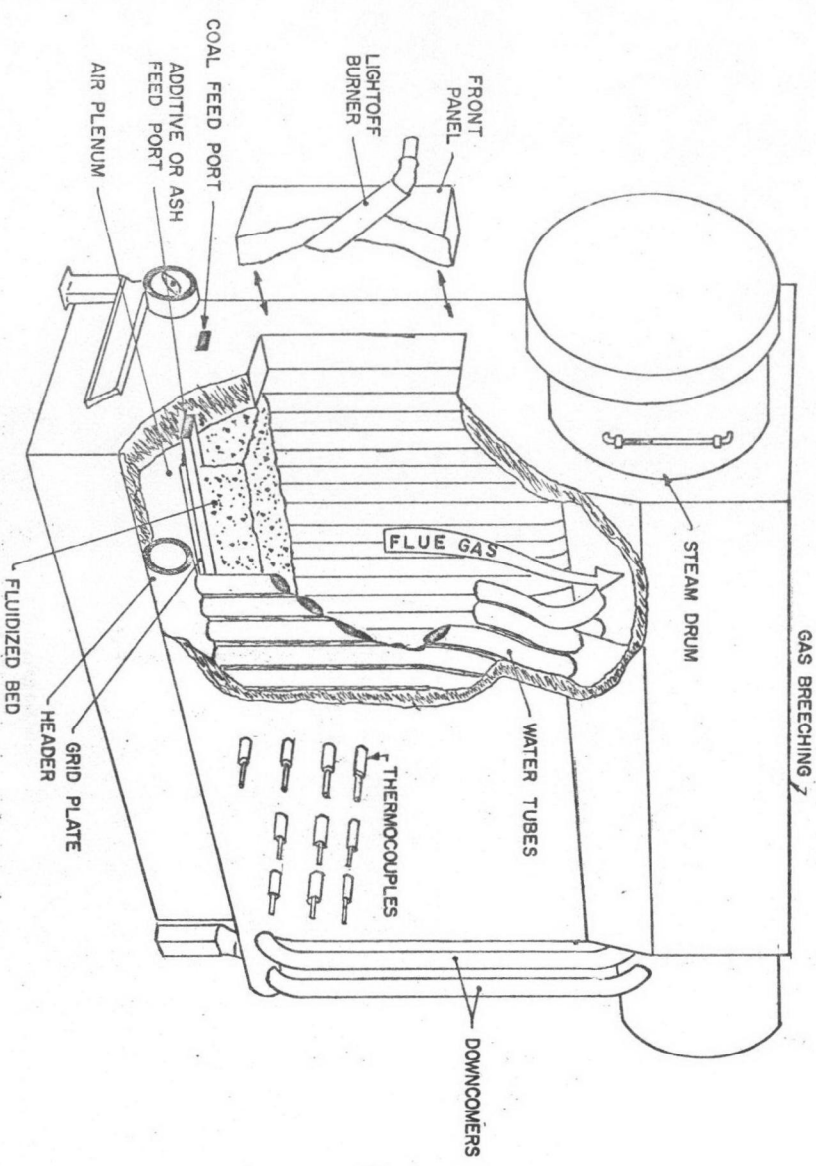


FIGURE 8. FLUIDIZED-BED MODULE (FBM) INTERNAL CONSTRUCTION

The combustion space is accessible through a water-cooled panel at the front of the unit. The panel contains a premix gas burner used to fire the bed. The burner directs a flame downward onto the front of the bed. Two pneumatic feed ports are provided below the access panel, one for the coal feed tube and the other for the additive or fly-ash feed tube. The tubes are extended into the bed area to discharge the solids at points shown in Figure 9.

From a plenum at the base of the unit, air is directed upward through a grid and into the bed area. The grid consists of a mild steel plate containing buttons of the same spacing and design used in the FBC operation. The bed material used in the FBM tests was the same -7 +14 mesh sintered ash. The static bed depth varied from 12 to 20 inches. Thermocouples were mounted throughout the bed as shown in Figure 8. Detailed specifications of the FBM are presented in Appendix A, Enclosure 2.

In operation, the bed is raised to the ignition point of coal by use of the gas burner. Combustion of the coal begins in the vicinity of the burner flame and propagates rapidly throughout the bed. Firing with a coal input of 800 lb/hr, the FBM produces 200 psig steam at the rate of 5000 lb/hr. The energy not absorbed by the waterwalls leaves this test rig as hot products of combustion. In a commercial unit, the energy of these gases would be extracted in a conventional gas-to-surface convection bank.

A schematic drawing of the FBM test system is shown in Figure 10. Air from an external forced-draft fan passes through the air preheater (or bypass) and into the FBM plenum. Coal feed is controlled by the rotation of a star feeder which drops the coal into a pneumatic feed tube at the injection port. A supply of coal is maintained automatically in a small hopper above the feeder by screw feed from a larger hopper. Sorbent materials were screw fed to the injection port at a rate controlled by a variable speed screw drive. Ash recirculation is accomplished by pneumatic transport of fly ash from the dust collector through a star feeder control.

Flue gas from the FBM is mixed with ambient air in the ducting above the unit to reduce temperature before it enters the air preheater. As the flue gas passes through the air preheater, a portion of the fly ash

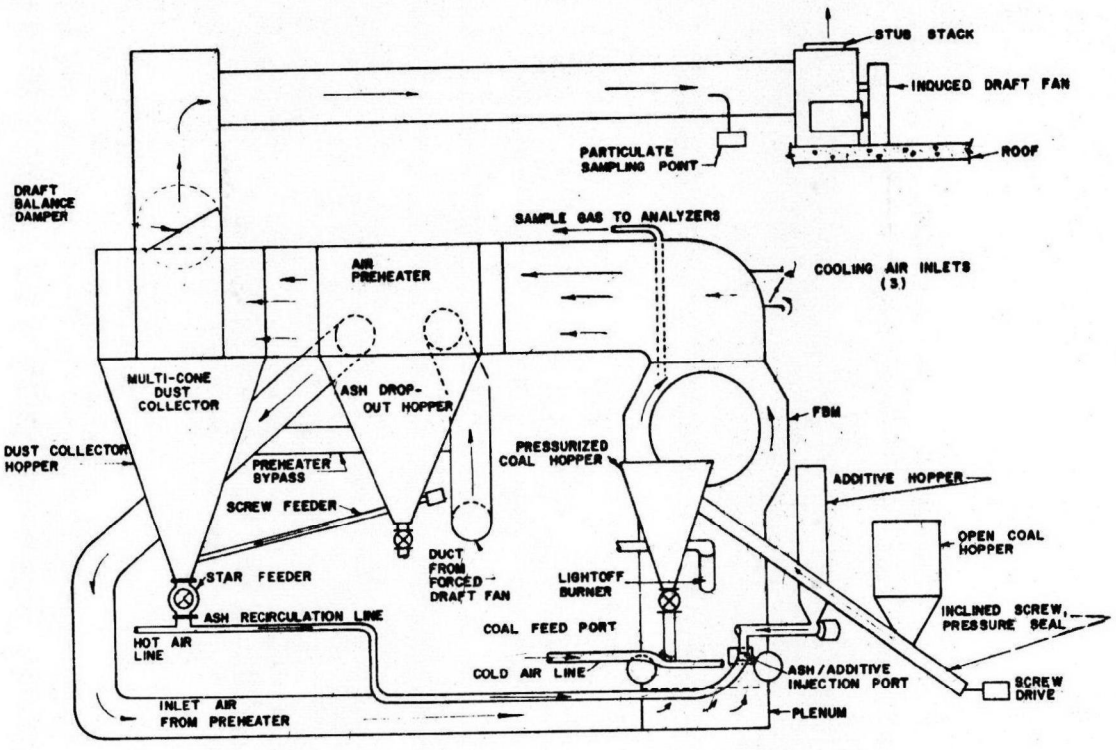


FIGURE 10. SCHEMATIC OF THE FBM TEST SYSTEM

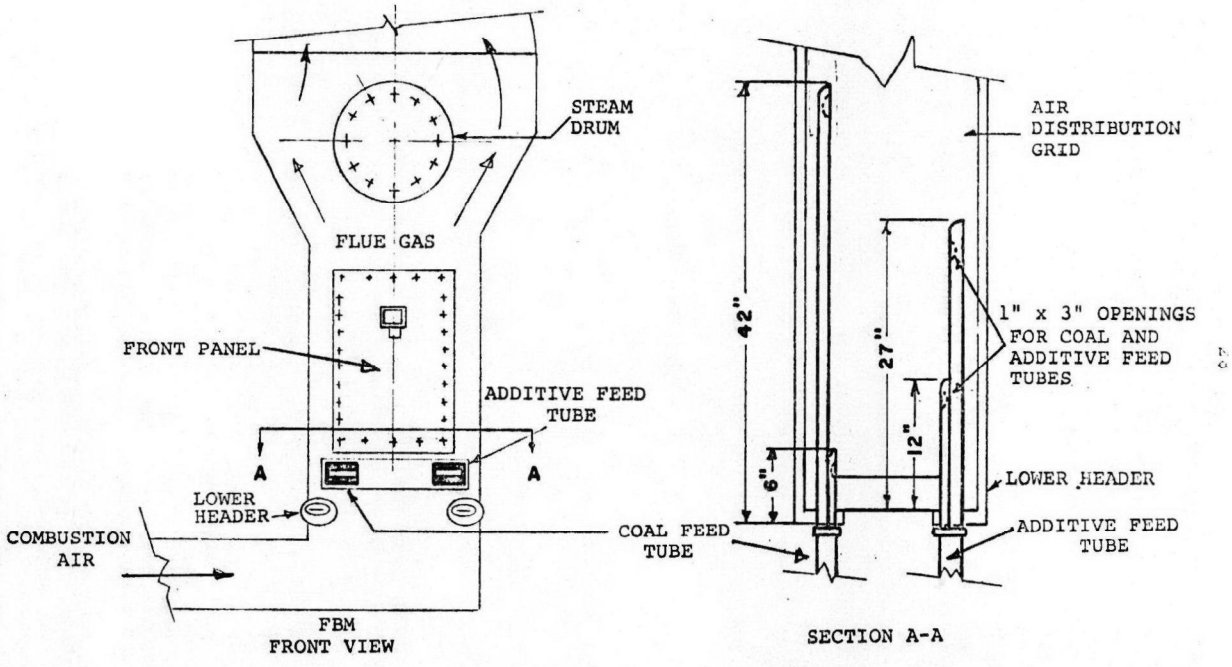


FIGURE 9. SCHEMATIC OF THE FBM SHOWING THE ARRANGEMENT OF PNEUMATIC COAL AND ADDITIVE FEED TUBES

drops out and is collected in the hopper shown. The bulk of the fly ash is retained in a multicone collector downstream. During recirculation, the ash in the first hopper is screw fed into the collector hopper. From the collector the gas flows through a long duct to an induced draft fan and then to atmosphere. A balance damper is provided in the ducting to control pressure in the combustion chamber. A photograph of the system is presented as Figure 11.

The test procedures for both the FBC and FBM operations involved igniting the bed and stabilizing the combustion at the desired bed temperature until steady-state conditions prevailed. Steady state was assumed when the Bailey Meter and the SO_2 IR analyzer indicated constant values of oxygen and sulfur dioxide in the flue gas. At steady state the sorbent feed was initiated or some other operating condition varied and the effect on emission observed. A period of 30 minutes, at least, was allowed for a new steady-state condition after an operating condition change.

The working bed depth was greater in the FBM tests than in the FBC tests because a greater depth was necessary to maintain the same ratio of cooling surface to bed plan area. With a given air rate and fuel rate the bed temperature will be approximately the same in systems having different dimensions if this ratio is kept constant. A derivation of the heat balance equation to show the relationship is presented in Appendix A, Enclosure 3.

Under equivalent conditions the bed depth requirement with change in bed cross section follows the simple relation:

$$\frac{d_2}{d_1} = \frac{l_2 w_2 (l_1 + w_1)}{l_1 w_1 (l_2 + w_2)}$$

where d, l and w are the depth, length and width respectively of the beds 1 and 2.

This relationship applies fairly well to the FBC and FBM. Let the FBC be bed 1 so that l_1 = 16", w_1 = 12" and at 1550°F bed temperature, d_1 = 12". In the FBM, bed 2, l_2 = 72", w_2 = 17.5 (the distance between tube banks). Applying the equation to predict d_2,

$$d_2 = \frac{12" (72" \times 17.5)(16" + 12")}{(16" \times 12")(72" + 17.5)} = 24.6"$$

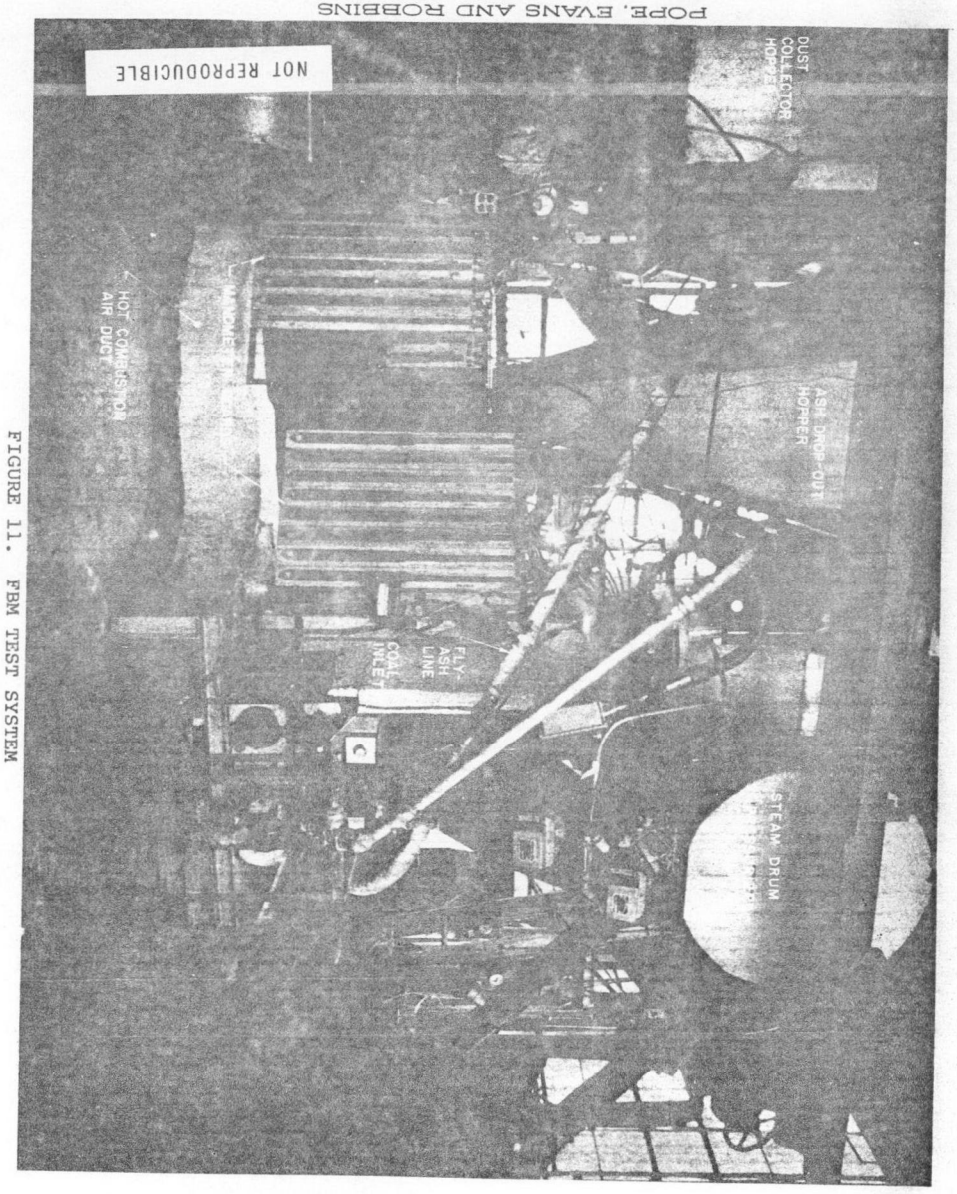


FIGURE 11. FBM TEST SYSTEM

In the FBM experiments it was found that a 24" deep bed did operate at 1550°F firing the same coal at the same rate, per unit bed plan area, as in the comparable FBC experiment. It is not suggested that the relation would apply to scale-up of very large beds.

Emissions were monitored in the FBM tests with the same procedures used in the FBC tests. Emissions were monitored without sorbent addition, with coarse sorbent addition, and with fine sorbent addition. Most of the tests were performed using fine sorbent addition, low bed temperature and a 3% oxygen concentration in the flue gas, which are the conditions found in the pilot scale to favor sulfur dioxide control. The sorbents used included both the 1337 dolomite and 1359 limestone ground to a -325 mesh particle size and the hydrated forms of these which occur naturally in a -325 mesh size. Precalcined sorbents were not tested in the FBM because of poor performance in the FBC tests. Ohio #8 Pittsburgh Seam coal, washed and unwashed, was used in the tests except for one test involving a low sulfur E. Kentucky coal.

Recirculation of fly ash was employed as a test condition by feeding the fly ash from the collector to the sorbent injection port as shown in Figure 8. The rate determined by the feeder was 80% - 90% of the input ash. In two tests, steam was injected into the inlet air at approximately 400 lb/hr.

5.3 METHOD OF SORBENT FEED

Three methods of sorbent feed were employed on the FBC during the course of the test program. The first involved screw-feeding the sorbent into the pneumatic line used to carry the coal feed into the unit. This method, pictured in Figures 12 and 13, employs a long inclined screw feeder and a variable speed drive. The assembly was designated the #1 feeder system.

A second system, designated the #2 feeder, was fabricated for injecting the sorbent at two points remote from the coal-feed injection port. The feeder system consisted of a lock hopper for the sorbent mounted on a short screw feeder as shown in Figures 14 and 15. The outlet of the feeder was connected to a pneumatic feed system which divided the sorbent flow between two injection tubes. The orientation of injection ports is shown in Appendix A, Enclosure 4.

The lock hopper in this system was necessary to counter the static pressure at the bottom of the bed. In the #1 feeder system and the coal feed system, this pressure differential is borne effectively by the inclined screw.

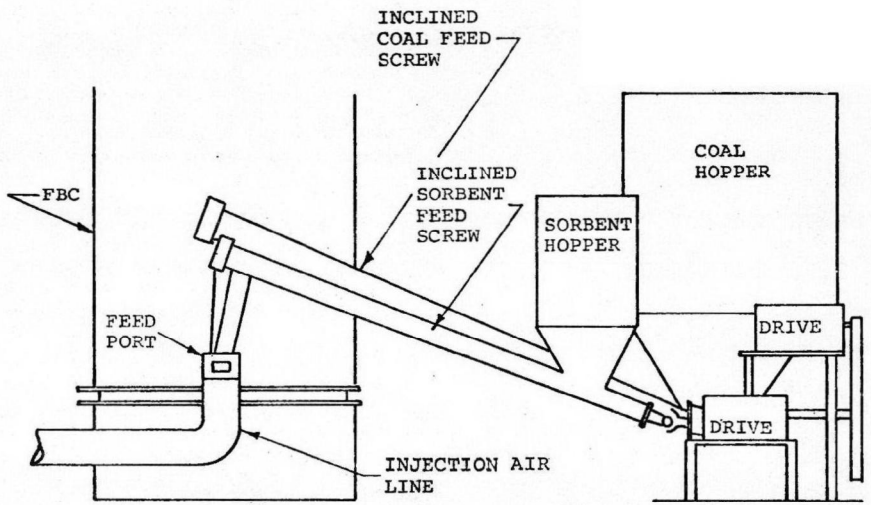


FIGURE 12. SCHEMATIC OF THE NO. 1 SORBENT FEED SYSTEM FOR THE FBC

FIGURE 13. FBC OPERATING WITH NO. 1 SORBENT FEED SYSTEM

FIGURE 14. FBC OPERATING WITH NO. 2 SORBENT FEED SYSTEM

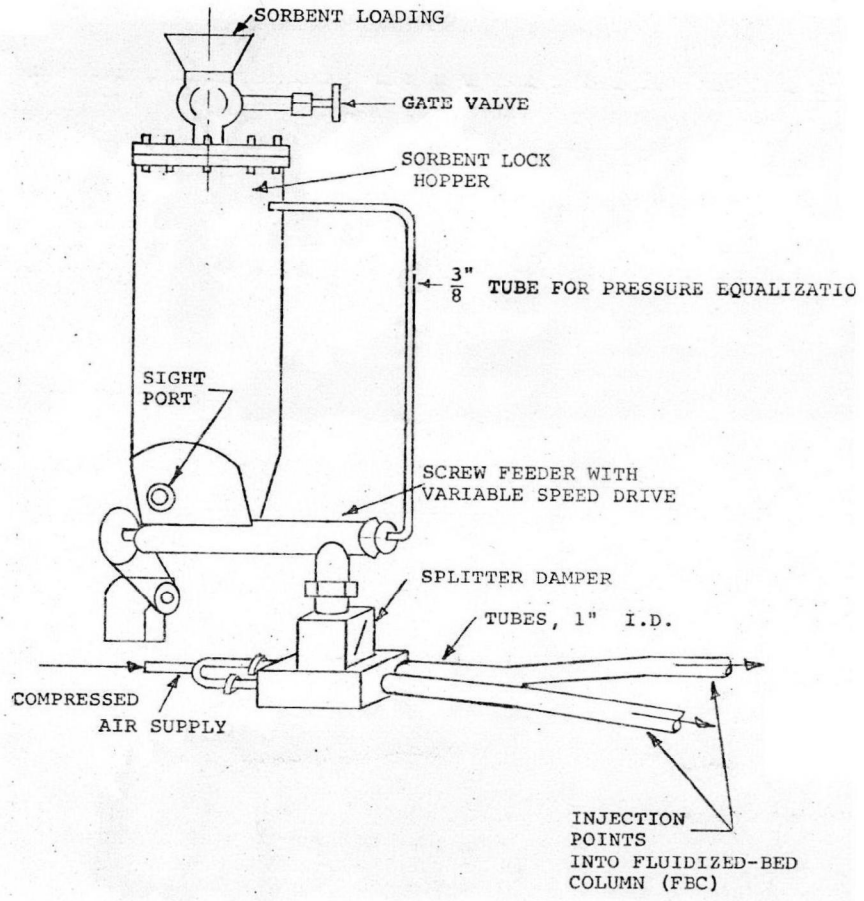


FIGURE 15. SCHEMATIC OF THE NO. 2 SORBENT FEED SYSTEM FOR THE

A third method of sorbent feed involved premixing the sorbent with the coal and feeding the mixture through the coal feed screw. A brief series of tests was conducted to determine the most effective system for sulfur dioxide control.

A combined feeder system was developed in an effort to study the effect of distribution based on the favorable results of the #2 feeder. The system consisted of a four-point injection configuration, shown in Figure 16, with provision for controlling the sorbent flow into one, two or all four sides of the FBC without change in sorbent mass flow.

An attempt to achieve the ultimate in sorbent distribution was made by injecting the fine material into the inlet air duct for distribution through the grid buttons. Although the sorbent particle size is much smaller than the button port diameter, the sorbent agglomerated and plugged the buttons rapidly.

In the FBM test series only one method of sorbent feed was used--that of feeding the sorbent into the bed at two points opposite the two coal feed ports. This feed arrangement is shown in the schematic of Figure 9 and in the photograph of Figure 17.

5.4 INSTRUMENTATION

Emissions of sulfur dioxide, nitric oxide and hydrocarbons were monitored continuously with the instrumentation pictured in Figure 18. Infrared analyzers (Beckman 215) were used to monitor sulfur dioxide and nitric oxide. Hydrocarbons were detected with a flame ionization analyzer (Beckman 109A) using methane as the reference gas. The signal output of each of these units was displayed on strip chart recorders shown at the right side of Figure 18.

The gas transfer system used with these analyzers is sketched in Figure 19. The system permitted rechecking of calibrations on any of the three units at any time during the test by switching from sample gas to reference and zero gases at the rotameter valves. The sample gas was drawn from the flue gas stream through a sintered stainless steel filter and conditioned to remove water. The sample gas was again filtered before entry into the analyzers to prevent possible contamination of the optical cells and the hydrogen burner.

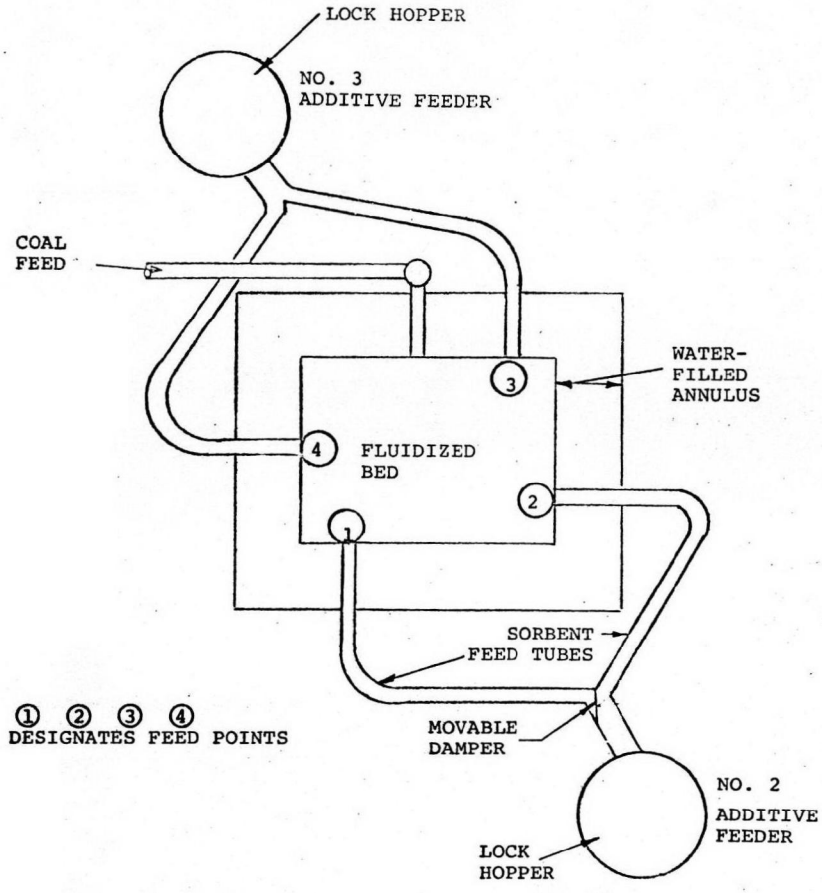
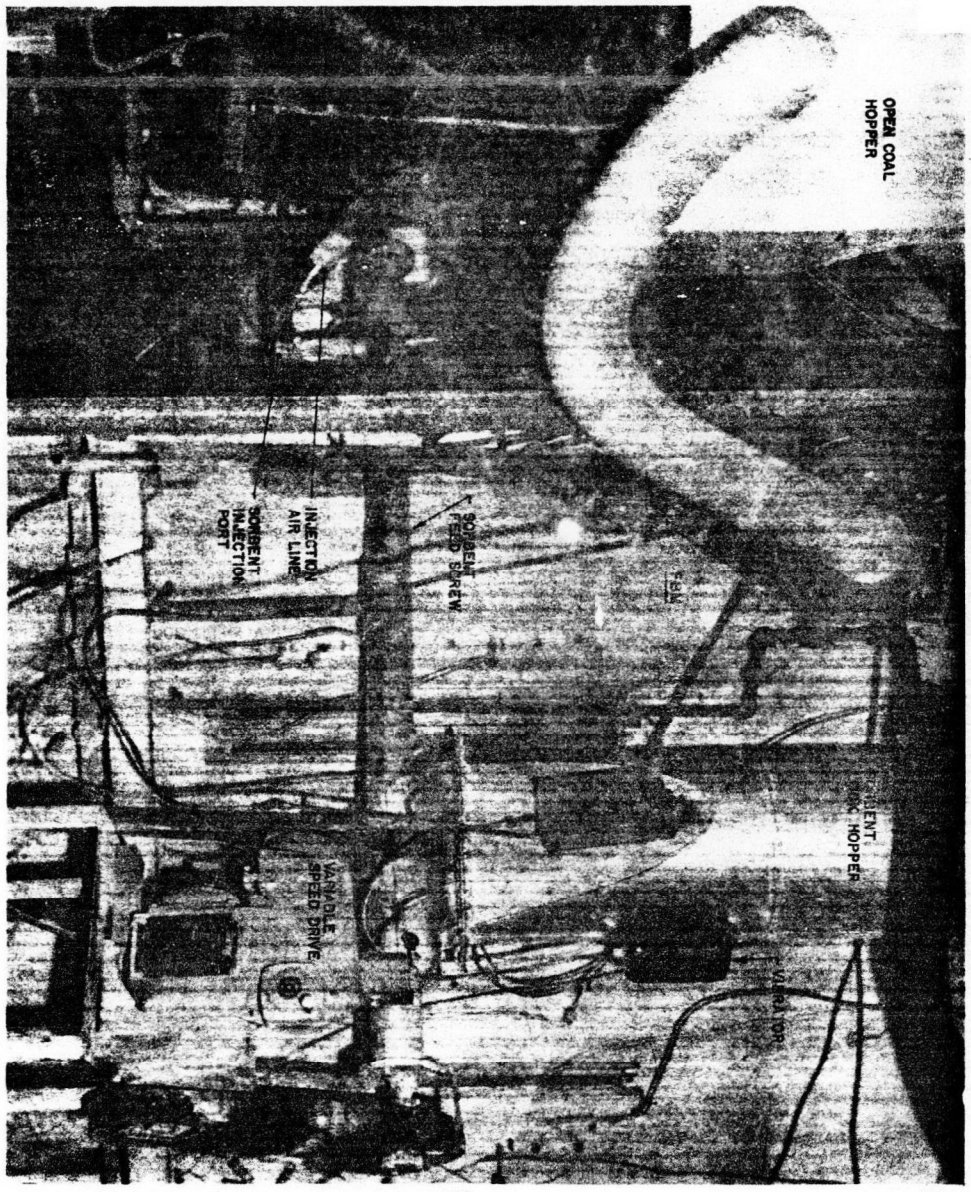


FIGURE 16. VERTICAL CROSS SECTION OF FBC SHOWING SORBENT FEEDERS AND FEED POINTS



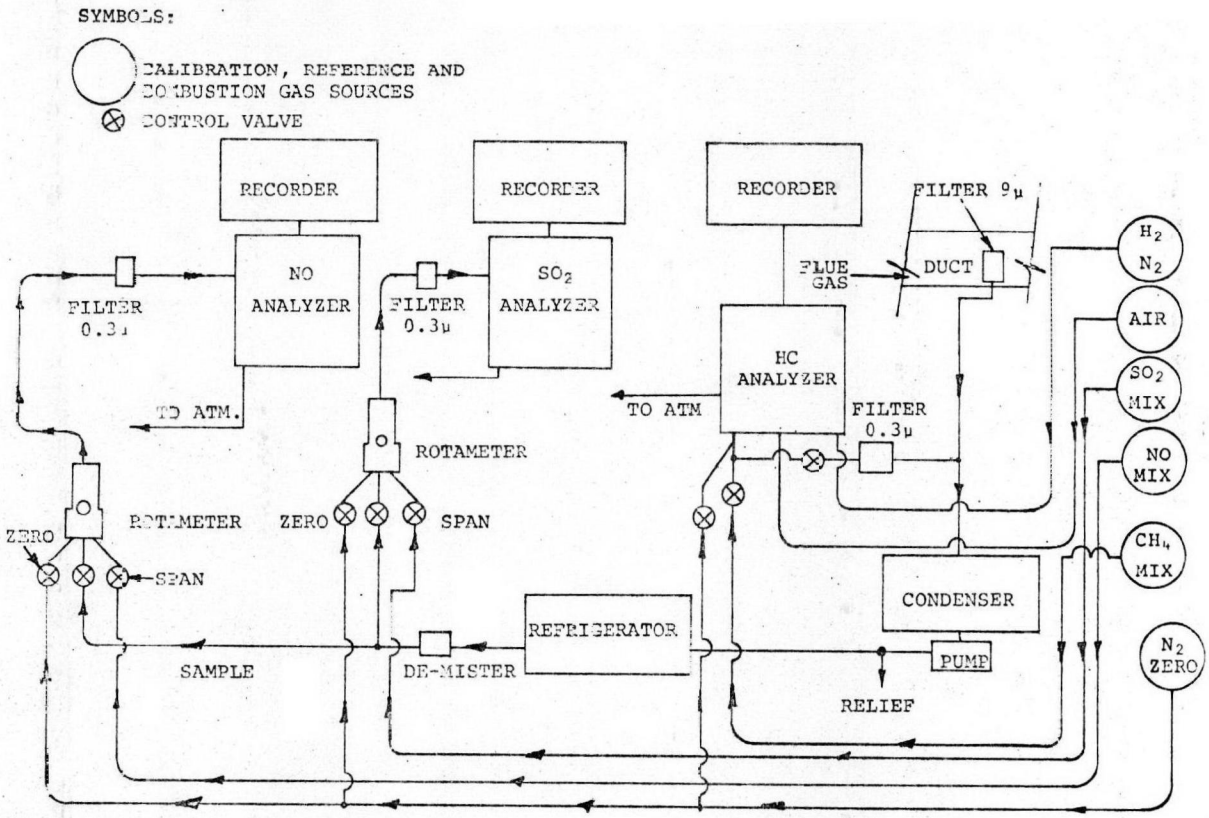


FIGURE 19. SCHEMATIC OF GAS TRANSFER SYSTEM FOR CONTINUOUS MONITORING OF SULFUR DIOXIDE, NITRIC OXIDE AND HYDROCARBONS

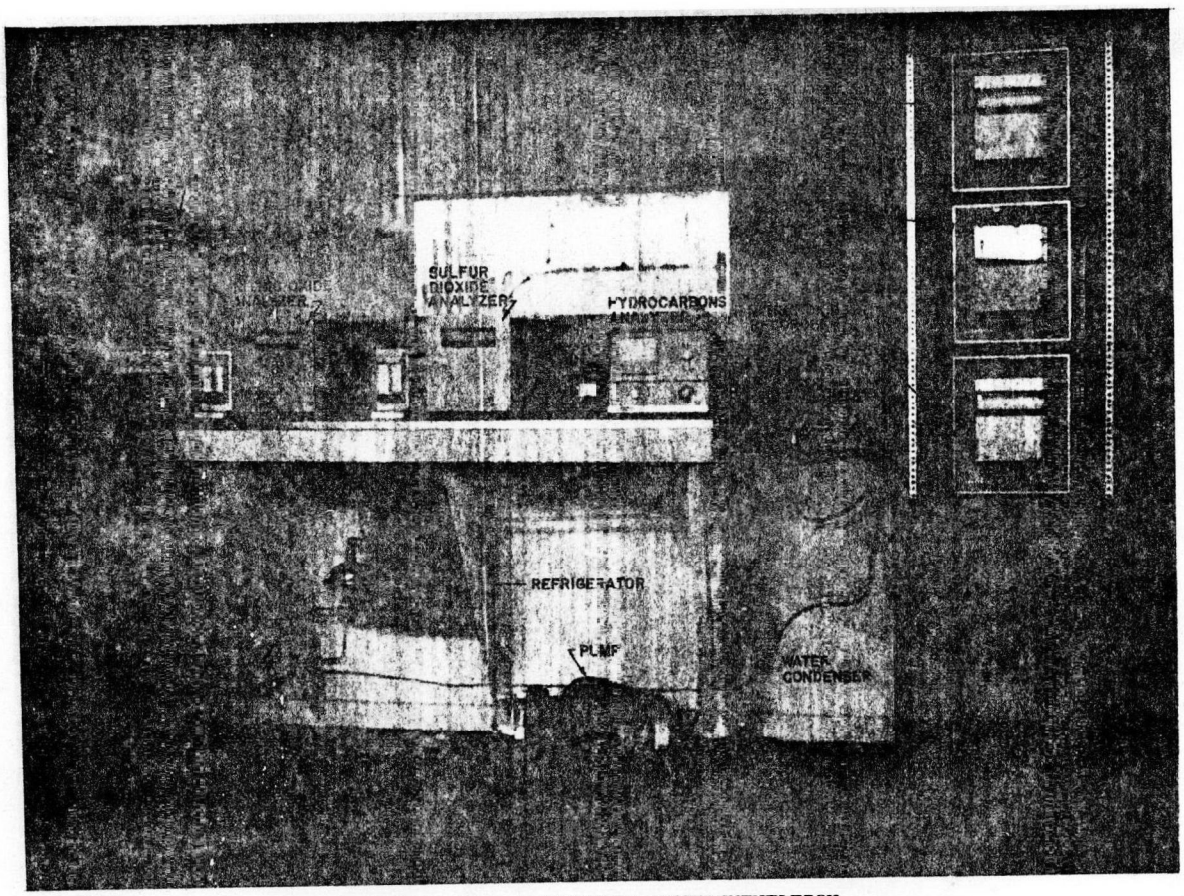


FIGURE 18. ANALYTICAL INSTRUMENTATION

The FBC gas sample was drawn into the instrument room from the FBC exhaust duct which extended overhead. In sampling the FBM flue gas, special precautions were necessary because of infiltration of dilution air in the duct above the unit and the poor instrument response which would result from drawing a small sample a long distance (~60 feet) from unit to instrument room. A system was devised to draw a large gas sample from within the FBM (at the gas passage around the steam drum), pass it through a dust collector, and then through a loop above the instrument room. The sample tube was a 3" pipe with sections screw-fitted and welded. The system was driven with an I.D. fan located at the discharge to atmosphere. A schematic drawing of the system is shown in Figure 20.

Periodic samples were taken from the flue gas to determine SO_3, SO_2, and NO_x by wet chemical analysis. The sulfur oxides analytical system consisted of a hydrogen peroxide absorption train preceded by a sulfur trioxide condenser shown in Appendix A, Enclosure 5. The sulfuric acid in each part of the system was determined by titration with barium perchlorate using thorin as the indicator. The nitrogen oxides analytical system consisted of the standard phenoldisulfonic acid procedure, using a Beckman Model B spectrophotometer for optical density measurement.

Particulate emissions were monitored with an isokinetic probe system shown in Appendix A, Enclosure 6. The probe design permits equalization of internal and external static pressures to match the sampling velocity with the stream velocity. Locations of sampling points in the FBC and FBM test systems were indicated in Figures 6 and 10 respectively.

A Bailey oxygen analyzer (Type OC1530A) was used as an operating device to indicate the oxygen concentration in the flue gas. During a test period, the air input rate was held constant and the coal rate adjusted to maintain the oxygen concentration at the desired value. The Bailey instrument was calibrated periodically with O_2, N_2 and CO_2 mixtures and found to be very reliable. The flue gas oxygen was also verified using the standard Orsat technique which determined also carbon dioxide and carbon monoxide.

Temperatures in the bed and at various other points in the system were recorded on a Honeywell Multipoint recorder. A multiple switch panel was used to connect

POPE EVANS AND ROBBINS

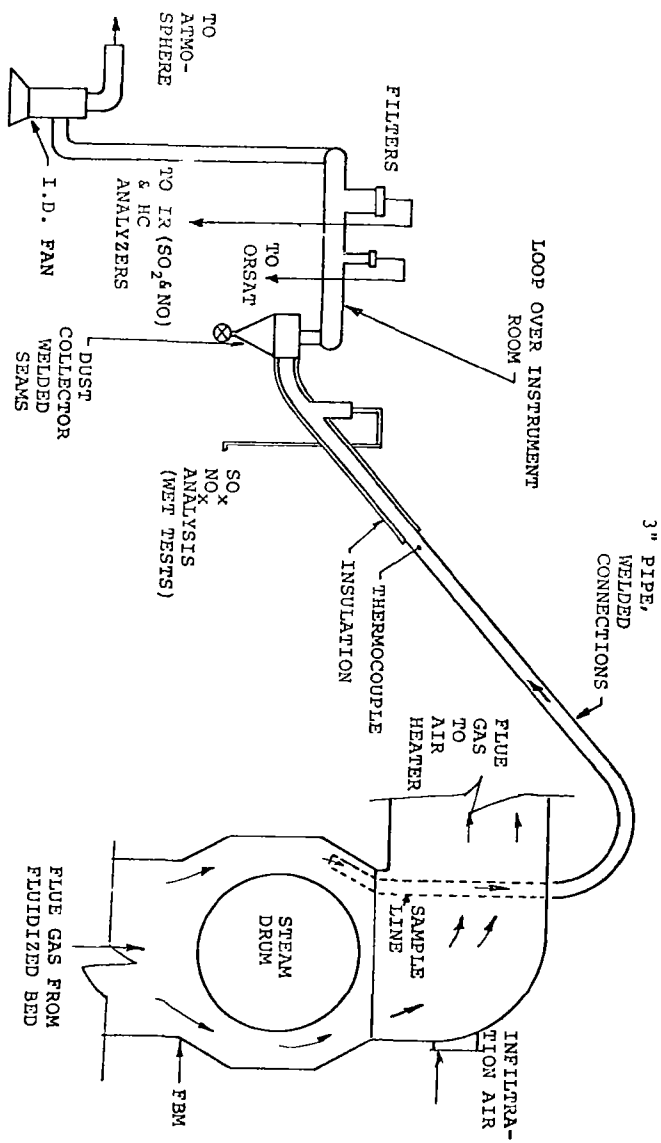


FIGURE 20. SCHEMATIC OF THE FBM GAS SAMPLING SYSTEM

the recorder input to either the FBC or FBM systems as required. Locations of thermocouples in the systems are indicated in Appendix A, Enclosures 7 and 8.

The infrared analyzers and the hydrocarbon analyzers were calibrated with gas mixtures supplied by vendors. The concentration of the active components in the calibration gases was checked after delivery to the laboratory. The methane mixture was analyzed by the National Bureau of Standards--a report is shown in Appendix A, Enclosure 9. This gas, containing 1265 ppm CH_4, was used to calibrate a second methane mixture before it was depleted.

The sulfur dioxide calibration gas was analyzed with a peroxide absorption train. Gas concentrations of 3906 ppm and 2530 ppm were used in the program. Analysis of the first calibration gas supply indicated a value of 2530 ppm as shown in Enclosure 10. Analysis of the nitric oxide calibration gas is shown in Enclosure 11.

The output signal of the infrared sulfur dioxide analyzer varies in a nonlinear manner with SO_2 concentration. The calibration curve provided with the instrument was checked by precision dilution of the known calibration gas. The curve was found to be correct except for a slight deviation at the low end of the range. The calibration curve and check points are shown in Appendix A, Enclosure 12. The calibration curve was used without correction since the deviation is not more than 1% of full scale.

The calibration curve for the nitric oxide I.R. analyzer is shown in Appendix A, Enclosure 13. The contribution of water vapor to the signal output is significant with this analyzer. The water vapor correction determined by the supplier (180 ppm) was checked by testing a dry gas in the analyzer for comparison with a moist sample. A correction of 200 ppm was noted and incorporated in the data reduction. The range of this unit is 0-1000 ppm NO.

5.5 MATERIALS

5.5.1 Coals. Two coals selected for the test program consisted of an unwashed high sulfur coal containing 4.5% sulfur and 10.7% ash, and the same coal after washing. The washed product contained 2.6% sulfur and 7.2% ash. The coal was mined from the #8 Pittsburgh seam at the Georgetown mine, Cadiz, Ohio. Proximate and ultimate analysis of each coal is shown in Appendix A, Enclosures 14 and 15. A comparatively high content of iron oxide in the ash is reported. One other coal, an East Kentucky, Pike County, low sulfur and low nitrogen coal,

was used in one test to compare the effect of nitrogen content on nitric oxide formation.

5.5.2 Sorbent Materials. Two limestone-based SO_2 control additives were studied in the raw, calcined and hydrate forms with a range of particle sizes from -7 +14 to -325 mesh. These additives consisted of a dolomite containing about 53% calcium carbonate and 46% magnesium carbonate (designated 1337) and a limestone containing 97% calcium carbonate (designated 1359). Analyses of these are given in Appendix A, Enclosure 16. The dolomite was supplied by the Dolite Company, Gibsonburg, Ohio, and the limestone by the M.J. Grove Company, Frederick, Maryland.

5.5.3 Bed Material. For the most part, the starting bed material consisted of sintered coal ash ground to a -7 +14 mesh. The sintered ash was procured from a local deposit and from the operation of the FBM in previous work. On occasion, the ash was obtained from the Anacostia power plant located nearby. Attempts to fluidize heavier bed materials such as limestone pointed up the need for special consideration.

The particle size range was selected to facilitate fluidization during the light-off procedure and also to preclude the possibility of serious elutriation losses at the operating bed temperature and superficial velocity. The flue gas velocity-particle size range for various material densities is shown in Appendix A, Enclosure 17. The particle density of the sintered ash is ~120 lb/cu ft and normal operating superficial velocity 12-14 ft/sec.

The FBC was operated successfully with a bed of the high calcium limestone (1359). Details of this effort are discussed in Section 6.9.

ENCLOSURE 1. FBC SPECIFICATIONS

1. Air Supply

Two centrifugal fans in series for 300 cfm at 30" w.g. connected to a smooth 4" diameter conduit 20' long. Air flow is controlled with a gate valve and monitored with pitot pressure, static pressure and temperature measurements.

2. Plenum

Mild Steel, 1/4" thickness, 21" x 18" x 12" outside dimensions with 8" diameter air inlet.

3. Water Column

Mild Steel, 1/4" thickness, 24" x 20" x 36" outside dimensions with 16" x 12" x 36" inside dimensions.

A. Wall on inlet air side contains:

a) One nominal 3" diameter pipe for lightoff burner

b) One nominal 1" diameter instrument port.

B. Left wall (facing air inlet) contains:

a) One nominal 2" diameter pipe with valve for removal of bed material.

b) Eight nominal 1" diameter instrument ports at various levels.

c) One nominal 1" diameter water outlet.

d) One nominal 2" diameter pressure relief port.

C. Right wall (facing air inlet) contains:

a) One rectangular 2" x 1" coal feed port.

b) One nominal 3/4" diameter cooling water inlet.

D. Wall opposite the air inlet contains:

Three nominal 1-1/2" diameter ports.

ENCLOSURE 1. (Continued)

4. Air Distribution Grid

The grid contains 130 stainless steel air distribution buttons spaced on 1-1/4" centers each containing eight drilled ports, .087" diameter. The air is discharged downward at an angle of 15° to the horizontal.

5. Water-cooled Hood

The hood is a truncated pyramid 24" x 20" at the bottom and 17" x 17" at the top with a height of 24" and a flue opening 12" diameter. Material is #10 gauge mild steel. One 4" diameter observation port is provided with 1" diameter water ports and a 2" diameter pressure relief port.

6. Flue System

From the FBC-1 hood, the flue system is run in 12" diameter #10 gauge steel pipe to the induced draft fan. From the fan the pipe is continued at 6" diameter again #10 gauge steel. All connecting sections are welded.

7. Dust Collector

The collector contains two 8" diameter centrifugal collector units with a dust hopper, rotary feeder and a valve for fly ash removal.

ENCLOSURE 2. FBM SPECIFICATIONS

1. Air Supply

   One centrifugal fan at 2500 cfm at 50" w.g. connected to 12-inch square duct which expands to full width of plenum at inlet. Air is controlled by means of a damper and monitored by an orifice.

2. Plenum

   Mild steel, ¼" thickness, 72" x 20¼" x 12" inside dimensions with a 6' x 1' air inlet.

3. Boiler Construction

   a. Single 20" steam drum
   b. Dual 6" lower headers
   c. 2½" risers on 4" centers for side walls
   d. 4" downcomers (external)
   e. 5'4" distance from grid to uninsulated bottom of steam drum
   f. Combustion space = 53 ft^3
   g. Projected heating surface = 80 ft^2
   h. Average direct contact surface = 30 ft^2
   i. Boiler capacity = 5000 lbs/hr excluding convection heat transfer; 7000 lbs/hr including convection heat transfer
   j. 8.75 ft^2 of bed area
   k. Heat release rate: 800,000 to 1,200,000 Btu/ft^2hr
   l. Pressure rating: 300 psi design, 200 psi normal operating

ENCLOSURE 2. (Continued)

4. Air Distribution Grid

   The grid contains 815 stainless steel air distribution buttons spaced on 1¼" centers, each containing eight drilled ports, .087" diameter. The air is discharged downward at an angle of 15° to the horizontal.

5. The flue system is fitted with three air infiltrators for temperature quenching, and a two-pass, 104 tube (1" x 6'), 600° air preheater; this is followed by a dust collector, which exits to a 16" duct. The system is drawn by a 4000 cfm, 5" w.g. static pressure, induced draft fan.

6. Dust Collector and Fly Ash Reinjection

   The dust collector contains twelve 10-inch diameter centrifugal collector units with a dust hopper, a 4" Allen-Sherman-Hoff rotary feeder for fly-ash reinjection and a valve for fly-ash removal.

7. Coal Input

   700 - 900 lbs per hour

ENCLOSURE 3. DERIVATION OF THE BED DEPTH RELATIONSHIP FROM BALANCE OF HEAT AND MASS

Heat balance on the system may be expressed as follows:

(1) $K_1 G_o \Delta H = G_o Cp_m (T_B - T_o) + hA_s (T_B - T_w)$

Heat release = flue gas loss + heat removed.

where G_o = mass flow of coal and air through the system lbs/hr

ΔH = heat content of fuel BTU/lb

K = constant

Cp_m = mean heat capacity of the flue gas

T_B = bed temperature °F

T_o = reference temperature °F

h = radiant plus convective heat transfer coefficient BTU/hr ft^2 °F

A_s = effective bed cooling surface ft^2

T_W = cooling surface wall temperature °F

Equation (1) may be restated as follows:

(2) $K_2 u A_c \rho \Delta H = u A_c \rho Cp_m (T_B - T_o) + hA_s (T_B - T_w)$

where u = superficial velocity ft/sec

A_c = bed cross section ft^2

ρ = gas density lbs/ft^3

by dividing (2) by the 1st term and rearranging

(3) $$\frac{hA_s (T_B - T_w)}{K_2 u A_c \rho \Delta H} = 1 - \frac{Cp_m (T_B - T_o)}{K_2 \Delta H}$$

The first term is the fraction of the total heat which is removed from the bed and is seen to be constant for fuel type and bed temperature.

ENCLOSURE 3. (Continued)

If u, h and T_B are to be constant for two systems of different size, then the ratio of cooling surface to bed cross-sectional area must be constant (A_s/A_c). The effect on bed depth is seen from the area ratio:

(4) $$\frac{A_s}{A_c} = \frac{d(2l + 2w)}{lw} = \text{constant}$$

for beds of varying dimension

(5) $$\frac{d_1}{d_2} = \frac{l_1 w_1 (l_2 + w_2)}{l_2 w_2 (l_1 + w_1)}$$

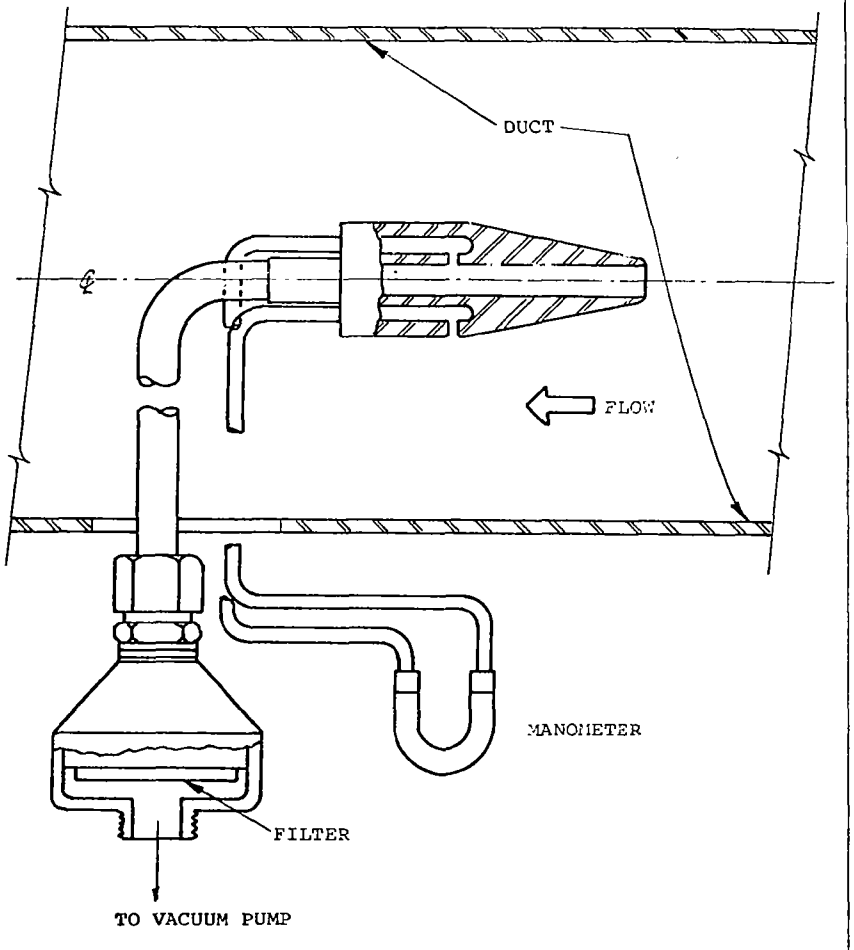
where d, l, and w are the respective depth, length and width of the two systems

This analysis assumes that the effective bed cooling surface is proportional to the bed depth. This is not strictly true, since radiation losses in a vertical direction from the bed are independent of bed depth. Another source of error is that use of the linear dimensions, l and w, does not account for the additional heat transfer surface of the round tubes which actually make up the walls of the FBM. At minimum bed temperature the respective bed depths are 10-12 inches and 20-24 inches.

VIEW THROUGH A-A

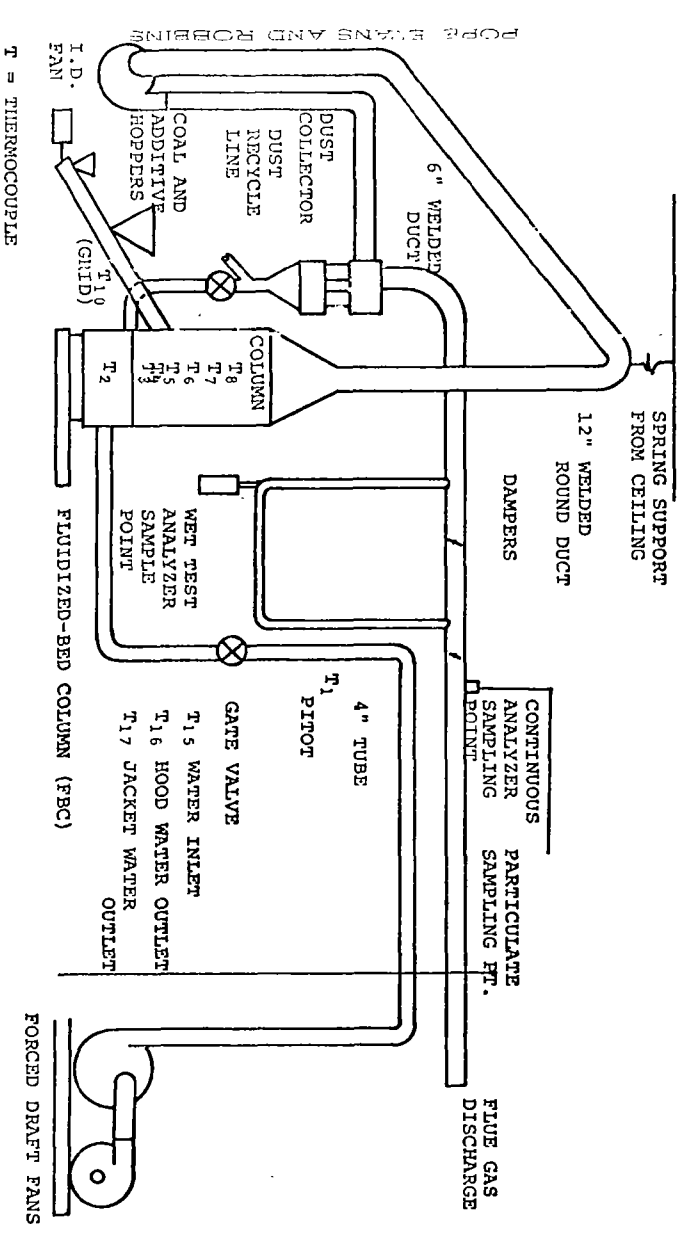
ENCLOSURE 4. SCHEMATIC CROSS SECTION OF THE #2 SORBENT FEED SYSTEM

ENCLOSURE 5. SCHEMATIC OF SULFUR TRIOXIDE CONDENSER

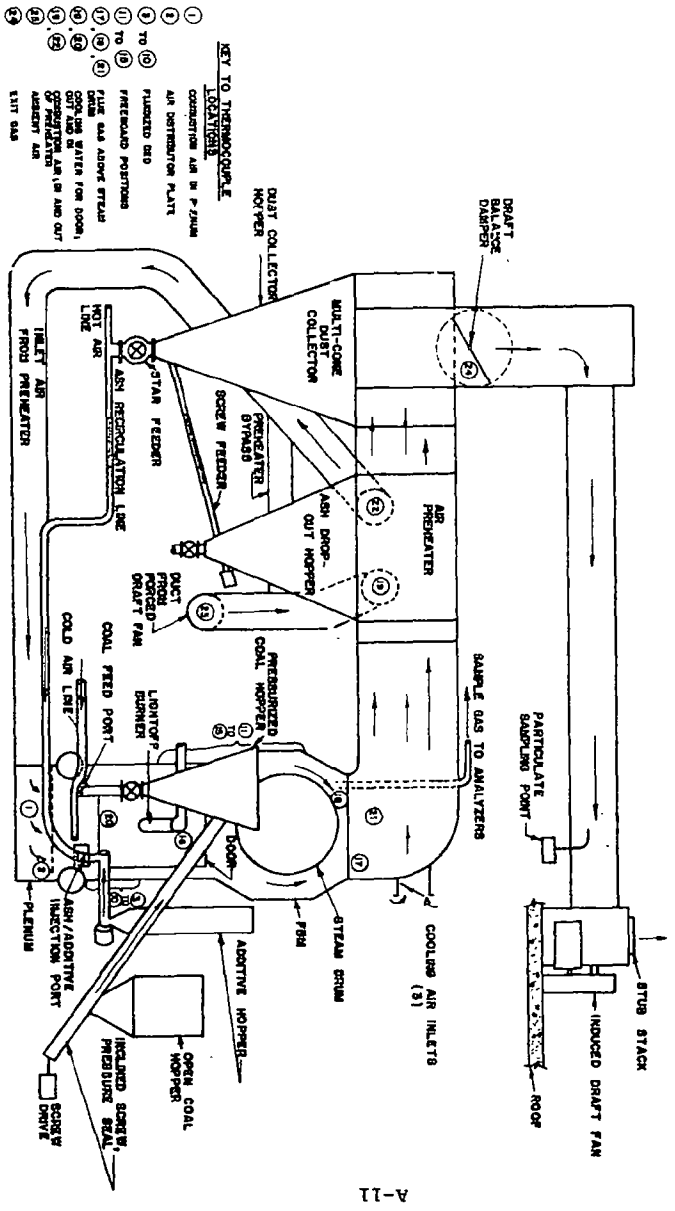


ENCLOSURE 6. ISOKINETIC PROBE FOR PARTICULATE EMISSIONS DETERMINATIONS

POPE EVANS AND ROBBINS



ENCLOSURE 7. SCHEMATIC ARRANGEMENT OF FBC AIR AND FLUE GAS DUCTING SHOWING GAS SAMPLING POINTS AND THERMOCOUPLE LOCATIONS



ENCLOSURE 8. SCHEMATIC ARRANGEMENT OF THE FBM TEST SYSTEM SHOWING GAS SAMPLING POINTS AND THERMOCOUPLE LOCATIONS

COPY OF LETTER

ENCLOSURE 9. ANALYSIS OF HYDROCARBONS ANALYZER CALIBRATION GAS

U.S. DEPARTMENT OF COMMERCE
National Bureau of Standards
Washington, D.C. 20234

REPORT OF ANALYSIS

Analysis of Dry Nitrogen for Methane

Requested by

Pope, Evans and Robbins
Attn: E. B. Robison
515 Wythe Street
Alexandria, Virginia 22314

Cylinder No. 42845T, containing dry nitrogen and methane, was analyzed for methane using a flame ionization detector. The instrument responses from standards containing 3241, 1032, 980, and 196 ppm methane were compared with the response from cyl. no. 42845T in order to determine its methane concentration. The concentration of methane in cyl. no. 42845T was found to be 1265 ± 16 ppm methane (based on ten comparisons with the standards).

(Signed)
Julian M. Ives, Chemist

(Signed)
John K. Taylor, Chief
Microchemical Analysis Section
Analytical Chemistry Division

October 4, 1968

69-79

Test #196806

POPE EVANS AND ROBBINS

ENCLOSURE 10. SULFUR DIOXIDE CALIBRATION GAS ANALYSIS

The gas was passed slowly thru two absorption columns each containing a solution of 1.5% H_2O_2 freshly prepared. Sample volume was determined by change in pressure in a tank of known volume and corrected for standard condition. The solutions were then boiled to remove peroxide and titrated with 1/10 N sodium hydroxide. Four tests were made with the following results:

Test No.	Tank Volume, liters	Pressure, mmHg Initial	Pressure, mmHg Final	ΔP, mmHg	Gas Temp., °F	Sample Volume corrected to STP liters
1	34.71	113	315	202	70	9.24
2	34.71	315	535	220	70	10.05
3	34.71	535	725	190	70	8.69
4	34.71	130	345	215	70	9.85

Titration of the solutions with 0.0985N sodium hydroxide yielded the following data and computed results:

Test No.	NaOH Solution Volume, ml. Scrubber #1	Scrubber #2	Blank	SO_2 Concentration ppm
1	19.10	0.70	0.0	2530
2	20.70	0.76	0.0	2530
3	17.50	0.65	0.0	2520
4	20.70	0.80	0.0	2540
			Avg	2530

The sodium hydroxide solution was standardized against a potassium acid phthalate solution.

Concentrations were computed from the relation:

$$SO_2 \text{ ppm} = \frac{12.05\ (V)N}{V_s}$$

where V = titer volume ml corrected for blank
N = titer normality (Equivalents/liter)
V_s = sample volume at STP-liters

ENCLOSURE 11. NITRIC OXIDE CALIBRATION GAS ANALYSIS

Analysis of the nitric oxide calibration gas was made with the Phenol-disulfonic acid procedure. Four gas samples were taken in flasks containing a small quantity of H_2O_2 solution and allowed to stand overnight. The solutions were processed as required and the absorbances of the final solution read on a Beckman Model B spectrophotmeter. The concentrations were determined from a calibration curve prepared by similar treatment of KNO_3.

The following data were taken during the tests:

Sample No.	Flask Volume, ml	Pressure, mmHg Initial	Pressure, mmHg Final	ΔP, mmHg	Temp., °F	Volume at STP (70°F) ml
1	1970	32	763	731	70	1900
2	1972	31	759	728	72	1885
3	1969	35	759	724	75	1860
4	1972	33	758	725	75	1870

Sample No.	Absorbance %	Equivalent mg NO_2	Concentration ppm NO
1	25.0	2.83	780
2	25.0	2.83	786
3	24.7	2.77	780
4	24.7	2.77	776
		Avg	780.5

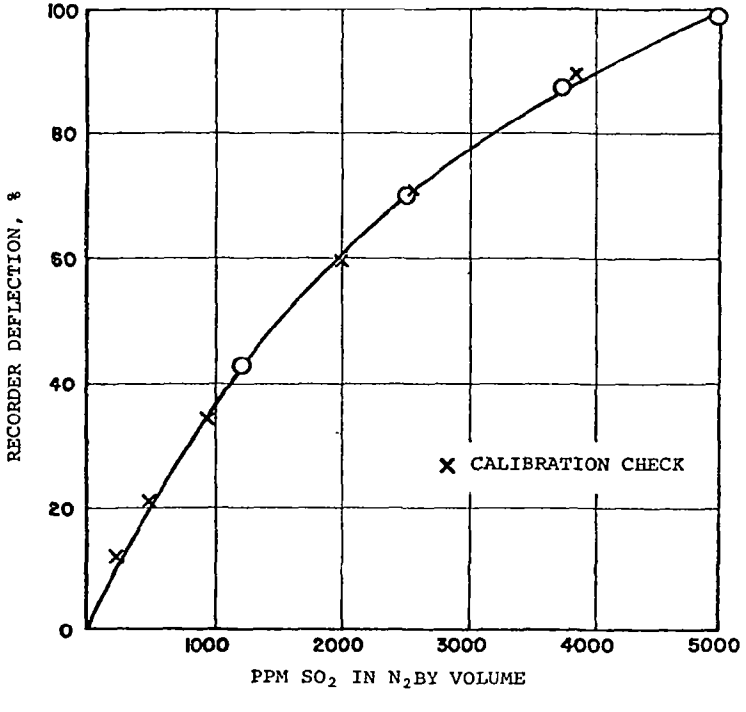
The concentration of NO was determined from the relation:

$$\text{ppm } NO_2 \text{ (or NO)} = \frac{(5.24 \times 10^5)(C)}{V_s}$$

where C = concentration of NO_2 mg
V_s = gas sample volume at 70°F and 760 mmHg

BECKMAN INSTRUMENTS, INC.
MODEL NO. 215 A

APPLICATION: SULFUR DIOXIDE, RANGE: 0-5000 PPM BY VOLUME
AMPLIFIER NO.: 200065, DETECTOR NO.: 1243 A
ZERO GAS: NITROGEN, CALIBRATION PRESSURE: ATMOSPHERIC
SPAN GAS: ANALYZED CYLINDER ~2650 PPM SO_2
CALIBRATION PRESSURE: ATMOSPHERIC
SAMPLE PRESSURE: ATMOSPHERIC

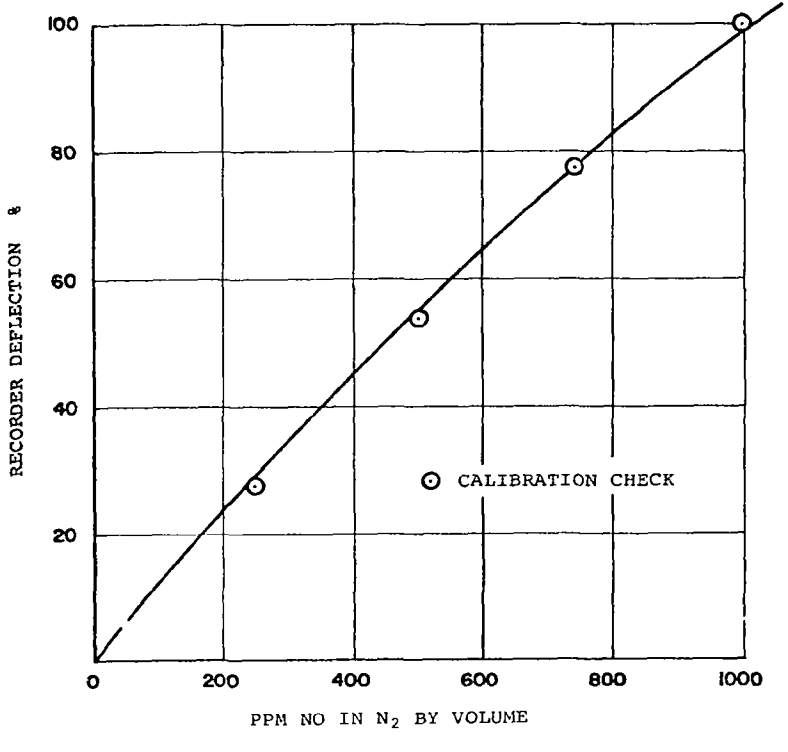


ENCLOSURE 12. CALIBRATION CURVE FOR SULFUR DIOXIDE INFRARED ANALYZER

POPE, EVANS AND ROBBINS

BECKMAN INSTRUMENTS, INC.
MODEL NO. 215 A

APPLICATION: NITRIC OXIDE, RANGE: 0-1000 PPM BY VOLUME
AMPLIFIER NO.: 200066, DETECTOR NO. 1436 A
ZERO GAS: NITROGEN, CALIBRATION PRESSURE: ATMOSPHERIC
SPAN GAS: ANALYZED CYLINDER ~900 PPM NO
CALIBRATION PRESSURE: ATMOSPHERIC
SAMPLE PRESSURE: ATMOSPHERIC



ENCLOSURE 13. CALIBRATION CURVE FOR NITRIC OXIDE INFRARED ANALYZER

POPE, EVANS AND ROBBINS

ENCLOSURE 14. ANALYSES OF "PERFECT EIGHT" UNWASHED 4.5% SULFUR COAL

Source: Pittsburgh #8 Seam, Georgetown Mine
Cadiz. (Harrison City) Ohio

		% Weight	
1.	ULTIMATE ANALYSIS	As Rec'd	Dry
	Moisture	6.01	
	Carbon	66.21	70.45
	Hydrogen	4.57	4.86
	Nitrogen	2.50	2.66
	Chlorine	0.05	0.05
	Sulfur	4.45	4.73
	Ash	10.73	11.42
	Oxygen (diff)	5.48	5.83
		100.00	100.00
2.	PROXIMATE ANALYSIS	As Rec'd	Dry
	% Moisture	6.01	
	% Ash	10.73	11.42
	% Volatile	36.49	38.82
	% Fixed Carbon	46.77	49.76
		100.00	100.00
	BTU	12157	12934
	% Sulfur	4.45	4.73
3.	SULFUR FORMS		
	% Pyritic Sulfur		2.92
	% Sulfate Sulfur		0.08
	% Organic Sulfur		1.73
	% Total Sulfur		4.73
4.	FUSION	REDUCING ATMOSPHERE	
	Initial Def. (ID)	1980°F	
	Softening (H=W)	2125°F	
	Softening (H=1/2W)	2160°F	
	Fluid Temp. (FT)	2270°F	
5.	ASH ANALYSIS		
	Silica (SiO_2)	43.64%	
	Iron Oxide (Fe_2O_3)	25.68%	
	Titania (TiO_2)	xxxxx	
	Alumina (Al_2O_3)	25.02%	
	Manganese Oxide (Mn_3O_4)	xxxxx	
	Lime (CaO)	2.06%	
	Magnesia (MgO)	trace	
	Alkalies (Na_2O / K_2O by diff.)	2.36%	
	Sulfur Trioxide (SO_3)	1.24%	
	Phosphorous pentoxide (P_2O_5)	xxxxx	
		100.00%	

ENCLOSURE 15. ANALYSES OF "PERFECT EIGHT" WASHED 2.6% SULFUR COAL

Source: Pittsburgh #8 Seam, Georgetown Mine
Cadiz. (Harrison City) Ohio

	As Received	Dry
Moisture	5.00%	--
Volatile	37.30	39.30%
Fixed Carbon	50.50	53.10
Ash	7.20	7.60
	100.00%	100.00%
BTU	13,000	13,680
Sulfur	2.60%	2.80%

Size Consist

1½" Modified at Mines	
1½" x 1¼"	16.75%
1¼" x 3/4"	44.85
3/4" x 3/8"	21.89
3/8" x 1/8"	11.19
1/8" x 0	3.32

ASH ANALYSIS

Silica (SiO_2)	43.64%
Iron Oxide (Fe_2O_3)	25.68
Titania (TiO_2)	xxxxxx
Alumina (Al_2O_3)	25.02
Manganese Oxide (Mn_3O_4)	xxxxxx
Lime (CaO)	2.06
Magnesia (MgO)	trace
Alkalies (Na_2) / K_2O by diff.)	2.36
Sulfur Trioxide (SO_3)	1.24
Phosphorous pentoxide (P_2O_5)	xxxxxx
	100.00%

ULTIMATE ANALYSIS

	As Rec'd.	Dry
Moisture	5.00%	--
Carbon	70.36	74.06%
Hydrogen	5.17	5.44
Nitrogen	0.99	1.03
Oxygen	8.73	9.19
Sulfur	2.55	2.66
Ash	7.20	7.62
	100.00%	100.00%

FUSION TEMPERATURE OF ASH

	Reducing Atmosphere	Oxidizing Atmosphere
Initial Deformation	2,020°F	2,365°F
Fusion (Softening)	2,120°F	2,440°F
Fluid Temperature	2,240°F	2,530°F

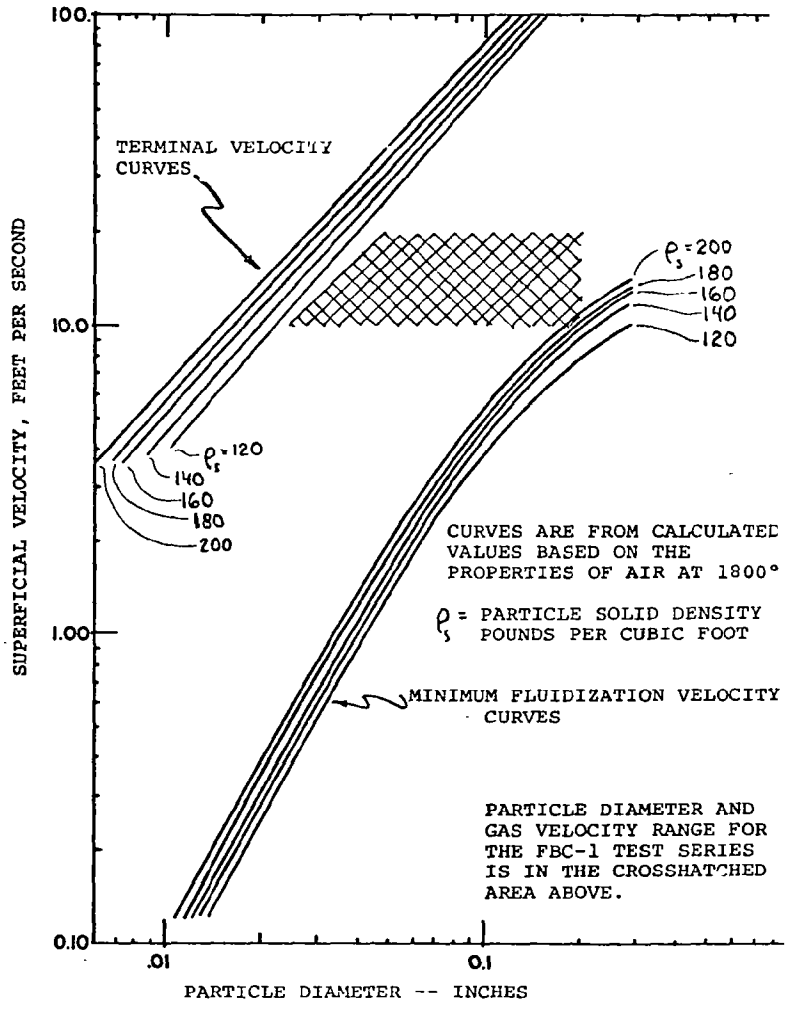
Hardgrove Index 58-61

Free Swelling Index 4-1/2

ENCLOSURE 16. ANALYSIS OF SORBENTS AFTER IGNITION*

CONSTITUENT	DOLOMITE (1337) % By Wt.	LIMESTONE (1359) % By Wt.
CaO	55	97
MgO	43	1.2
Fe_2O_3	0.33	0.22
SiO_2	0.92	1.07
Al_2O_3	0.15	0.29
LOSS ON CALCINATION	47.4	43.6

*Analysis provided by the National Air Pollution Control Administration



ENCLOSURE 17. TERMINAL VELOCITY AND MINIMUM FLUIDIZATION VELOCITY VS PARTICLE DIAMETER

ENCLOSURE 18. ESTIMATION OF ELUTRIATING PARTICLE SIZE FOR THE 1359 LIMESTONE

The smallest particle size that would be retained in the bed was estimated from the intermediate law which is applicable for the test conditions, i.e., Reynolds number between 2 and 500. The particle size follows the relation[1]

$$D_p = \left[\frac{U_t \mu^{0.428} \rho_f^{0.285}}{g^{0.714} (\rho_s - \rho_f)^{0.714}} \right]^{0.875}$$

Where:

D_p is the particle diameter in inches

U_t the superficial gas velocity ft/sec taken as 14.0

μ the gas viscosity, lb/ft sec taken as 2.9×10^{-5}

ρ_f the gas density, lb/ft^3 taken as .020 at 1600°F

g the gravitational constant, 32.2 ft/sec^2

ρ_s the particle density, lb/ft^3 taken as 162.0

Accordingly:

$$D_p = 62.7 \left[\frac{(14)(2.9 \times 10^{-5})^{0.428} (0.02)^{0.285}}{(32.2)^{0.714} (162 - 0.02)^{0.714}} \right] = 0.022 \text{ in.}$$

For this particle size the Reynolds number is:

$$N_R = \frac{D_p U_t \rho_f}{\mu} = \frac{(0.022/12)(14)(0.02)}{2.9 \times 10^{-5}} = 17.9$$

which value falls in the applicability range of the law.

[1] Adapted from Leva, Max: "Fluidization," McGraw Hill Book Company, Inc., New York, 1959

ENCLOSURE 19. NITRIC OXIDE EQUILIBRIUM CONCENTRATIONS FOR THE FLUIDIZED-BED ENVIRONMENT

The equilibrium constant, K, defined as

$$P_{NO} / (P_{N_2} \cdot P_{O_2})^{0.5}$$

is related to the free energy change, ΔG, and is given directly in the JANAF tables

T, °K	T, °F	Log_{10} K	K
1100	1520	-3.633	0.00023
1200	1700	-3.275	.00053
1300	1880	-2.972	.00107
1400	2060	-2.712	.00194
1500	2240	-2.487	.00319

With air at 1500°K, (2240°F) for example,

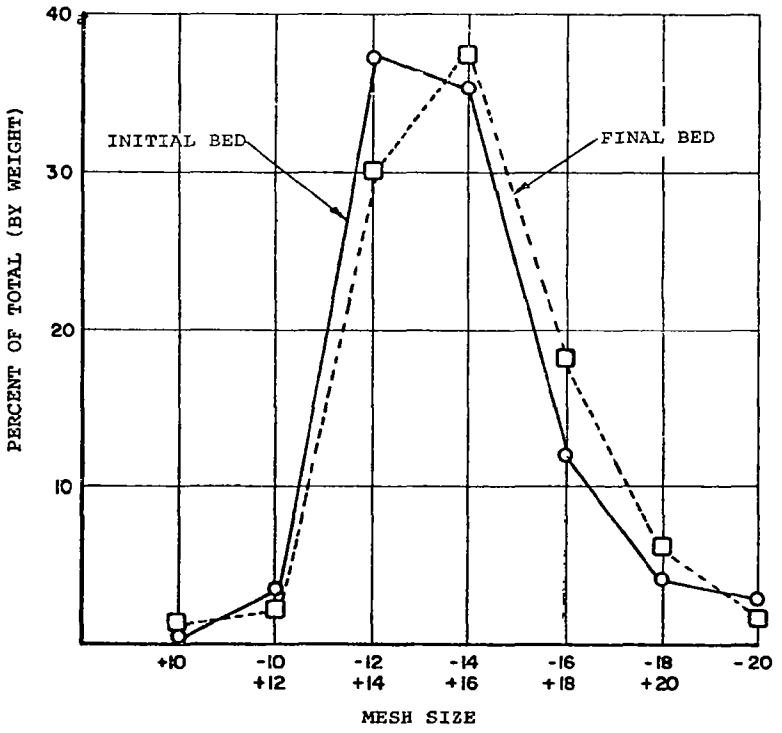
$$\text{PPM(NO)} = 10^6 P_{NO} = 10^6 \left(P_{N_2} \cdot P_{O_2} \right)^{0.5} K =$$

$$10^6 (0.2 \times 0.8)^{0.5} (.0032) = 1276 \text{ ppm}$$

Similarly,

T, °F	Equi. NO ppm
1500	92
1700	222
1880	429
2060	775

If P_{O_2} is reduced to 0.05 corresponding to a possible FBM condition, P_{NO} is reduced by a factor $(.05/.2)^{0.5}$ or to half the value shown above.



ENCLOSURE 20. PARTICLE SIZE DISTRIBUTION OF 1359 LIMESTONE BED BEFORE AND AFTER FLUIDIZED-BED COMBUSTION

ENCLOSURE 21. INTEGRATED SULFUR BALANCE FOR SORPTION-DESORPTION OF THE LIMESTONE BED DURING FBC TEST 114

For the absorption phase the total sulfur absorbed by the bed is:

S_B = Input sulfur - fly ash loss - emission

or $S_B = \int G_c S_c dt - \int S_F G_F dt - k \int M C_{SO_2} G_c dt$

where: S_B = total sulfur in the bed lbs

G_c = coal rate, lb/hr

S_c = sulfur content in coal, lb/lb

t = time, hours

S_F = sulfur content in fly ash, lb/lb

G_F = fly-ash rate, lb/hr

k = constant = $\frac{\text{dry mole flue gas}}{\text{lb coal}} \times 10^{-6}$

C_{SO_2} = concentration of SO_2 in flue gas, ppm

M = molecular wt. of sulfur = 32

The sulfur retained in the bed during the 4.28 hour absorption period was computed as follows:

(A) Sulfur input = $\int_0^t G_c S_c dt = 63.0 \times \frac{3.09}{100} \times 4.28 = 8.35$ lbs

(B) Fly-ash loss = $\int_0^t G_F S_F dt = 16 \times \frac{1.8}{100} \times 4.28 = 1.23$

(C) Emission loss = $KMG_c \int_0^t C_{SO_2} dt = \frac{.326\ (32)(63)(3200)}{10^6} =$ 2.11 lbs

ENCLOSURE 21. (Continued)

The integral $\int C_{SO_2} dt$ represents the area under the curve during the absorption period.

The sulfur retained in the bed is

$$S_B = A - (B + C)$$

(D) $= 8.35 - (1.23 + 2.11) = 5.01$ lbs

$$\% \text{ retained in bed} = \frac{5.01}{8.35} = 60.2\%$$

During the regeneration phase (t = .65 hours), the sulfur loss from the bed is:

Recovered sulfur = emission - (input - fly ash)

$$S_R = kMG_c \int_0^t C_{SO_2} dt - \int_0^t G_c S_c dt + \int_0^t G_F S_F dt$$

(E) Emission sulfur $= \frac{.326\ (32)(63)(7950)}{10^6} = 5.25$ lbs

The value 7950 ppm hours was determined from the area under the concentration curve (Figure 2) during regeneration.

ENCLOSURE 21. (Continued)

(F) Input = 63 $(\frac{3.09}{100})(.65)$ = 1.27 lbs

(G) Fly ash = 16 $(\frac{1.9}{100})(.65)$ = .21 lbs

(H) Sulfur recovered is E - F + G = 5.25 - 1.27 + 21 = 4.19 lbs

(I) Sulfur retained in bed after regeneration

$$= \text{bed mass} \times S_c = 49\ (\frac{1.0}{100}) = .49 \text{ lbs}$$

(J) Total of H and I = 4.68 lbs

Percent of sulfur recovered from bed

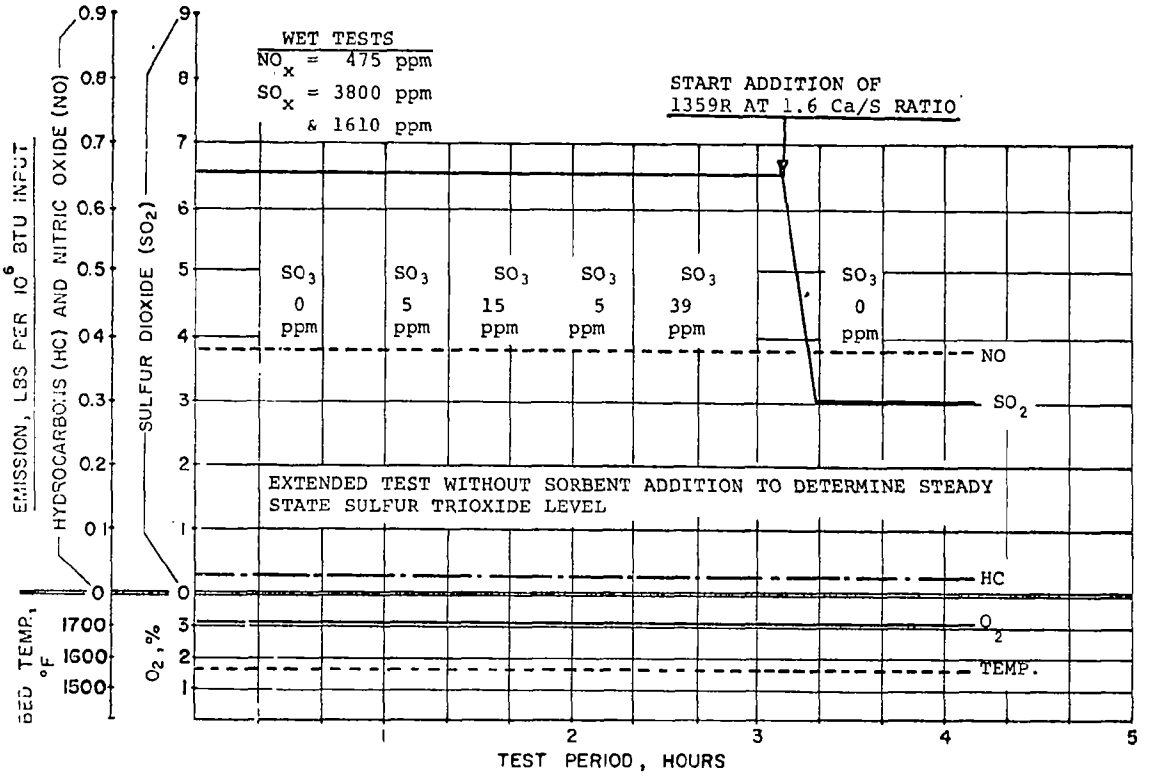
$$= \frac{4.19}{4.68} \times 100 = 89.8\%$$

Sulfur unaccounted for = A + F - (B + C + E + G + I)

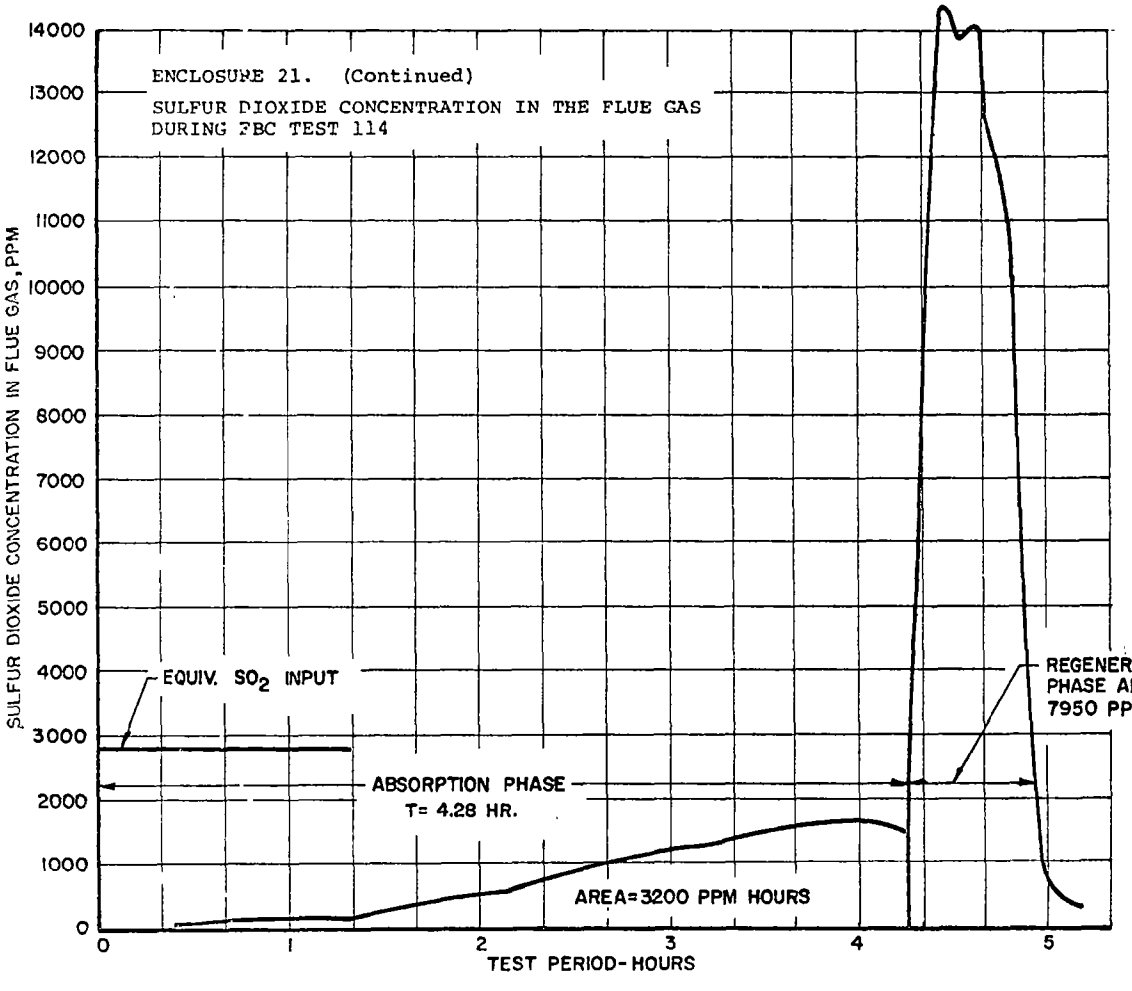
$$= 8.35 + 1.27 - [1.23 + .21 + 5.25 + 2.11 + .49]$$

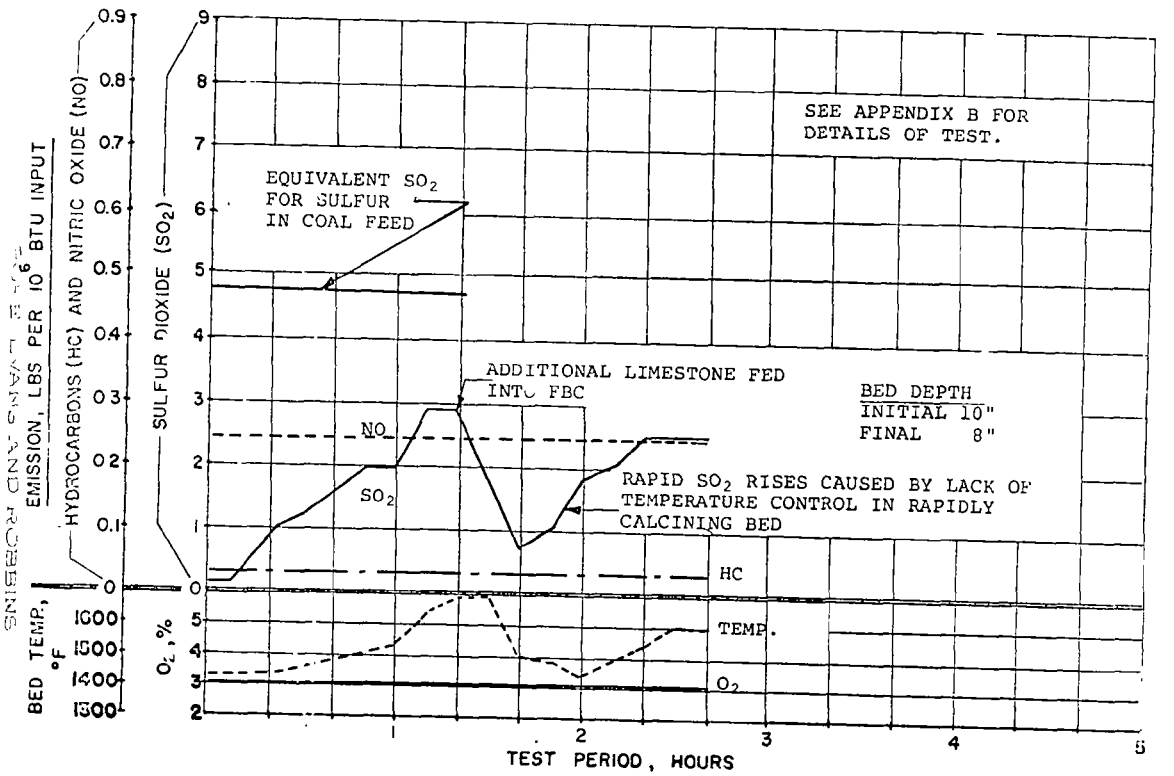
$$= .33 \text{ lbs}$$

$$\% \text{ unaccounted} = \frac{.33}{9.62} = 3.5\%$$

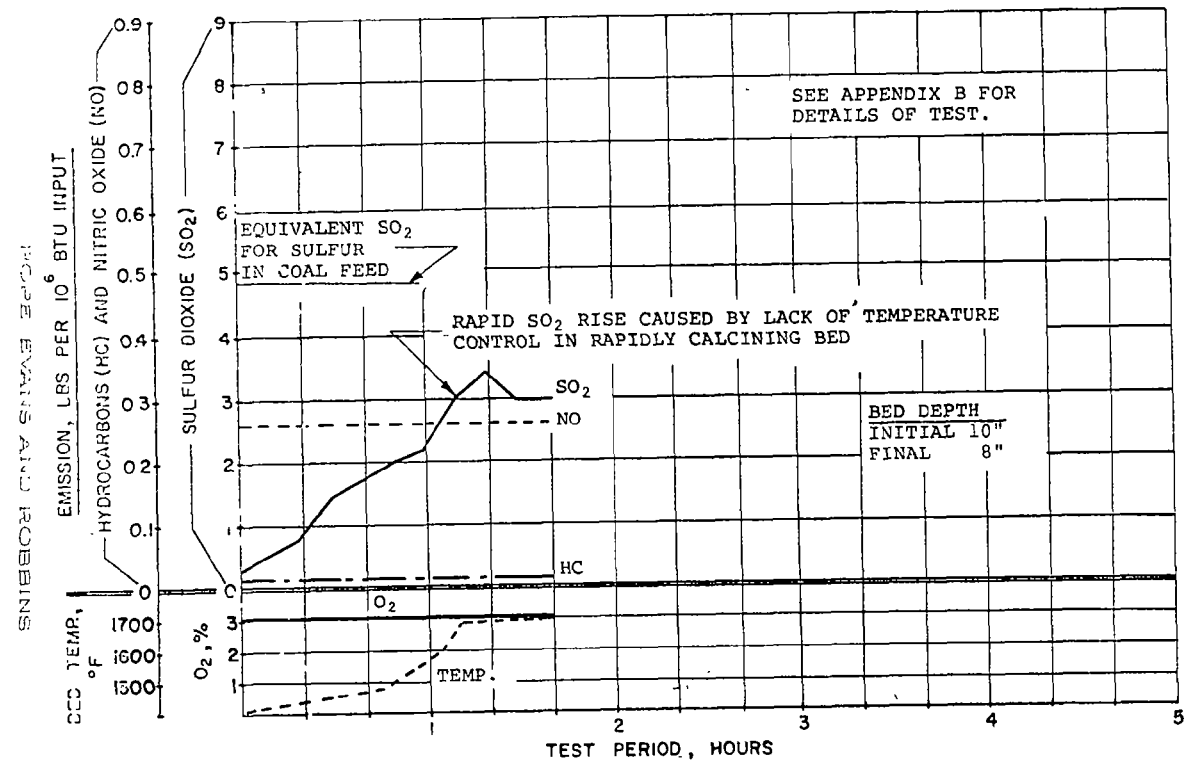


ENCLOSURE 22. EMISSIONS DURING FBC TEST 63 FOR SULFUR TRIOXIDE EMISSION BURNING A 4.5% SULFUR COAL

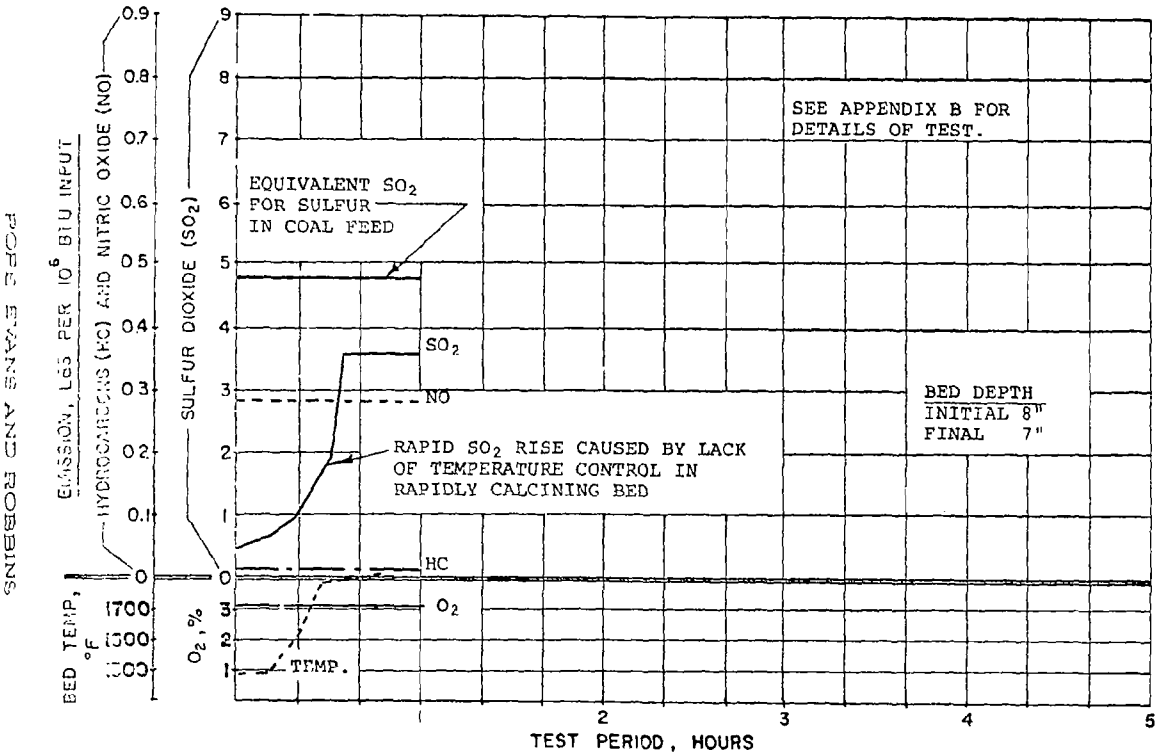




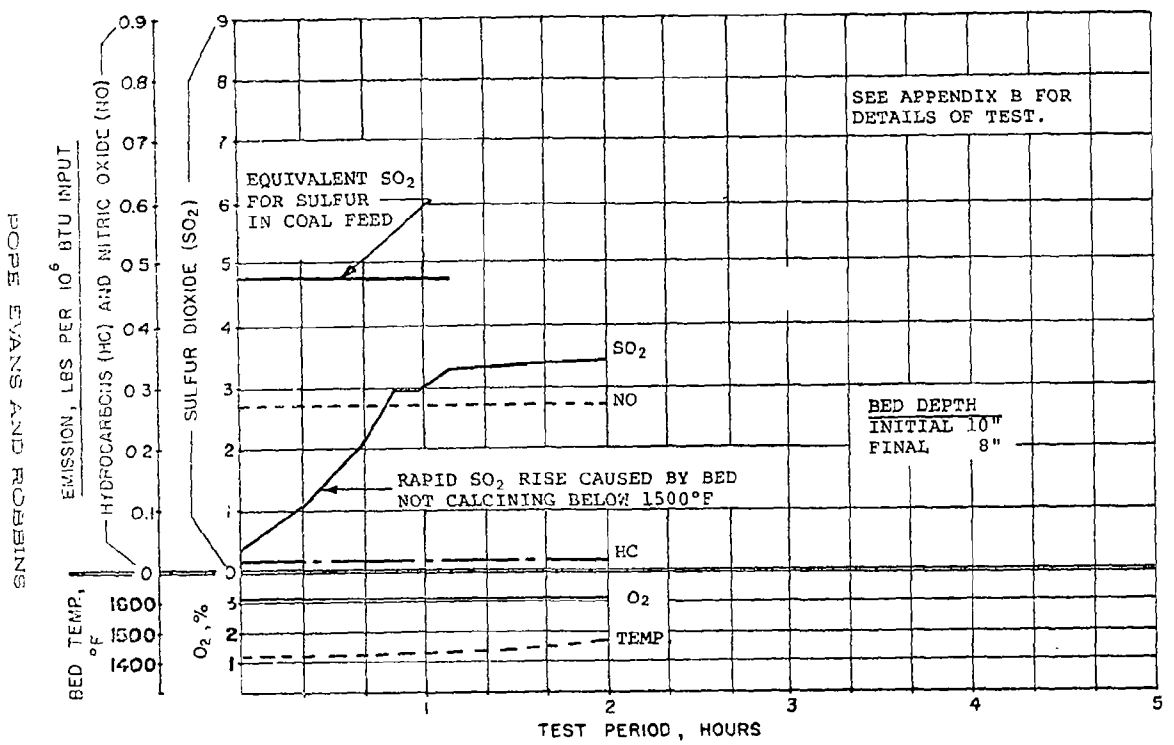
ENCLOSURE 24. EMISSIONS DURING FBC TEST 102 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED



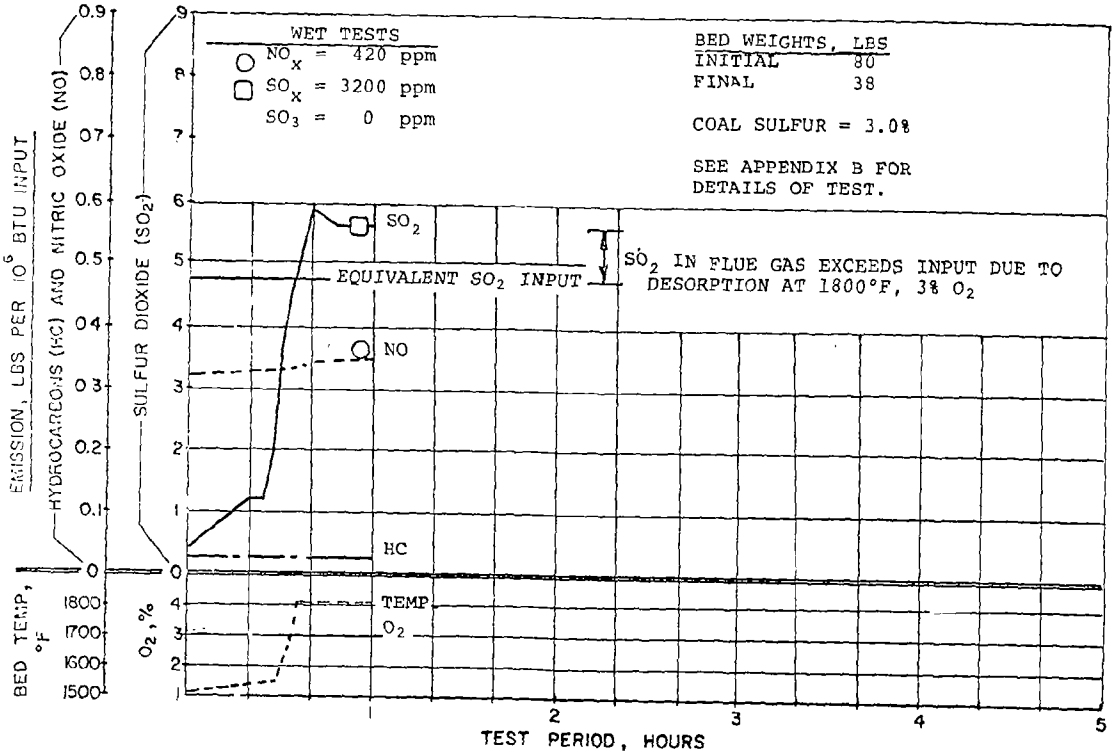
ENCLOSURE 23. EMISSIONS DURING FBC TEST 101 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED



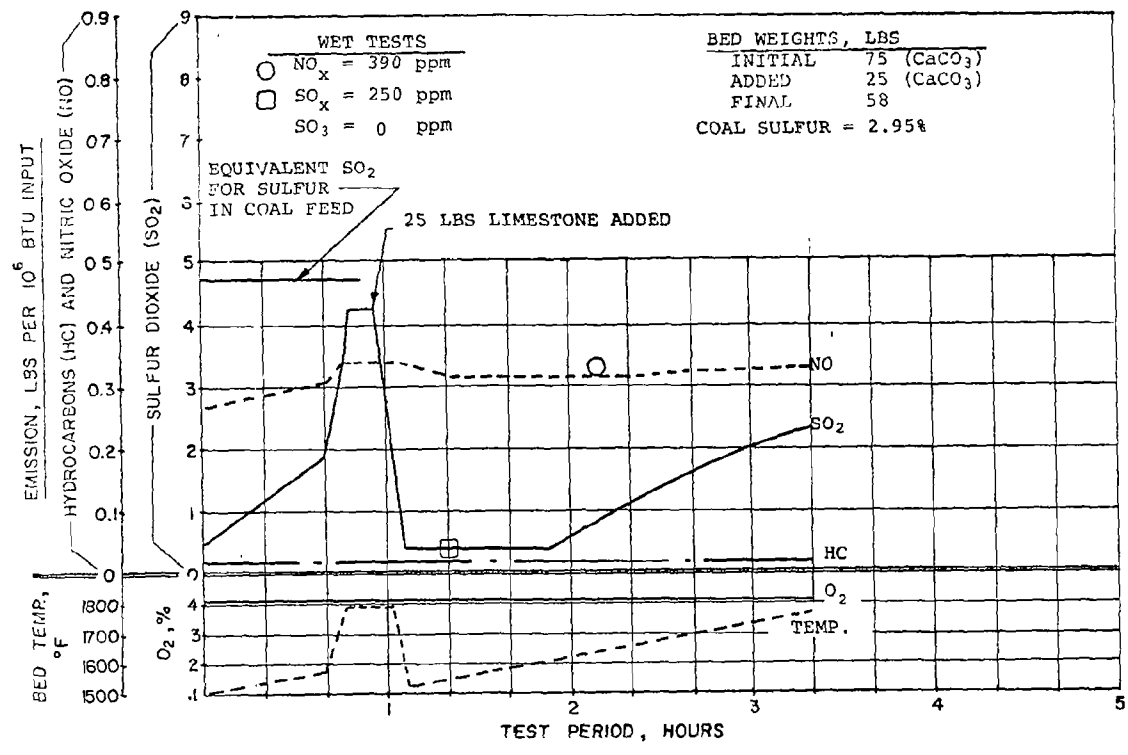
ENCLOSURE 26. EMISSIONS DURING FBC TEST 105 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED



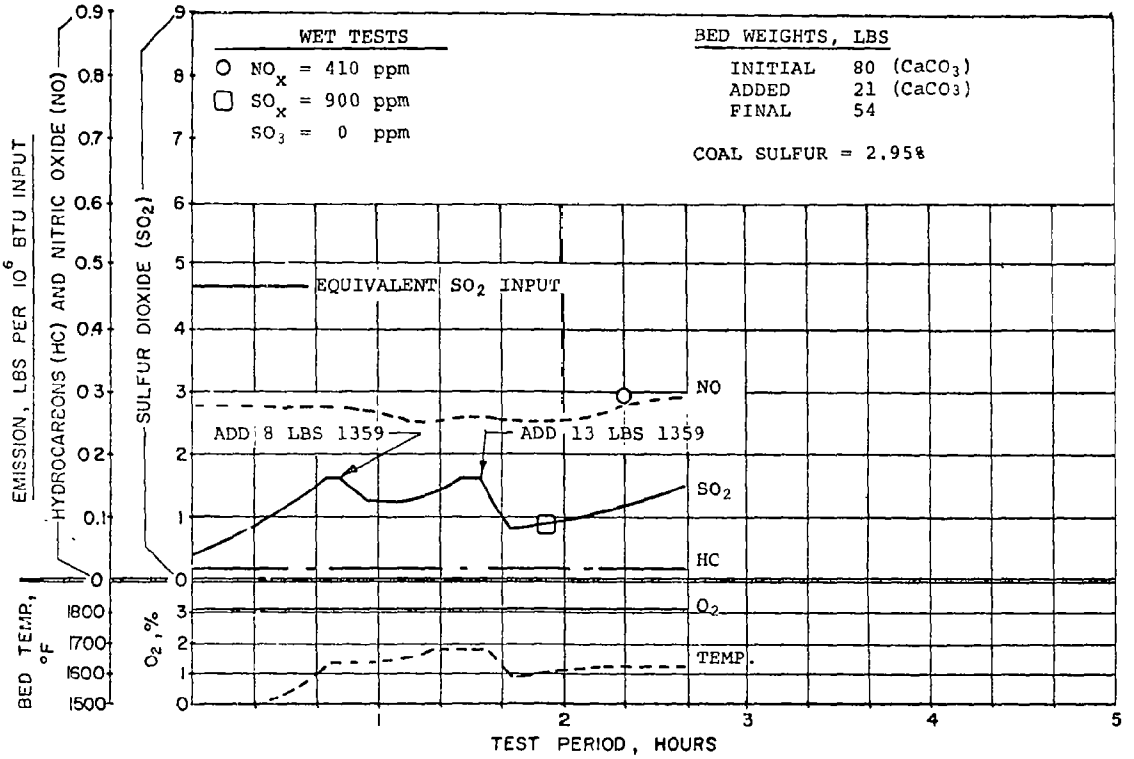
ENCLOSURE 25. EMISSIONS DURING FBC TEST 104 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED



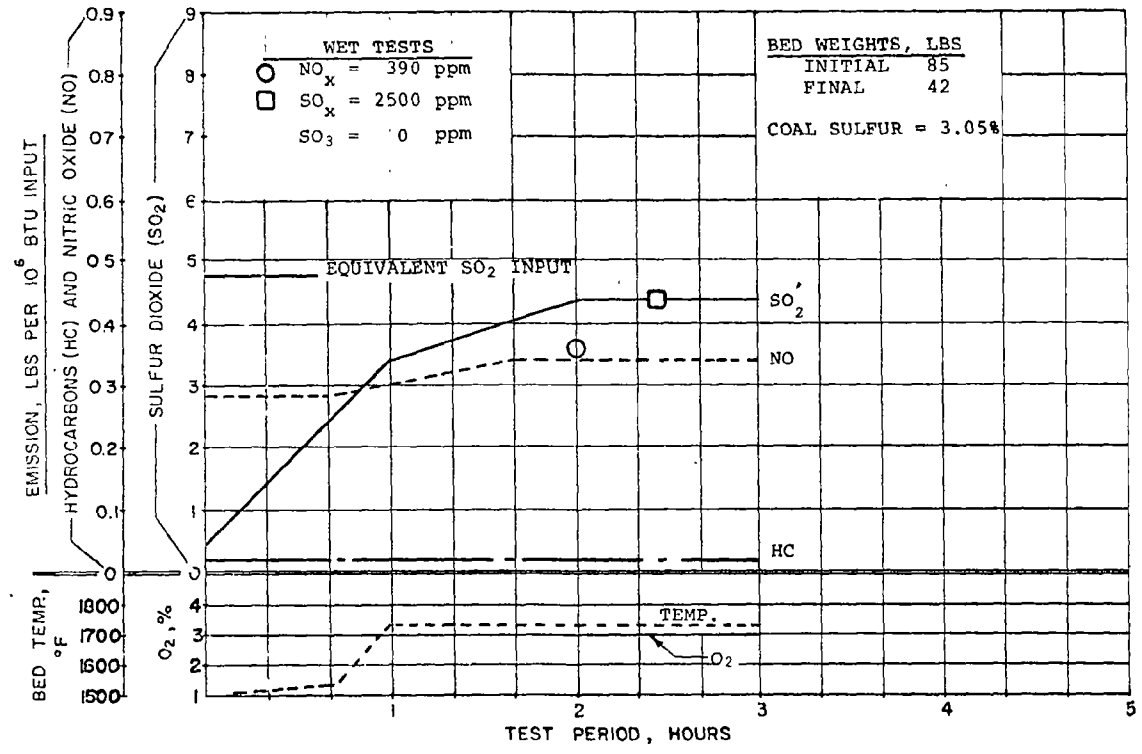
ENCLOSURE 28. EMISSIONS DURING FBC TEST 107 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED



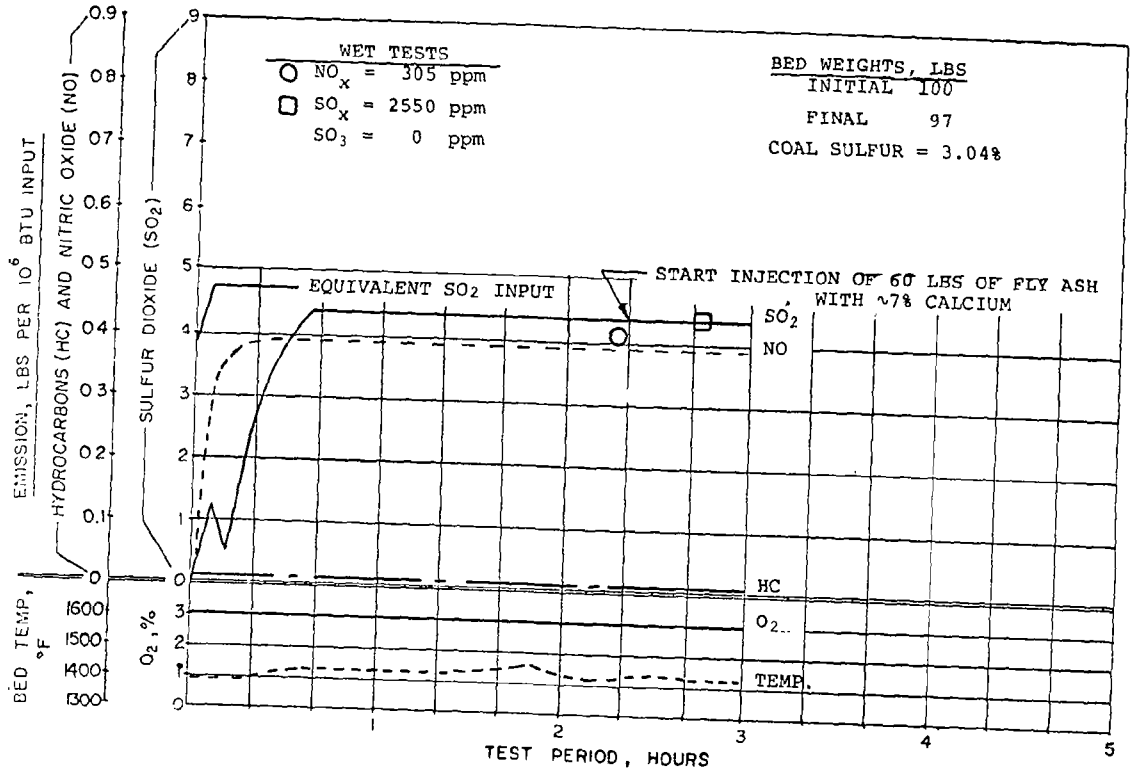
ENCLOSURE 27. EMISSIONS DURING FBC TEST 106 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED



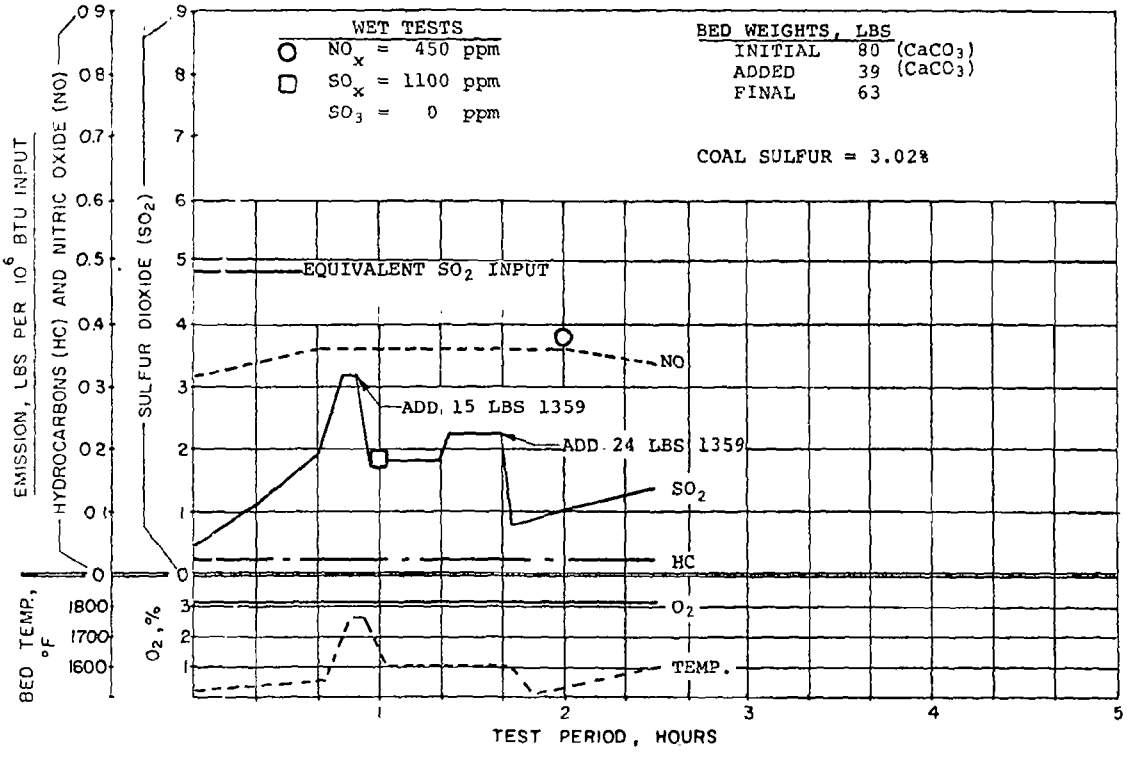
ENCLOSURE 30. EMISSIONS DURING FBC TEST 109 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED



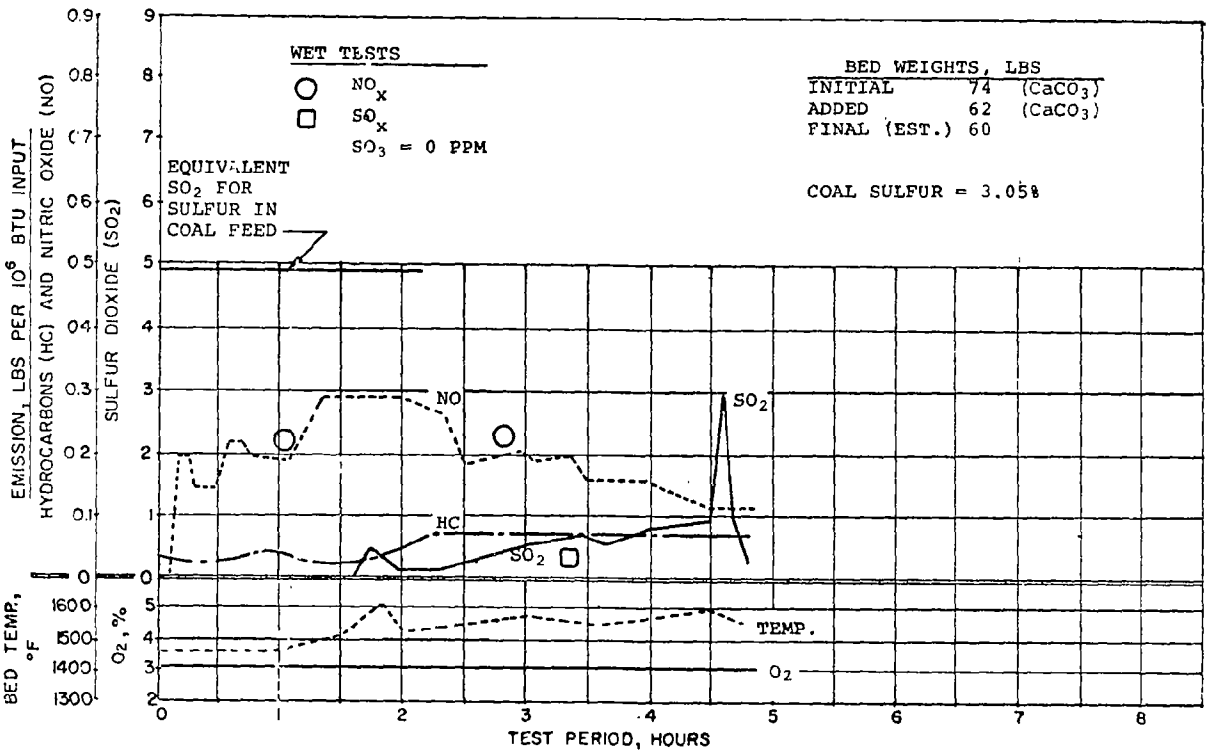
ENCLOSURE 29. EMISSIONS DURING FBC TEST 108 BURNING A MEDIUM SULFUR COAL IN A 1369 LIMESTONE BED



ENCLOSURE 32. EMISSIONS DURING FBC TEST 111 BURNING A MEDIUM COAL IN A 1359 LIMESTONE BED

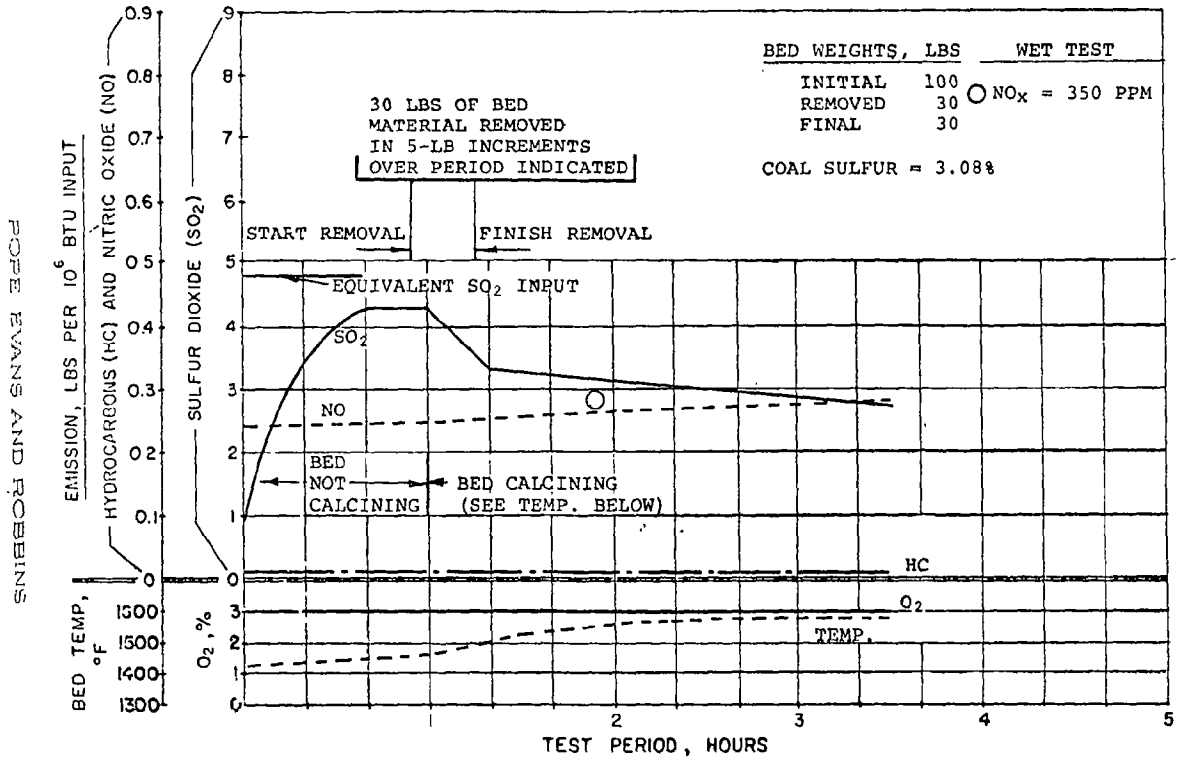


ENCLOSURE 31. EMISSIONS DURING FBC TEST 110 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED



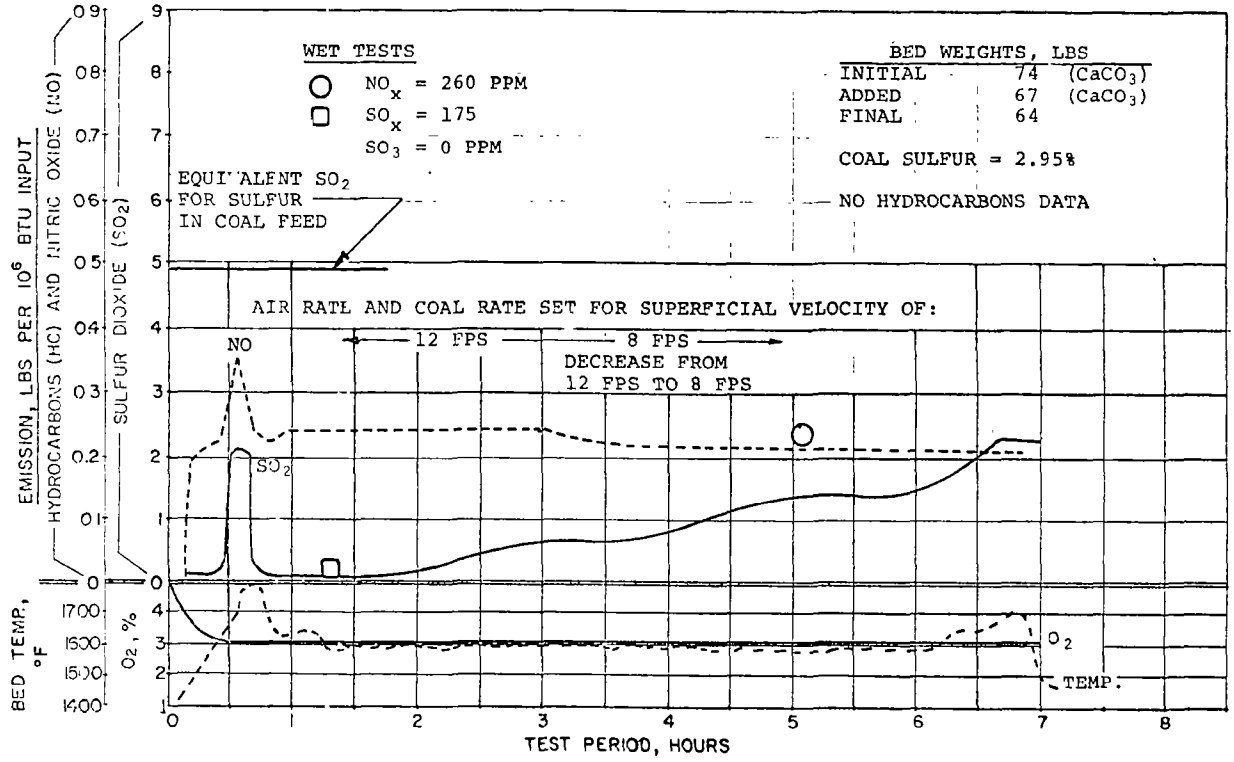
ENCLOSURE 34. EMISSIONS DURING FBC TEST 113 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED

A-40

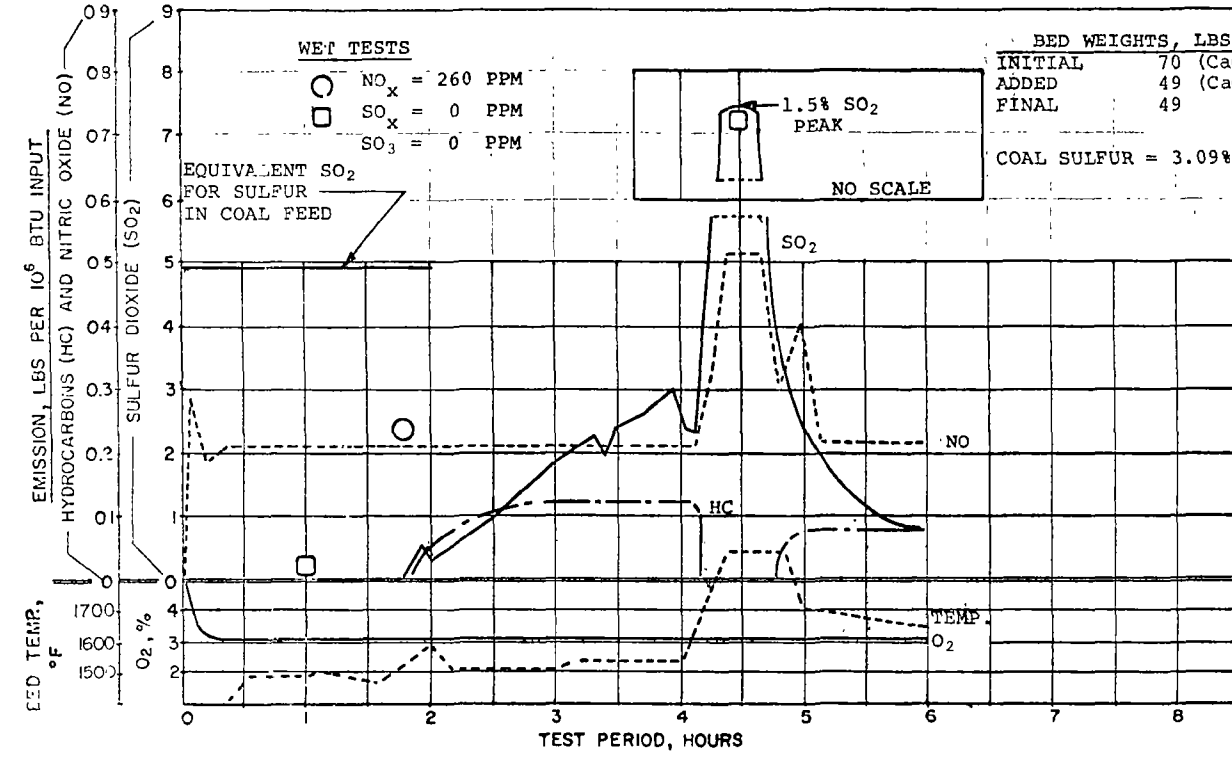


ENCLOSURE 33. EMISSIONS DURING FBC TEST 112 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED

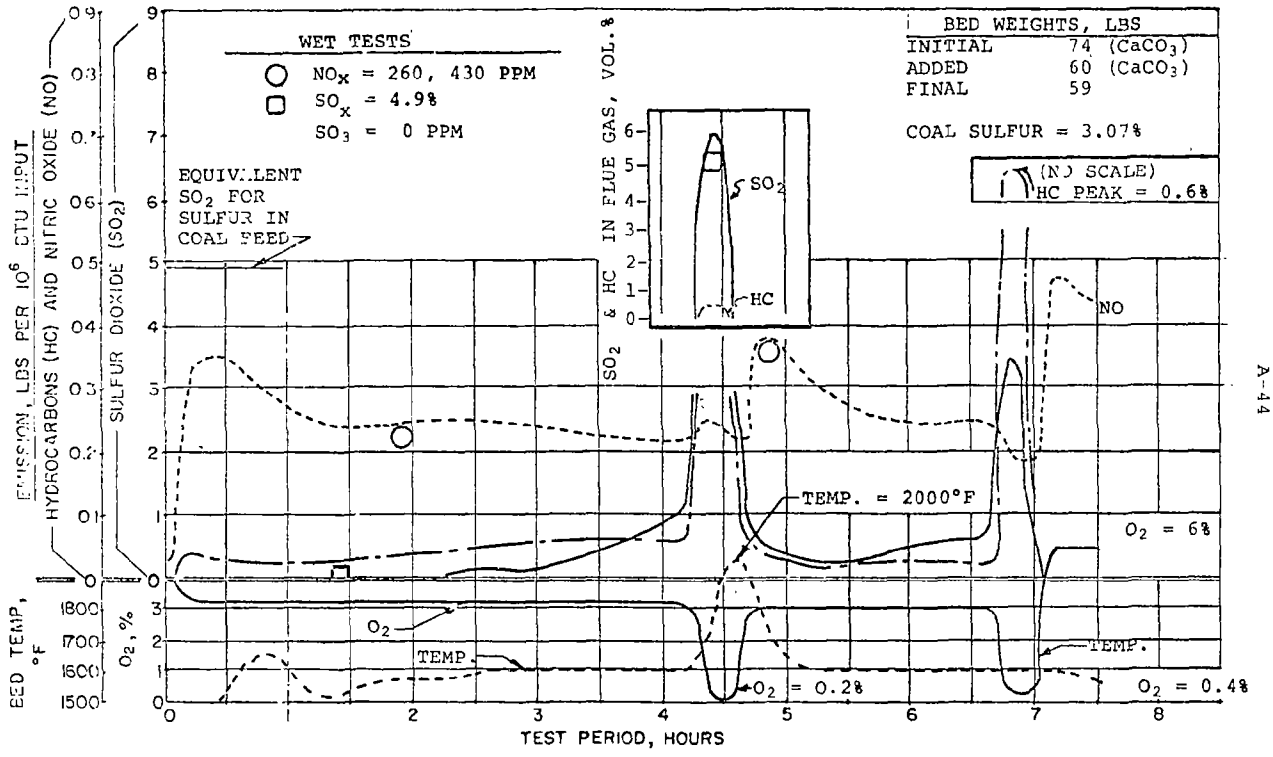
A-39



ENCLOSURE 36. EMISSIONS DURING FBC TEST 115 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED WITH CHANGE IN GAS VELOCITY

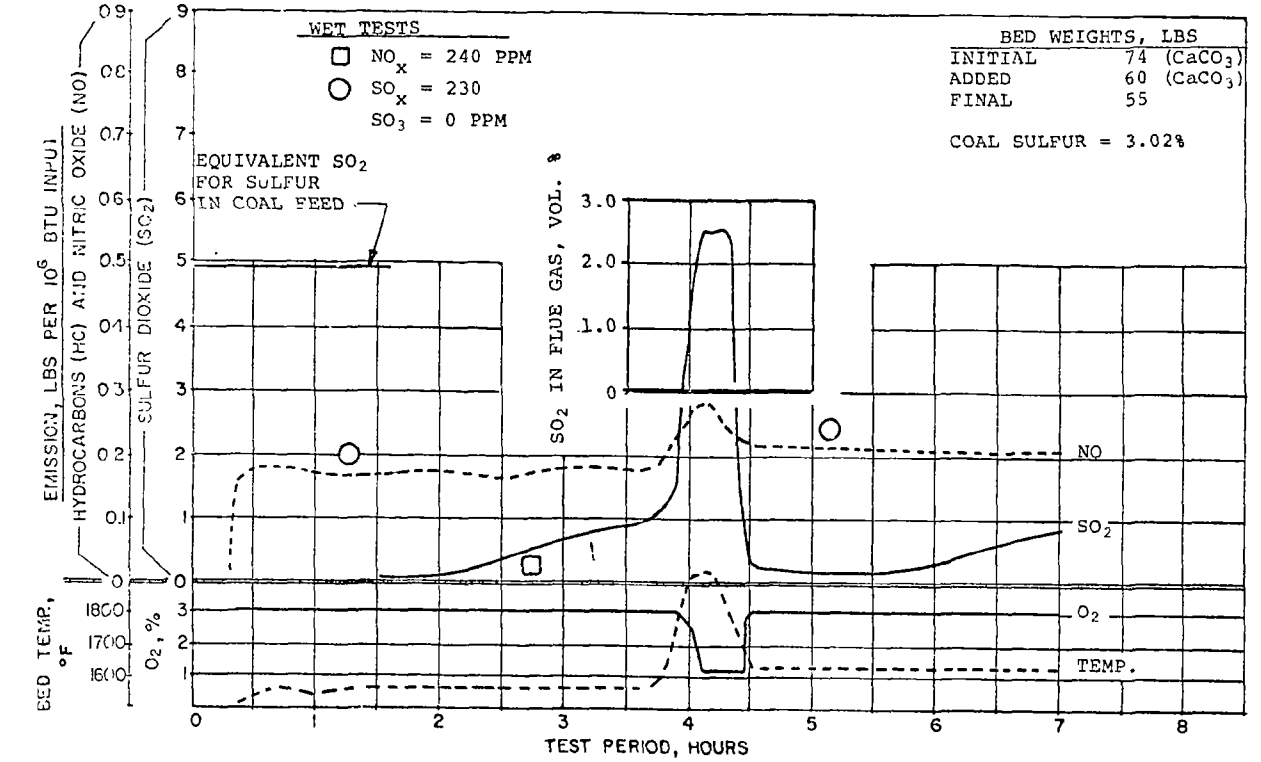


ENCLOSURE 35. EMISSIONS DURING FBC TEST 114 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED WITH REGENERATION



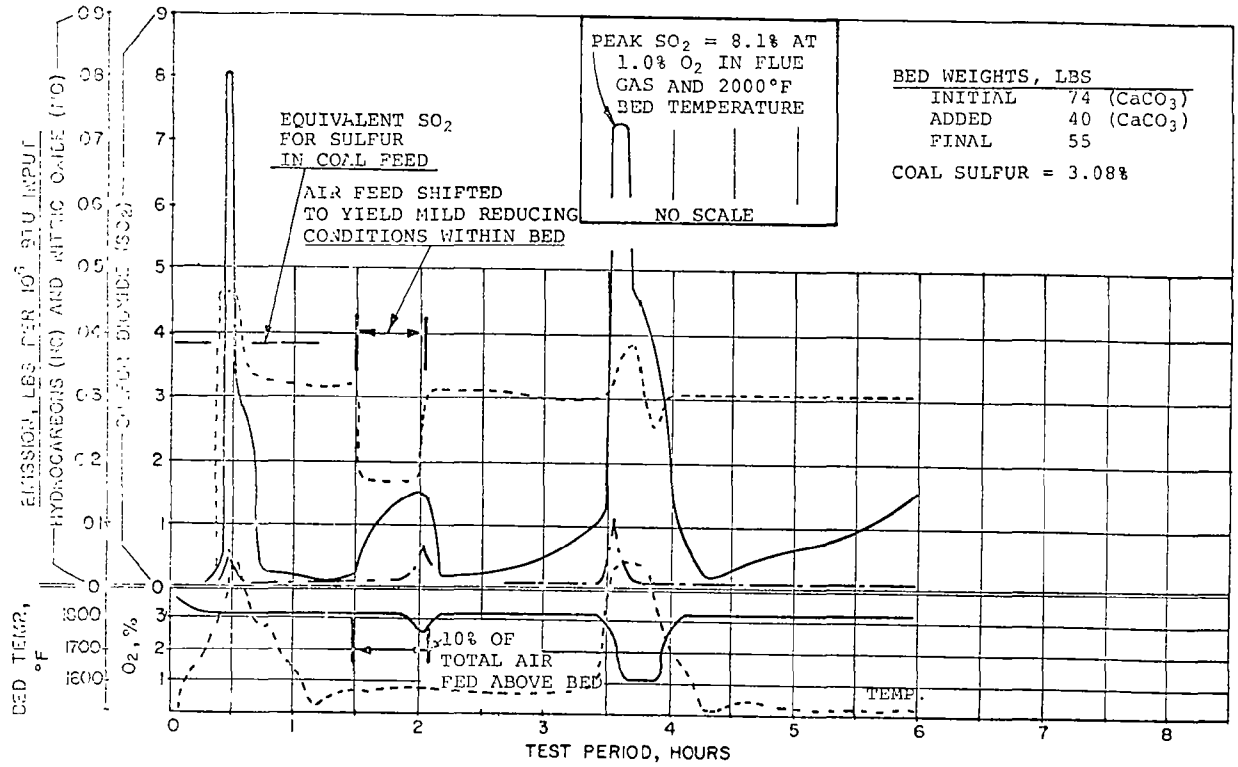
ENCLOSURE 38. EMISSIONS DURING FBC TEST 117 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED WITH REGENERATION

A-44

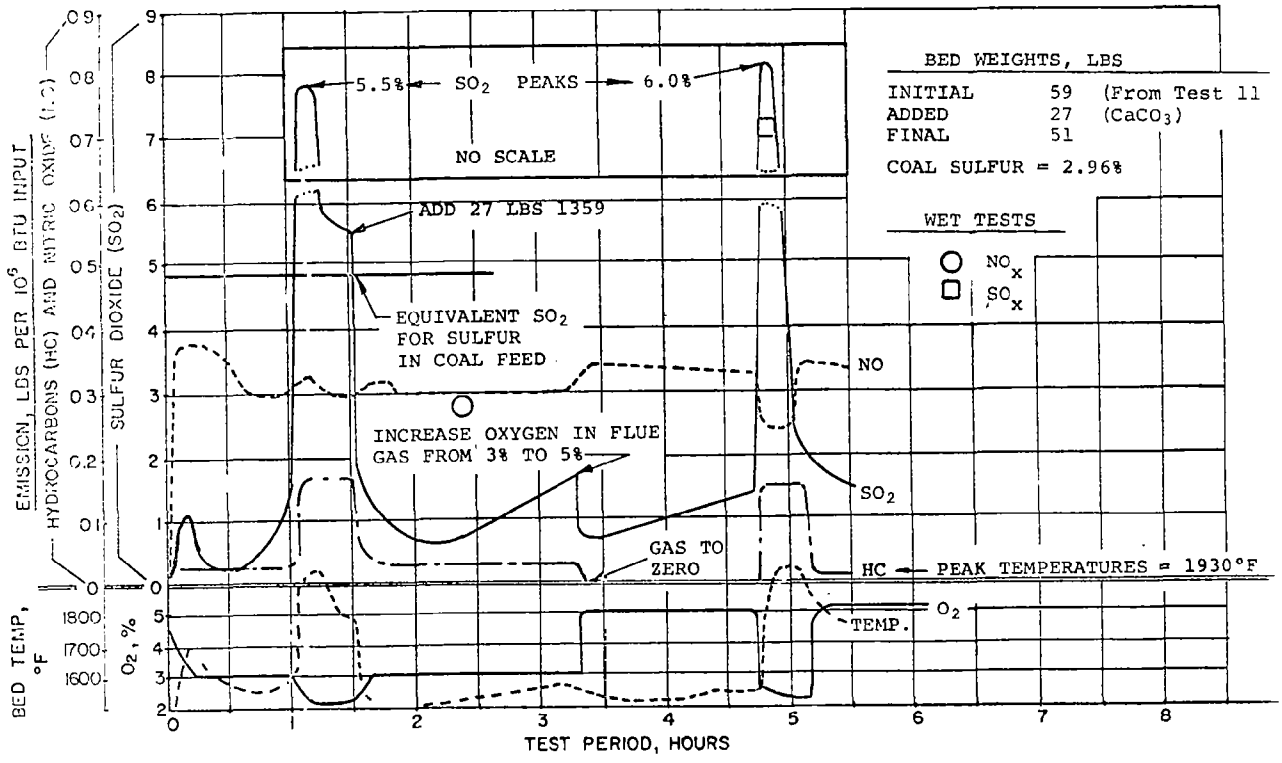


ENCLOSURE 37. EMISSIONS DURING FBC TEST 116 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED WITH REGENERATION

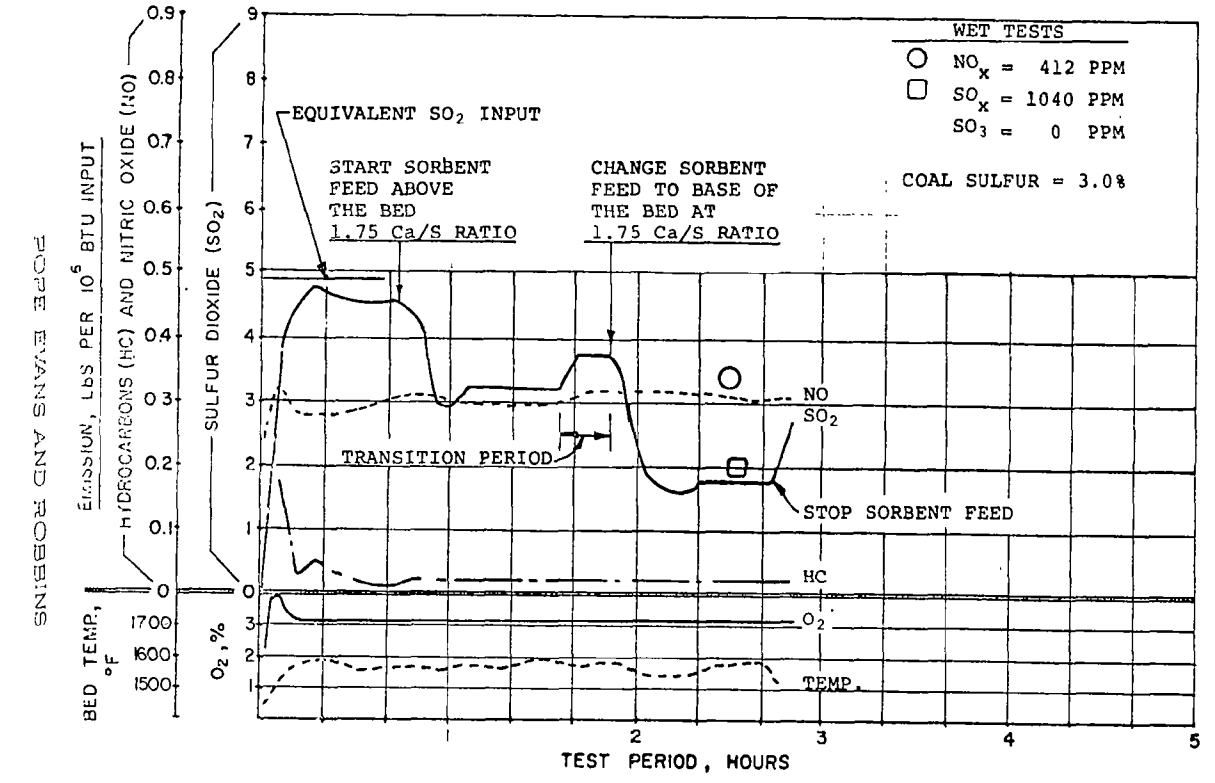
A-43



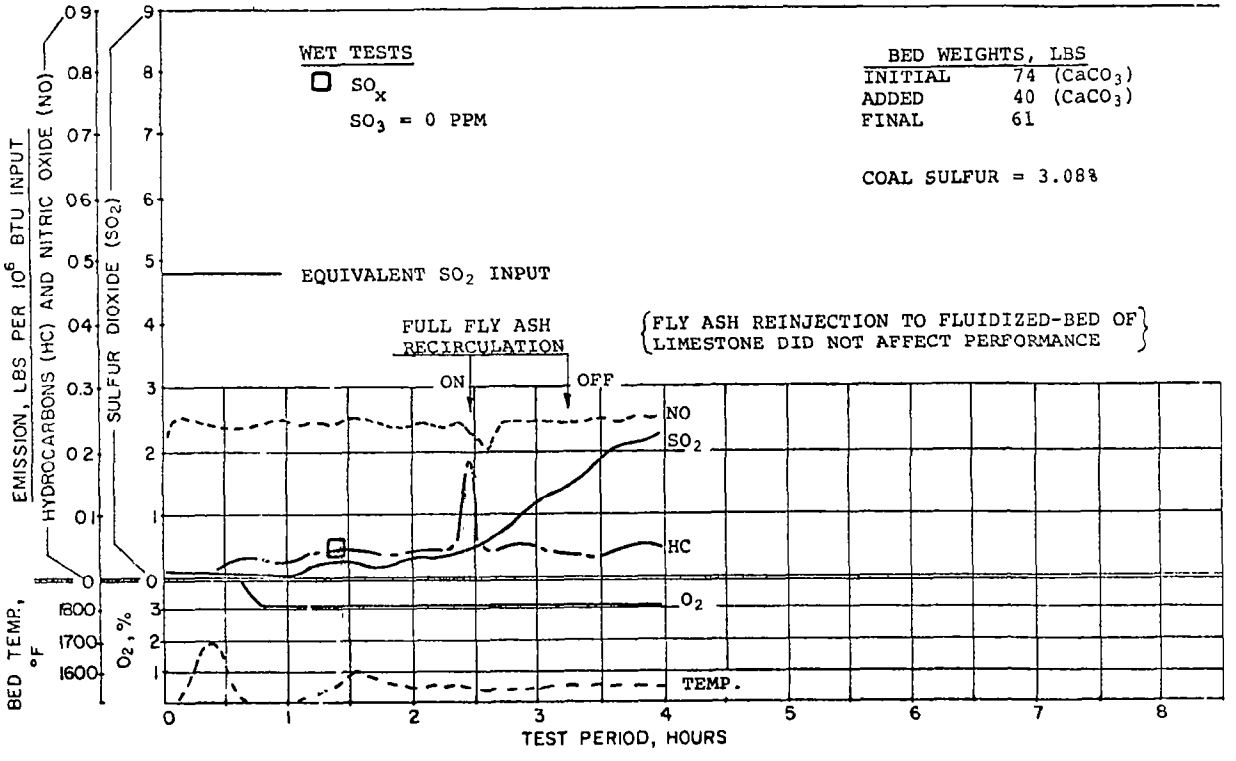
ENCLOSURE 40. EMISSIONS DURING FBC TEST 119 BURNING A MEDIUM SULFUR COAL IN A LIMESTONE BED WITH BED REDUCING



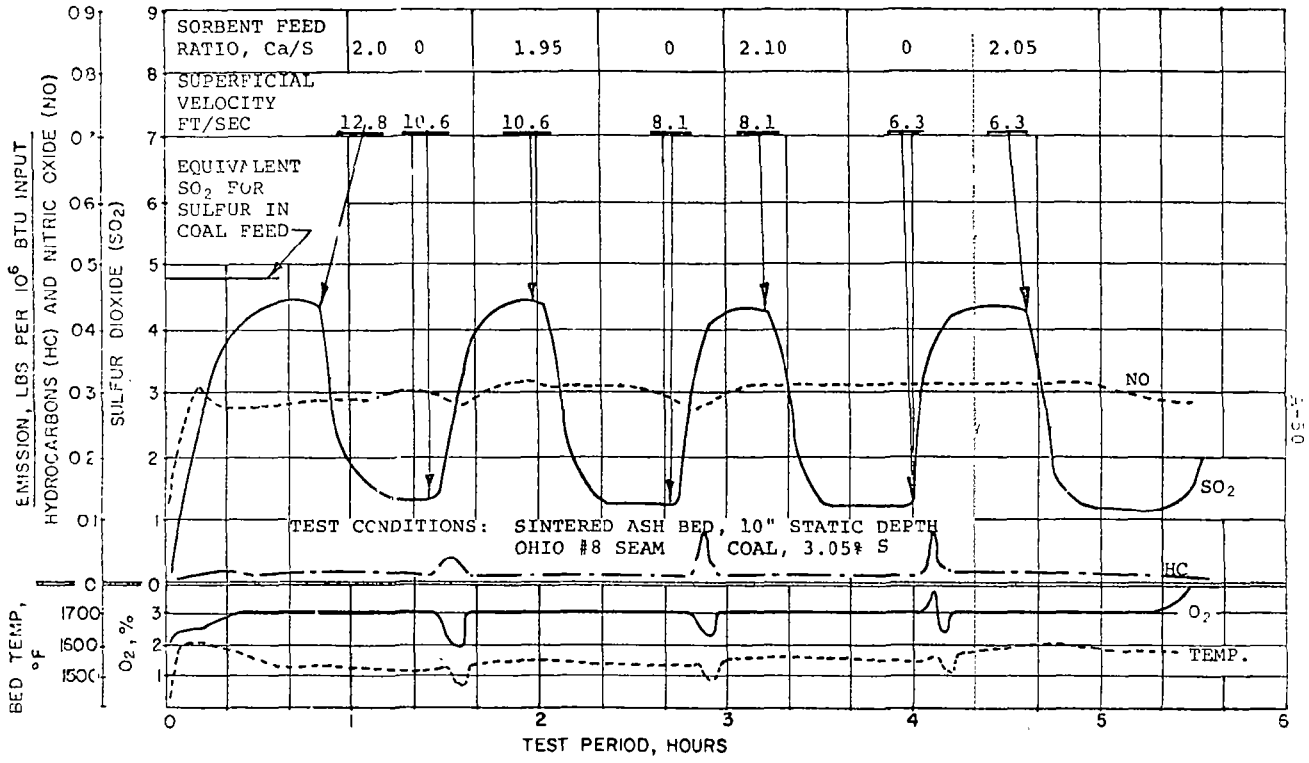
ENCLOSURE 39. EMISSIONS DURING FBC TEST 118 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED WITH CHANGE IN FLUE GAS O_2 CONTENT



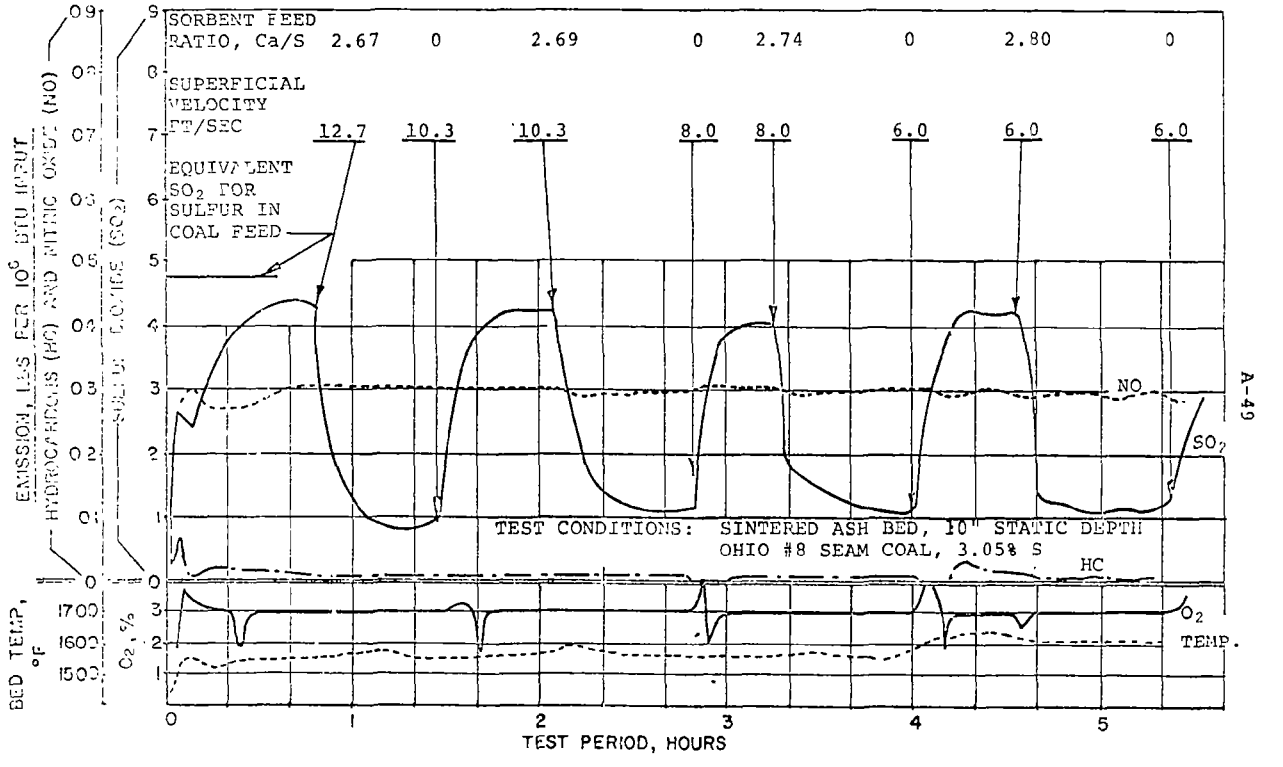
ENCLOSURE 42. EMISSIONS DURING FBC TEST 75 BURNING A MEDIUM SULFUR COAL WITH INJECTION OF -325 MESH 1359R LIMESTONE ABOVE AND AT BASE OF BED AT CONSTANT Ca/S RATIO



ENCLOSURE 41. EMISSIONS DURING FBC TEST 120 BURNING A MEDIUM SULFUR COAL IN A 1359 LIMESTONE BED WITH ASH RECIRCULATION



ENCLOSURE 44. EMISSIONS DURING FBC TEST 77 BURNING A MEDIUM SULFUR COAL WITH INJECTION OF -325 MESH 1359R LIMESTONE WITH CHANGE IN SUPERFICIAL VELOCITY



ENCLOSURE 43. EMISSIONS DURING FBC TEST 76 BURNING A MEDIUM SULFUR COAL WITH ADDITION OF -325 MESH 1359R LIMESTONE WITH CHANGE IN SUPERFICIAL VELOCITY

APPENDIX B

TEST DATA FBC AND FBM

POPE EVANS AND ROBBINS

POPE EVANS AND ROBBINS

ENCLOSURE 45. PROPOSED DESIGN ARRANGEMENT FOR MINIMIZING PARTICULATE EMISSION CONTROL COSTS

TABLE B-1 FBM TEST CONDITIONS AND DATA

Test No	Coal Type and Composition	Bed Material	Static Bed Depth, inches	Test Cond No	Coal Rate, lb/hr	Bed Temp, °F	Air Rate, lb/hr	Flue Gas Oxygen (Control), %	Sorbent Data: Type and Size[a]	Sorbent Data: Injection Rate, lb/hr	Sorbent Data: Mole Ratio, Ca/S	Gas Analysis, Concentration in volume percent or ppm as indicated: I R SO_2, ppm[b]	Wet Tests: SO_2, ppm	Wet Tests: SO_3, ppm	I R NO, ppm	PDS MO_x, ppm[c]	Hydrocarbons, ppm	Grast Analysis, %: CO_2	Grast Analysis, %: O_2	Grast Analysis, %: CO	SO_2 Reduction (I R), %	Sorbent Utilization,[d] %	Fly Ash Data: Carbon Content %	Fly Ash Data: Recirculation[e]	Fly Ash Data: Emission, lb/hr	Remarks
FBM 1	E Ky Pike County 1% Sulfur 8% Ash	Ash	20	1	880	1540	7200	0 5	None	0	0	680			200		4600	17 5	0 2	0 5	0	-	66			Test of effect of vary-ng excess air
				2	800	1540	7200	3.0		0	0	600			220		3250	16 2	2 8	0.0	0	-				
				3	800	1540	7200	6 0		0	0	500			260		300	13 8	5 8	0.0	0	-	45			
				4	800	1540	7200	2 0		0	0	620			240		1600	16 5	2.1	0 0	0	-				
FBM 2	Change from E Ky to Ohio #8 Seam unwashed	Ash	12	1	830	1800	7400	1 0		0	0	700			250		450	17 0	1 0	0.3	0	-	55.2			Coal transition
			20	2	800	1650	7400	3.0		0	0	3400			300		200	15 5	2 9	0 0	0	-	59 0			
			20	3	800	1600	7400	3.5	1337R	370	1 75	1550			300		120	14 9	3 6	0 0	56.5	31 2				
			20	4	800	1500	7400	4 0	-7 +14	0	0	800[f]			340		90	14 9	3 6	0 0	0	-	43 6			*50 lbs limestone added as slug
FBM 3	Ohio #8 Seam unwashed 4 5% Sulfur 10 7% Ash	Ash	20	1	830	1620	7600	3 8		0	0	3600			380		50	12 6	3 2	0 0	0	-	56 5			
				2	870	1540	7600	3.8	1337R -7 +14	320	1 4	2050			360		50	13 2	3 8	0 0	43 0	30 7	49 0			Gas sample system leak
FBM 4	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	15	1	830	1760	7600	3 2		0	0	3800	3340	409	360	134	60	13 8	4 0	0 0	0	-	64 8			
				2	830	1740	7600	3.2	1337R	254	1.15	3700			340		60	14 2	3 3	0 0	2 6	1 7				
				3	830	1720	7600	3.2	-7 +14	355	1.60	3300			340		60	14 5	3 5	0 0	13 2	8.3				
				4	830	1720	7600	3.2		480	2 20	2700			300		60	14 5	3 2	0 0	29 0	13 2	37 4	Yes		
FBM 5	Ohio #8 Seam unwashed 4 5% Sulfur 10.7% Ash	Ash	20	1	720	1600	6800	3 8		0	0	3800	3500	248	280	162	120	14 3	3 2	0.0	0	-	69 2			
				2	720	1570	6800	3.0		0	0	3000			280		80	14 7	3 1	0 0	21 0	18.6	45 0			
				3	720	1570	6800	3 4	1337R	220	1.13	3000			280		50	14.2	3 5	0.0	37 0	22.4	43 6			
				4	720	1570	6800	3 0	-7 +14	317	1.65	2400			280		100	14 5	3 0	0.0	48 0	20.0	43 4			
										460	2 40	2000	1850	102												
FBM 6	Lost Ignition - Discarded																									
FBM 7	Ohio #8 Seam unwashed 4 5% Sulfur 10 7% Ash	Ash	20	1	780	1620	7500	3 0		0	0	3900	3640	285	280	189	200	13.8	3.0	0.0	0	-	54 7			
				2	780	1480	7500	3.0	NHL *	175	1.11	2800			260		150	14 8	3.2	0.0	28 2	25.4				*Natural mine limestone
				3	780	1480	7500	2.3	-7 +14	265	1 70**	2150			220		390	15 2	2.8	0 0	45 0	26.5	42 1			72% $CaCO_3$
				4	780	1480	7500	3 0		175	1.11*	2400	2150	125	160		180	14 8	3.0	0 0	39.0	35.0	43 8			** Steam added to air inlet
FBM 17	Ohio #8 Seam unwashed 4.5% Sulfur 10 7% Ash	Ash	20	1	750	1600	7800	3 0		0	0	3900	3870	35	280						0	-	65.2			
				2	820	1580	7800	3.0	1359B*	63	.72	2800	2690	0	240	285	180	15 0	3.0	0.0	28 2	39.2	50 0			* 1359 Hydrate
				3	800	1580	7800	3.0		84	98	2100	2180	0	200		180	15 3	3 0	0.0	41 0	42.0	46.7			-325 Mesh particle size
				4	840	1490	7800	3 0		84	93**	1800	1820	0	200		180	15 5	3 2	0 0	54 0	58.0	42 0			** 5 5% Steam added to air inlet
																										Bed banked between Cond. 2 and 3
FBM 18	Ohio #8 Seam washed 2.6% Sulfur 7 2% Ash	Ash	13	1	830	1870	7800	1.0	None	0	0	2800			280		2100	16 5	0.8	0.4	0	-	62.0			Flue gas O_2 variation
				2		1840	7800	2.0		0	0	2650			300		470	15.8	2 0	0.0	0	-	53.7			
				3		1800	7800	3.0		0	0	2500			320	325	180	15.2	2.9	0.0	0	-	54.0			
				4	800	1770	7800	4.0		0	0	2200			340		60	16.6	4.0	0.0	0	-	48 2			
FBM 19	Not a limestone injection test																									
FBM 20	Ohio #8 Seam washed 2 6% Sulfur 7 2% Ash	Ash	13	1	800	1770	7800	3 0		0	0	2200	2130	25	340		180	15.0	2.8	0.0	0	-	55 0			
				2	880	1810	7800	3 0	1337M	112	1 17	1650	1590	0	260	280	180	14.9	2.8	0.0	25.0	21.4	47.0			
				3	900	1770	7800	3 0	-325	144	1 46	1300			238		180	14.9	2 9	0.0	40.0	27.5	43.5			
				4*	900	1710	7800	3.0		144	1.46	900			230		180	15.0	3 0	0.0	59.0	40.8	45.8			* 450 lb/hr Steam added to air inlet
FBM 21	Ohio #8 Seam washed 2.6% Sulfur 7 2% Ash	Ash	19	1	880	1680	7800	3.0		0	0	2250	2200	35	280	305	260	14.9	3.0	0.0	0	-	58.7		12.2	
				2	880	1680	7800	3.0	1337M	132	1.37	1000	850	0	280		260	15 2	3 0	0 0	56 0	41.0	44.3		13 8	
				3*	880	1680	7800	3.0	-325	132	1.37	850						15 2	3 0	0.0	62.0	44.0	46 0			*450 lb/hr Steam added to air inlet
FBM 22	Ohio #8 Seam washed 2 6% Sulfur 7 2% Ash	Ash	20	1	925	1650	7800	3.0		0	0	2300	2250	47	280	275	250	15 2	2.9	0.0	0	-	49 8		11.6[f]	
				2	915	1650	7800	3.0	1337M -325	145	1 46	750	680	0	280		250	14.8	2.6	0 0	68.0	45 0	42.7		14.2	
FBM 23	Ohio #8 Seam washed 2.6% Sulfur 7 2% Ash	Ash	19	1	800	1630	7400	3.0		0	0	2300			300	280	240	14.9	3 0	0.0	0	-	56 1	No	13.9[f]	
				2	800	1550	7400	3.0	1337M -325	282	2 4	800			300		240	16 2	2 8	0 0	65 3	27.2	42 8	No	15.3	
FBM 24	Ohio #8 Seam washed 2 6% Sulfur 7.2% Ash	Ash	22	1	760	1600	7400	3 0		0	0	2600	2400	17	260	285	210	15 2	2 8	0.0	0	-	54 0		9 3	
				2	800	1600	7400	3 0	1337M	280	2 4	450	560	0	260		210	16 5	3 0	0.0	81.8	36 1			13.6	
				3	800	1600	7400	3.0	-325	260	2.2	500	620	0	260		210	16 0	3.0	0.0	77.4	35.0	40 3			
				4	800	1600	7400	3.0		260	2.2	400			260		210	15 8	2 9	0 0	83 5	37 0	39 6	Yes		
FBM 25	Ohio #8 Seam unwashed 4 5% Sulfur 10.7% Ash	Ash	24	1	820	1550	7400	3.0		0	0	3750	3840	32	240	305	230	15.5	2 8	0.0	0	-	59 3		7.8	
				2	840	1550	7400	3 0	1337R	378	1 7	1100	1210	0	240		230	16 9	3.0	0 0	71 5	42 0	38 4		14.3	
				3	840	1520	7400	3.0	-325	378	1 7	950			240		230	16 5	3.2	0.0	74 2	43 8	36 7			
FBM 26	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	20	1	800	1660	7400	3 0		0	0	3750	3740	27	220		285	14.9	3 0	0 0	0	-	61 0		12.4	
				2	800	1660	7400	3.0	1337M	360	1 7	1350	1350	0	220		285	15 8	3 0	0 0	64 2	37 8	42.5		16 8	
				3	800	1580	7400	3.0	-325	400	1.9	1100	1160	0	220		285	16.0	3 2	0 0	70 9	37 3	37.8			
FBM 27	Ohio #8 Seam unwashed 4.5% Sulfur 10 7% Ash	Ash	20	1	765	1630	7400	3 0		0	0	3700	3630	35	320	320	260	14 8	3.2	0 0	0	-	55 4		10.5	
				2	765	1570	7400	3.0	1359B -325	220	2 0	930	980	0	320		260	16 1	2 9	0.0	74 0	37 0	43.8		16.5	

[a] Sorbent Data: Type and Size
[b] Gas Analysis, Concentration in volume percent or ppm as indicated: I R SO_2, ppm
[c] PDS MO_x, ppm
[d] Sorbent Utilization, %
[e] Fly Ash Data: Recirculation
[f]

TABLE B-1 (Continued)

Gas Analysis, Concentration in Volume Percent or ppm as Indicated

Test No	Coal Type and Composition	Bed Material	Static Bed Depth, inches	Test Cond No	Coal Rate, lb/hr	Bed Temp, °F	Gas Rate, lb/hr	Flue Gas O2 (Dry) %	Sorbent Data: Type and Size[a]	Sorbent Injection Rate, lb/hr	Sorbent Ratio, Ca/S	Wet Tests SO2, ppm	Wet Tests SO3, ppm	I R SO2, ppm[b]	PDS SO2, ppm[c]	Hydrocarbons, ppm	Orsat Analysis, % CO2	Orsat Analysis, % O	Orsat Analysis, % CO	SO2 Reduction (I R), %	Sorbent Utilization,[d] %	Fly Ash Data: Carbon Content %	Fly Ash Data: Recirculation[e]	Fly Ash Data: Emission, lb/hr	Remarks
FBM 28	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	20	1	745	1600	7400	3.0		0	0	2650	2820	0	260		260	15.5	2.9	0.0	0	-	51.3		8.9
				2	745	1600	7400	3.0	1359R	150	2.4	800	810	0	240		240				71.6	29.8	40.9		12.4
				3	745	1600	7400	3.0	-325	150	2.2	950	1020		260		260				64.9	29.4	43.4		
FBM 29	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	20	1	720	1600	7200	3.0		0	0	3770	3730	20	240	300	250	14.9	3.2	0.0	0	-	52.0		12.1
				2	720	1600	7400	3.0	1359R	175	1.7	1500	1480	0	240		250				60.0	35.2	51.8		
				3	720	1600	7400	3.0	-325	220	2.0	1000	1080		240		250				73.5	36.6	45.4		14.7
				4	720	1600	7400	3.0		220	2.0	1000			240		250				73.5	36.6		Yes	
FBM 30	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	20	1	740	1620	7400	3.0		0	0	2570	2540		250	285	260	15.2	3.1	0.0	0	-	57.2		7.7
				2	760	1620	7400	3.0	1359H	66	1.4	1290	1340		250		260				50.0	35.7	51.2		11.4
				3	760	1620	7400	3.0	-325	86	1.8	1020	1030		250		260				60.4	33.3	49.8		
FBM 31	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	20	1	820	1620	7600	3.0		0	0	2600	2630		270	305	350	15.3	2.9	0.0	0	-	62.7		11.2
				2	770	1620	7400	3.0	1359H	63	1.3	1200	1250		270		350				53.8	41.4	53.2		10.9
				3	800	1620	7600	3.0	-325	80	1.6	1000	1080		270		320				61.5	38.4	49.9		
FBM 32	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	20	1	700	1610	7400	3.0		0	0	2650	2690		290	310	250	14.8	3.1	0.0	0	-	50.9		10.9
				2	720	1610	7400	3.0	1359R	97	1.6	1070	1050		290		250				61.9	38.6	47.2		13.7
				3	720	1610	7400	3.0	-325	108	1.8	970	970		290		250				64.9	36.0	38.8		
				4	720	1610	7400	3.0		108	1.8	780			290		250				70.5	39.1	42.3	Yes	

NOTES

[a] Type Limestone identification and numbering system of Bituminous Coal Research, Inc
C = calcined by supplier, H = hydrated by supplier, R = raw stone

[b] Size: U S Standard sieve size

[c] Sorbent Utilization defined as

$$\% \text{ Utilization} = \frac{\% \text{ SO}_2 \text{ Reduction in gas}}{\text{Stoichiometric Ratio, Ca/S}}$$

[c] PDS designates Phenoldisulfonic Acid Method

[d] Gas analysis by Infrared Analyzer

[e] Fly ash not recirculated unless indicated by "Yes"

TABLE B-2 FBC TEST CONDITIONS AND DATA

Test No	Coal Type and Composition	Bed Material	Static Bed Depth, inches	Test Cond No	Coal Rate, lb/hr	Bed Temp, °F	Air Rate, lb/hr	Flue Gas Oxygen (Control), %	Sorbent Data: Type and Size[a]	Sorbent Data: Injection Rate, lb/hr	Sorbent Data: Ratio, Ca/S	Gas Analysis, Concentration in volume percent or ppm as indicated: Inlet SO_2, ppm[b]	Wet Tests: Outlet SO_2, ppm	Wet Tests: SO_3, ppm	Inlet NO, ppm	Inlet CO, ppm	Hydrocarbons, ppm	Orsat Analysis, %: CO_2	Orsat Analysis, %: O_2	Orsat Analysis, %: CO	SO_2 Reduction (Inlet), %	Sorbent Utilization,[d] %	Fly Ash Data: Carbon Content %	Fly Ash Data: Recirculation[e]	Fly Ash Data: Emission, lb/hr	Remarks
FBC 1	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash -8 +16	6	1	110	1750	990	SHAKE DOWN																		Shakedown Test
FBC 2	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash -8 +16	7	1 2 3 4 5	103 103 103 103 103	1750	880	1.0 3.0 1.0 3.0 5.0	None	0 0 0 0 0	0 0 0 0 0	4900* 4500 4900 4530 4000	4000	254	320 390 110 410 440		1250 150 1500 100 0	15.6	0.9	0.6	0 0 0 0 0	- - - - -	50.5			
FBC 3	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Serpentine Rock -24 +20	7	1 2	80	1550 1550	940	1.0 1.0	1337R -8 +16	0 24.0	0 1.15	5000 2300	3550 3400		420 395		850 850	16.5	1.0	0.0	0 61.6	- 53.5	35.6 33.0			
FBC 4	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Serpentine Rock -14 +20	7	1 2	80	1550 1450	940	1.0 3.0	1337R -8 +14	0 26	0 1.23	2650 1800	2630 1600		390 330		650 100	16.2 15.4	0.8 2.6	0.2 0.0	0 32	- 26.0	47.5 34.0			Continuation of FBC 3 SO_2 Calibration Gas incorrect, data discarded
FBC 5	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	7	1 2 3	105 105 105	1800 1800 1800	1000 1000 1000	1.0 1.0 3.0	1337R -7 +14	0 32.3 30.7	0 1.16 1.1	4450 3700 3150	3700 3750 3030	420 210 150	220 220 320		1000 500 50	16.0 17.0	1.0 1.0	0.0 0.2	0 17.0 23.2	- 15.4 21.0				
FBC 6	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	7	1 2 3 4	110 110 110 110	1700 1700 1700 1550	960 960 960 960	1.0 1.0 1.0 3.0	1337R -7 +14	0 35.0 35.0 35.0	0 1.2 1.2 1.2	4650 3800 3600 2600	3850 3650 3250 2450	420 110 50 120	350 330 380 460		500 500 490 50	16.1 17.2 17.5 15.5	1.0 1.0 1.2 3.1	0.0 0.0 0.0 0.0	0 18.5 22.6 39.0	- 15.4 18.8 36.6	48.2 23.6 32.0 30.0	Yes Yes		
FBC 7	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	6	1 2	115 115	1820 1900	1050 1050	1.0 2.0	None	0 0	0 0	4800 4450	4300	300	320 220			15.9 15.2	1.0 2.0	0.0 0.0	0 0	- -	49.5			
FBC 8	E. Ky Pike County 1% Sulfur 7% Ash 1.6% H_2	Ash	6	1 2 3 4	100 100 100 100	1800 1700 1600 1790	1000 1000 1000 1000	1.0 5.0 5.0 1.0	None	0 0 0 0	0 0 0 0	720 540 540 700	620 582 385 369	120 50 130 215	330 370 370 260			16.8 13.0 13.8 17.0	1.0 4.8 4.8 1.0	0.0 0.0 0.0 0.0	0 0 0 0	- - - -	83.5 73.3 74.1 62.4			Test for effect of coal nitrogen on NO emission
FBC 9	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	6	1 2 3	100 126 148	1800 1800 1800	1180 1130 1200	1.0 1.0 1.0	1337R -7 +14	0 46.0	0 1.17	4550 3850	3900 3650	450 205	250 210			15.8 16.8 16.0	1.0 0.4 1.0	0.2 0.6 0.0	0 0 15.0	- - 12.8	68.8 60.0 40.0			Transition to overbed air
FBC 10	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	7	1 2 3 4	99 104 107 100	1750 1750 1750 1720	870 940 940 880	1.0 1.0 1.0 1.0	1337R -7 +14	0 0 31.3 38.9	0 0 1.1 1.2	4650 4650 3600 3150	3850 2860 2650	150 80 95	320 240 290 360	200 146 130 342	410 430 430 390	16.0 16.4 16.8 17.0	1.2 1.0 1.2 1.0	0 0.2 0 0.2	0 0 22.5 32.2	- - 29.4 26.8	62.0 55.0 37.0 32.0			Reducing bed Oxidising bed
FBC 11	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	6	1 2 3 4	87 92 91 91	1700 1680 1650 1600	940 940 940 940	3.0 3.0 3.0 3.0	1337R -7 +14	0 27.9 49.2 62.0	0 1.14 2.04 2.56	4350 3150 1900 1500	4020 2900 1580 1490	410 290 210 130	280 320 320 330	170 127 85 94	30 30 30 30	15.2 15.6 16.2 16.2	2.4 2.6 2.6 3.0	0 0 0 0	- 27.6 56.3 65.5	- 24.2 27.7 25.6	56.0 - - 28.0			
FBC 12	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	9	1 2	114 105	1520 1540	1065 1080	2.0 2.0	1337R -7 +14	0 30.5	0 1.09	4500 2650	4300 0	170 0	280 280	48 50	230 250				0 41.0	- 37.8	65.0			
FBC 13	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	10	1 2 3	106 106 106	1500 1480 1480	1050 1050 1050	3.0 3.0 3.0	1337R -7 +14	0 22.3 40.6	0 .79 1.44	4350 3000 2000	4100 0 2050	170 0 60	285 285 285	390 370 409	90 100 80	14.6 15.3 16.2	3.0 2.9 2.4	0 0 0	0 31.0 54.0	- 39.2 37.4	69.5 39.1 37.3			
FBC 14	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash, Sleeve*	10	1 2 3 4	110 110 110 110	1900 1880 1840 1800	1050 1050 1050 1050	3.0 3.0 3.0 4.0	1337R -7 +14	0 32.2 50.0 50.0	0 1.1 1.7 1.7	4200 3750 3320 3000	0 0 0 0	0 0 0 0	260 260 235 330		%eglig %eglig %eglig %eglig				0 10.7 21.0 28.6	- 9.7 12.2 16.8				*Sleeve installed so as to reduce heat loss to waterwalls
FBC 15	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash, Sleeve	10	1 2	63 63	1720 1700	700 700	2.0 2.0	1337R -7 +14	0 16.8	0 1.0	4000 3100	0 0	0 0	190 200		230 120				0 22.5	- 22.5				
FBC 16	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash, Sleeve	10	1 2 3	66 66 66	1660 1620 1620	720 720 720	2.0 3.0 2.0	1337R -7 +14	0 17.5 28.8	0 1.0 1.6	4200 3450 2900	4000 3250 2780	180 145 120	350 330 290	124 121 119	200 250 160	15.8 16.2 16.3	2.2 2.0 2.0	0 0 0	0 17.8 31.0	- 17.8 19.4	48.2 31.1 31.5			
FBC 17	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Lost Ignition - Discarded																								
FBC 18	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash Coal Lumps	Ash, Sleeve	14	1 2 3	86 84 83	1760 1760 1670	800 800 800	2.0 2.0 2.0	1337R -7 +14	0 22.3 38.8	0 1.0 1.75	3600 3150 2500	3440 3050 2370	298 72 85	250 280 260	116 115 37	120 80 20	15.2 16.1 16.7	2.0 2.0 2.0	0 0 0	0 12.5 30.5	- 12.5 17.4	55.5 35.8 33.0			Notable NO decline. Small clinker formation when sorbent feed began

Gas Analysis, Concentration in volume percent or ppm as indicated

TABLE B-2 (Continued)

Test No.	Coal Type and Composition	Bed Material	Static Bed Depth, inches	Test Cond No	Coal Rate, lb/hr	Bed Temp, °F	Air Rate, lb/hr	Flue Gas Oxygen (Control),	Sorbent Data: Type and Size[a]	Sorbent Data: Injection Rate, lb/hr	Sorbent Data: Mole Ratio, Ca/S	Gas Analysis, Concentration in volume percent or ppm as indicated: I R SO_2,[b] ppm	Wet Tests: SO_2, ppm	Wet Tests: SO_3, ppm	I R. NO, ppm	PDS NO_x,[c] ppm	Hydrocarbons, ppm	Orsat Analysis, %: CO_2	Orsat Analysis, %: O_2	Orsat Analysis, %: CO	SO_2 Reduction (I R), %	Sorbent Utilization,[d] %	Fly Ash Data: Carbon Content %	Fly Ash Data: Recirculate[e]	Emission, lb/hr	Remarks
FBC 19	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	118	1540	1120	3.0		0	0	3900	3960	37	260	123	85	15.0	2.9	0	0	-				
				2	119	1540	1120	3.0	1359R	18.2	1.15	3300			250	110	95	15.8	2.8	9.2	15.4	13.4				
				3	110	1510	1100	3.0	-7 +14	30.8	1.93	2830			250	125	90	16.6	2.8	0	27	14.0				
				4	110	1480	1100	3.0		30.8	1.93	2350	2250	62	220		90	16.2	2.8	0	40	20.7		Yes		
FBC 20	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	118	1740	1070	3.0		0	0	3800	3870	40	280	310	70	14.8	2.4	0	0	-		52.3		
				2	107	1700	1070	3.0	1359R	17.8	1.15	3350	3140	10	250	300	90	15.6	3.2	0	11.8	10.2		45.0		
				3	112	1740	1040	3.0	-7 +14	24.7	1.47	2750	2820	40	220	280	85	15.8	3.2	0	26.6	18.1		51.1		
				4	110	1650	1070	3.0		32.2	2.0	2500			220		90	15.8	3.1	0	34.2	17.1				
				5	110	1640	1070	3.0		32.7	2.0	2400			220		70	15.0	3.0	0	39.5	19.7		Yes		
FBC 21	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	110	1530	1100	3.0		0	0	3800	3750	150	225		80	13.7	3.1	0	0	-		60.9		
				2	115	1520	1100	3.0	1359R	14.8	9	3300	3200	70	240		60	15.2	2.8	0	13.2	14.7		56.0		
				3	112	1530	1100	3.0	-7 +14	25.6	1.57	2920	2900	10	210		60	15.9	3.2	0	23.1	14.1		41.2		
				4	112	1500	1100	3.0		35.0	2.0	2900	0	0	200		60				23.6	12.8				
				5	120	1520	1100	3.0		35.0	2.0	2607	0	0	220		60				31.5	15.6		Yes		
FBC 22	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	7	1	116	1650	1100	3.0		0	0	3800	3560	120	320	112	80	13.8	2.9	0	0	-		60.1		
				2	112	1650	1100	3.0	1359R	16.3	1.0	3300	3108	54	320	114	80	15.2	2.3	0	15.4	15.4		59.7		
				3	112	1640	1100	3.0	-7 +14	31.2	1.9	2720	2510	21	270	115	20	15.6	3.0	0	30.8	16.2		45.4		
FBC 23	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	101	1620	1080	3.0		0	0	4180			330	500	40	15	2.8	0	0	-	23	Yes		
				2	107	1600	1080	3.0	1337C	11.8	1.78	2950			300		40	15.2	2.8	0	31.8	24.8	23.0	Yes		
				3	103	1580	1080	3.0	-7 +14	20	3.17	2700	2550	120	260	131	40	14.8	2.3	0	35.6	16.4	25.7	Yes		
				4	108	1580	1080	3.0		32.6	2.54	2950	1980	137	260		40	15	2.7	0	33	20	27.0	Yes		
				5	108	1580	1080	3.0		21	2.17[a]	1720			240	133	160	14.2	3.0	0	59	27.2		Yes	[a]Water injected into gas space above bed	
FBC 24	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	7	1	99	1680	1150	3.0		0	0	3500	3580	0	410	670	120	15	2.8	0	0	-	42.6	Yes		
				2	97	1680	1120	3.0	1337C	17.2	1.13	2730	2850	0	320		90	16	2.6	0	16.1	14.4	33.9	Yes		
				3	87	1680	1120	3.0	-7 +14	26.2	1.73	1610	2260	0	280	582	40	15.5	2.8	0	31.2	18.0	31.8	Yes		
				4	101	1680	1350	3.0		34.0	2.26	2250			260	525	20	16	2.8	0	35.6	15.7		Yes		
				5	101	1680	1350	3.0		-	-[a]	2400			260		150	14.4	3.5	0	31.4	-		Yes	[a]Water injected into gas space. Sorbent feed data unavailable	
FBC 25	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	126	1750	1080	3.0		0	0	3650	3600	100	200		40	14.8	2.8	0	0	-	47.9			
				2	120	1720	1050	3.0	1337C	20.4	1.15	3350	3400	70			40	15.1	2.8	0	8.2	7.1	27.9			
				3	125	1720	1080	3.0	-7 +14	29.8	1.6	3180	3150	50	195		40				13.9	8.1	24.3			
FBC 26	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	123	1710	1100	3.0		0	0	3800			370		40	13.9	3.2	0	0	-	43.7			
				2	116	1700	1100	3.0	1359C	13.7	1.5	3300			330		40	15	2.9	0	13.2	11.4	40.3			
				3	123	1700	1090	3.0	-7 +14	22.6	2.3	2950			330		40	14.8	3	0	22.4	10.3	39			
				4	119	1695	1120	3.0		32.2	3.4	2370			288		40	14.8	3.2	0	37.5	11.0	36.4			
FBC 27	Northern W. Va Coal 5% Sulfur	Ash	8	1	130	1540	1090	3.0		0	0	4300	4150	185	290	305	120	12.4	3.8	0	0	-	44.5			
				2	125	1580	1090	3.0	KHL[a]	35.2	1.3	3880			310		90	13	2.9	0	9.8	7.6	52.5			
				3	126	1550	1098	3.0	-7 +14	38	1.4	2750			290		90	15.5	2.6	0	36	25.4	47.9			
FBC 28	Northern W. Va. Coal 5% Sulfur	Ash	6	1	122	1760	1100	3.0		0	0	4520	4250	230	280	165	250	14.2	3.0	0	0	-	38.8			
			6	2	116	1760	1100	3.0	KHL[a]	18.8	.8	4100			330		250	15.6	3.0	0	9.3	11.6	32.5			
			6	3	115	1748	1100	3.0	-7 +14	42.2	1.75	3900			300		120	16.5	2.8	0	13.7	7.8	28.2			
			8	4	108	1640	1108	3.0		42.4	1.75	3000			280		90	15	2.8	0	33.6	19.2	51.3			
FBC 29	Injection of Hydrate into Plenum - Buttons Plugged No Data Obtained																									
FBC 30	Premix of 1359 Hydrate with Coal - Coal Feeder Plugged No Data Obtained																									
FBC 31	Premix of 1359 Hydrate with Coal - Coal Feeder Plugged No Data Obtained																									
FBC 32	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	118	1580	1050	3.0		0	0	3700	3650	100	240	270	50	14.8	2.7	0	0	-	48.7		No. 1 Feeder	
				2	110	1580	1050	3.0	1359R -325	13	1.1	2200	2200	20	240	265	50	14.8	2.8	0	40.5	16.8	28.4			
FBC 33	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	9	1	111	1600	1050	3.0		0	0	3750			340		50	13.7	3.3	0	0	-	48.8		No. 1 Feeder	
				2	106	1608	1050	3.0	1337B -325	21.4	1.05	2000			380		50	14.5	3.0	0	46.7	44.5	19.7			
FBC 34	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	9	1	105	1560	1050	3.0		0	0	3550	3500	117	320		50	13.8	3.0	0	0	-	38.2		No. 2 Feeder	
				2	109	1560	1050	3.0	1359R 56% -200	12.8	.81	2550	2530	0	320		50	14.3	3.1	0	29.2	34.8	31.8			
FBC 35	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	9	1	119	1560	1050	3.0		0	0	3600	3250	213	360	380	100	14.8	2.7	0	0	-	46.5			
				2	123	1560	1050	3.0	1359R	13.4	1.04	2200	0	0	360		100	15.2	3.0	0	39.0	37.5	30.0		No. 1 Feeder	
				3	114	1560	1050	3.0	-325	13.4	1.1	1950	1798	88	360	370	100	15.2	3.0	0	46.0	42	38.6		No. 2 Feeder	
				4	113	1560	1050	3.0		25.9	2.15	1408	0	0	360		100	15.1	3.1	0	61.2	28.4	38.0		No. 1 and No. 2 Feeders	
FBC 36	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash (1/8" x 0)	Ash	9	1	115	1540	1080	3.0		0	0	3550	3440	118	380		100	13.5	3.6	0	0	-	44.5			
				2	111	1540	1080	3.0	1359R	13.8	1.16	1700	1570	147	380		100	14.6	2.6	0	52.2	44.3	32.4		No. 1 Feeder	
				3	113	1580	1080	3.0	-325	16.5	1.36	1700	0	0	360		100	14.5	3.0	0	51.2	37.9	19.2		No. 2 Feeder	
				4	115	1580	1080	3.0		30.4	2.55	800	0	0	380		100				78.0	31.4	57.4		No. 1 and No. 2 Feeders	

A

2

TABLE B-2 (Continued)

Test No.	Coal Type and Composition	Bed Material	Static Bed Depth, inches	Test Cond No.	Coal Rate, lb/hr	Bed Temp °F	Air Rate, lb/hr	Flue Gas Oxygen (Control),	Sorbent Data: Type and Size[a]	Sorbent Data: Injection Rate, lb/hr	Sorbent Data: Mole Ratio, Ca/S	Gas Analysis, Concentration in volume percent or ppm as indicated: I R SO_2,[b] ppm	Wet Tests: SO_2, ppm	Wet Tests: SO_3, ppm	I R NO, ppm	PDS NO_x,[c] ppm	Hydrocarbons, ppm	Orsat Analysis, %: CO_2	Orsat Analysis, %: O_2	Orsat Analysis, %: CO	NO_x Reduction (I R),	Sorbent Utilization,[d]	Fly Ash Data: Carbon Content	Fly Ash Data: Recirculation[e]	Fly Ash Data: Emission, lb/hr	Remarks
FBC 37	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	116	1640	1020	3.0		0	0	3400	3230	163	390	380	50	15.2	2.6	0	0	-	28.7			No 1 Feeder
				2	104	1640	1020	3.0	1359#	14.8	1.34	3000	1800	87	385	380	50	15.0	3.0	0	41.2	30.8	13.5			No. 2 Feeder
				3	111	1620	1020	3.0	-325	15.8	1.34	2250	2200	44	370	320	50	14.9	2.9	0	34.0	25.2	6.5			No 1 and No 2 Feeders
				4	115	1630	1020	3.0		30.5	3.84	1500			370		50	15.2	2.8	0	56.0	31.2	7.0			
FBC 38	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	8	1	86	1760	890	3.0	1359#	0	0	3500	3330	262	375	380	50				0	-				No 1 Feeder
				2	86	1760	890	3.0	-325	16.0	1.75	1470			375		50				57.9	33.1				
				3	103	1720	1020	3.0		16.0	1.65	1780	1650	63	375		50				50.9	35.2				
				4	118	1725	1170	3.0		16.0	1.28	1230			375		50				50.0	39.2				
				5	129	1760	1350	3.0		16.0	1.16	2030			375		50				42.0	35.2				
FBC 39	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	9½	1	120	1540	1080	3.0	1337#	0	0	3400	3400	85	360	385	50	13.5	3.5	0	0	-	33.0			No 1 Feeder
				2	116	1540	1080	3.0	-325	30.6	1.38	1600			360		50	14.8	3.0	0	53.0	38.6	24.6			No. 2 Feeder
				3	117	1580	1080	3.0		42.0	1.87	400	288	52	364		50	14.8	2.9	0	88.5	47.4	27.6			Premixed
				4	118	1540	1080	3.0		26.5[d]	1.17	1650	1250	131	358		50	14.6	2.8	0	51.5	44.0	27.0			
FBC 40	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	7 to 6½	1	111	1760	1080	3.0	1337#	0	0	3550			375	390	50	14.5	2.8	0	0	-	22.0			No 1 Feeder
				2	105	1740	1080	3.0	-325	31.2	1.55	1250	1230	122	375		50	14.8	3.0	0	64.7	41.8	6.3			No 2 Feeder
				3	109	1720	1080	3.0		43.7	2.08	650			375		50	14.2	2.6	0	83.9	39.4	31.7			*Premixed with coal. 22.5 pounds limestone per 100 pounds coal
				4	-	1760	1080	3.0		22.5[d]	1.17	1800	1640	139	375		50	14.5	2.9	0	49.2	42.0	6.0			
FBC 41	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	10 to 9½	1	120	1560	1080	3.0	1359#	0	0	3450	3400	100	375	350	50	14.9	2.9	0	0	-	34.4			No. 1 Feeder
				2	111	1560	1080	3.0	-325	40.6	3.4	350	150	0	375		50	14.9	2.7	0	90	76.6	20.0			No. 2 Feeder
				3	114	1600	1080	3.0		27.4	2.24	1000	620	0	375		50	15.0	2.9	0	71.0	31.8	16.0			*Premixed with coal. 22.5 pounds limestone per 100 pounds coal
				4	-	1600	1080	3.0		22.5[d]	2.10	900	900	0	375	398	50	13.2	3.0	0	74.0	35.5	9.1			
FBC 42	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	9½	1	116	1600	1080	3.0	1359#	0	0	3350	3250	150	375	380	50	14.9	3.0	0	0	-	35.4			
				2	112	1640	1080	3.0	-325	19.7	1.45*	1150			375		50	14.8	2.9	0	65.5	39.6	20.5			
				3	-	1600	1080	3.0		30.0**	2.8**	600			375		50	14.8	2.9	0	82.2	29.4	10.9			*Fed from Port A only. Then fed from Port B only. **Premixed with coal 30 pounds limestone per 100 pounds coal
FBC 43	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	10 to 10½	1	116	1580	1080	3.0		0	0	3400	3150		378	110	50	14.8	3.0	0	0	-	24.8			
				2	-	1580	1080	3.0	1359# 65% -200	30.0*	2.07	2050			375	80	50	15.5	3.0	0	39.7	19.2	22.4			Premixed with coal 30 pounds limestone per 100 pounds coal
FBC 44	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	12	1	129	1620	1080	1.0	No Additive			2650	1950	130	360	380	400	16.0	1.2	0.2	0	-				
				2	118	1590	1080	2.0				2250	1600	200	370	380	300	15.8	2.0	0	0	-				
				3	112	1630	1080	3.0				2000	800	0	380	395	100	14.9	3.1	0	0	-				
				4	124	1740	1080	1.0				3400			305		860	16.3	1.3	0	0	-				
			8	5	116	1730	1080	2.0				2250			368		150	16.0	1.6	0	0	-				
				6	106	1700	1080	3.0				3000			400		50	13.5	3.0	0	0	-				
				7	118	1680	1080	1.0				2500			380		840	15.9	.8	0.4	0	-				
			5	8	112	1620	1080	2.0				2200			340		150	16.0	1.6	0	0	-				
				9	104	1780	1080	3.0				2000			380		50	16.0	2.7	0	0	-				
FBC 45	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	10	1	105	1630	1080	3.0		0	0	2000			388	270	50	14.8	3.2	0	0	-				No 1 Additive Feeder
				2	102	1600	1050	3.0	1359#	31.8	5.2	50			340	160	50	14.2	3.1	0	96.0	18.4	32.8			No 2 Additive Feeder
				3	106	1670	1050	3.0	-325	28.6	4.3	300	240	0	330	50	50	15.4	3.0	0	85.0	19.8	38.5			
FBC 46	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	11	1	115	1580	1050	3.0		0	0	2200	1970	0	360	140	50	14.8	3.0	0	0	-	52.0		1.9	
				2	108	1600	1050	3.0	1359#	24.0	3.6	270			380	130	50	15.0	3.0	0	85.0	23.6	26.9		4.2	No. 1 Additive Feeder
				3	105	1640	1050	3.0	-325	37.2	2.6	550	480	0	360	160	50	14.5	2.7	0	75.0	28.8	39.4			No 2 Additive Feeder
				4	-	1650	1050	3.0		13.0*	2.1	760			360	80	50				65.5	31.2	21.0			*Premixed with coal. 13.0 pounds limestone per 100 pounds coal
FBC 47	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	6	1	102	1780	1050	3.0		0	0	2200	2100	30	360		50	14.8	2.6	0	0	-	46.8			
				2	103	1780	1050	3.0	1359#	12.8	2.0	1500	1440	0	360		50	15.0	3.0	0	31.8	15.9	21.6			No 1 Additive Feeder
				3	106	1850	1050	3.0	-325	9.6	1.4	1700	1610	0	360		50	15.0	2.8	0	22.7	18.2	25.8			No 2 Additive Feeder
				4	-	1800	1050	3.0		16.5*	2.6	1400	1350	0	360		50	14.6	2.6	0	36.3	13.9	19.8			*Premixed with coal. 16.5 pounds limestone per 100 pounds coal
FBC 48	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	12	1	108	1570	1050	3.0		0	0	2000	1950	32	330		50	15.0	3.0	0	0	-	48.2		1.2	
				2	104	1585	1050	3.0	1337#	16.0	0	800	780	0	330		50	14.5	3.2	0	60.0	42.8	33.3			No 1 Additive Feeder
				3	110	1600	1050	3.0	-325	19.4	1.6	500			330		50	14.5	2.6	0	75.0	48.9	29.3		2.6	No 2 Additive Feeder
				4	-	1590	1050	3.0		12.7*	1.16	1000			330		50	14.7	2.6	0	59.0	43.0	21.8			*Premixed with coal. 12.7 pounds limestone per 100 pounds coal
FBC 49	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	6	1	100	1780	1050	3.0		0	0	2350			380	405	50	15.0	2.8	0	0	-	23.9			No. 1 Additive Feeder
				2	101	1760	1050	3.0	1337#	17.3	1.6	1400			380		50	14.7	3.0	0	40.6	25.2	22.4			No. 2 Additive Feeder
				3	112	1770	1050	3.0	-325	18.0	1.46	1550			380		50	14.4	2.8	0	38.0	24.3	39.6			*Premixed with coal. 12.7 pounds limestone per 100 pounds coal
				4	-	1790	1050	3.0		12.7*	1.16	1550			380		50	14.2	3.2	0	34.0	29.2	19.5			
FBC 50	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	12	1	105	1560	1050	3.0		0	0	1900	1880	40	370	395	50	14.5	3.0	0	0	-	46.8			
				2	101	1570	1050	3.0	1337#	14.7	1.35	800	860	0	370		50	14.5	2.9	0	57.9	43.5	53.5			No 1 Additive Feeder
				3	110	1580	1050	3.0	-325	18.6	1.55	450	490		370		50	15.0	3.0	0	76.4	49.0	31.6			No 1 Additive Feeder
				4	-	1570	1050	3.0		16.0*	1.46	930	910		370		50				51.0	35.0	34.9			*Premixed with coal 16.0 pounds limestone per 100 pounds coal
FBC 51	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	12	1	109	1560	1050	3.0		0	0	2100	2080	31	380	365	50	14.8	3.4	0	0	-				No 1 Additive Feeder
				2	102	1580	1050	3.0	1337#	15.6	1.4	1100	1050	0	380		50	15.2	3.2	0	47.5	34.0				No 2 Additive Feeder
				3	109	1600	1050	3.0	-325	19.0	1.6	700	650		380		50	15.2	3.0	0	66.7	41.7				*Premixed with coal 16 pounds limestone per 100 pounds coal
				4	-	1600	1050	3.0		16.0*	1.46	1110	1100		380		50	15.5	3.2	0	47.5	32.8				
FBC 52	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	12	1	111	1540	1050	3.0		0	0	2000	1890	45	375	398	50	15.0	2.9	0	0	-			1.9	
				2	105	1570	1050	3.0	1337#	21.8	1.9	700	690	0	375		50	14.8	3.1	0	65.0	34.2			4.7	No. 1 Additive Feeder
				3	111	1600	1050	3.0	-325	19.6	1.6	600	660		375		50	14.8	2.9	0	70.0	43.5			4.6	No. 2 Additive Feeder
				4	-	1690	1050	3.0		18.1*	1.65	900	830		375		50	15.0	3.2	0	55.0	33.3				*Premixed with coal. 18.1 pounds limestone per 100 pounds coal
FBC 53	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	11	1	110	1560	1050	3.0		0	0	2050	2000	15	378	398	50				0	-				
				2	107	1595	1050	3.0	1337#	31.5	2.0	550	460	0	378	75	50				73.5	36.7				No. 2 Feeder
				3	105	1570	1050	3.0	-325	18.7	1.2	1050									48.0	40.0				

1

2

TABLE B-2. (Continued)

Test No	Coal Type and Composition	Bed Material	Static Bed Depth, inches	Test Cond No	Coal Rate, lb/hr	Bed Temp., °F	Air Rate, lb/hr	Flue Gas Oxygen (Control),	Sorbent Data: Type and Size[a]	Sorbent Data: Injection Rate, lb/hr	Sorbent Data: Mole Ratio, Ca/S	Gas Analysis, Concentration in volume percent or ppm as indicated: I.R. SO_2,[f] ppm	Wet Tests: SO_2, ppm	Wet Tests: SO_2, ppm	I.R. NO, ppm	I.R. FDS NO_x, ppm[c]	Hydrocarbons, ppm	Orsat Analysis, %: CO_2	Orsat Analysis, %: O_2	Orsat Analysis, %: CO	SO_2 Reduction (I.R.),	Sorbent Utilization,[d]	Fly Ash Data: Carbon content	Fly Ash Data: Recirculation[e]	Emission, lb/hr	Remarks
FBC 54	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	11	1 2 3	110 110 -	1570 1565 1565	1050 1050 1050	3.0 3.0 3.0	1359R 65% -200	0 27.7 16.8*	0 3.0 2.0	2150 1100 1300	2000 980 1320	95	360 360 360		50 50 50	14.6 14.2 15.2	2.8 2.6 2.8	0 0 0	.0 48.9 39.5	- 16.3 19.7			1.5 2.9	*No. 2 Feeder *Premixed with coal. 16.8 pounds limestone per 100 pounds coal
FBC 55	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	10	1 2 3 4	110 112 116 118	1580 1620 1580 1530	1050 1050 1050 1050	1.0 2.0 3.0 4.0	None	0 0 0 0	0 0 0 0	2800 2400 2130 2050	2400 2320 2290 2180	0 0 41 52	320 380 390 420		1530 200 50 0	16.5 15.9 15.1 14.0	5 2.1 2.9 4.0	4 0 0 0	0 0 0 0	- - - -			1.4 1.6	
FBC 56	Ohio #8 Seam unwashed 4.4% Sulfur 10.7% Ash	Ash	10	1 2	118 118	1580 1560	1050 1050	3.0 3.0	1337R -325	0 35.0	0 1.12	3550 1750	3450 1700	175 6	380 380		50 50	14.8 15.5	3.2 3.0	0 0	0 49.0	- 43.6			2.0 5.8	
FBC 57	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	10	1 2	116 113	1590 1570	1050 1050	3.0 3.0	1337R -325	0 37.6	0 1.25	3550 1400	3560 1400	32 3	380 380	360	50 50	15.- -	3.1	6 -	0 60.7	- 43.5			2.5 6.8	
FBC 58	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	10	1 2 3	111 111 112	1570 1570 1570	1050 1050 1050	3.0 3.0 3.0	1337R -325	0 46.0 46.0	0 1.57 1.57	3600 1000 1300	3620 950	37 0	378 378 378	395	50 50 50	14.8 15.7 15.5	2.9 3.1 3.1	0 0 0	0 72.2 72.2	- 45.4 45.4	 Yes		1.9 7.8 6.9	
FBC 59	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	10	1 2 3	109 109 109	1590 1590 1590	1080 1080 1080	3.0 3.0 3.0	1359R -325	0 37.7 37.7	0 2.33 2.33	2350 650 650	630	20 0	380 380 380	385 385	50 50 50	15.2	3.4	0	0 72.4 72.4	- 30.7 30.7	 Yes		2.1 6.5 5.8	
FBC 60	Ohio #8 Seam washed 2.6% Sulfur 7.2% Ash	Ash	10	1 2 3	110 110 110	1580 1550 1550	1080 1080 1080	3.0 3.0 3.0	1359R -325	0 18.5 12.0	0 2.0 1.3	2300 900 1250	2500 850	25 0	348 340 340	293	50 50 50	14.5 15.2 14.8	3.2 3.0 3.0	0 0 0	0 60.8 46.0	- 30.4 45.0	52.0 43.0		1.3 4.2 -	
FBC 61	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	10	1 2 3	120 118 118	1590 1540 1520	1080 1080 1080	3.0 3.0 3.0	1359R -325	0 21.4 21.4	0 1.25 1.25	3550 1900 1900	1900 1850	52 0 0	300 300 300	305	50 50 50	14.8 15.4 15.7	3.1 2.9 2.8	0 0 0	0 45.4 45.4	- 37.3 37.3	35.0 33.0	 Yes	1.5 3.9 4.9	
FBC 62	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	12	1 2 3 4	119 120 120 110	1580 1550 1520 1520	1080 1080 1080 1080	3.0 3.0 3.0 3.0	1359 R -325	0 28.0 28.0 41.5	0 1.6 1.6 2.6	3550 1500 1500 830	3500 1600 1550	62 0 0	350 350 350 350	370 370 [b]	50 50 50 50				0 57.8 57.8 77.0	- 36.0 36.0 29.8	 Yes		2.4 4.8 4.7	
FBC 63	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	12	1 2	115 117 119 113 119 120	1540 1550 1560 1560 1560 1560	 1080 1080 1080 1080 1080	3.0 3.0 3.0 3.0 3.0 3.0	 1359R -325	0 0 0 0 0 27.8	0 0 0 0 0 1.6	2800 3800 3800 3800 3800 1400	- 3800 3780 3690 3740 1410	- 5 15 5 38 0	 460 460 460 460 460	 475	40 40 40 40 40 40	15.5 15.2 15.4 15.4 15.5 16.0	2.5 2.5 2.6 2.7 2.6 2.6	0 0 0 0 0 0	0 0 0 0 0 57.0	- - - - - 35.6	48.0 33.0		2.6	SO_2 Test
FBC 64	Ohio #8 Seam unwashed 4.5% Sulfur 10.7% Ash	Ash	12	1 2 3	121 120 117	1550 1550 1550	1080 1080 1080	3.0 3.0 3.0	1359C -325	0 25.0 14.2	0 2.6 1.5	3750 2200 2560	3700 2280 2620	28 0 0	400 400 408	390	40 40 40	15.2 - -	2.9	0	0 41.2 30.8	- 15.8 21.1	 		1.9 3.2 2.5	
FBC 65	Ohio #8 Seam unwashed 3.0% Sulfur 10.7% Ash	Ash	10	1 2 3 4	119 113 116 114	1540* 1540 1540 1540	1080 1080 1080 1080	3.0 3.0 3.0 3.0	1359R -12 +14 -14 +16 -16 +18	0 28.0 28.0 29.0	0 2.6** 2.6 2.6	2500 1870 1870 1700	2390 1870	0	340 340 340 340		50 50 50 50				0 28.0 28.0 28.0	- 10.8 10.8 13.1	- 70.0 70.0 65.0			*Temperature was varied during test. **See Text for description of tests.
FBC 66	Ohio #8 Seam unwashed 3.0% Sulfur 10.7% Ash	Ash	10	1 2 3 4	113 113 116 117	1550* 1550 1550 1550	1080 1080 1080 1080	3.1 3.1 3.1 3.1	1359R -20 +30 -40 +50 -325	0 26.9 28.2 28.2	0 2.4 2.6 2.6	2500 1700 1940 760	2600 1680 1700 650		400 408 408 400	- - - 320	50 50 50 50				0 32.0 34.0 72.0	- 13.2 9.3 27.7	65.5 48.0 48.6 42.6			*Temperature was varied during test **See Text for description of tests.
FBC 67	Ohio #8 Seam unwashed 2.9% Sulfur 10.7% Ash	Ash	18	1 2 3 4	65 69 69 65	1520* 1570 1520 1520	620 620 620 620	2.9 2.9 2.9 3.0	1359R -20 +30 -40 +50 -325	0 16.0 16.0 16.7	0 2.5** 2.5 2.6	2450 1250 1600 550	2450 - - -		400 380 460 460		50 50 50 50				0 49.0 34.6 77.0	- 19.6 13.8 30.8	- 65.7 63.8 58.8			*Temperature was varied during test. **See Text for description of tests.
FBC 68	Ohio #8 Seam unwashed 2.95% Sulfur 10.7% Ash	Ash	10	1 2 3 4	62 63 63 61	1790* 1730 1800 1780	620 620 620 620	3.0 3.0 3.0 3.0	1359R -325 -40 +50 -20 +30	0 15.5 16.8 16.2	0 2.6** 2.8 2.7	2550 1700 2230 2150			348 380 380 380		50 50 50 50				0 11.0 12.5 15.5	- 11.9 4.5 5.7	56.2 57.8 42.6 52.5			*Temperature was varied during test. **See Text for description of tests.
FBC 69	Ohio #8 Seam unwashed 3.01% Sulfur 10.7% Ash	Ash	18	1 2 3 4 5	65 65 65 65 65	1800* 1760 1760 1790 1730	620 620 620 620 620	3.0 3.0 3.0 3.0 3.0	1359R -20 +30 -40 +50 -100 +200 -325	0 17.0 16.4 17.0 15.8	0 2.7** 2.6 2.7 2.5	2560 2150 2230 1700 1700			480 480 440-480 490		50 50 50 50 50				- 15.1 11.0 32.0 32.0	- 5.6 4.2 11.9 12.8	66.8 53.9 46.0 47.2 56.2			*Temperature was varied during test. **See Text for description of tests.
FBC 70	Ohio #8 Seam unwashed 3.02% Sulfur 10.7% Ash	Ash	10 10 10 10 10	1 2 3 4 5	65 62 65 65 69	1620* 1550 1820 1550 1800	620 620 620 620 620	3.0 3.0 3.0 3.0 3.0	1359R -100 +200	0 16.8 17.6 17.6 18.1	0 2.8** 2.8 2.8 2.8	2530 1750 2100 1250 1900			480 480 450 440 460		50 50 50 50 50				0 30.0 13.7 50.0 23.3	- 10.7 4.8 17.9 8.3	50.0 58.1 59.0 - -			*Temperature was varied during test. **See Text for description of tests.
FBC 71	Ohio #8 Seam unwashed 3.0% Sulfur 10.7% Ash	Ash	10	1 2 3 4 5	66 65 66 65 65	1650* 1650 1630 1670 1660	620 620 620 620 620	2.9 2.9 2.9 3.0 3.0	1359R -20 +30 -40 +50 -100 +200 -325	0 16.3 16.7 16.3 16.3	0 2.6** 2.8 2.7 2.6	2500 1950 2100 2000 1350	- - - - 1400		370 380 380 340 370	 375	50 50 50 50 50				0 22.0 18.6 22.5 49.0	- 8.5 7.2 8.7 18.8	56.3 48.5 37.0 42.0 49.0			*Temperature was varied during test. **See Text for description of tests.

TABLE B-2 (Continued)

Test No.	Coal Type and Composition	Bed Material	Static Bed Depth, inches	Test Cond No.	Coal Rate, lb/hr	Bed Temp., °F	Air Rate, lb/hr	Flue Gas Oxygen (Control), %	Sorbent Data: Type and Size[a]	Sorbent Data: Injection Rate, lb/hr	Sorbent Data: Mole Ratio, Ca/S	Gas Analysis, Concentration in volume percent or ppm as indicated: I.R. SO_2, ppm[b]	Wet Tests: SO_2, ppm	Wet Tests: SO_3, ppm	I.R. NO, ppm	PDS NO_x, ppm[c]	Hydrocarbons, ppm	Orsat Analysis, %: CO_2	Orsat Analysis, %: O_2	Orsat Analysis, %: CO	SO_2 Reduction (I.R.), %	Sorbent Utilization,[d] %	Fly Ash Data: Carbon Content %	Fly Ash Data: Recirculation[e] %	Emission, lb/hr	Remarks
FBC 72	Ohio #8 Seam washed 3.04% Sulfur 10.7% Ash	Ash	18	1		1700*	620	3.0	1359R	0	0	2800			410		20	-	-	-	0	-				*Temperature was varied during test **See Text for Description of tests.
				2	71.5	1600	620	3.0	-20 +30	18.2	2.6**	1500			400		20	15.0	3.0	0.0	46.4	17.8	42.0			
				3	80.5	1640	620	3.0	-40 +50	21.2	2.7	1800			460		20	-			35.7	13.2	46.5			
				4	78.0	1640	620	3.0	-100 +200	19.8	2.6	1600			430		20	-			42.8	16.5	44.0			
				5	71.0	1660	620	3.0	-325	19.5	2.8	1100			460		20	-			60.7	21.9	45.0			
FBC 73	Ohio #8 Seam washed 2.9% Sulfur 10.7% Ash	Ash	10	1		1570	1050	3.0	1359R	0	0	2500			460		20	15.4	3.0	0.0	0	-				Additive Feed 1 side (Cond. 2)
				2	103	1550	1050	3.0	-325	19.2	2.0	1050	1050		450	450	20	15.4	3.0	0.0	58.0	29.0	30.3			Additive Feed 2 sides (Cond 3)
				3	116	1550	1050	3.0		21.7	2.0	1100	1100		450		20	-			56.0	28.0	26.8			Additive Feed 4 sides (Cond 4)
				4	107	1550	1050	3.0		20.0	2.0	1100			450		20	-			56.0	28.0	26.4			Additive Feed 2 sides opposite (Cond 5)
				5	108	1550	1050	3.0		20.2	2.0	1100			450		20	-			56.0	28.0	28.4			Additive Feed 4 sides (Cond 6)
				6	114	1550	1050	3.0		21.9	1.0 + 2.0	650					20	-			74.0	24.7	-			1.0 Ratio 2 sides, 2.0 Ratio 2 sides, Opposite (3 total)
FBC 74	Ohio #8 Seam washed 3.0% Sulfur 10.7% Ash	Ash	10	1	115	1550	1050	2.9	1359R	0	0	2400			420		20	-			0	-	58.0			
				2	117	1570	1050	2.9	-325	22.6	2.0	1050			420		20	14.8	2.8	0.0	56.3	28.3	35.6			*Additive premixed with coal (coal weight includes limestone)
				3	150*	1550	1050	3.0		24.5	2.0*	1580	1580		470	450	20	-			34.2	17.1	38.0			Dual Additive Feed
FBC 75	Ohio #8 Seam washed 3.0% Sulfur 10.7% Ash	Ash	10	1	110	1560	1050	3.0	1359R	0	0	2680			380		20	15.2	3.0	0.0	0	-				*Point of injection varied Above bed feed (into gas space) during Cond 2 Base of bed feed during Cond. 3
				2	116	1580	1050	3.0	-325	11.2*	1.75	1800			380		20	-	-	-	32.8*	18.75*				
				3	116	1570	1050	3.0		11.2	1.75	1040	1030	0	380	412	20	14.8	3.2	0.0	61.2*	35.0				
FBC 76	Ohio #8 Seam washed 3.05% Sulfur 10.7% Ash	Ash	10	1	115	1550	1050	2.9	1359R	0	0	2590			380		24				0	-				Superficial Velocity Change
				2	115	1550	1050	3.0	-325	29.0	2.67	530			380		24	14.9	3.1	0.0	79.5	29.8				
				3	92	1560	840	3.0		23.0	2.69	590			380		24	-	-	-	77.5	28.7				
				4	72	1580	630	3.0		18.0	2.76	650			380		24	-	-	-	75.0	27.2				
				5	52	2000	480	3.0		16.0	2.80	710			380		24	-	-	-	72.8	25.9				
FBC 77	Ohio #8 Seam washed 3.05% Sulfur 10.7% Ash	Ash	10	1	110	1520	1050	3.0	1359R	0	0	2650			360		24				0	-				Superficial Velocity Change
				2	110	1510	1050	3.0	-325	21.4	2.0	760			380		24	15.0	3.3	0.0	71.4	35.7				
				3	92	1530	840	3.0		17.2	1.95	710			390		24	-	-	-	73.2	37.6				
				4	70	1550	630	3.0		14.0	2.10	710			390		24	-	-	-	73.2	34.8				
				5	54	1580	480	3.0		10.6	2.05	650			380		24	-	-	-	75.4	36.8				
FBC 101	Ohio #8 Seam washed 3% Sulfur -1/4 x 0	1359R -10 +20	10	1	107	1400 - 1700	1100	3.0	**	**	**	*			120		50	16.0	2.9	0.0	**	**	55.2			*See Appendix A, Enclosure 23. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
												For Tests 101-120 SO_2 Data pertain to a single point in time					50	17.0	2.8	0.0			48.1			
																	50	15.7	3.0	0.0			56.0			
																	50	15.5	3.0	0.0						
FBC 102	Ohio #8 Seam washed 3% Sulfur -1/4 x 0	1359R -10 +20	10	1	107	1420 - 1680	1100	3.0	**	**	**	*			300		50	15.8	2.9	0.0	**	**	61.2			*See Appendix A, Enclosure 24. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
																	50	16.4	3.0	0.0			50.9			
																	50	14.6	2.8	0.0			50.0			
																	50	14.4	3.0	0.8						
																	50	15.8	3.1	0.9						
FBC 103*	Ohio #8 Seam washed 3% Sulfur -1/4 x 0	1359R -10 +20	10	1	105	1480 - 1810	1100	3.0	**	**	**				340		50	14.3	3.0	0.0	**	**				*This test was for information purposes only; it should not be used for comparison
																	50	17.2	2.6	0.0						
																	50	18.0	2.5	0.0						
																	50	15.8	2.2	0.0						
FBC 104	Ohio #8 Seam washed 3% Sulfur -1/4 x 0	1359R -10 +20	10	1	107	1420 - 1470	1100	3.0	**	**	**	*			340		50	14.7	3.1	0	*	**	54.3			*See Appendix A, Enclosure 25. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
																	50	14.0	2.6	0	*	**	55.1			
																	50	14.5	2.3	0	*	**	49.6			
																	50	14.7	2.5	0						
FBC 105	Ohio #6 Seam washed 3% Sulfur -1/4 x 0	1359R -10 +20	8	1	105	1480 - 1820	1100	3.0	**	**	**	*			340		50	18.9	2.1	0	*	**	44.6			*See Appendix A, Enclosure 26. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
																	50	16.7	3.0	0	*	**	42.2			
FBC 106	Ohio #8 Seam washed 2.95% Sulfur -1/4 x 0	1359R -10 +20	8	1	105	1500 - 1790	1100	4.0	**	**	**	*					50	19.8	3.8	0	*	**	45.5**			*See Appendix A, Enclosure 27. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
				2	105	1520 - 1680	1100	4.0					-	-			50	20.1	4.3	0	*	**	43.8			
				3	105	1600 - 1780	1100	4.0									50	18.7	4.1	0	*	**	33.6			
FBC 107	Ohio #8 Seam washed 3.0% Sulfur -1/4 x 0	1359R -10 +20	8	1	105	1500	1100	3.0	**	**	**	*	-	0	390	-	50	18.0	3.0	0	*	**	54.8			*See Appendix A, Enclosure 28. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form
				2	105	1800	1100	3.0					1200	-	420	420	50	16.4	3.2	0.1	*	**	44.0 35.8 33.8 31.2 27.9			
FBC 108	Ohio #8 Seam washed 3.05% Sulfur -1/4 x 0	1359R -10 +20	8	1	115	1500 - 1550	1100	3.0	**	**	**	*	-	-	340	-	50	17.9	2.9	0	*	**	59.2			*See Appendix A, Enclosure 29 **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form
				2	115	1720	1100	3.0					-	-	390	390	50	16.9	2.8	0	*	**	55.5			
				3	115	1720	1100	3.0					2500	0	370	-	50	17.1	3.2	0	*	**	47.6			
FBC 109	Ohio #8 Seam washed 2.95% Sulfur -1/4 x 0	1359R -10 +20	8	1	108	1470 - 1620	1100	3.0	**	**	**	*	-	-	350	-	50	19.1	3.3	0	*	**	57.0**			*See Appendix A, Enclosure 30. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form. *** 8 lbs 1359R added at end of condition
				2	108	1620 - 1680	1100	3.0					-	-	330	-	50	18.7	3.0	0	*	**	52.4**			
				3	108	1600 - 1620	1100	3.0					900	-	360	410	50	18.3	2.8	0	*	**	53.8			***13 lbs 1359R added at end of condition
FBC 110	Ohio #8 Seam washed 3.02% Sulfur -1/4 x 0	1359R -10 +20	8	1	106	1520 - 1560	1100	3.0	**	**	**	*	-	-	410	-	50	18.9	3.0	0	*	**	58.2			*See Appendix A, Enclosure 31. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form
				2	95	1600	1100	3.0					1100	0	440	450	50	17.6	3.1	0	*	**	56.9			
				3	95	1600	1100	3.0					-	-	410	-	50	17.2	2.5	0	*	**	46.1			
FBC 111	Ohio #8 Seam washed 3.04% Sulfur -1/4 x 0	1359R -10 +20	10	1	110	1380	1100	3.0	**	**	**	*	-	-	290	-	50	15.5	2.0	0.2	*	**	66.0			*See Appendix A, Enclosure 32. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form. Added 60 lbs fly ash
				2	105	1420	1100	3.0					-	-	290	305	50	15.7	2.3	0.1	*	**	69.1	Yes		
				3	105	1420	1100	3.0					2550	0	290	-	50	14.7	2.4	-	*	**	76.2			
FBC 112	Ohio #8 Seam washed 3.08% Sulfur -1/4 x 0	1359R -10 +20	10	1	98	1430 - 1450	1100	3.0	**	**	**	*	-	-	300	-	20	18.1	3.0	0	*	**	54.0			*See Appendix A, Enclosure 33. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form. Removed 30 lbs of bed in 5-lb increments
				2	93	1450 - 1540	1100	3.0					-	-	335	350	20	17.5	3.1		*	**	47.6			
				3	93	1540 - 1580	1100	3.0					-	-	370	350	20 20	17.8 17.8	3.1 3.1		*	**	50.1			

TABLE B-2 (Continued)

Test No.	Coal Type and Composition	Bed Material	Static Bed Depth, inches	Test Cond No.	Coal Rate, lb/hr	Bed Temp., °F	Air Rate, lb/hr	Flue Gas Oxygen (Control), %	Sorbent Data: Type and Site[a]	Sorbent Data: Injection Rate, lb/hr	Sorbent Data: Mole Ratio, Ca/S	Gas Analysis, Concentration in volume percent or ppm as indicated: I.R. SO_2, ppm[b]	Wet Tests: SO_2, ppm	Wet Tests: SO_3, ppm	I R NO, ppm	PDS NO_x, ppm[c]	Hydrocarbons, ppm	Orsat Analysis, %: CO_2	Orsat Analysis, %: O_2	Orsat Analysis, %: CO	SO_2 Reduction (I.R.), %	Sorbent Utilization,[d] %	Fly Ash Data: Carbon Content %	Fly Ash Data: Recirculation[e]	Fly Ash Data: Emission, lb/hr	Remarks
FBC 113	Ohio #8 Seam washed 3.05% Sulfur	1359R -10 +20	10	1	84	1450 - 1600	650	3.0	**	**	**	*	-					18.3	3.1	0	**	**	53.5			*See Appendix A, Enclosure J4. **Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
				2	53		650	3.0					400	0				18.2	2.9		**	**	42.1			
				3	60		650	3.0					-					17.2	3.0		**	**	49.8			62 lbs 1359R added
FBC 114	Ohio #8 Seam washed 3.09% Sulfur	1359R -10 +20	10	1	52	1450 - 1900	650	3.0	**	**	**	*	0	0				18.8	2.6	0	**	**	48.7			49 lbs 1359R added *See Appendix A, Enclosure J5.
				2	67		720	3.0					-	-	285	260		17.0	2.9		**	**	44.2			
				3	75		720	3.0					-	-	285	260		17.0	3.0		**	**	62.3			**Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form
																							64.9			
																							53.5			
FBC 115	Ohio #8 Seam washed 2.95% Sulfur	1359R -10 +20	10	1	47.4	1790	650	3.0	**	**	**	*	-	-				18.0	3.0	0	**	**	42.7			67 lbs 1359R added
				2	47.4	1580	580	3.0					175	0	270	260		16.0	3.1		**	**	60.4			Gas velocity decreased from 12 to 9 fps
				3	47.4	1580	500	3.0					-	-				16.8	3.4		**	**	62.1			*See Appendix A, Enclosure J6.
																							63.5			**Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
FBC 116	Ohio #8 Seam washed 3.02% Sulfur	1359R -10 +20	10	1	61	1550	650	3.0	**	**	**	*	-	-	220	260		19.2	3.0	0	**	**	55.1			*See Appendix A, Enclosure J7.
				2	61	1920	650	1.0					230	-	350			14.1	1.2	.2	**	**	56.9			**Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form
				3	61	1610	650	3.0					-	-	250			17.2	3.1	0	**	**	52.6			60 lbs bed material added
FBC 117	Ohio #8 Seam washed 3.07% Sulfur	1359R -10 +20	10	1	57	1600	650	3.0	**	**	**	*	0	0	280	260		18.7	2.9	0	**	**	44.6			*See Appendix A, Enclosure J8.
				2	56	2000	650	.2					4.9%	-	280			15.2	.2	4	**	**	44.6			60 lbs bed material added
				3	62	1600	650	3.0					-	-	440	430		17.8	2.9	0	**	**	58.2			**Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
				4	62	1600	650	.4					-	-	310			15.0	1	2	**	**	57.6			
FBC 118	Ohio #8 Seam washed 2.96% Sulfur	1359R -10 +20	10	1	40	1580	650	3.0	**	**	**	*	-	-		360	50	18.9	2.9	0	**	**	55.7			*See Appendix A, Enclosure J9.
				2	47	1930	650	.2					-	-	350	260	420	14.2	0	.1	**	**	49.1			**Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
				3	58	1550	650	5.0					5.5%	-	-	410	0	18.7	4.8	0	**	**	65.0			
				4	79	1930	650	.2					-	-	-	310	420	13.1	.1	.2	**	**	50.9			
FBC 119	Ohio #8 Seam washed 3.08% Sulfur	1359R -10 +20	10	1		2060	650	3.0	**	**	**	*			380		<50	18.8	3.1	0	**	**	39.2			*See Appendix A, Enclosure 60.
				2		1580	650	3.0							380			17.2	3.2	0	**	**	41.2			**Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
				3		2000	650	1.0							380			14.9	9	0.2	**	**	43.9			40 lbs 1359R added
				4		1510	650	3.0							380			15.6	2.6	0	**	**	52.6			
FBC 120	Ohio #8 Seam washed 3.08% Sulfur	1359R -10 +20	10	1	60	1700	650	6.0	**	**	**	*			310		<100	20.5	2.9	0	**	**	54.7	Yes		*See Appendix A, Enclosure 61.
				2	56	1550	650	3.0							310			15.2	3.2		**	**	54.3	Yes		40 lbs 1359R added
																		14.5	3.2		**	**	59.7	Yes		**Fluidized Bed of Limestone. No sorbent cross flow. Tests were transient in nature and results are presented in graphical form.
																							61.8	Yes		
																							54.7	Yes		

NOTES

[a] Limestone identification and numbering system of Bituminous Coal Research, Inc.: C = calcined by supplier, H = hydrated by supplier, R = raw stone

[b] Size U S Standard sieve size

[c] Sorbent Utilization defined as: % Utilization = $\frac{\%\ SO_2\ \text{Reduction in gas}}{\text{Stoichiometric Ratio, Ca/S}}$

[d] PDS designates Phenoldisulfonic Acid Method

[e] Gas analysis by Infrared Analyzer

[f] Fly ash not recirculated unless indicated by "Yes"

A

B

TABLE C-1. SULFUR BALANCE DATA

FBC Test No.	Cond. No.	SO_2 Sulfur Emission Lbs/M Btu	Fly ash Rate Lbs/M Btu	Fly ash Sulfur Fraction	Fly ash Sulfur Lbs/M Btu	Total Emission and Fly ash Sulfur Lbs/M Btu	Sulfur Input (4.5% S) Lbs/M Btu	Unaccounted Sulfur Lbs/M Btu
32	1	3.15	19.0	.012	0.23	3.38	3.7	.32
	2	1.90	29.1	.057	1.94	3.84	3.7	-.14
33	1	3.20	18.0	.014	0.25	3.45	3.7	.25
	2	1.75	35.0	.057	1.99	3.74	3.7	-.04
34	1	3.05	18.5	.025	.46	3.51	3.7	.19
	2		28.6	.041	1.17	3.37	3.7	.33
35	1	3.06	17.6	.018	.32	3.32	3.7	.38
	2	1.90	26.2	.053	1.38	3.33	3.7	.37
	3	1.70	27.6	.052	1.44	3.14	3.7	.56
	4	1.25	35.4	.039	1.45	3.70	3.7	.00
36	1	3.00	18.2	.024	.44	3.44	3.7	.26
	2	1.50	28.8	.063	1.81	3.31	3.7	.39
	3	1.50	30.4	.069	2.09	3.59	3.7	.11
	4	.75	39.6	.060	2.37	3.12	3.7	.58
37	1	2.80	19.1	.017	.32	3.12	3.7	0.58
	2	1.75	30.1	.063	1.89	3.64	3.7	0.06
	3	1.90	30.1	.051	1.54	3.44	3.7	0.26
	4	1.30	41.2	.055	2.27	3.59	3.7	0.23
39	1	2.70	18.5	.021	.39	3.09	3.7	0.69
	2	1.30	35.6	.056	2.00	3.30	3.7	0.40
	3	.30	30.5	.039	1.19	2.49	3.7	1.21
	4	1.38	32.0	.058	1.85	3.23	3.7	0.47
40	1	2.90	18.8	.023	.43	3.33	3.7	0.37
	2	1.05	37.8	.053	2.00	3.05	3.7	0.65
	3	.60	41.5	.023	.95	1.55	3.7	2.15
	4	1.50	32.4	.047	1.95	3.45	3.7	0.25

APPENDIX C

SULFUR BALANCE DATA FBC AND FBM

POPE EVANS AND ROBBINS

TABLE C-2. FBC SULFUR BALANCE DATA

FBC Data: Rates in pounds per hour

TEST NO. 46 — Additive 1359 H — Coal Sulfur Content 2.50

Test Condition	1	2	3	4
Additive St. Ratio	0.0	3.6	2.6	2.1
Sulfur input	2.72	2.57	2.77	2.75
Sulfur emission	2.27	0.44	0.65	0.84
Sulfur in fly ash	0.38	1.30	1.62	1.34
Sulfur retained in bed	-	0.80	0.40	0.40
Input less output	0.07	0.03	0.10	0.17

TEST NO. 47 — Additive 1359 H — Coal Sulfur Content 2.60

Test Condition	1	2	3	4
Additive St. Ratio	0.0	2.0	1.4	2.6
Sulfur input	2.65	2.70	2.75	2.78
Sulfur emission	2.40	1.64	1.91	1.52
Sulfur in fly ash	0.24	0.57	0.76	1.04
Sulfur retained in bed	-	0.35	0.17	.17
Input less output	0.01	0.14	- 0.09	0.05

TEST NO. 48 — Additive 1337 H — Coal Sulfur Content 2.50

Test Condition	1	2	3	4
Additive St. Ratio	0.0	1.4	1.6	1.16
Sulfur input	2.70	2.58	2.75	2.62
Sulfur emission	2.23	0.84	0.67	1.16
Sulfur in fly ash	.32	1.08	1.55	1.18
Sulfur retained in bed	-	0.70	0.35	0.35
Input less output	0.15	- 0.04	0.13	-0.02

TABLE C-1. (Continued)

FBC Test No.	Cond. No.	SO_2 Sulfur Emission Lbs/M Btu	Fly ash Rate Lbs/M Btu	Fly ash Sulfur Fraction	Fly ash Sulfur Lbs/M Btu	Total Emission and Fly ash Sulfur Lbs/M Btu	Sulfur Input (4.5% S) Lbs/M Btu	Unaccounted Sulfur Lbs/M Btu
41	1	2.79	16.8	.021	.35	3.05	3.7	0.65
	2	.70	45.7	.062	2.84	3.44	3.7	0.26
	3	.80	35.7	.071	2.53	3.33	3.7	0.37
	4	.70	34.5	.067	2.32	3.02	3.7	0.68
42	1	2.70	17.9	.022	.39	3.09	3.7	0.61
	2	1.00	31.8	.072	2.30	3.30	3.7	0.40
	3	.90	30.5	.070	2.14	3.05	3.7	0.65

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour

TEST NO. 49 — Additive 1337 H — Coal Sulfur Content 2.60 %

Test Condition	1	2	3	4
Additive St. Ratio	0	1.6	1.4	1.16
Sulfur input	2.62	2.73	2.73	2.80
Sulfur emission	2.52	1.54	1.53	1.53
Sulfur in fly ash	.22	.73	.95	1.05
Sulfur retained in bed	-	.35	.20	.14
Input less output	-0.12	0.13	.05	.08

TEST NO. 50 — Additive 1337 H — Coal Sulfur Content 2.50 %

Test Condition	1	2	3	4
Additive St. Ratio	0	1.3	1.5	1.46
Sulfur input	2.56	2.62	2.58	2.67
Sulfur emission	2.10	.92	.61	1.11
Sulfur in fly ash	0.32	1.15	1.41	1.14
Sulfur retained in bed	-	.70	.40	.30
Input less output	0.14	-0.15	- .04	0.13

TEST NO. 51 — Additive 1337 H — Coal Sulfur Content 2.60 %

Test Condition	1	2	3	4
Additive St. Ratio	-	1.4	1.6	1.46
Sulfur input	2.52	2.46	2.72	2.62
Sulfur emission	2.19	1.25	.88	1.27
Sulfur in fly ash	.22	.80	1.40	1.14
Sulfur retained in bed	-	.40	.30	.15
Input less output	0.11	0.01	0.14	0.06

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour

TEST NO. 52 — Additive 1337 H — Coal Sulfur Content 2

Test Condition	1	2	3	4
Additive St. Ratio	0	1.9	1.6	1.65
Sulfur input	2.58	2.67	2.70	2.68
Sulfur emission	2.15	0.71	0.68	0.98
Sulfur in fly ash	0.38	1.20	1.55	1.30
Sulfur retained in bed	-	0.70	0.40	0.30
Input less output	0.05	0.06	0.17	0.10

TEST NO. 53 — Additive 1337 H — Coal Sulfur Content 2

Test Condition	1	2	3	4
Additive St. Ratio	0	1.6	-	-
Sulfur input	2.68	2.60		
Sulfur emission	2.35	1.53		
Sulfur in fly ash	.36	.57		
Sulfur retained in bed	-	.60		
Input less output	-0.03	-0.10		

TEST NO. 54 — Additive 1359 R — Coal Sulfur Content 2.5

Test Condition	1	2	3
Additive St. Ratio	0	3.0	2.0
Sulfur input	2.72	2.80	2.72
Sulfur emission	2.39	1.28	1.51
Sulfur in fly ash	0.28	0.78	0.70
Sulfur retained in bed	-	0.60	0.40
Input less output	0.05	0.14	0.11

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour

TEST NO. 56 Additive 1337 R

Coal Sulfur Content 4.40 %

Test Condition	1	2
Additive St. Ratio	0	1.12
Sulfur input	4.72	4.53
Sulfur emission	3.96	1.93
Sulfur in fly ash	.59	1.80
Sulfur retained in bed	--	0.70
Input less output	0.17	0.10

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour

TEST NO. 57 Additive 1337 R

Coal Sulfur Content 4.3 %

Test Condition	1	2
Additive St. Ratio	0.00	1.30
Sulfur input	4.80	4.80
Sulfur emission	4.15	1.68
Sulfur in fly ash	.40	2.36
Sulfur retained in bed	-	0.60
Input less output	0.25	0.16

TEST NO. 58 Additive 1337 R

Coal Sulfur Content 4.3 %

Test Condition	1	2
Additive St. Ratio	0.0	1.60
Sulfur input	4.75	4.75
Sulfur emission	4.22	1.20
Sulfur in fly ash	.45	2.55
Sulfur retained in bed	-	0.70
Input less output	0.08	0.30

TEST NO. 59 Additive 1337 R

Coal Sulfur Content 2.6 %

Test Condition	1	2
Additive St. Ratio	0.0	2.3
Sulfur input	2.83	2.83
Sulfur emission	2.82	0.83
Sulfur in fly ash	.15	1.44
Sulfur retained in bed	-	0.40
Input less output	-.14	0.16

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour

TEST NO. 60 — Additive 1359 R — Coal Sulfur Content 2.6 %

Test Condition	1	2
Additive St. Ratio	0.0	2.0
Sulfur input	2.85	2.85
Sulfur emission	2.65	1.11
Sulfur in fly ash	.1	1.35
Sulfur retained in bed	-	.35
Input less output	0.10	0.04

TEST NO. 61 — Additive 1359 R — Coal Sulfur Content 4.3 %

Test Condition	1	2
Additive St. Ratio	0.0	1.25
Sulfur input	5.35	5.10
Sulfur emission	4.75	2.39
Sulfur in fly ash	.38	1.93
Sulfur retained in bed	-	.70
Input less output	0.22	0.08

TEST NO. 62 — Additive 1359 R — Coal Sulfur Content 4.2 %

Test Condition	1	2
Additive St. Ratio	0.0	1.6
Sulfur input	5.00	5.00
Sulfur emission	4.42	1.90
Sulfur in fly ash	.42	2.40
Sulfur retained in bed	-	.55
Input less output	0.16	0.15

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour

TEST NO. 63 — Additive 1359 — Coal Sulfur Content 4.3 %

Test Condition	1	2
Additive St. Ratio	0.0	1.6
Sulfur input	4.90	4.90
Sulfur emission	4.50	2.03
Sulfur in fly ash	.30	2.20
Sulfur retained in bed	-	.55
Input less output	0.10	0.12

TEST NO. 64 — Additive 1359 R — Coal Sulfur Content 4.4 %

Test Condition	1	2	3
Additive St. Ratio	0.0	2.6	1.5
Sulfur input	5.30	5.30	5.30
Sulfur emission	4.60	2.70	3.14
Sulfur in fly ash	.30	2.00	1.60
Sulfur retained in bed	-	.50	.30
Input less output	0.40	0.10	0.26

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour*

FBC Test No. 106

Test Condition	1	2	3
Flue gas output	1.70	.27	1.25
Fly ash output	.41	.37	.33
Bed retention	.95	2.4	1.6
Total output	3.06	3.04	3.18
Input	3.12	3.12	3.12
Input-output	.06	.08	-.06

FBC Test No. 107

Test Condition	1	2
Flue gas output	.60	3.75
Fly ash output	.35	.39
Bed retention	2.20	-.90
Total output	3.15	3.24
Input	3.16	3.16
Input-output	.01	-.08

FBC Test No. 108

Test Condition	1	2	3
Flue gas output	1.45	2.5	2.7
Fly ash output	.42	.28	.32
Bed retention	1.1	.28	.08
Total output	2.97	3.06	3.10
Input	3.13	3.16	3.13
Input-output	.16	.10	.03

*Tests 106-120 run with bed of 1359 limestone calcined in place. Rates are in pounds of sulfur per hour.

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour*

FBC Test No. 109

Test Condition	1	2	3
Flue gas output	.9	.9	.82
Fly ash output	.25	.19	.23
Bed retention	1.9	2.0	2.2
Total output	3.05	3.09	3.25
Input	3.18	3.18	3.18
Input-output	.13	.09	-.07

FBC Test No. 110

Test Condition	1	2	3
Flue gas output	1.02	1.28	.81
Fly ash output	.32	.25	.29
Bed retention	1.7	1.6	2.0
Total output	3.04	3.13	3.00
Input	3.18	3.18	3.18
Input-output	.14	.05	.18

FBC Test No. 111

Test Condition	1	2	3
Flue gas output	2.52	2.52	2.52
Fly ash output	.02	.45	.39
Bed retention	.50	.02	.02
Total output	3.04	2.99	2.93
Input	3.17	3.17	3.17
Input-output	.13	.18	.24

*Tests 106-120 run with bed of 1359 limestone calcined in place. Rates are in pounds of sulfur per hour.

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour*

FBC Test No. 112

Test Condition	1	2	3
Flue gas output	2.74	2.1	1.8
Fly ash output	.17	.36	.32
Bed retention	.2	.5	.8
Total output	3.11	2.96	2.92
Input	3.00	3.00	3.00
Input-output	-.11	.04	.08

*Tests 106-120 run with bed of 1359 limestone calcined in place. Rates are in pounds of sulfur per hour.

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour*

FBC Test 113

Test time hrs	1	2	3	4
Flue gas output	0.0	0.08	.6	.7
Fly ash output	0.35	0.28	.32	.38
Bed retention	1.56	1.64	.90	.80
Total output	1.91	2.00	1.82	1.88
Input	1.98	1.98	1.98	1.98
Input - Output	.07	-.02	.16	.10

FBC Test 114

Test time hrs	1	2	3	4	4.3*
Flue gas output	0.0	.39	0.85	1.15	10.4
Fly ash output	0.27	.25	0.35	.27	.30
Bed retention	1.75	1.47	.7	.5	-8.8
Total output	2.02	2.11	1.90	1.92	1.8
Input	1.95	1.95	1.95	1.95	1.95
Input - Output	-.07	-.16	.05	.03	.15

*Regeneration

FBC Test 115

Test time hrs	1	2	3	4	5	6
Flue gas output	.08	.16	.63	.45	-	.8
Fly ash output	.26	.40	.34	.14	-	.25
Bed retention	1.65	1.33	.84	.99	-	.44
Total output	1.99	1.89	1.81	1.58	-	1.49
Input	1.95	1.95	1.95	1.38		1.38
Input - Output	-.04	.06	.14	.20		.11

*Tests 106-120 run with bed of 1359 limestone calcined in place. Rates are in pounds of sulfur per hour.

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour*

FBC Test 116

Test Time hrs	1	2	3	4.2**	5	6
Flue gas output	0.00	.08	.42	17.5	-	.16
Fly ash output	.45	.37	.30	.27	-	.26
Bed retention	1.48	1.41	1.20	-16.0	-	1.57
Total output	1.83	1.86	1.82	1.77		1.99
Input	1.96	1.96	1.96	1.96		1.96
Input-Output	.13	.10	.04	.19		-.03

FBC Test 117

Test Time hrs	1	2	3	4.4**	5	6
Flue gas output	0.00	0.00	.16	25.5	.41	-
Fly ash output	.42	.54	.40	.6	.24	-
Bed retention	1.47	1.40	1.40	-24.4	1.18	-
Total output	1.89	1.94	1.96	1.7	1.83	
Input	1.85	1.85	1.85	1.85	1.85	
Input-Output	-.04	-.09	-.11	.15	.02	

FBC Test 118

Test Time hrs	1**	2	3	4	4.8**	6
Flue gas output	20.5	.16	.47	.37	21.5	-
Fly ash output	.2	.25	.22	.15	.15	-
Bed retention	-19.0	1.34	1.05	1.30	-19.8	-
Total output	1.7	1.75	1.74	1.82	1.85	
Input	1.8	1.80	1.80	1.80	1.80	
Input-Output	.1	.05	.06	-.02	-.05	

* Tests 106-120 run with bed of 1359 limestone calcined in place. Rates are in pounds of sulfur per hour.

** Regeneration

TABLE C-2. (Continued)

FBC Data: Rates in pounds per hour*

FBC Test 119

Test Time hrs	1	2	3	4	5	6
Flue gas output	.19	.65	.20	-	.25	.62
Fly ash output	.30	.26	.14	-	.20	.18
Bed retention	1.41	.90	1.44	-	1.30	.96
Total output	1.90	1.81	1.78		1.75	1.76
Input	1.88	1.88	1.88		1.85	1.88
Input-Output	-.02	.09	.10		.10	.12

FBC Test 120

Test Time hrs	1	2	3	
Flue gas output	.00	.05	.50	
Fly ash output	.36	.25	.02	(Recirculation)
Bed retention	1.56	1.60	1.52	
Total output	1.92	1.90	2.04	
Input	2.02	2.02	2.02	
Input-Output	.10	.12	-.02	

*Tests 106-120 run with bed of 1359 limestone calcined in place. Rates are in pounds of sulfur per hour.

TABLE C-3. FBM SULFUR BALANCE DATA

FBM Data: Rates in pounds per hour

TEST NO. 17 — Additive 1359 H — Coal Sulfur Content 4.3 %

Test Condition	1	2	3
Additive St. Ratio	0.0	.72	.84
Sulfur input	32.2	32.2	34.2
Sulfur emission	26.3	20.5	19.0
Sulfur in fly ash	1.6	9.3	10.5
Sulfur retained in bed	—	2.8	2.2
Input less output	2.3	- .4	2.5

TEST NO. 20 — Additive 1337 H — Coal Sulfur Content 2.6 %

Test Condition	1	2	3
Additive St. Ratio	0.0	1.17	1.46
Sulfur input	19.0	23.0	23.4
Sulfur emission	17.4	15.1	12.6
Sulfur in fly ash	1.0	7.0	8.7
Sulfur retained in bed	—	1.2	.6
Input less output	0.6	-0.3	1.5

TEST NO. 21 — Additive 1337 H — Coal Sulfur Content 2.5 %

Test Condition	1	2
Additive St. Ratio	0.0	1.37
Sulfur input	22.3	24.0
Sulfur emission	19.7	9.1
Sulfur in fly ash	.8	10.4
Sulfur retained in bed	—	2.5
Input less output	1.8	2.0

TABLE C-3. (Continued)

FBM Data: Rates in pounds per hour

TEST NO. 22 — Additive 1337 H — Coal Sulfur Content 2.5 %

Test Condition	1	2
Additive St. Ratio	0.0	1.46
Sulfur input	23.1	22.9
Sulfur emission	22.4	7.0
Sulfur in fly ash	.3	11.8
Sulfur retained in bed	—	2.8
Input less output	0.4	1.3

TEST NO. 23 — Additive 1337 R — Coal Sulfur Content 2.5 %

Test Condition	1	2
Additive St. Ratio	0.0	2.4
Sulfur input	20.8	20.8
Sulfur emission	20.3	6.8
Sulfur in fly ash	.8	11.9
Sulfur retained in bed	—	1.8
Input less output	-0.3	0.3

TEST NO. 24 — Additive 1337 R — Coal Sulfur Content 4.3 %

Test Condition	1	2	3
Additive St. Ratio	0.0	2.4	2.2
Sulfur input	18.9	19.9	19.9
Sulfur emission	18.2	4.0	4.4
Sulfur in fly ash	.8	13.6	13.5
Sulfur retained in bed	—	1.8	.9
Input less output	-0.1	0.5	1.1

TABLE C-3. (Continued)

FBM Data: Rates in pounds per hour

TEST NO. 25 — Additive 1337 R — Coal Sulfur Content 4.3 %

Test Condition	1	2
Additive St. Ratio	0.0	1.8
Sulfur input	35.0	35.2
Sulfur emission	30.5	9.7
Sulfur in fly ash	1.8	21.6
Sulfur retained in bed	—	2.8
Input less output	2.7	1.1

TEST NO. 26 — Additive 1337 R — Coal Sulfur Content 4.4 %

Test Condition	1	2	3
Additive St. Ratio	0.0	1.7	1.9
Sulfur input	39.6	39.6	39.6
Sulfur emission	34.5	12.6	10.4
Sulfur in fly ash	1.7	21.0	24.0
Sulfur retained in bed	—	4.5	2.7
Input less output	3.4	1.5	2.5

TEST NO. 27 — Additive 1359 R — Coal Sulfur Content 4.3 %

Test Condition	1	2
Additive St. Ratio	0.0	2.0
Sulfur input	33.2	33.2
Sulfur emission	28.3	7.9
Sulfur in fly ash	2.1	17.8
Sulfur retained in bed	—	5.6
Input less output	2.8	1.9

TABLE C-3. (Continued)

FBM Data: Rates in pounds per hour

TEST NO. 28 — Additive 1359 R — Coal Sulfur Content 2.

Test Condition	1	2	3
Additive St. Ratio	0.0	2.4	2.2
Sulfur input	20.6	20.3	20.3
Sulfur emission	20.9	6.8	7.7
Sulfur in fly ash	0.8	11.0	11.1
Sulfur retained in bed	—	1.8	.9
Input less output	-1.1	0.7	0.6

TEST NO. 29 — Additive 1359 R — Coal Sulfur Content 4.

Test Condition	1	2	3
Additive St. Ratio	0.0	1.7	2.0
Sulfur input	30.9	31.0	31.0
Sulfur emission	27.2	11.0	7.7
Sulfur in fly ash	2.4	14.2	17.6
Sulfur retained in bed	—	4.5	3.6
Input less output	1.3	1.3	2.1

TEST NO. 30 — Additive 1359 H — Coal Sulfur Content 2.6

Test Condition	1	2	3
Additive St. Ratio	0.0	1.4	1.8
Sulfur input	22.4	22.6	22.6
Sulfur emission	22.0	11.0	8.9
Sulfur in fly ash	.6	9.0	11.5
Sulfur retained in bed	—	2.0	1.2
Input less output	-0.2	0.6	1.0

TABLE C-3. (Continued)

FBM Data: Rates in pounds per hour

TEST NO. 3 Additive 1359 H

Coal Sulfur Content 2.6 %

Test Condition	1	2	3
Additive St. Ratio	0.0	1.3	1.6
Sulfur input	20.8	20.8	20.8
Sulfur emission	20.0	10.3	8.8
Sulfur in fly ash	0.2	8.2	9.5
Sulfur retained in bed	-	1.6	0.8
Input less output	0.6	0.7	1.7

TEST NO. 32 Additive 1359 R

Coal Sulfur Content 2.6 %

Test Condition	1	2	3
Additive St. Ratio	0.0	1.6	1.8
Sulfur input	18.7	19.0	19.0
Sulfur emission	18.5	7.5	6.6
Sulfur in fly ash	0.2	9.3	10.5
Sulfur retained in bed	-	1.6	0.8
Input less output	0.0	0.6	1.1

PB 198 413

BIBLIOGRAPHIC DATA SHEET

1. Report No. APTD-0655
2.
3. Recipient's Accession No.

4. Title and Subtitle
Characterization and Control of Gaseous Emissions From Coal-Fired Fluidized-Bed Boilers

5. Report Date October 1970
6.

7. Author(s)
E. B. Robison, A. H. Bagnulo, J. W. Bishop, and S. Ehrlich

9. Performing Organization Name and Address
Pope, Evans and Robbins
Consulting Engineers
A Division of Perathon Incorporated
Alexandria, Virginia 22314

8. Performing Organization Rept. No.
10. Project/Task/Work Unit No.
11. Contract/Grant No.

12. Sponsoring Organization Name and Address
Division of Process Control Engineering
National Air Pollution Control Administration
Environmental Health Service - Public Health Service
Department of Health, Education, and Welfare
Washington, D. C.

13. Type of Report & Period Covered
Interim
14.

15. Supplementary Notes

16. Abstracts
Results are presented from a test program to characterize the air pollution emissions from the combustion of coal in a fluidized bed combustion and to assess the potential of fluidized-bed combustion for air pollution control. These emissions were monitored under a comparatively large number of different conditions. Efforts were made to reduce emissions of oxides of sulfur by the use of limestone-based sorbents and to determine the conditions most favorable for the reduction. Emissions of sulfur dioxide, nitric oxide and hydrocarbon were monitored continuously with periodic samples taken for measurement of particulates and wet test determination of SO_x and NO_x. When conditions most favorable for air pollution control were established on a pilot scale, the conditions were reproduced in tests with the fluidized-bed boiler module.

17. Key Words and Document Analysis. 17a. Descriptors
Fluidized bed processors
coal
Combustion gases
Limestone
Sulfur Dioxide
Nitrogen dioxide
Hydrocarbons

17b. Identifiers/Open-Ended Terms

17c. COSATI Field/Group 13/B

18. Availability Statement
Unlimited

19. Security Class (This Report) UNCLASSIFIED
20. Security Class (This Page) UNCLASSIFIED
21. No. of Pages 80
22. Price

FORM NTIS-35 (10-70)
USCOMM-DC 40329-P71

6. RESULTS OF PILOT SCALE (FBC) TESTS - SINTERED ASH BED

6.1 SULFUR DIOXIDE EMISSION WITH COARSE ADDITIVES

Initial tests to investigate the SO_2 control potential of the fluidized-bed combustor were carried out with the 1359 limestone and 1337 dolomite sorbent materials ground and screened to a -7 +14 mesh, a size roughly that of the bed material. This size was selected in an attempt to increase the residence time of the particle in the bed and thus increase the sulfur capture. The sorbents were used in the raw state and as calcined by the supplier. The effects of bed temperature, bed depth, sorbent feed rate and excess air (as determined by the oxygen content in the flue gas) were investigated initially in order to determine the optimum operating conditions for sulfur retention. Three tests were conducted with reducing conditions in the bed. 90% to 95% of the input sulfur is emitted as sulfur dioxide without sorbent addition.

The reductions in SO_2 emissions observed in the FBC with the coarse 1337 dolomite are shown in Table I and the corresponding data for the 1359 limestone in Table II. Sulfur dioxide reductions observed in the FBC tests are plotted as a function of stoichiometric ratio in Figures 21 and 22 respectively. Sorbent utilization percentages are given in Tables I and II; these are obtained by calculating the average portion of calcium in the sorbent feed which reacts with sulfur. Stoichiometric ratios were computed on the basis of 4.5% sulfur in the coal and the calcium content of the sorbent. The magnesium fraction in the dolomite was assumed to be chemically inert. The stoichiometric ratio, designated the Ca/S ratio in the tables, is the ratio of moles of calcium in the sorbent fed to moles of sulfur in the coal.

Comparison of results, presented in Tables I and II, indicates that the dolomite is more effective in sulfur capture than the high calcium limestone, based on the calcium alone, and ignoring the magnesium fraction. Sorbent utilization values of up to 35 to 40% were observed with the dolomite whereas the limestone utilization was limited to a maximum of about 20%. One contributing factor may have been the friability of the dolomite. The dolomite tended to decrepitate during calcination in the bed and was elutriated. The limestone, on the other hand, tended to build up in the bed. The dolomite, in breaking up, could have exposed more surface per unit mass for the sulfur reaction and combined with sulfur before leaving the bed. The limestone, in retaining its particle size, would expose less reactive surface per unit mass.

TABLE I. SULFUR DIOXIDE REDUCTIONS OBSERVED WITH ADDITION OF COARSE (-7 +14 MESH) 1337 DOLOMITE TO THE FBC BURNING A 4.5% SULFUR COAL

Test No.	Dolomite State	Bed Depth In.	Bed Temp. °F	Flue Gas O_2 %	Ca/S Ratio	SO_2 Concen. ppm Initial	Final	% SO_2 Reduction	% Sorbent Utilization
FBC 5	Raw	7	1800	1.0	1.10	4450	3700	17.0	15.4
	Raw	7	1800	3.0	1.10	4100	3150	23.2	21.0
FBC 6	Raw	7	1700	1.0	1.20[1]	4650	3600	22.6	18.8
			1550	3.0	1.20	4250	2600	39.0	32.5
FBC 9	Raw	7	1800	1.0	1.17[1]	4550	3850	15.0	12.8
FBC 10	Raw	7	1750	1.0	1.10[1]	4650	3600	22.5	20.4
	Raw	7	1750	1.0	1.20		3150	32.2	26.8
FBC 11	Raw	6	1680	3.0	1.14	4350	3150	27.6	24.2
	Raw	6	1650	3.0	2.04		1900	56.3	27.7
	Raw	6	1600	3.0	2.56		1500	65.5	25.6
FBC 12	Raw	9	1540	2.0	1.09	4500	2150	52.2	47.8
FBC 13	Raw	10	1500	3.0	0.79	4350	3000	31.0	39.2
	Raw	10	1480	3.0	1.44		2000	54.0	37.4
FBC 14	Raw	10	1880	3.0	1.10	4200	3750	10.7	9.7
	Raw	10	1860	3.0	1.70		3320	21.0	12.2
	Raw	10	1800	4.0	1.70		3000	28.6	16.8
FBC 15	Raw	10	1700	2.0	1.00	4000	3100	22.5	22.5
FBC 16	Raw	10	1620	2.0	1.00	4200	3450	17.8	17.8
	Raw	10	1620	2.0	1.60		2900	31.0	19.4

[1]Reducing conditions in bed

TABLE II. SULFUR DIOXIDE REDUCTIONS OBSERVED WITH ADDITION OF COARSE (-7 +14 MESH) 1359 LIMESTONE TO THE FBC BURNING A 4.5% SULFUR COAL

Test No.	Limestone State	Bed Depth In.	Bed Temp. °F	Flue Gas O_2 %	Ca/S Ratio	SO_2 Concen. ppm Initial	SO_2 Concen. ppm Final	% SO_2 Reduction	% Limestone Utilization
FBC 19	Raw	8	1540	3.0	1.15	3900	3300	15.4	13.4
	Raw	8	1540	3.0	1.93		2850	27.0	14.0
	Raw	8	1540	3.0	1.93[1]		2350	40.0	20.7
FBC 20	Raw	8	1700	3.0	1.15	3800	3350	11.8	10.2
	Raw	8	1740	3.0	1.47		2750	27.6	18.8
	Raw	8	1650	3.0	2.00		2500	34.2	17.1
	Raw	8	1650	3.0	2.00		2300	39.5	19.7
FBC 21	Raw	8	1520	3.0	0.90	3800	3300	13.2	14.7
	Raw	8	1530	3.0	1.57		2920	23.1	14.7
	Raw	8	1500	2.0	2.00		2900	23.6	11.8
	Raw	8	1520	3.0	2.00		2600	31.5	15.8
FBC 22	Raw	8	1650	3.0	1.00	3900	3300	15.4	15.4
	Raw	8	1540	3.0	1.90		2720	30.8	16.2
FBC 26	Calcined	8	1700	3.0	1.50	3800	3300	13.2	8.8
	Calcined	8	1700	3.0	2.30		2950	22.4	9.7
	Calcined	8	1700	3.0	3.40		2370	37.5	11.0

[1]=Ash Recirculation

TABLE I. (Continued)

Test No.	Dolomite State	Bed Depth In.	Bed Temp. °F	Flue Gas O_2 %	CA/S Ratio	SO_2 Concen. ppm Initial	SO_2 Concen. ppm Final	% SO_2 Reduction	% Sorbent Utilization
FBC 23	Calcined	8	1600	3.0	1.28[2]	4180	2850	31.8	24.8
	Calcined	8	1580	3.0	2.17[2]		2700	35.6	16.4
	Calcined	8	1580	3.0	2.54[2]		2050	51.0	20.0
	Calcined	8	1580	3.0	2.17[2,3]		1720	59.0	27.2
FBC 24	Calcined	7	1680	3.0	1.13[2]	3500	2930	16.3	14.4
	Calcined	7	1680	3.0	1.73[2]		2410	31.2	18.0
	Calcined	7	1680	3.0	2.26[2]		2250	35.6	15.7
FBC 25	Calcined	6	1720	3.0	1.15	3650	3350	8.5	7.4
	Calcined	6	1720	3.0	1.60		3150	12.9	8.1

[2] = Ash Recirculation

[3] = With water injection

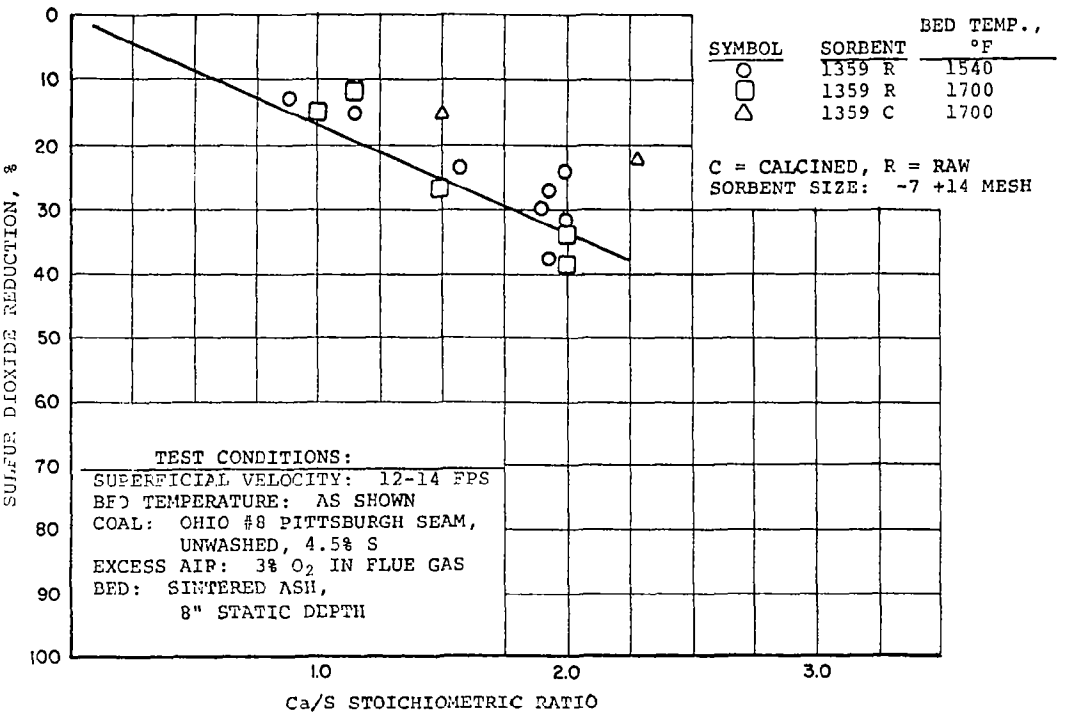


FIGURE 22. REDUCTION OF SULFUR DIOXIDE EMISSION FROM THE FBC BURNING A 4.5% S COAL WITH COARSE 1359 LIMESTONE ADDITION

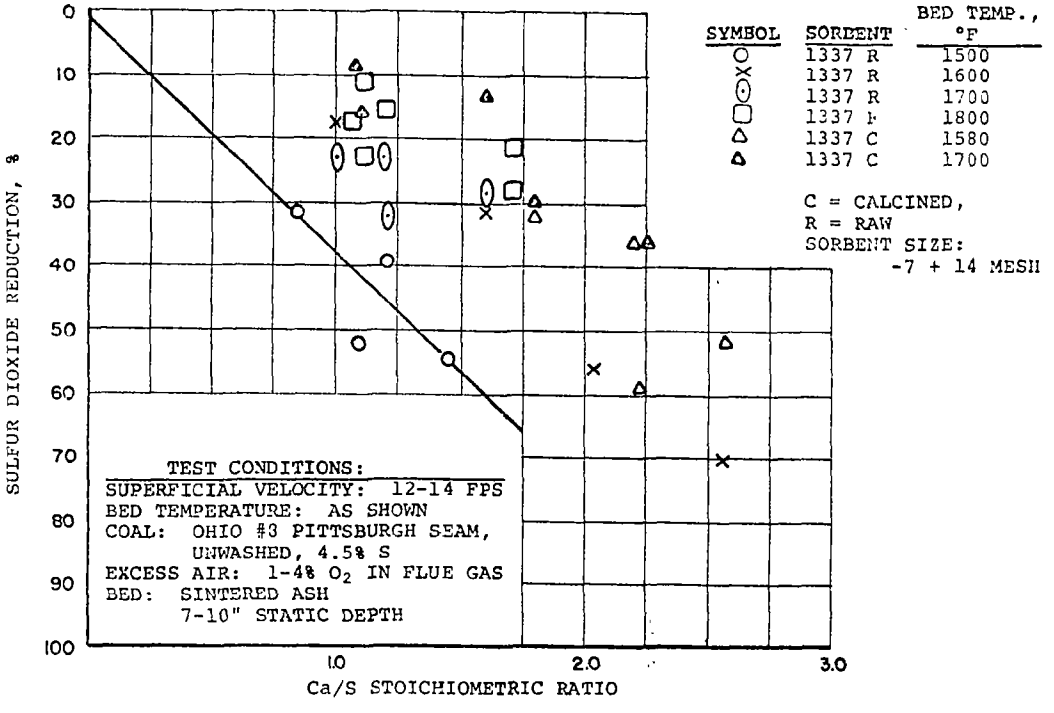


FIGURE 21. REDUCTION OF SULFUR DIOXIDE EMISSION FROM THE FBC BURNING A 4.5% S COAL WITH COARSE 1337 DOLOMITE ADDITION

Sulfur retention and sorbent utilization are seen to increase slightly with flue gas oxygen content. An increase was observed in FBC Test 5, 10 and 14 in which other factors were held fairly constant. In Test No. 5, conducted at a bed temperature of 1800°F and a Ca/S ratio of 1.1 the sorbent utilization was increased from 15.4% to 21.0%, when the oxygen content was increased from 1% to 3%. Test No. 10 was initiated with reducing conditions in the bed, i.e., with less than stoichiometric air passing through the bed. The balance to make up the 1% oxygen concentration in the flue gas was supplied by overbed air. The sorbent utilization increased from 20.4 to 26.8% when the bed condition was changed from reducing to oxidizing. The results of Test No. 14 indicated an increase in sorbent utilization from 12.2 to 16.8% with increase in oxygen although the improvement may have been partly due to 60°F drop in bed temperature. This result is reasonable inasmuch as oxygen is required to retain sulfur in a more stable form according to the relation:

$$CaO + SO_2 + \frac{1}{2} O_2 \rightarrow CaSO_4$$

In all subsequent FBC tests the oxygen concentration in the flue gas was maintained at 3% to improve sulfur capture but more importantly to limit hydrocarbons emission, as discussed in Section 6.5.

Sulfur retention and sorbent utilization increase with decrease in bed operating temperature to the lower end of the operating range. This effect is evident from the results of FBC Tests 5, 6 and 11 for the 1337 dolomite. Under otherwise similar conditions the sorbent utilization changed from 21.0 to 24.2 to 32.5 for respective temperatures of 1800°F, 1680°F and 1550°F. A similar effect is noted in Table II for the 1359 limestone.

The effect of bed depth is less well defined because of variation in other parameters. Interpolation of reductions and Ca/S ratios for Test No. 13 as shown in Figure 23 indicates a reduction of 45% at a ratio of 1.2 with a 10-inch bed. Also shown is a reduction of 39% observed in Test No. 6 conducted at this ratio and a 7-inch bed depth. The effects of bed depth and temperature were similar with injection of fine sorbents as indicated in Section 6.3.

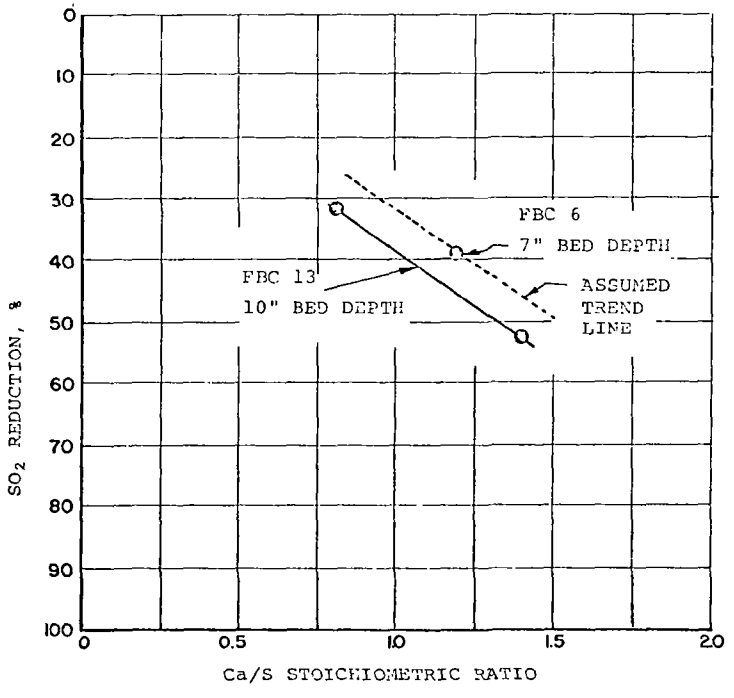


FIGURE 23. INTERPOLATION OF 10-INCH BED DEPTH DATA FOR COMPARISON WITH 7-INCH BED DEPTH DATA

The calcined 1337 dolomite was less effective than the raw stone as recorded in Table I for FBC Tests No. 23, 24, and 25. The performance of the 1359 limestone in the -7 +14 particle size was likewise poor as shown in Table II. A deep mined limestone from Northern West Virginia was tested with a 5% sulfur coal from the same mining area. The limestone, containing 72% $CaCO_3$, and screened to the -7 +14 mesh size, effected an SO_2 reduction of 36% at a Ca/S ratio of 1.4 (25.4% utilization). The test conditions were 1550°F bed temperature and 3.0% O_2 in the flue gas. Data for this test (FBC 27) and others are summarized in Appendix B. Sulfur balances are shown in Appendix C.

6.2 SULFUR DIOXIDE EMISSION WITH FINELY DIVIDED SORBENTS

The investigation of sulfur dioxide emission control by sorbent injection was redirected to the use of fine sorbents in an effort to increase the reactive surface of the sorbent for greater desulfurization.

The tests were conducted in the FBC with the following considerations with respect to variables:

Sorbent Type: Two sorbents were tested, the 1337 dolomite and the 1359 limestone.

Sorbent State: The raw stone of each of the two sorbents was ground to a -325 mesh particle size for the test series. Both sorbents were also tested in the hydrated form which is commercially available in a -325 mesh particle size. Both the calcium and magnesium fractions of the dolomites were hydrated. The sorbents are designated 1337R, 1359R, 1337H and 1359H to distinguish the raw and hydrated forms respectively. One test was run with precalcined limestone designated 1359C.

Sorbent Particle Size: The effect of variation in particle size from -12 to -325 mesh was tested with the 1359R limestone. Except for these tests, reported in Section 6.3, the sorbent particle size was -325 mesh.

Sorbent Feed Rate: The sorbent feed rate was varied in the range of 1 to 3 stoichiometric ratio based on the calcium content of the sorbent and the sulfur content in the coal.

Sorbent Feed System: Three methods of sorbent feed were employed as discussed in some detail in Section 5.3. These were (1) addition of sorbent at the coal feed port [#1 Feeder], (2) injection of sorbent at two points away from the coal feed port [#2 Feeder] and (3) premixing the sorbent and coal in the hopper. The #2 Feeder system was modified for four-point feed in a test of sorbent distribution. One test was conducted with sorbent injection above the bed for comparison.

Coal Sulfur Content: The tests were conducted with Ohio #8 Pittsburgh seam coal, unwashed and washed containing respectively 4.5% and 2.6% sulfur.

Flue Gas Oxygen Content: The oxygen content in the flue gas was held constant at 3% since previous results, noted in Section 6.5, indicated this value to be minimum for control of hydrocarbons emission. Higher values contribute to loss in thermal efficiency.

Bed Temperature and Depth: The bed temperature was varied in the range of 1500°F to 1800°F to investigate the temperature effect with fine sorbent particles. The bed depth was adjusted to the greatest value consistent with bed temperature. A test series was conducted to investigate the independent effects of bed temperature depth and particle size. The series is discussed in Section 6.3.

Ash Recirculation: Fly ash was recirculated on a number of tests as a final test condition. The rate was 80% of the collected ash.

Superficial Gas Velocity: The tests were conducted with the superficial gas velocity held constant within the range of 12 to 14 fps in most tests. The effect of superficial velocity was investigated as discussed in Section 6.3.

The test results indicated a marked improvement in sulfur dioxide reduction and sorbent utilization with the fine sorbent as compared to the coarse sorbents under similar conditions. The improvement was most pronounced with the use of the 1359 limestone. A comparison is presented in Figure 24 showing the effect of particle size change with both sorbents in the raw state. Test conditions included a 1500°F - 1600°F temperature

FIGURE 24. COMPARISON OF SULFUR DIOXIDE REDUCTIONS OBSERVED WITH THE COARSE AND FINE SORBENT AND A 4.5% SULFUR COAL

range, 3% oxygen in the flue gas, a 14 fps superficial velocity and a 4.5% sulfur coal in each case.

The reductions observed with the fine sorbents, both raw and hydrated, while burning the 4.5% sulfur coal are shown in Table III. One test of precalcined 1359 limestone is included. The reductions are plotted as a function of Ca/S ratio in Figure 25 for bed temperature in the range of 1500°F - 1600°F. The trends show the 1337 dolomite, again, to be more reactive than the 1359 limestone when the magnesium fraction of the dolomite is considered inert. The trend indicates further that the hydrated form of the sorbents is as reactive as the corresponding fine raw sorbent. The most favorable single observation was made with the dolomite hydrate, designated 1337H. The reduction was 88% at a Ca/S ratio of 1.8, the corresponding utilization being 47.2%. The average dolomite utilization based on the trend line containing both the hydrate and raw form data is indicated to be ~45%.

The data trend in Figure 25 for the 1359 fine limestone indicates a similar reactivity for both the hydrate and raw forms, a lesser reduction than observed with the dolomite (65% at a Ca/S ratio of 1.8) and utilization decreasing with increasing Ca/S ratio. The utilization varies from 40% at a ratio of 1.0 to 28% at a ratio of 3.0. By comparison, the coarse stone utilization (Table II) did not exceed 20%.

The precalcined form of the 1359 limestone, designated 1359C in Table III and Figure 25, was less effective than the same stone in the raw or hydrated forms. This result may have been due to the possibility that the supplier's conditions for calcination may not have produced as "soft" a calcine as the 1500°F fluidized-bed environment. The limestone was calcined by the supplier to optimize hydration, but the conditions were reported to be proprietary and were not released.

Sulfur dioxide reductions observed with injection of sorbents during combustion of a 2.6% sulfur coal are summarized in Table IV. The percent reductions, as a function of the Ca/S, ratio, were approximately the same for this medium sulfur coal as for the 4.5% sulfur coal. The points, taken at bed temperatures in the range of 1500°F - 1600°F, are plotted in Figure 26.

TABLE III. (Continued)

Test No.	Bed Depth In.	Bed Temp. °F	Flue Gas O_2 %	Sorbent	Ca/S Ratio	Feed System	SO_2 Concen. ppm Initial	SO_2 Concen. ppm Final	Reduction	Utilization
FBC 41	10	1560	3.0	1359H	3.40	#1	3450	350	90.0	26.5
	10	1600	3.0	1359H	2.24	#2		1000	71.0	31.7
	10	1600	3.0	1359H	2.10	PREMIX		900	74.0	35.2
FBC 42	9	1640	3.0	1359H	1.65	#2	3350	1150	65.5	39.6
	9	1600	3.0	1359H	2.80	PREMIX		600	82.0	29.3
FBC 56	10	1580	3.0	1337R	1.12	#2	3550	1750	49.0	43.6
FBC 57	10	1570	3.0	1337R	1.25	#2	3550	1400	60.7	43.5
FBC 58	10	1570	3.0	1337R	1.57	#2	3600	1000	72.2	46.0
FBC 61	10	1540	3.0	1359R	1.25	#2	3550	1900	46.5	37.3
FBC 62	10	1550	3.0	1359R	1.6	#2	3550	1500	57.8	36.0
FBC 64	12	1550	3.0	1359C	2.6	#2	3750	2200	41.2	15.8
	12	1550	3.0	1359C	1.5	#2	3750	2560	30.8	21.1

TABLE III. SULFUR DIOXIDE REDUCTION OBSERVED WITH FINE SORBENT ADDITION TO COMBUSTION OF A 4.5% SULFUR COAL IN THE FLUIDIZED BED

Test No.	Bed Depth In.	Bed Temp. °F	Flue Gas O_2 %	Sorbent	Ca/S Ratio	Feed System	SO_2 Concen. ppm Initial	SO_2 Concen. ppm Final	% SO_2 Reduction	% Sorbent Utilization
FBC 32	8	1580	3.0	1359H	1.10	#1	3650	2200	40.5	36.8
FBC 33	9	1600	3.0	1337H	1.05	#1	3750	2000	46.7	44.5
FBC 39	9	1540	3.0	1337H	1.38	#1	3400	1600	53.0	38.4
	9	1580	3.0	1337H	1.87	#2		400	88.2	47.2
	9	1540	3.0	1337H	1.17	PREMIX		1650	51.2	43.6
FBC 40	7	1740	3.0	1337H	1.55	#1	3550	1250	64.7	41.8
	7	1720	3.0	1337H	2.08	#2		650	81.9	39.2
	7	1760	3.0	1337H	1.17	PREMIX		1800	49.2	41.1
FBC 32	8	1580	3.0	1359H	1.10	#1	3700	2200	40.5	36.8
FBC 35	9	1560	3.0	1359H	1.04	#1	3600	2200	39.0	37.5
	9	1560	3.0	1359H	1.10	#2		1950	46.0	42.0
	9	1560	3.0	1359H	2.15	#1 + #2		1400	61.2	28.5
FBC 36	9	1540	3.0	1359H	1.16	#1	3550	1700	52.2	45.0
	9	1580	3.0	1359H	1.36	#2		1700	52.2	38.4
	9	1580	3.0	1359H	2.55	#1 + #2		800	77.5	30.4
FBC 37	8	1640	3.0	1359H	1.34	#1	3400	2000	41.2	30.8
	8	1620	3.0	1359H	1.34	#2		2250	34.0	25.2
FBC 38	8	1760	3.0	1359H	1.75	#1	3500	1470	57.9	33.1
	8	1720	3.0	1359H	1.45	#1		1780	49.2	33.9
	8	1720	3.0	1359H	1.28	#1		1800	48.8	38.0
	8	1760	3.0	1359H	1.16	#1		2030	42.0	36.2

TABLE IV. SULFUR DIOXIDE REDUCTION OBSERVED WITH FINE SORBENT ADDITION TO COMBUSTION OF A 2.6% SULFUR COAL IN THE FLUIDIZED BED

Test No.	Bed Depth In.	Bed Temp. °F	Flue Gas O_2 %	Sorbent	Ca/S Ratio	Feed System	SO_2 Conc. ppm Initial	SO_2 Conc. ppm Final	% SO_2 Reduction	% Sorbent Utilization
FBC 46	11	1600	3.0	1359H	3.60	#1	2200	270	87.7	24.4
	11	1640	3.0	1359H	2.60	#2		550	75.0	28.8
	11	1650	3.0	1359H	2.10	PREMIX		760	65.5	31.2
FBC 47	6	1780	3.0	1359H	2.00	#1	2200	1500	31.8	15.9
	6	1800	3.0	1359H	1.40	#2		1700	22.7	16.2
	6	1800	3.0	1359H	2.60	PREMIX		1400	36.3	13.9
FBC 48	12	1585	3.0	1337H	1.40	#1	2000	800	60.0	42.8
	12	1600	3.0	1337H	1.60	#2		500	75.0	46.9
	12	1590	3.0	1337H	1.16	PREMIX		1000	50.0	43.0
FBC 49	6	1780	3.0	1337H	1.60	#1	2350	1400	40.4	25.2
	6	1770	3.0	1337H	1.46	#2		1550	34.0	23.3
	6	1790	3.0	1337H	1.16	PREMIX		1550	34.0	29.2
FBC 50	12	1570	3.0	1337H	1.35	#1	1880	860	54.2	40.1
	12	1580	3.0	1337H	1.55	#2		490	73.9	47.7
	12	1570	3.0	1337H	1.46	PREMIX		910	51.6	35.4
FBC 51	12	1580	3.0	1337H	1.40	#1	2100	1100	47.5	34.0
	12	1600	3.0	1337H	1.60	#2		700	66.7	41.7
	12	1600	3.0	1337H	1.46	PREMIX		1110	47.5	32.5

POPE EVANS AND ROBBINS

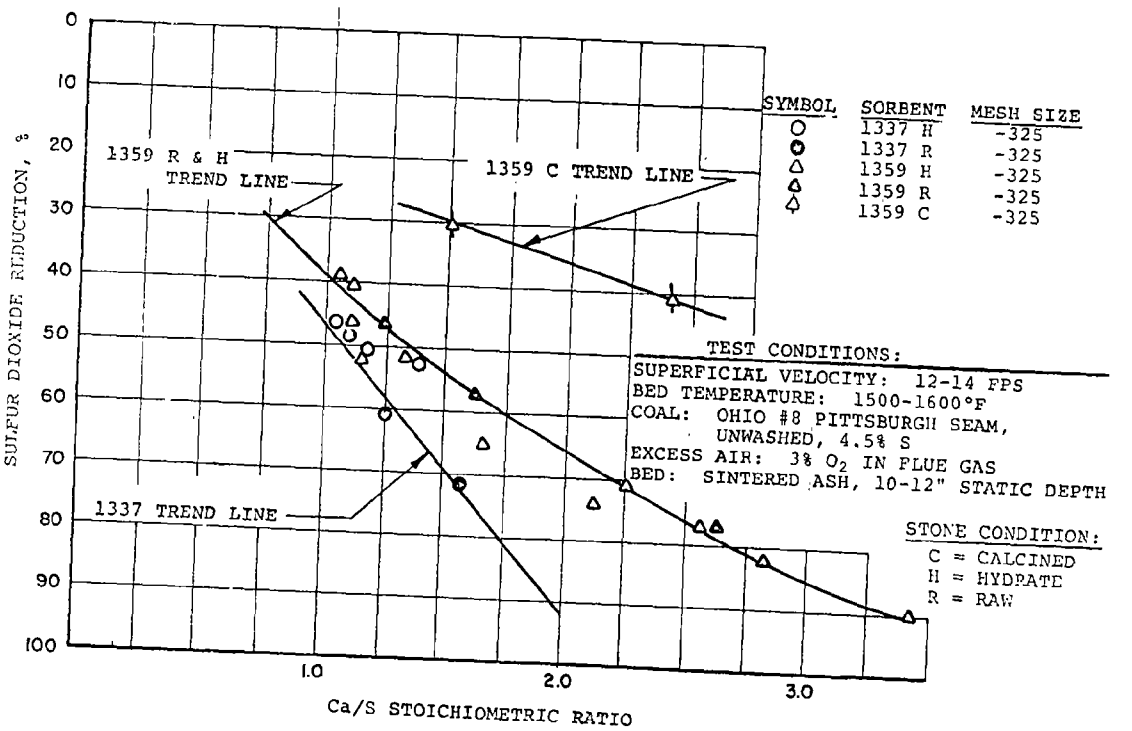


FIGURE 25. SULFUR DIOXIDE REDUCTION WITH FINE SORBENT ADDITION TO THE FBC BURNING A 4.5% SULFUR COAL

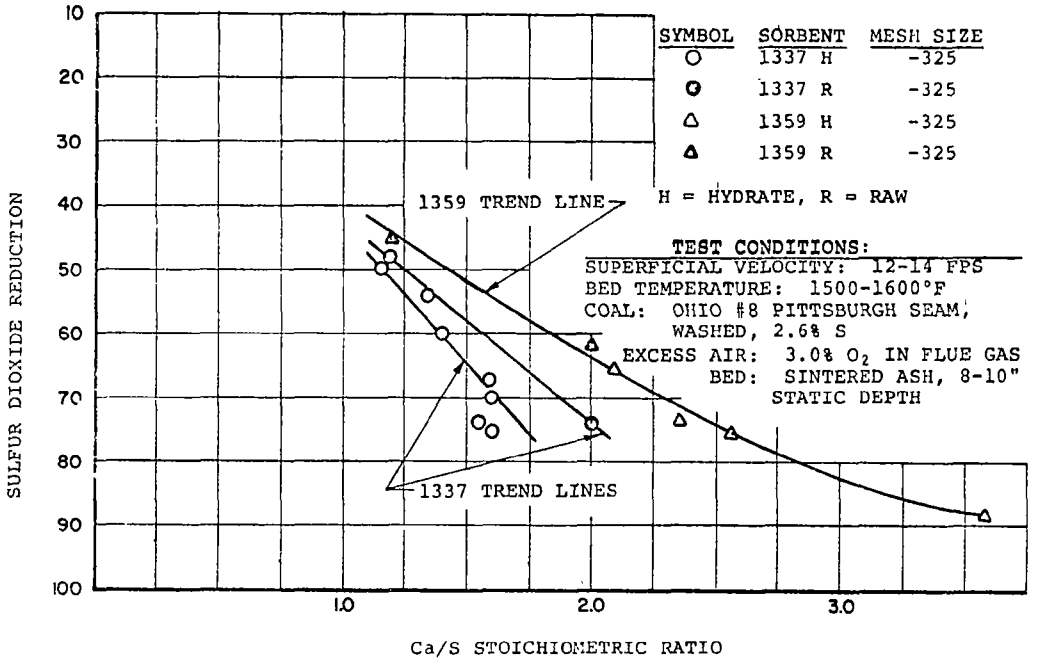


FIGURE 26. SULFUR DIOXIDE REDUCTION WITH FINE SORBENT ADDITION TO THE FBC BURNING A 2.6% SULFUR COAL

TABLE IV. (Continued)

Test No.	Bed Depth In.	Bed Temp. °F	Flue Gas O_2 %	Sorbent	Ca/S Ratio	Feed System	SO_2 Conc. ppm Initial	SO_2 Conc. ppm Final	% SO_2 Reduction	% Sorbent Utilization
FBC 53	11	1560	3.0	1337R	2.00	#2	2050	550	73.5	36.7
	11	1560	3.0	1337R	1.20	#2	2050	1060	48.0	40.0
FBC 59	10	1590	3.0	1359R	2.33	#2	2350	650	72.4	31.0
FBC 60	10	1550	3.0	1359R	2.33	#2	2300	900	60.8	30.4
	10	1550	3.0	1359R	1.20	#2	2300	1250	46.0	35.0

POPE EVANS AND ROBBINS

TABLE V. DATA SUMMARY FOR SO_2 REDUCTION VS. 1359 LIMESTONE PARTICLE SIZE, BED DEPTH AND TEMPERATURE

FBC Test No.	Bed Depth Inches	Bed Temp. °F	Particle Size Microns	Stoich. Ratio	SO_2 Conc. ppm Initial	Final	% SO_2 Reduction
65	10	1540	1680	2.6	2500	1870	28.0
		1530	1410	2.6	2500	1870	28.0
		1530	1000	2.6	2500	1700	34.0
66	10	1530	840	2.4	2500	1700	34.0
		1530	420	2.6	2500	1940	24.0
		1550	44	2.6	2500	760	72.0
67	18	1520	840	2.5	2450	1250	49.0
		1580	420	2.5	2450	1600	34.6
		1550	44	2.6	2400	550	77.0
68	10	1770	840	2.6	2550	2150	15.5
		1810	420	2.8	2500	2230	12.5
		1770	44	2.7	2500	1700	31.0
69	18	1770	840	2.7	2500	2150	15.1
		1750	420	2.6	2500	2230	11.0
		1700	149	2.7	2500	1700	32.0
		1750	44	2.5	2500	1700	32.0
70	10	1520	149	2.8	2500	1750	30.0
	10	1850	149	2.8	2500	2100	13.7
	18	1550	149	2.8	2500	1250	50.0
	18	1830	149	2.8	2500	1900	23.3
71	10	1600	840	2.6	2500	1750	30.6
		1620	840	2.6		1800	28.2
		1650	840	2.6		1950	22.0
		1670	840	2.6		2050	18.0
	10	1670	420	2.8	2580	2100	18.6
		1690	420	2.8		2250	13.0
	10	1630	149	2.7	2580	2000	22.5
		1620	149	2.7		1900	26.2
		1670	149	2.7		2100	18.6
	10	1660	44	2.6	2620	1350	49.0
		1670	44	2.6		1450	45.0

TABLE V. (Continued)

Cst o.	Bed Depth Inches	Bed Temp. °F	Particle Size Microns	Stoich. Ratio	SO_2 Conc. ppm Initial	Final	% SO_2 Reduction
72	18	1700	840	2.6	2530	1800	29.0
		1640	840	2.6		1650	34.7
		1610	840	2.6		1500	41.0
	18	1660	420	2.7	2530	2000	21.0
		1650	420	2.7		1900	25.0
		1640	420	2.7		1850	26.9
	18	1700	149	2.6	2530	1900	25.0
		1640	149	2.6		1700	33.0
		1610	149	2.6		1600	36.7
	18	1660	44	2.8	2530	1200	52.5
		1650	44	2.8		1100	56.5

The data trends indicate that the fine dolomite (1337) is again more reactive than the fine limestone (1359). The dolomite hydrate proved to be somewhat more reactive than the fine, raw stone. Utilization of the fine raw dolomite is 42% and 37% at respective Ca/S ratios of 1.0 and 2.0.

The 1359 limestone hydrate appears to be as reactive as the fine raw stone when used with the 2.6% sulfur coal. Utilization of the fine raw stone indicated by the trend is 39%, 32% and 27% for respective Ca/S ratios of 1.0, 2.0, and 3.0.

A comparison of the method of sorbent feed into the FBC failed to point up a clear advantage for any particular method of sorbent feed although, in general, the most favorable observations were made with the #2 feed system.

Test data for the series are summarized in Appendix B. Sulfur balances are presented in Appendix C.

6.3 TESTS FOR INDEPENDENT EFFECTS OF BED TEMPERATURE, BED DEPTH, SORBENT PARTICLE SIZE, SORBENT DISTRIBUTION, AND SUPERFICIAL VELOCITY

A statistical experiment was conducted to establish the separate effects of bed temperature, bed depth and sorbent particle size on the desulfurization reaction in the fluidized bed. The information provided by the experiment was intended to form a basis on which to estimate the necessity for fine grinding and to establish the relative advantage of more massive beds which must be supported by added fan power.

The experiment was conducted in the FBC after modification to permit control of bed temperature with a movable internal cooling surface. The modification is described in Section 5.1. A sintered ash bed, sized -7 +14 mesh, was fired with Ohio #8 seam washed coal which, in this case, contained 3% sulfur. The 1359 limestone was selected as the sorbent because of its apparent durability observed in previous tests. The sorbent was injected with the #2 Feeder system described in Section 5.3. The sorbent feed rate was controlled as closely as possible to a stoichiometric ratio of 2.6, a ratio estimated to yield an 80% SO_2 reduction with the -325 mesh particle size.

The 1359 limestone was prepared in seven sizes ranging from 12 mesh to -325 mesh. These size groups were -12 +14, -14 +16, -18 +20, -20 +30, -40 +50, -100 +200 and -325 U.S. Standard Mesh. The particle sizes represented by the largest screen size in these ranges

correspond to 1680, 1410, 1000, 840, 420, 149 and 44 microns respectively. The first three were tested at a single test condition for reference, while the last four were tested over the range of temperature and bed depth. The elutriation particle size, i.e., the smallest particle size that remains in the bed at the 14 fps superficial gas velocity was 30 mesh. This size was estimated from the intermediate law as shown in Appendix A, Enclosure 18.

The bed depth was varied at two levels--10 inches and 18 inches. The bed temperature was varied in three levels, one value at the extreme ends of the operating range 1500°F - 1800°F and one intermediate temperature.

The results of the test, indicating the reduction in SO_2 with sorbent particle size, bed temperature and depth are summarized in Table V and plotted in Figure 27.

The data trends suggest the following conclusions:

a. A sorbent ground to pass through a 200 mesh screen can be expected to be much less effective than sorbent ground finer so as to pass through a 325 mesh screen.

b. The reduction-particle size curve appears to pass through a minimum reduction in the particle size range of -40 +50 mesh (420 microns). Such a minimum might occur from loss in bed residence time without a compensating increase in reactive surface.

c. Increase in bed depth (and residence time) is less effective with the -325 mesh particle than with larger sizes. In every case the advantage of increased residence time declines as the bed temperature is raised from 1550°F to 1800°F.

d. All particle sizes are more effective in sulfur capture at bed temperatures of 1550°F than at 1800°F. This result is consistent with thermodynamic equilibrium data reported by others[1] and with performance observed in the regeneration of limestone beds (Section 6.9).

[1]Battelle Memorial Institute, "Fundamental Study of Sulfur Fixation by Lime and Magnesia," June 30, 1966

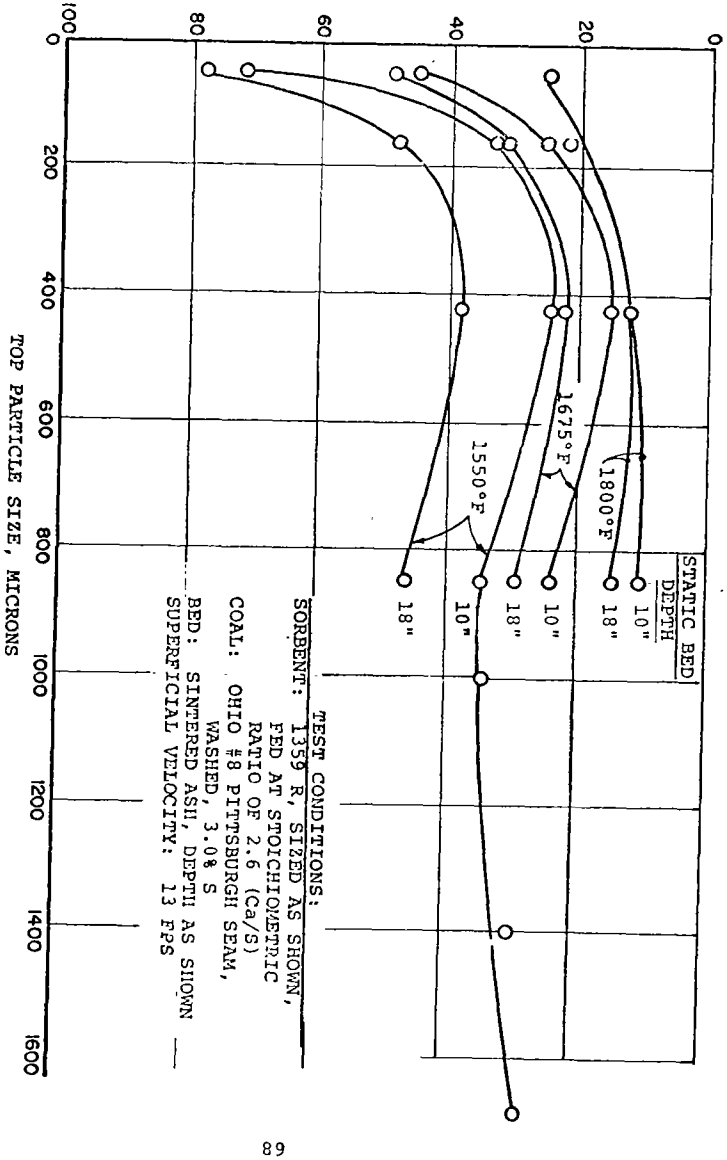


FIGURE 27. VARIATION IN SULFUR DIOXIDE REDUCTION WITH SORBENT PARTICLE SIZE, BED DEPTH AND BED TEMPERATURE

Other data are summarized in Appendix B. Emissions of nitric oxide and hydrocarbons were the same as observed in previous tests (Sections 6.5 and 6.6).

SORBENT DISTRIBUTION TEST

A test procedure was carried out to investigate the possibility of improving the desulfurization efficiency by better distribution of the sorbent around the periphery of the FBC. For the test procedure a second two-point feeder was placed on the side of the FBC opposite to the first as shown in Figure 16. The two feeders were then connected to the FBC by pneumatic tubes which would permit injection of sorbent on one, two and all four sides. The feeders were calibrated precisely and the feed rates adjusted to maintain a constant sorbent feed into the bed as the number of injection points was increased. The 1359 limestone, in a -325 mesh size, was injected into the FBC at first one, then two, and then four sides with a constant rate of 2.0 stoichiometric ratio.

Emissions monitored during the test are shown in Figure 28. The results failed to show an improvement with increase in the number of injection ports. At the end of the test the Ca/S ratio was increased to 3 to check for a possible defect in the instrumentation which might have prevented a variation. The decline in SO_2 emission at the higher ratio indicates normal functioning of the instrument.

The results indicate that single-point injection in the FBC is adequate to effect the optimum SO_2 reductions for the bed volume. For the larger bed volume in the FBM, the results suggest that distribution may not be a problem. The two-point injection appeared to be adequate in the FBM but a similar distribution test was not made.

In a subsequent test the -325 mesh limestone was injected above the bed for a comparison of the SO_2 control effectiveness with the inbed injection. Test conditions were otherwise the same as employed in the distribution tests. The coal and sorbent feed rates were held constant as the sorbent feed was diverted from above the bed to the base of the bed.

The results showed a marked loss in effectiveness of capture when feeding the sorbent above the bed as compared to the usual in or below the bed feeding.

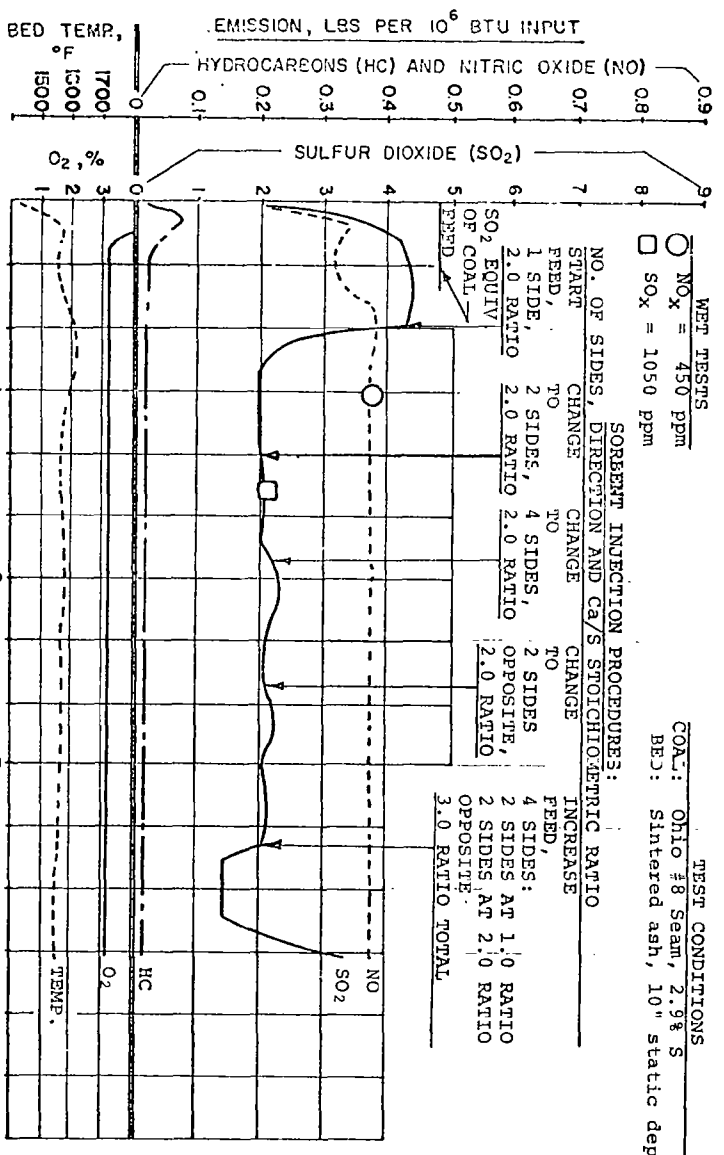


FIGURE 28. EMISSIONS FROM FBC TEST NO. 73 BURNING A MEDIUM SULFUR COAL AND INJECTING -325 MESH, 1359 LIMESTONE ON 1, 2 AND 4 SIDES

The results are summarized as follows:

Test Condition	Ca/S Ratio	SO_2 Conc., ppm	Reduc., %	Limestone Utilization, %
No sorbent input	--	2600	--	--
Sorbent, above bed	1.75	1750	29	16.6
Sorbent, base of bed	1.75	1000	62	35.4

Emission curves for this test (FBC 75) are presented in Appendix A, and other data are summarized in Appendix B.

SUPERFICIAL GAS VELOCITY TESTS

Superficial gas velocity is defined as the flue gas velocity which would exist in the combustion unit at the operating temperature without the fluidized-bed material. This parameter is directly related to heat release rate, a factor which marks a principal advantage of the fluidized-bed combustion process over other methods of firing. Operation of the fluidized-bed boiler at less than maximum heat release rate (and maximum gas velocity) would not be beneficial unless an advantage with respect to sulfur emission control could be demonstrated. This control should improve with increased sorbent residence time afforded by a reduction in gas velocity.

Tests were conducted in the FBC to investigate the effect on sulfur dioxide emission when the superficial gas velocity was reduced from 13 to 6 feet per second without change in the stoichiometric sorbent feed rate. The finely divided limestone was injected into the bed through the #2 feeder at a Ca/S ratio of 2.7. As the coal and air rates were reduced to effect the lower superficial velocities, the sorbent feed was reduced in proportion to maintain the Ca/S ratio. A sintered ash bed, 10 inches deep, was operated at 1550°F with 3% oxygen in the flue gas. The test was repeated at a Ca/S ratio of 2.0, a position on the curve where the sorbent utilization is greater. The test results, summarized in Table VI, indicate little or no improvement in SO_2 reduction or limestone utilization when the superficial velocity was decreased. Emission curves for these tests (FBC 76 and 77) are included in Appendix A.

6.4 SULFUR TRIOXIDE EMISSION

Sulfur trioxide formation is favored by the low operating temperature of the fluidized-bed combustor according to thermodynamic equilibrium theory, but its formation is slow in the absence of a catalyst because of the high activation energy of the dioxide. Gas samples taken from the stack at 600°F and cooled in a condenser at 140°F indicated small concentrations of 30 to 50 ppm in a field of 3800 ppm sulfur dioxide. The sulfur trioxide disappeared completely with sorbent injection.

A six-hour test was conducted in the FBC without sorbent injection to determine if the low concentrations resulted from residual sorbent in the test system from a previous test. The test indicated a value of 39 ppm sulfur trioxide after six hours of operation and the concentration was not increasing. Emission curves for the test (FBC 63) are shown in Appendix A, Enclosure 22. Tests for sulfur trioxide emission during the limestone bed tests failed to show any sulfur trioxide in the flue gas.

6.5 HYDROCARBONS EMISSION

The FBC test results showed that emission of hydrocarbons is primarily controlled by oxygen concentration in the flue gas and in turn by the excess air supplied to the combustion. Although the fluidized-bed combustion process can be operated at 1% oxygen in the flue gas (5% excess air) without evolution of visible smoke, concentrations of hydrocarbons were found to be high.

Typical variation in hydrocarbons concentration with flue gas oxygen concentration is shown in Figure 29. Four curves are plotted to show the variation with bed depth. Point temperatures are indicated. The percentage excess air corresponding to the oxygen content of the flue gas is also shown in the figure.

The data show that at 1% oxygen concentration in the flue gas the hydrocarbons concentration, measured as methane, may vary from 400 to 1500 ppm. When the oxygen concentration is increased, the hydrocarbons concentration is reduced to ~50 ppm at 3% and to 0 ppm at 4% oxygen.

The variation of hydrocarbons concentration with bed depth and temperature appears to indicate that low emission is favored more by high bed temperature than by deep beds. These results are not consistent, however. The results of many subsequent FBC tests, conducted at 3% O_2 in the flue gas, indicate that bed temperature and

TABLE VI. DATA SUMMARY FOR THE SUPERFICIAL VELOCITY TESTS

Test No.	Coal Rate lb/hr	Superficial Velocity fps	1359R Limestone Rate lb/hr	Stoich. Ratio	SO_2 Conc. ppm	SO_2 Reduction %	Calculated Limestone Utilization %	Particulate Emission lb/hr
76								
1	109	12.7	--	--	2550	--	--	1.0
2	109	12.7	29.0	2.67	550	78	29.2	3.1
3	89	10.3	23.0	2.69	600	76	28.2	2.4
4	68	8.0	18.0	2.76	500	80	29.0	2.0
5	52	6.0	14.0	2.80	600	76	27.1	2.0
77								
1	110	12.8	--	--	2600	--	--	1.1
2	110	12.8	21.4	2.00	780	70	35.0	2.1
3	92	10.6	17.2	1.95	740	72	37.0	1.9
4	70	8.1	14.0	2.10	700	73	34.8	1.8
5	54	6.3	10.6	2.05	700	73	35.6	1.6

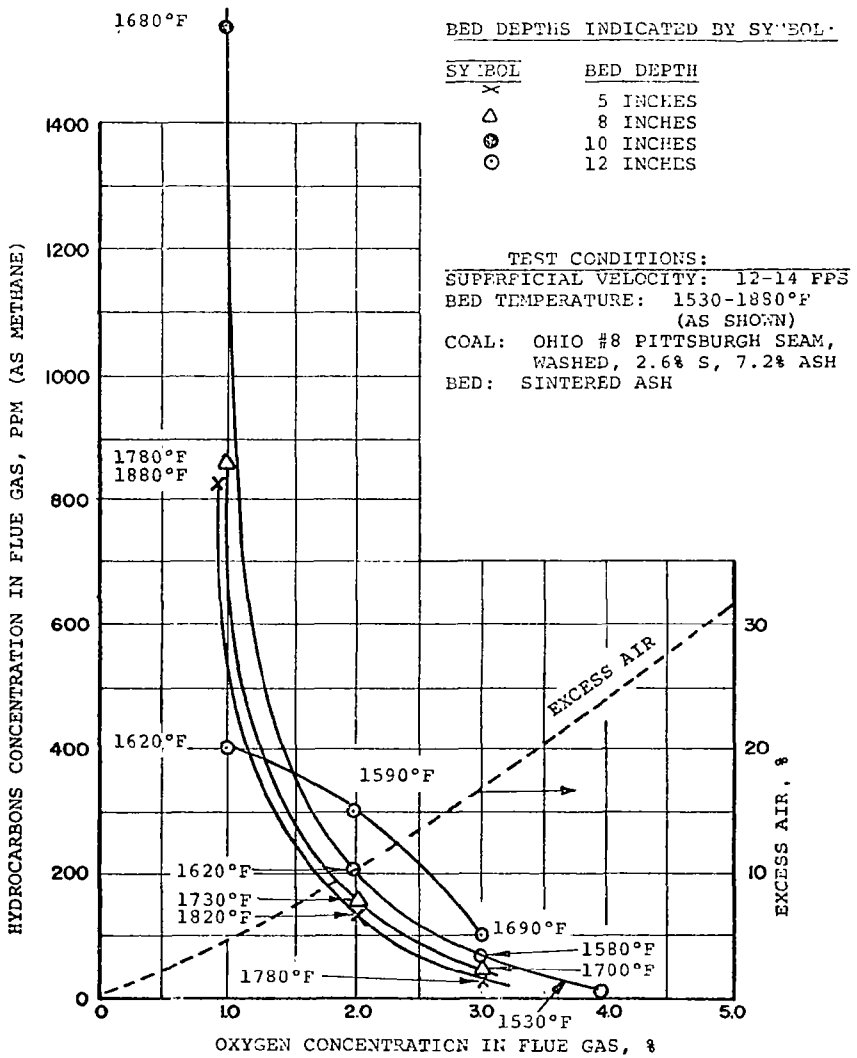


FIGURE 29. HYDROCARBONS VARIATION WITH FLUE GAS OXYGEN CONCENTRATION IN THE FBC OPERATION

bed depth have a negligible effect in comparison with the oxygen concentration, and that hydrocarbons can be limited to ∿50 ppm at this value.

A flue gas oxygen content of 3% corresponds to an excess air rate of approximately 17% above the stoichiometric requirement. Any air in excess of the stoichiometric requirement will result in a thermal loss chargeable to the boiler since heat transfer surface cannot be economically provided to recover heat by cooling flue gas below about 250°F. In addition, excess air removes heat from the bed which must be recovered, in part, by convective heat transfer surface which is less effective than in-bed heat transfer surface. For this, as well as other reasons, e.g. larger fans, increasing the excess air requirement increases the capital cost of the boiler system.

The thermal loss due to excess air is partially compensated for by the energy released by burning the hydrocarbons to the 50 ppm level. So, for example, where the excess air is increased from 5% to 17% and the flue gas exits at 400°F, an efficiency loss of about 0.8% is incurred due to excess air while the hydrocarbons, assumed to be methane, drop from 800 ppm to 50 ppm. The combustion of the hydrocarbons releases an additional 110 Btu per pound of coal fed for an efficiency gain of ∿0.9%. In this example the optimum operating point might be around 10-12% excess air, if maximum thermal efficiency were the only goal. Most of the tests in the program were made at 17% excess air, primarily because of the hydrocarbon emission. Concentrations of 50 ppm at this level correspond to ∿.02 lbs of methane per MBtu input. This emission would appear to be favorable in comparison with conventional boilers, but data on the latter operating at the same excess air level are lacking. Further work on coal feeding systems may provide a basis for lower excess air operation without an increase in hydrocarbons.

Injection of sorbent materials into the bed does not increase hydrocarbons emissions in steady-state operation. Sorbent injection at rates as high as 60 lbs per 100 lbs of coal failed to show a significant increase in hydrocarbons emission.

Hydrocarbons generated by low excess air or reducing conditions in the bed can be burned effectively by injection of air above the bed in sufficient quantity to make up the 3% oxygen content in the flue gas. This result is discussed further in Sections 6.8 and 6.9.

Carbon monoxide concentrations in the flue gas from the FBC may be as much as 0.5% at 1% oxygen content but are negligible at the 3.0% oxygen level.

TEST CONDITIONS:
SUPERFICIAL VELOCITY: 14 FPS
BED TEMPERATURE: 1750°F
BED: 8" STATIC DEPTH
COAL: OHIO #8 PITTSBURGH SEAM, UNWASHED
SORBENT FEED RATE: ZERO

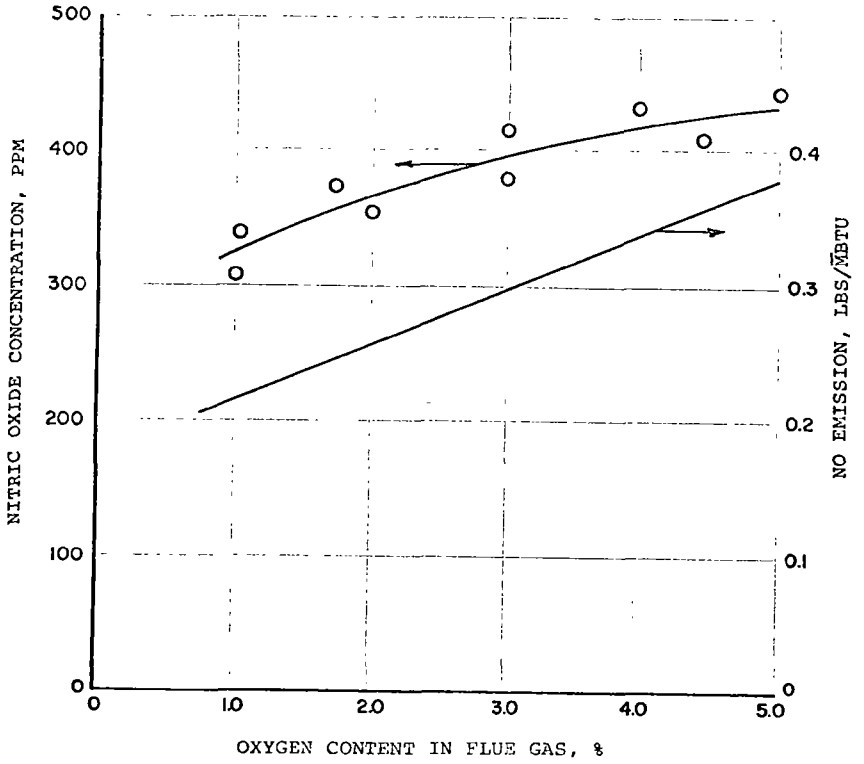


FIGURE 30. TYPICAL VARIATION IN NITRIC OXIDE CONCENTRATION WITH OXYGEN CONTENT IN THE FLUE GAS FROM THE FBC

6.6 OXIDES OF NITROGEN EMISSION

Emission of nitric oxide from the FBC was found to vary with the oxygen concentration in the flue gas as determined by the excess air rate. Nitric oxide in the flue gas was found to increase from 320 ppm at 1.0% oxygen content to 440 ppm at a 5.0% oxygen content. This variation is shown in Figure 30 together with the emission in terms of pounds of NO per MBtu input. The emission at 3% oxygen content is 0.30 lbs per MBtu input.

In a number of tests conducted at 3% oxygen content in the flue gas the nitric oxide concentration varied from 220 ppm to 470 ppm with no apparent correlation with bed temperature. Data points observed when burning a 4.5% S, 2.5% N_2 coal with 3% O_2 in the flue gas are shown in Figure 31. Theoretical curves are also presented in the figure to show the thermodynamic equilibrium concentrations of nitric oxide that should exist for the oxygen concentrations that exist across the bed, i.e., 20% O_2 in the inlet air and 3.0% O_2 in the flue gas, and for the range of temperatures investigated. The shaded area in the figure is the area in which the data would theoretically be expected to fall. For the method used to produce the theoretical curves see Appendix A, Enclosure 19.

The figure shows that NO concentration should not exceed 100 ppm at a bed temperature of 1550°F. The fact that concentrations of 300 to 400 ppm were observed suggests the presence of local temperatures around the coal higher than those observed by the bed thermocouples. Another possibility is that nitrogen in the coal may play a role in the reaction. One test conducted with two coals of different nitrogen contents is discussed in the FBM test results (Section 7.3).

Nitric oxide emissions from the FBC appeared to be unrelated to bed depth at the 3% oxygen concentration level. Variation in bed depth during FBC Test 44 produced the following results:

Bed Depth	5 in.			8 in.			12 in.		
O_2 Conc., %	1	2	3	1	2	3	1	2	3
NO Conc., ppm	280	340	380	305	360	400	360	370	380

As a rule, the use of sorbent materials was observed to have little or no effect on nitric oxide emission. Steady state concentration values were found to decrease and increase with sorbent injection. In two instances, however, a definite reduction was observed.

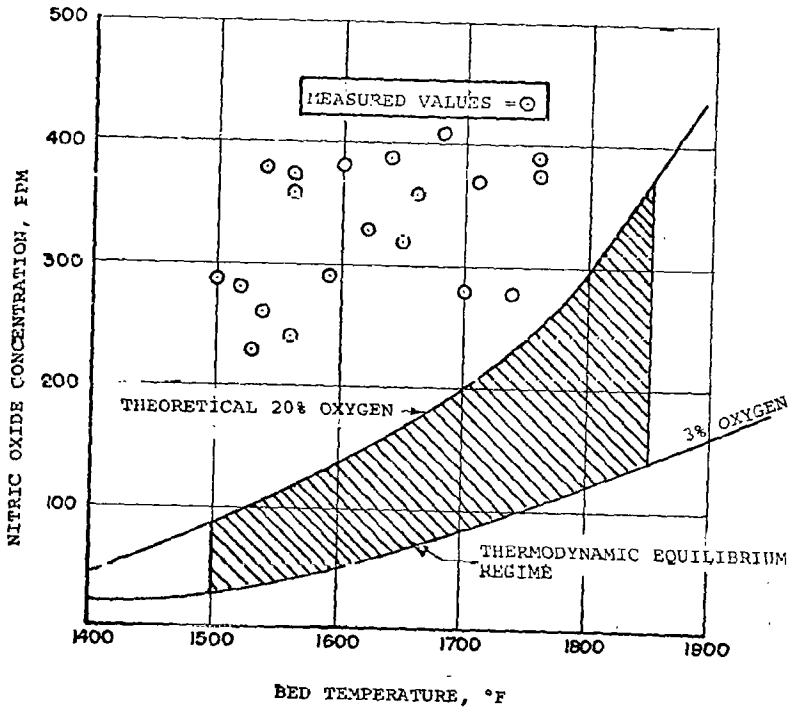


FIGURE 31. MEASURED VALUES OF NITRIC OXIDE CONCENTRATION IN THE FLUE GAS AT 3% OXYGEN AND VARIOUS BED TEMPERATURES SHOWN WITH THEORETICAL EQUILIBRIUM VALUES FOR THE TEMPERATURE - O_2 CONTENT REGIME

During FBC Test 18 conducted with the unwashed coal, a sintered ash bed 14 inches deep and operating at 1750°F with 2% O_2 in the flue gas, the NO concentration was reduced from 250 ppm to 60 ppm when 1337 raw dolomite in a -7 +14 mesh size was injected through the #1 feeder at a Ca/S ratio of 1.75. A careful examination of the instrumentation failed to reveal a defect which might have caused the reduction. The reduction was real but its cause undetermined. A similar effect was observed in FBC Test 25.

Although the discussion has been directed to the emission of nitric oxide, NO, the results are applicable to total oxides emission, NO_X. Tests to determine all the oxides by the phenoldisulfonic acid procedure indicated approximately the same concentrations as the infrared absorption unit which is sensitive to NO only. Concentrations of the oxides of nitrogen higher than nitric oxide are estimated by difference to vary in the range of 10 to 30 ppm.

On the average, the nitric oxide emission from the FBC is approximately 0.30 lbs/MBtu input at the 3% oxygen content in the flue gas. The corresponding concentration is 375 ppm.

6.7 PARTICULATE EMISSION

Most of the ash from the coal burned in the FBC was elutriated as fly ash from the bed and collected in a cyclone. The location of the cyclone in the test assembly is shown in Figure 6. Isokinetic samples taken downstream of the cyclone indicated that up to 10% of the fly ash was discharged from the system. This high particulate loss reflects principally the poor collection efficiency of the test cyclone with the fine ash.

When the finely divided sorbents were added to the system, the particulate emission was increased. Typical emission data with and without sorbent addition are summarized as follows:

Computed Ash Input lb/hr	1359 Limestone Input lb/hr	Fly ash Collected, lb/hr	Fly ash Discharged lb/hr
12.8	0	22.0	1.5
12.9	0	23.2	2.4
12.6	21.4	41.0	3.9
12.9	28.0	43.4	4.9

Although the particulate emission is increased by sorbent addition, the results show that the bulk of the sorbent is retained in the collector despite the -325 mesh particle size.

The particulate emission from the FBM cyclone during a similar sorbent test was counted by microscope for particle size distribution. The results are discussed in Section 7.5.

The fact that the total fly-ash rate is larger than the computed ash input, without sorbent addition, is due to the presence of unburned carbon in the fly ash. The fly-ash carbon content may vary from 45% to 60%. When sorbent is added to the system, the fly-ash carbon content is reduced to about 30% apparently from dilution with the spent sorbent.

The energy lost from unburned carbon in the fly ash amounts to about 10% of the input energy. Recovery of this energy through the use of the Carbon-Burnup Cell concept is now under investigation. The energy can be recovered to some extent with recirculation of the fly ash through the combustor. Recirculation of the fly ash containing spent sorbent improved sulfur capture in some instances but the results were inconsistent.

6.8 OPERATION AT REDUCING CONDITIONS

Three tests were made in the FBC with the bed at slightly reducing condition. The reducing conditions were produced by stabilizing the combustion at 1% oxygen content in the flue gas and then decreasing the air rate by 10% with constant coal feed. Since the 1% flue gas oxygen content corresponds to 5% excess air (From Figure 29), an air rate reduction of 5% would effect stoichiometric conditions. A reduction of an additional 5% in the air rate produces a 5% deficiency of oxygen in the bed. After the 10% air reduction, air was supplied above the bed to reestablish the oxygen concentration in the flue gas at 1%.

For the effect on sulfur control, 1337R dolomite was added at a Ca/S ratio of 1.1 during the reducing condition. When the sulfur dioxide concentration dropped to a lower steady-state level, the operation was reverted to the oxidizing condition without change in the coal or sorbent feed rates. Nitric oxide and hydrocarbons were monitored continuously.

The results of the tests are summarized in Table VII. Sulfur dioxide reduction is shown to improve with the oxidizing bed (31.2% vs 21.4%) whereas NO reduction is favored by reducing conditions in the bed (240 ppm vs 320 ppm concentrations). Hydrocarbons concentrations appear to be greater with reducing conditions, but the difference observed may have been due to very small changes in the oxygen content. The rapid variation of hydrocarbons emission with flue gas oxygen at the 1% level was discussed in Section 6.5.

Subsequent tests conducted at 3% O_2 in the flue gas indicated a more effective reduction in NO emission with reducing conditions. At this oxygen level, hydrocarbons can be consumed with overbed air. These points are discussed in Section 6.9.

6.9 FBC OPERATION WITH A LIMESTONE BED

6.9.1 General

The FBC was operated with a bed consisting entirely of 1359 limestone instead of inert ash. Emissions were monitored from the combustion of a washed #8 Pittsburgh Seam coal in the bed, and the parameters affecting sulfur retention were investigated. Removal of sulfur retained in the bed was also studied. The overall heat transfer coefficient was determined for comparison with the value observed with the sintered ash beds.

Initial attempts to fire a bed of limestone in the FBC led to problems in bed temperature control. The weight loss and endothermic heat requirement of calcination and the rapid heat removal combined to create an unstable situation. When the bed became calcined, the bed temperature increased causing attrition losses. The loss of bed in turn reduced the heat removal rate and further increased the temperature. The operation could probably have been stabilized by trial-and-error addition of limestone. It was decided, however, that an independent means of temperature control would solve the problem and provide a desirable control capability during the investigation. The independent temperature control was effected with a sleeve installed in the FBC to retard the heat transfer through the walls and a movable coil installed in the bed. This modification was discussed in Section 5.1 and the coil and sleeve arrangement shown in Figure 7. The bed temperature was controlled by adjusting the vertical position of the coil in the bed.

TABLE VII. DATA SUMMARY FOR OPERATION AT REDUCING CONDITIONS*

FBC Test No.	6	9	10
Bed Temp. °F	1800	1800	1750
SO_2 Reduction, %[1]			
Reducing Bed	17.8	16.9	21.4
Oxidizing Bed	22.6	NA	31.2
NO Concentration, ppm			
Reducing Bed	280	220	240
Oxidizing Bed	330	NA	320
HC Concentration, ppm			
Reducing Bed	Erratic	NA	560
Oxidizing Bed	500	NA	435

[1]Addition of 1337R dolomite -7 +14 mesh at 1.1 ratio with 4.5% Sulfur coal

* Flue gas oxygen content 1.0% for all conditions

See text for further description of test conditions.

The sleeve and coil arrangement permitted the use of deeper beds, i.e., 20" as compared to 10" in the open unit. It was found convenient to ignite the bed with a 10" depth and then add an additional quantity, even though the whole of the bed could have been ignited and stabilized. Positioning the coil provided the fine adjustment of temperature.

The 1359 limestone was selected because of its apparent durability and screened to a -10 +20 mesh particle size. This particle size selection is somewhat smaller than the sintered ash bed size (-7 +14 mesh) because of the greater density of the raw limestone (2.6 vs 1.8 specific gravity). The particle size distribution is shown in Appendix A, Enclosure 20. Typically 75 lbs of the limestone made up the original charge with an additional 60 lbs added for an initial raw bed weight of 135 lbs and a depth of 16 to 17 inches.

6.9.2 Sorption of Sulfur

The tests were made with Ohio #8 (Pittsburgh) Seam washed coal containing about 3% sulfur. The coal was fired at a rate of ~65 lbs per hour into a bed having an initial weight of ~135 lbs. No sorbents were added other than the bed limestone. The superficial velocity was maintained at the same level employed in the sintered bed operation, i.e., 12-14 fps. Concentrations of sulfur dioxide, nitric oxide and hydrocarbons in the flue gas were monitored continuously and spot samples were taken for sulfur trioxide and oxides of nitrogen. All sorption tests were conducted at 3% oxygen in the flue gas unless otherwise noted.

The bed operating temperature was found to be important with respect to sulfur retention. At 1400°F the limestone did not calcine and consequently did not retain sulfur. At 1900°F the retention was minimal as expected from previous work (Section 6.3, Figure 27). The temperature range of 1500°F - 1600°F appeared to be most favorable for sulfur sorption.

The results of sorption tests indicated that sulfur in the coal could be sorbed almost completely for a period of two to three hours after which time sulfur dioxide began to appear in quantity in the flue gas. This behavior is illustrated in Figure 32 which shows the emissions monitored during FBC Test 113, one of the best of the program. The figure also shows the sulfur input in equivalent sulfur dioxide emission. The bed

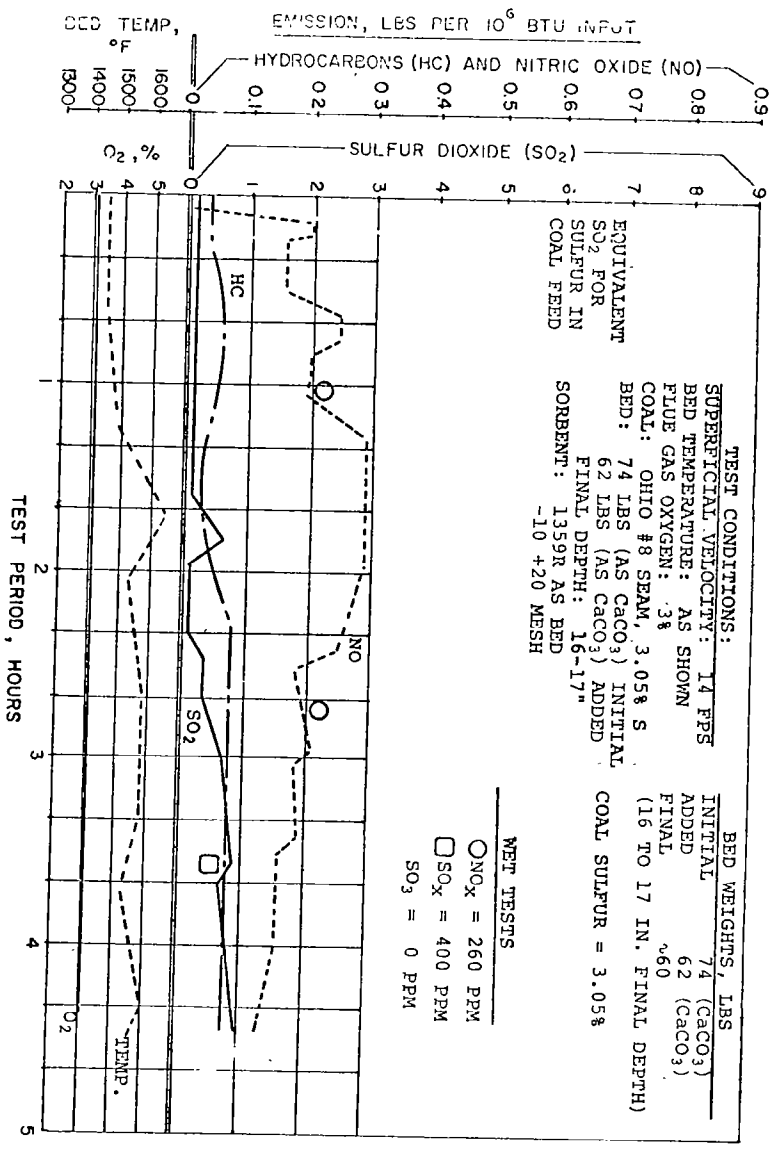


FIGURE 32. EMISSIONS DURING FBC TEST 113 BURNING A MEDIUM SULFUR COAL IN A FLUIDIZED-BED OF 1359 LIMESTONE

temperature curve shows a variation in the range of 1500 F to 1600 F. The oxygen content was fixed at 3%. The initial bed weight was 136 lbs and the depth 17 inches. Emission curves for other tests are presented in Appendix A.

The variation in sulfur and calcium contents of the bed for Test 113 is shown in Figure 33 together with the sulfur and calcium contents of the fly ash. The sulfur content of the bed had increased to 7.4 wt. % at end of the test. This value indicates that 16% of the bed limestone had been utilized in sulfur capture. The increase in calcium content of the bed is due to the weight lost in calcination.

The sulfur contents of the fly ash indicate that a small part of the sulfur is retained in the fly ash. The rate of sulfur flow in the system was indicated to be the following:

SULFUR RATE, LBS/HR

Test Time, Hours	1	2	3	4
Flue gas output	0.00	0.08	0.60	0.70
Fly-ash output	0.35	0.28	0.32	0.38
Bed retention	1.56	1.64	0.90	0.80
Total output	1.91	2.00	1.82	1.88
Input	1.98	1.98	1.98	1.98
Input less output	0.07	-.02	0.16	0.10

The fact that sulfur in the fly ash remained relatively constant suggests that this sulfur is contained in the fly-ash particle core and is not affected by the bed reaction.

Attrition loss of the bed material was found to be high during the calcination phase but comparatively low afterward. During calcination, 5% to 7% of the calcium in the bed was lost per hour. The calcium loss during subsequent sorption was reduced to a rate of 2% to 4% of the initial calcium charge. These values approximate the loss during regeneration to be discussed in Section 6.9.3.

Loss of unburned carbon in the fly ash during the limestone bed tests indicated substantially the same loss observed with the sintered ash bed, i.e., 9% to 12% of the input energy. Typical variation in fly-ash carbon loss is shown in Figure 34. Heat transfer measurements in the limestone bed indicated a coefficient of 47.0 Btu/ft^{2}hr°F, about the same coefficient observed in the sintered ash bed operation.

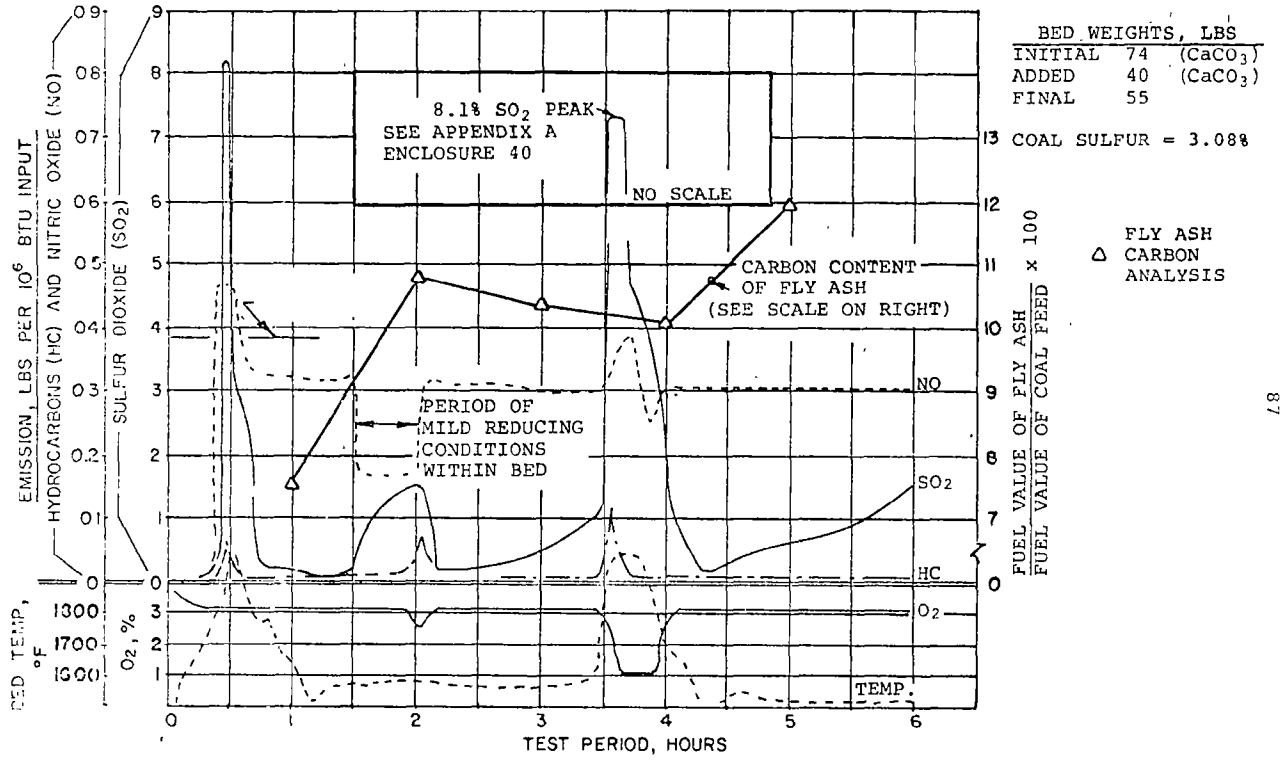


FIGURE 34. EMISSIONS DURING FBC TEST 119 BURNING A MEDIUM SULFUR COAL IN A LIMESTONE BED WITH MILD REDUCING CONDITIONS AND REGENERATION

POPE, EVANS AND ROBBINS

87

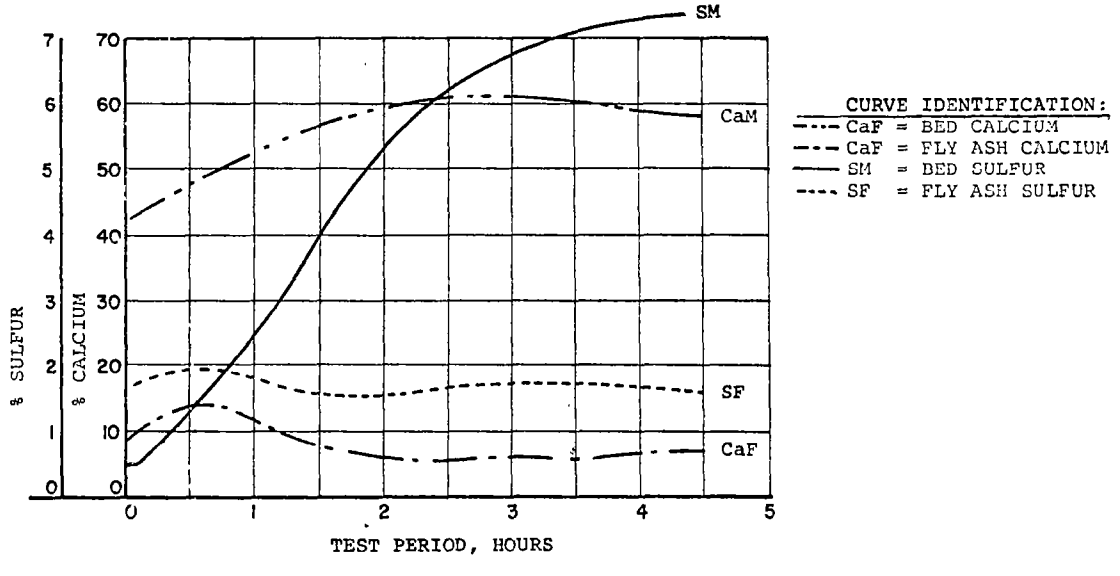


FIGURE 33. VARIATION OF SULFUR AND CALCIUM IN BED AND FLY ASH DURING FBC TEST 113

86

TABLE VIII. DATA SUMMARY FOR THE 1359 LIMESTONE BED TESTS

FBC Test No.	SO_2 Input Equiv. * lbs/MBTU	Bed Mass Initial lbs	Bed Mass Final lbs	Init. Bed Depth in.	Flue Gas O_2 %	Bed Temp. °F	Time for 20% SO_2 Breakthrough hrs.	NO Emission lbs/MBTU	HC Emission lbs/MBTU	Test Condition Remarks
111	4.8	100	97	14.0	3.0	1420	0.25	0.38	.02	Sorption - bed did not calcine
113	4.8	136	60	19.5	3.0	1500 to 1600	4.5	0.15 to 0.28	.04 .06	Sorption
114	4.9	119	49	17.0	3.0	1490 to 1530	2.5	0.21	0.0 0.12	Sorption
"	4.9	--	--	--	3.0	1920	--	0.52	0.0	Desorption 1.5% Peak SO_2 conc.
115	4.8	141	64	20.0	3.0	1580	Est 3.7	0.24	N.A.	Sorption-20% breakthrough delayed .6 hrs by lower velocity (8 fps)
"	3.2	--	--	--	--	1580	4.3	--	--	
116	4.8	134	55	19.2	3.0	1550	3.7	0.18	N.A.	Sorption 1st cycle
"	4.8	--	--	--	1.0	1920	--	0.28	N.A.	Desorption 2.5% Peak SO_2
"	4.8	--	--	--	3.0	1610	2.6	0.21	N.A.	Sorption 2nd cycle
117	4.9	134	59	19.2	3.0	1580	4.1	0.24	0.05	Sorption 1st cycle
"	4.9	--	--	--	0.2	1980	--	0.24	1.7	Desorption 6.0% Peak SO_2
"	4.9	--	--	--	0.3	1600	--	0.18	2.5	Desorption at low temp.

*Symbol MBTU indicates Million Btu.

Results of the stable limestone bed tests are summarized in Table VIII showing the time observed for 20% breakthrough of sulfur dioxide above the bed and other data. The sorption of sulfur is shown, by the comparison, to be favored by increase in the bed mass (and depth), increase in the oxygen content of the flue gas and by a low bed temperature in the range of 1500°F to 1600°F. At lower temperatures, sorption may be limited by failure of the bed to calcine. Reduced sorption at the higher temperature is consistent with results of previous tests (Section 6.3). Recirculation of fly ash did not appear to improve the sorption rate, and reducing conditions in the bed seriously lowered the sorption efficiency of the bed. Lowering the superficial velocity from 12 to 8 fps delayed the 20% breakthrough as might be expected, since the input sulfur is proportional to the superficial velocity.

6.9.3 Desorption of the Sulfated Bed

The difference in sulfur retention in the bed with variation in temperature and oxygen level suggested the possibility that sulfur retained at the favorable conditions could be released by changing either temperature or oxygen level or both. Desorption of sulfur might effectively "regenerate" the bed for further sorption.

The "regeneration" procedure was first carried out in FBC Test 114 with increase in temperature only and no change in the oxygen content of the flue gas. The procedure involved sorption in a bed weighing 119 lbs for four hours at a temperature of 1520°F. The temperature was then increased to 1920°F.

A plot of emissions monitored during the test is shown in Figure 35. Variation of calcium and sulfur contents in the bed is shown in Figure 36. The results show that during regeneration sulfur was released from the bed at a rate sufficient to produce an SO_2 concentration of 1.5% (15,000 ppm) above the bed. At the same time, the sulfur content in the bed decreased from 6% to ~0.8%. A rigorous sulfur balance employing integration of the sorption, input and desorption curves indicated that about 90% of the sulfur sorbed in the bed was released during the period of higher temperature.

The details of the sulfur balance are presented in Appendix A, Enclosure 21.

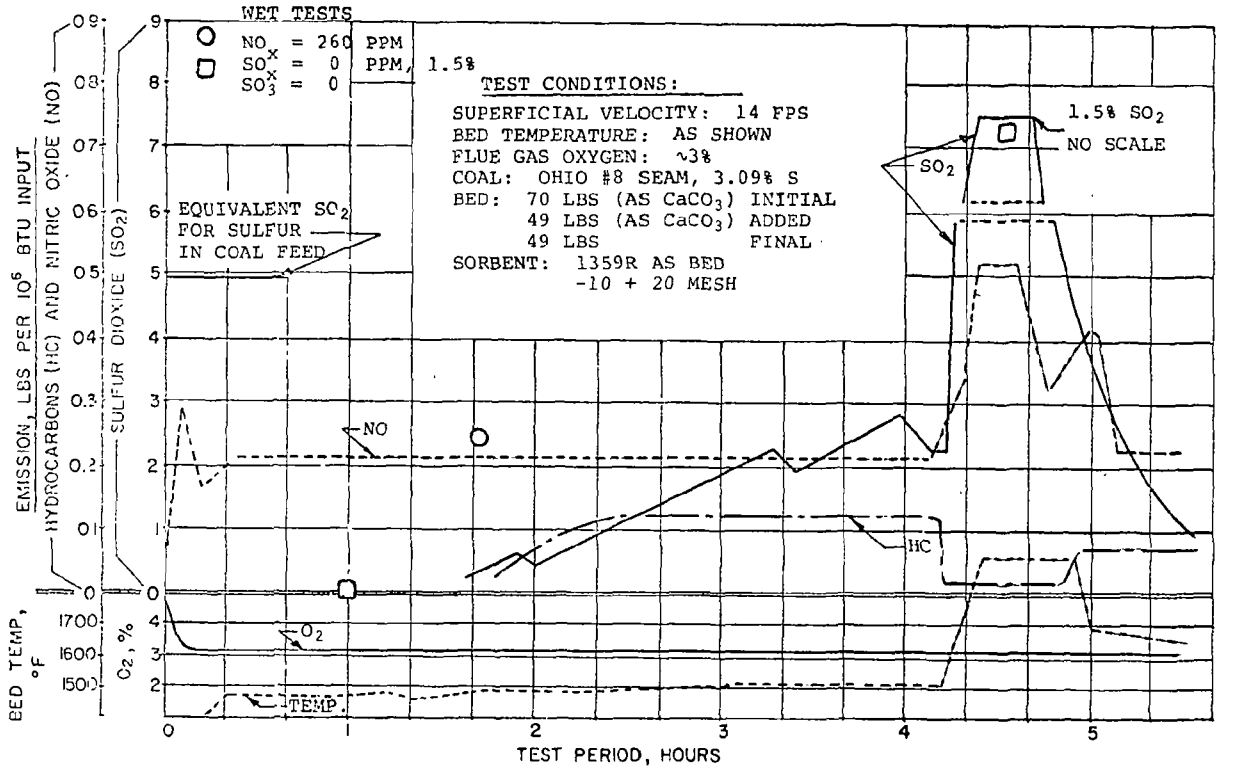


FIGURE 35. EMISSIONS DURING FBC TEST 114 BURNING A MEDIUM SULFUR COAL IN A FLUIDIZED-BED OF 1359 LIMESTONE

POPE, EVANS AND ROBBINS

TABLE VIII. (Continued)

FBC Test No.	SO_2 Input Equiv. lbs/MBTU	Bed Mass Initial lbs	Bed Mass Final lbs	Init. Bed Depth in.	Flue Gas O_2 %	Bed Temp. °F	Time for 20% SO_2 Breakthrough hrs.	NO Emission lbs/MBTU	HC Emission lbs/MBTU	Test Condition Remarks
118	4.7	59[1]	--	11	3.0	1600	0.9	0.28	0.02	Sorption
(continuation of 117)	4.7	--	--	--	0.1	1920	--	0.29	0.16	Desorption Peak SO_2 5.5%
"	4.7	86[2]	--	14	3.0	1550	1.3	0.29	0.02	Sorption
"	4.7	--	--	--	5.0	1550	2.5	0.33	0.00	Sorption 20% breakthrough delayed 1.2 hrs.
"	4.7	--	--	--	0.2	1930	--	0.25	0.15	Desorption Peak SO_2 6.0%
119	4.8	114	55	17	3.0[3]	1570	0.1[4]	0.16	0.02	Reducing condition in bed after 1 hr sorption
"	4.8	--	--	--	3.0	1570	3.4	0.30	0.02	Sorption 1st cycle
"	4.8	--	--	--	1.0	2000	--	0.38	0.05	Desorption Peak SO_2 8.1%
"	4.8	--	--	--	3.0	1520	1.5	0.30	0.02	Sorption 2nd cycle
120	4.8	114	51	17	3.0	1550	2.9	0.24	0.04	Sorption with 80% ash recirculation

Notes: [1] Final bed from 117

[2] Added 27 lbs limestone

[3] Made up with overbed air

[4] Time from start of reducing conditions

FIGURE 36. VARIATION IN BED AND FLY ASH SULFUR AND CALCIUM DURING TEST 114

This result appears to contradict thermodynamic data which show calcium sulfate to be stable at 1900°F under oxidizing conditions. When chemical analysis of the bed showed only small quantities of the more unstable sulfite in the bed before regeneration, it was concluded that local reducing conditions break down the sulfate according to the following relation:

$$Ca\,SO_4 + CO(\text{or } H_2) \rightarrow Ca\,O + SO_2 + CO_2 \text{ (or } H_2O)$$

From the sulfur and calcium analyses, the flow of sulfur was estimated as follows:

SULFUR RATE, LBS/HR

Test Time, hours	1	2	3	4	5
Flue gas output	0.00	0.39	0.85	1.15	10.4
Fly-ash output	0.27	0.25	0.35	0.27	0.3
Bed retention	1.75	1.47	0.70	0.50	-8.8
Total output	2.02	2.11	1.90	1.92	1.90
Input	1.95	1.95	1.95	1.95	1.95
Input less output	-.07	-0.16	0.05	0.03	0.05

This and sulfur balance data for other tests are included in Appendix C.

Before leaving Test 114 it should be noted that the initial weight of limestone in the bed was less than used previously so as to reduce the sorption time and to study the effect of bed mass (or depth). The results in Table VIII show that decreasing bed depth significantly decreases the time for 20% breakthrough of sulfur dioxide.

Subsequent tests, with simultaneous reduction of oxygen content and bed temperature increase, showed that sulfur could be desorbed from the bed more rapidly than with simple change in bed temperature. A concentration of 8.1% SO_2 was observed in regeneration during Test 119 when the oxygen level was reduced from 3% to 1% as the temperature was increased to 2000°F. This variation is shown in Figure 34.

The figure also shows the trend of emissions when reducing conditions were effected in the bed for a short period, i.e., with ~80% of the combustion air passing through the bed and the remainder of the air supplied above the bed to hold constant the 3% oxygen content.

The test showed that the bed could be desorbed with reducing conditions alone and without an increase in bed temperature. The notable effects on nitric oxide and hydrocarbons emissions are discussed in the next section.

The reactivity of the limestone bed throughout a number of sorption regeneration cycles could not be studied in detail, but it was apparent from a second cycle operation that the 20% breakthrough time is shortened on the second cycle. It was also apparent that the rate of increase in sulfur dioxide content in the flue gas, once some appears, is approximately the same as the first cycle rate.

Attrition loss of the bed material during regeneration appears to vary in the range of 2% to 4% (calcium) of the original charge per hour.

6.9.4 Other Emissions

The data summary in Table VIII shows the variation in nitric oxide and hydrocarbons emissions during the limestone bed test series. The nitric oxide emission at low bed temperature varied approximately in the same range observed with the sintered ash bed operation, i.e., .20 to .30 lb/MBtu but appeared to be more responsive to change in bed temperature. During the desorption phase of Test 114, the NO emission increased from 0.21 to 0.52 lb/MBtu with temperature increase from 1530°F to 1920°F at constant (3%) O_2 in the flue gas. The NO emission did not increase with temperature during Test 117, apparently because the O_2 content was reduced to 0.2%. The characteristic reduction in NO emission with lower O_2 content was discussed in Section 6.6.

When reducing conditions were created in the limestone bed during the low temperature sorption phase of Test 119, the NO emission showed a decrease from 0.30 to 0.16 lb/MBtu despite the 3% O_2 in the flue gas (supplied by overbed air). This result indicates that NO emission can be limited by a simple form of two-stage combustion. Unfortunately, this particular mode of operation did not favor sulfur sorption in the bed as indicated in Section 6.9.3.

The reducing conditions phase of Test 119 also pointed up the fact that hydrocarbons can be consumed with overbed air at the 3% oxygen level. At this value the hydrocarbon emission remained constant at 0.02 lb/MBtu in the change from oxidizing to reducing conditions. The emission varied in the range of 0.02 to 0.05 lb/MBtu at low temperature operations. The

0.12 value observed during Test 114 is thought to be an instrument error. When the temperature is increased during desorption, the hydrocarbons disappear except when the oxygen content is lowered simultaneously. At the lower oxygen levels, the hydrocarbons emission is sharply increased at any temperature.

Sulfur trioxide emission during the limestone bed operation was zero.

Particulate emission during the sorption phase of the limestone bed tests and the energy lost in unburned carbon are summarized as follows:

FBC Test No.	Fly ash Collected lbs/hr	Carbon Content %	Discharge to Atmos. lbs/hr	Coal Input lbs/hr	Energy Loss in Unburned Carbon % of input
113	14.0	46	1.6	64	11.7
114	15.0	43	1.4	65	12.0
115	15.6	42	1.5	65	11.6
116	14.9	42	1.2	64	11.2
117	16.6	39	1.7	64	11.5
118	14.4	43	1.2	62	12.4
119	14.8	38	1.5	65	10.7
120	12.8	47	1.3	61	11.9

The particulate emission was about the same as observed with the sintered ash bed operation and was less than that observed with a sintered ash bed and fine sorbent injection. The latter was discussed in Section 6.7. The energy lost in unburned carbon is about the same loss observed with the sintered ash bed operation.

7. RESULTS OF BOILER MODULE (FBM) TESTS

Procedures employed in the FBM tests involving both coarse and fine limestone injection are discussed in Section 5.2. In general, the test conditions selected were those observed to favor sulfur emission control during the FBC tests. Tests for sulfur trioxide emission from the FBM are not discussed separately in this section since results are comparable to low values observed in the FBC tests (Section 6.4). The method of gas sampling is discussed in 5.4 and the sampling system is shown in Figure 20. Variations in emissions during the course of the tests are shown in Appendix A. Complete data summaries are presented in Appendix B and sulfur balances in Appendix C.

7.1 SULFUR DIOXIDE EMISSION

Emission of sulfur dioxide from the FBM without sorbent addition indicates that 90% to 95% of the sulfur in the coal appears as sulfur dioxide in the flue gas. The remainder of the sulfur is held in the fly ash. This distribution of sulfur is shown in sulfur balances in Appendix C.

When coarse, raw 1337 dolomite was injected into the FBM while burning the 4.5% sulfur coal, the most favorable calcium utilization observed was 31.2%. The SO_2 reduction was 54.5%, the bed temperature 1600°F, the oxygen content 3.5%, the stoichiometric feed ratio 1.75, and the sorbent particle size -7 +14 mesh. These data and others pertaining to the coarse dolomite addition are summarized in Table IX. The results are comparable to values reported in Table I for the FBC under similar test conditions. The coarse 1359 limestone was not tested in the FBM because of its poor performance in the FBC, as indicated in Figure 22.

When the finely divided sorbents (-325 mesh) were added to the combustion of the 4.5% sulfur coal in the FBM, the sulfur dioxide reductions and calcium utilizations were found to equal those observed in the FBC. The results of the tests are summarized in Table X, and a comparison with FBC data trends is shown in Figure 37. The FBC Trend lines were reproduced from Figure 25.

The FBM results indicate a reduction of 74% at a Ca/S ratio of 1.7 with the 1337 raw dolomite. This reduction is exactly comparable to the FBC results as indicated in Figure 37. A reduction of 74% observed in the

TABLE IX. SULFUR DIOXIDE REDUCTIONS WITH ADDITION OF COARSE (-7 +14 MESH) 1337 DOLOMITE TO THE FBM BURNING A 4.5% SULFUR COAL

FBM Test No.	Dolomite State	Bed Depth in.	Bed Temp. °F	Flue Gas O_2 %	Ca/S Ratio	SO_2 Conc. ppm Initial	SO_2 Conc. ppm Final	% SO_2 Reduction	% Sorbent Utilization
2	Raw	20	1600	3.5	1.75	3400	1550	54.5	31.2
3	"	20	1540	3.8	1.40	3600	2050	43.0	30.7
4	"	15	1720	3.2	1.60	3800	3300	13.2	8.3
5	"	20	1570	3.0	1.13	3800	3000	21.0	18.6
	"	20	1570	3.4	1.65	"	2400	37.0	22.6
	"	20	1570	3.0	2.40	"	2000	48.0	20.0

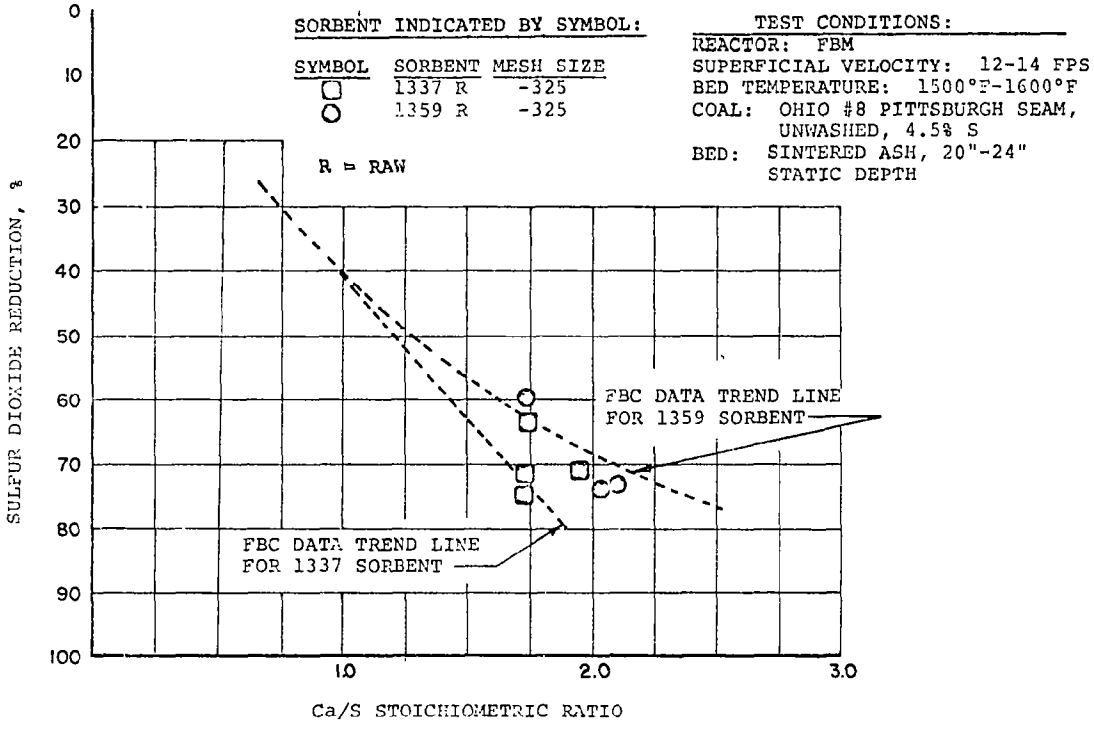


FIGURE 37. SULFUR DIOXIDE REDUCTION WITH SORBENT ADDITION TO THE FBM BURNING A 4.5% SULFUR COAL

TABLE X. DATA SUMMARY FOR INJECTION OF -325 MESH SORBENTS INTO THE FBM BURNING A 4.5% SULFUR COAL

FBM Test No.	Sorbent Type	Bed Depth in.	Bed Temp. °F	Flue Gas O_2 %	Ca/S Ratio	SO_2 Conc. ppm Initial	SO_2 Conc. ppm Final	% SO_2 Reduction	% Sorbent Utilization
25	1337R	24	1550	3.0	1.70	3750	1100	71.5	42.0
	"	"	1520	3.0	1.70	"	950	74.2	43.8
26	1337R	20	1660	3.0	1.70	3750	1350	64.2	37.8
	"	"	"	"	1.90	"	1100	70.9	37.3
27	1359R	20	1570	3.0	2.00	3700	950	74.0	37.0
29	1359R	20	1600	3.0	1.70	3730	1500	73.5	35.2
	"	"	"	"	2.00	"	1000	73.5	36.6

POPE, EVANS AND ROBBINS

FBM with the 1359 limestone at a ratio of 2.0 is indicated to be somewhat more favorable than the 70% reduction indicated in the FBC at this ratio.

Reductions in sulfur dioxide emission observed with the medium sulfur coal are summarized in Table XI and are compared with the FBC data trends in Figure 38. The comparison indicates that the -325 mesh 1359 raw limestone is as reactive in the FBM as in the FBC. The hydrated forms of both sorbents indicated a reactivity comparable to the raw stone.

7.2 HYDROCARBONS EMISSION

Emission of hydrocarbons from the FBM was observed to vary sharply with flue gas oxygen content in the same manner as noted in the FBC tests, but the general level of emission was somewhat higher. Concentrations varied as shown in Figure 39 from ~4600 ppm at 0.5% O_2 to 50 ppm at 4.0% O_2. At the 3% O_2 level maintained during the FBM tests, the concentration varied from 210 to 260 ppm. During the FBC tests, the concentration varied from 50 to 100 ppm (Section 6.5). An average concentration of 230 ppm for the FBM test operation corresponds to 0.10 lbs CH_4/MBtu input.

These results indicate that a 4% oxygen content in the flue gas would be necessary to limit hydrocarbon concentrations to 50 ppm (.02 lb CH /MBtu emission). The excess air requirement would be approximately 24% unless improvements in coal feeding methods are made.

Injection of sulfur control sorbents did not affect hydrocarbons emission.

Carbon monoxide emission, determined by Orsat analysis, was found to be nil at oxygen concentrations of 2% and higher. CO concentrations of 0.4% appeared in the flue gas when the oxygen content was reduced to 1%.

7.3 NITRIC OXIDE EMISSION

The concentration of nitric oxide in the flue gas from the FBM was observed to increase from 280 ppm to 340 ppm with increase in oxygen content from 1% to 4%. This variation, shown in Figure 40, is characteristic of the variation observed in the FBC but the concentrations are somewhat less (cf Figure 30). An average value of 275 ppm for the FBM tests compares favorably with an approximate average of 380 ppm for the FBC operation. These concentrations correspond to emission values of 0.22 and 0.30 lbs NO/MBtu respectively.

TABLE XI. DATA SUMMARY FOR INJECTION OF -325 MESH SORBENTS INTO THE FBM BURNING A 2.6% SULFUR COAL

FBM Test No.	Sorbent Type	Bed Depth in.	Bed Temp. °F	Flue Gas O_2 %	Ca/S Ratio	SO_2 Conc.ppm Initial	SO_2 Conc.ppm Final	% SO_2 Reduction	% Sorbent Utilization
21	1337H	19	1625	3.0	1.37	2200	850	60.0	44.0
22	1337H	20	1650	3.0	1.46	2250	680	68.0	45.0
23	1337R	19	1550	3.0	2.4	2300	800	65.0	27.2
24	1337R	22	1600	3.0	2.4	2400	450	81.8	36.1
	"	"	"	3.0	2.2	"	500	77.4	35.0
	"	"	"	3.0	2.2	"	400	83.5	37.0
28	1359R	20	1600	3.0	2.4	2850	800	71.6	29.8
	"	"	"	3.0	2.2	"	950	64.7	29.4
30	1359H	20	1620	3.0	1.4	2570	1290	50.0	35.7
	"	"	"	3.0	1.8	"	1020	60.4	33.3
31	1359H	20	1620	3.0	1.4	2600	1200	53.8	41.4
	"	"	"	3.0	1.8	"	1000	61.5	38.4
32	1359R	20	1610	3.0	1.6	2650	1020	61.9	38.6
	"	"	"	3.0	1.8	"	920	64.9	36.0
	"	"	"	3.0	1.8	"	780	70.5	39.1

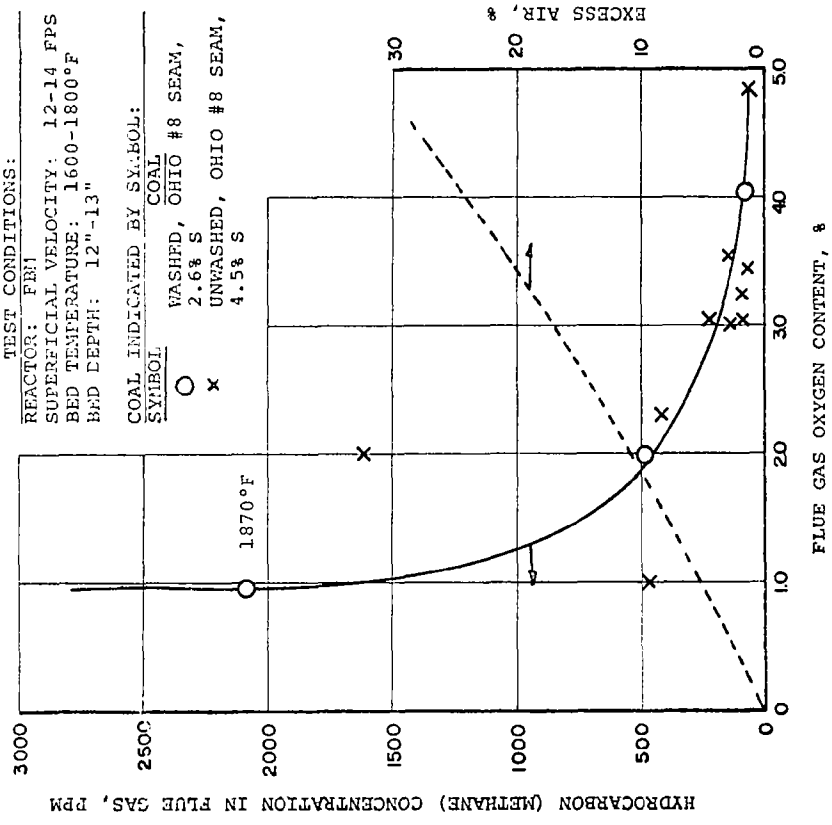


FIGURE 39. VARIATION IN HYDROCARBONS CONCENTRATION WITH FLUE GAS OXYGEN CONTENT IN THE FBM

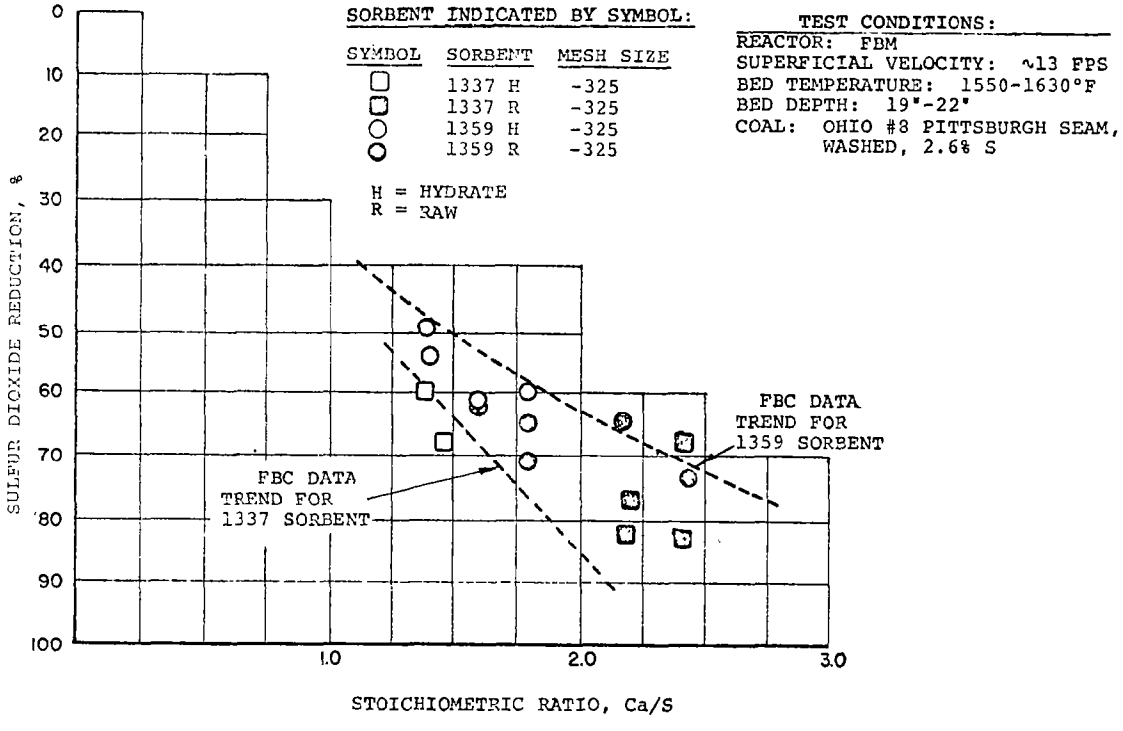


FIGURE 38. SULFUR DIOXIDE REDUCTION WITH SORBENT ADDITION TO THE FBM BURNING A 2.6% SULFUR COAL

TEST CONDITIONS:
REACTOR: FBM 20" x 72"
SUPERFICIAL VELOCITY: 12-14 FPS
BED TEMPERATURE: 1770-1840°F
BED: SINTERED ASH, STATIC DEPTH 13"
COAL: OHIO #8 PITTSBURGH SEAM, WASHED, 2.6% S

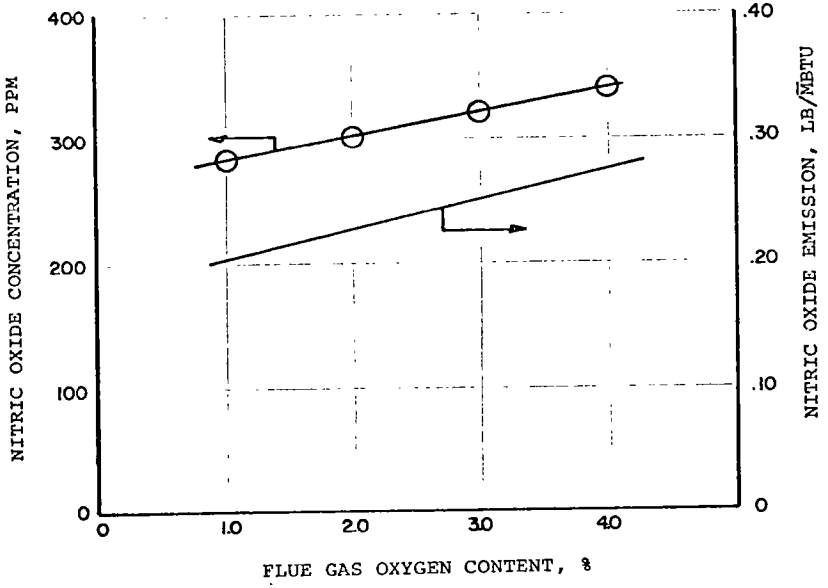


FIGURE 40. FBM - VARIATION OF NITRIC OXIDE CONCENTRATION WITH FLUE GAS OXYGEN CONTENT

Two tests conducted at a bed temperature of 1800°F indicate NO emissions as high as 0.28 lbs/MBtu but values observed at 1600°F varied from 0.18 to .24 lbs NO/MBtu. These results suggest a correlation with bed temperature, but scatter of additional data points at the higher temperature might prove otherwise.

Injection of sulfur control sorbents reduced the nitric oxide emission by about 30% in one test, but the results were not reproduced in subsequent tests with the same sorbents. In general, the nitric oxide emission was not affected by sorbent injection.

A small increase in NO emissions (5% - 10%) was noted during the transition from a low nitrogen coal (1.6% N_2) to one of higher nitrogen content (2.5% N_2). The increase was less than the data scatter, however, and the test was felt to be inconclusive. Nitrogen oxides concentrations, determined by the phenoldisulfonic acid procedure generally agreed with values of nitric oxide determined by infrared absorption. These are shown on the emission curves in Appendix A.

7.4 EFFECTS OF FLY ASH RECIRCULATION AND STEAM INJECTION

Recirculation of fly ash from the collector hopper to the base of the fluidized bed was tested as a means of improving sulfur capture. The procedure was discussed in Section 5.2 and the system shown in Figure 10. Recirculation was initiated during a steady-state reduction of sulfur dioxide with sorbent injection, and continued for one hour. The results are indicated as follows:

FBM Test No.	Coal Feed lbs/hr	Sulfur Cont. %	1359R Sorbent Feed lbs/hr	Fly Ash Recir. Rate	SO_2 Emission lbs/MBtu	SO_2 Reduc. %	Sorbent Util.
29	720	4.5	220	0.0	1.7	73.5	36.6
"	"	"	"	80%	1.7	73.5	36.6
32	720	2.6	108	0.0	1.8	65.0	36.0
"	"	"	"	80%	1.8	70.5	39.1

These results indicate little or no improvement in sulfur capture with 80% ash reinjection during a one-hour period. However, a one-hour period is not sufficient to achieve steady state in a recirculating mode and some improvement in utilization might have been found at steady state. Although 25 to 30 hours would be required to approach steady state, the marginal improvement in

the first hour, which would show the largest increment of improvement, suggests that the once-through material is essentially inert.

Continuous recirculation, with limestone injection, would cause infeasible dust loadings and would require sending some of the collected dust to waste so as to avoid "choking" the system.

Nitric oxide and hydrocarbons emission were not affected by recirculation.

Injection of steam into the inlet air during sorbent addition improved the sulfur capture but the improvement was probably due to a simultaneous decrease in bed temperature. The observations are summarized as follows:

FBM Test No.	Coal Sulfur Content %	Inlet Air Water Content % Vol.	Bed Temp. °F	Ca/S Ratio	SO_2 Emission lbs/MBtu	SO_2 Reduc. %	Sorb. Util.
20	2.6	0.5	1780	1.46	2.20	40	27
"	"	8.8	1700	1.46	1.50	58	40
21	2.6	0.5	1680	1.37	1.50	56	41
"	"	8.8	1600	1.37	1.25	62	44

The reduction in Test 20 appears to show a significant effect from water injection except for the fact that the bed temperature was also reduced. Test 21 shows that virtually the same reduction can be produced at the lower temperature without water injection. Nitric oxide and hydrocarbons emissions were unaffected by the water injection.

Since bed temperature can be adjusted readily with bed depth, there appears to be no advantage to water injection. A disadvantage would be a slight reduction in the boiler thermal efficiency.

7.5 PARTICULATE EMISSION

Isokinetic samples of particulate matter were drawn from the long duct above the FBM at a point just upstream of the induced draft fan as shown in Figure 10. The sample point is downstream of the FBM cyclones and the samples taken were proportional to the rate of particulate discharge to atmosphere. When fine sorbents were injected, the particulate emission increased as expected. Typical data are summarized as follows:

FBM Test No.	Type	Rate lbs/hr	Additive Type	Additive Rate lbs/hr	Fly Ash Collected lbs/hr	Fly-ash Emission lbs/hr	Collector Efficiency
27	Unwashed	760	No Additive		156	10.5	94
	Unwashed	760	1359R	220	332	16.5	94
28	Washed	745	No Additive		102	8.9	91
	Washed	745	1359R	150	230	12.4	94
29	Unwashed	720	No Additive		135	12.1	92
	Unwashed	720	1359R	175	295	14.7	95
31	Washed	800	No Additive		108	7.7	93
	Washed	800	1359H	65	200	11.4	94
32	Washed	720	No Additive		115	10.9	91
	Washed	720	1359R	97	180	13.7	93

One fly-ash sample taken from the cyclone discharge during the addition of sorbent in FBM Test 24 was analyzed for particle size distribution by microscopic count. The size distribution, shown in Figure 41, indicates that 90% (by number) of the material was smaller than 5 microns. Assuming spherical particles of equal density, only about 52% (by weight) of the particles were smaller than 5 microns. The sorbent was 1337R, -325 mesh fed at a rate of 260 lbs/hr with the washed coal at 800 lbs/hr. The particle size distribution of fly ash collected in the cyclone was not determined.

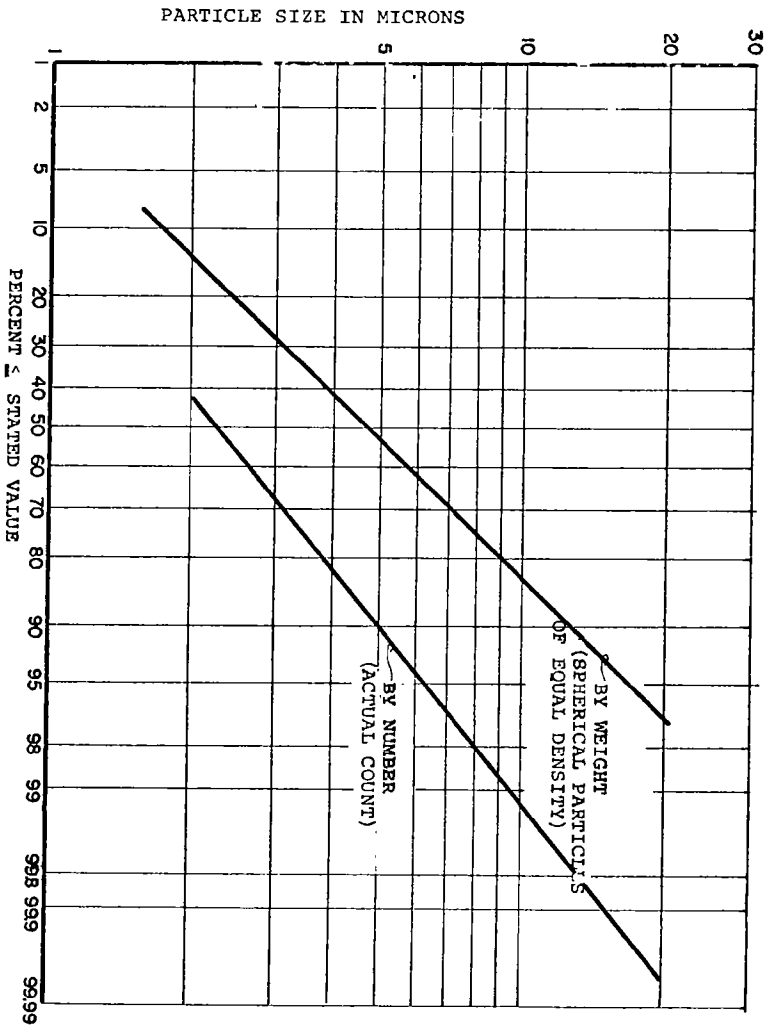


FIGURE 41. FBC COLLECTOR EMISSION. CUMULATIVE FREQUENCY DISTRIBUTION OF PARTICLE COUNT

8. DISCUSSION OF FBC AND FBM TEST RESULTS

On the basis of performance observed during the test program, the fluidized-bed boiler appears to offer pollution control advantages with respect to all three of the chemical pollutants studied, i.e., sulfur oxides, nitrogen oxides and hydrocarbons. On the other hand, control of particulate emission may be somewhat more difficult with injection of fine sorbents for sulfur emission control. Factors which relate to possible advantages in boiler maintenance are apparent. These considerations and the effects of dominant variables are discussed in the following paragraphs.

Emission of sulfur dioxide from combustion of coal in a fluidized bed contains 90% to 95% of the input sulfur. The balance is retained in the fly ash probably as a pyrite form. A very small amount of sulfur appears as sulfur trioxide in the flue gas.

In the control of sulfur dioxide emission, effectiveness of sorbent materials was seen to depend primarily on sorbent type, feed rate, particle size, bed operating temperature, oxygen content in the flue gas and, to a lesser extent, on bed depth. The effect of sorbent was shown in the comparison of reductions with the 1337 dolomite and the 1359 limestone. The dolomite proved to be superior on a Ca/S basis, i.e., when the magnesium fraction was discounted as a sorbent. On a weight basis, however, the 1359 limestone was more effective particularly when ground to a -325 mesh particle size.

The improvement in desulfurization, observed with increased stoichiometric feed ratio of the limestone, is accompanied by a decline in the sorbent utilization. Utilization of the finely divided, raw 1359 limestone, under the most favorable conditions was found to vary from 40% at a Ca/S ratio of 1.0 to 33% at a ratio of 2.0 and 28% at a ratio of 3.0. This result is consistent with decline in the driving force in the reaction, i.e., the sulfur dioxide concentration in the system. In terms of SO_2 emission reduction, the performance indicates that 80% of the sulfur emitted from a 4.5% sulfur coal could be captured with a Ca/S ratio of 2.7.

Grinding the sorbents to a fine particle size (-325 mesh) markedly improved sulfur capture (and the sorbent-utilization) despite the expectation that the residence time of fine particles in the fluidized bed would limit desulfurization. The improved utilization is apparently the result of greater reactive surface per unit mass of sorbent, and the ease of calcining the small particle

as the initial step in the desulfurization reaction. The attempt to find a particle size which would provide an optimum between residence time and reactive surface failed to show such an optimum. The small particle size (325 mesh) was more reactive than any larger size at constant bed depth.

Increase in bed residence time by increasing the bed depth from 10 to 18 inches indicated a small improvement in sulfur capture at the low bed temperature. This result and the failure to observe an optimum suggests that product shell diffusion is controlling even with small sorbent particles. This conclusion is further supported by the fact that increasing particle residence time by reducing superficial gas velocity did not show an improvement in sulfur capture.

The rapid improvement in desulfurization with reduction in particle size suggests that fine grinding may be necessary for effective utilization of the 1350 limestone. The corresponding lime hydrate, which occurs naturally in the fine state, was equally as effective as the fine raw stone but is considerably more expensive. Other, less durable limestones, may tend to decrepitate in the bed and mitigate the grinding requirement.

The reactivity of sorbents in the fluidized bed was found to be greater in every instance at a bed temperature of 1550°F than at 1800°F. This behavior is consistent with thermodynamic predictions for the reaction but equilibrium in the bed is improbable. It is inconsistent with kinetic considerations. A possible explanation is that the lower bed temperature produces a soft, highly porous calcine with minimum crystal growth. At temperatures below 1500°F the reactivity may be reduced by failure of the sorbent to calcine.

Operation of a fluidized-bed boiler at 1550°F instead of 1800°F does not mean that less heat is transferred out of the bed. The bed temperature is reduced from 1800°F to 1550°F by increasing the bed depth and hence the bed cooling contact surface. The fact that the gases leave the bed at a lower temperature means a lower heat loss in the gas and hence an even greater heat removal from the bed. The input energy is fixed by the superficial velocity range.

The low bed operating temperature should reduce boiler tube slagging.

Heat required to calcine the sorbent does not create a demand on the system since it is supplied, for the most part, by heat release from the desulfurization reaction. Standard reaction energies indicate that one pound of $CaCO_3$ would absorb 775 Btu in calcination but would release 1300 Btu if fully converted to sulfate. The energies balance, if the utilization is 37%--roughly the utilization observed in the test program. The sensible heat loss with sorbent feed will be small by comparison. However, the use of a sorbent must be considered in the design of the boiler since heating of the sorbent and calcination both take energy and hence tend to reduce bed temperature. This energy, removed from the bed in the form of a hot solid and hot CO_2, is recovered, in part, in the convection zones.

The fact that sulfur capture is favored by increase in excess air is readily apparent from the limestone bed investigation. This study clearly demonstrated that sulfur can be captured effectively for a period of time in a bed of limestone and then discharged from the bed by reducing excess air and increasing bed temperature. The sulfur release apparently follows the reaction:

$$C(\text{or } H_2) + CaSO_4 \rightarrow CaO + SO_2 + CO_2 \text{ (or } H_2O)$$

It was shown that sulfur release may occur with oxygen in the flue gas probably because of local reducing conditions in the bed. Mildly reducing conditions in the bed accelerate the sulfur release and effect higher sulfur concentrations in the flue gas.

Most significant is the fact that concentrations of sulfur dioxide in the off-gas from the bed during the regeneration period may be thirty times the untreated flue gas concentration. Concentration as high as 8.1% observed during regeneration markedly increases the feasibility of sulfur recovery. Concentrations in excess of 8.1% might be achieved by designing the regeneration region so as to minimize heat loss. This, in turn, would reduce the fuel and air requirement and so reduce dilution of SO_2 by CO_2 and N_2.

Recycle of the limestone through absorption and regeneration phase might provide the means for improving the effective limestone utilization beyond the present limit. This will depend on how well the reactivity is retained and the long-term attrition rates. Additional work in this area is indicated. Utilization per cycle might be increased by larger percentages of excess air.

The method of sorbent feed into the bed appears to be optional in the FBC with no clear advantage for any of the methods under study. The test of optimum sorbent distribution by injection on all four sides of the FBC unit showed the same sulfur control as single side injection. These results speak well of the mass transfer within the FBC bed. The two-point feed system used in the FBM appears to be as effective as any of the systems used in the FBC.

Failure to observe a consistent, beneficial effect from recirculation of spent sorbent in the fly ash suggests again the product shell limitation. Wetting the fly ash before recycle may improve the sorbent utilization by breaking down the particle as the core becomes hydrated. This procedure has not been tested.

Sulfur trioxide elimination with sorbent use is consistent with the active nature of the compound. Its absence could make electrostatic precipitation of fly ash more difficult unless the design of the system exploits the high carbon content of the primary fly ash.* On the other hand, boiler tube corrosion should be reduced.

Emission of hydrocarbons from the fluidized-bed boiler clearly precludes its operation at very low values of excess air (5%) but the advantage is noted that hydrocarbons can be eliminated with only moderate rates of excess air. The test results suggest that a 4% oxygen content in the flue gas will be necessary to prevent hydrocarbons emissions from the FBM operation. This oxygen content corresponds to an excess air rate of 24%--a value which compares favorably with values of 40 - 50% commonly used in coal fired industrial boilers.

The loss in energy from hydrocarbon emission would probably be as great as the heat saved by lower excess air operation, as estimated in Section 6.5

The lower excess air requirement for the FBC operation (17%) suggests a better distribution of volatile matter in the smaller bed. The potential seems to exist for decreasing the excess air requirement about 10% while still burning essentially all hydrocarbons and CO if the fuel distribution system is substantially improved. A fuel saving on the order of 1/2% would then be realized.

*See Appendix A, Enclosure 45.

Nitrogen oxides emission from the FBM were found to be less than the emission from the FBC (0.22 vs 0.30 lbs/MBtu). This result may be related to the higher hydrocarbons emission from the FBM, possibly by the reaction:

$$(CH_x) + 2NO \rightarrow N_2 + CO_2 + (H_2O)$$

The flue gas oxygen concentrations were 3.0% for both the FBM and FBC tests. This suggests that a hydrocarbon gas properly dispersed at the grid might reduce the NO emission without affecting the sulfur control functions.

The moderate sensitivity of nitric oxide emission to flue gas oxygen content suggests that the level can be reduced by lowering the average oxygen concentration in the bed, i.e., by reducing conditions. A NO reduction of 50% was observed in Test No. 119 with reducing conditions in the bed and overbed air to make up the 3% oxygen content (cf. Figure 36). Unfortunately, this mode of operation is not conducive to sulfur capture in the sulfate form. Hydrocarbons from the bed were effectively consumed by overbed air.

These results would indicate that nitric oxide emission can be reduced in a limestone bed without aggravating the hydrocarbons emission by two-stage combustion, i.e., by reducing conditions in the bed and an oxidizing environment above. It may be possible to capture sulfur as the sulfide in a cyclic operation under these conditions.

The nitric oxide emission from the FBM is favorable in comparison with emissions from other combustion units of equal size. The average value of 0.22 pounds/MBtu is less than reported values for most conventional boilers. A full scale boiler made up of modules according to the present concept may not be subject to the increase in NO_x emissions generally observed with increase in unit capacity.

Most of the fine sorbent added to the bed is collected in a single stage mechanical cyclone operating at 95% efficiency. Controlling emission of the remaining 5% may present a problem if subsequent tests show that 90% of the particulate is smaller than 5 microns when fine sorbents are used. The microscopic count showing this distribution applied to one sample. Additional data are needed for a firm conclusion regarding particulate emission control.

9. ECONOMIC ANALYSIS

9.1 GENERAL

9.1.1 Economic Evaluations in Industry. Investment decisions in the selection of steam and power generating equipment are made in a number of ways. The factors which are utilized vary from industry to industry and from company to company within an industry. Typically, however, a central steam supply is viewed as a long-term investment not subject to the same rapid pay-out demands as a process investment might be.

Whatever factors are applied, a rational technique of structuring the decision-making process is required. It is possible, for example, to apply the present-worth method. By this technique all capital and operating expenses are reduced to a single dollar figure, the "present worth" of all present and future expenses. A number of other investment appraisal techniques exist but present worth appears to be the most popular.

Making application of the present-worth method in a sophisticated manner requires that predictions be made as to the future cost of labor, the future cost of fuel, etc.; and, when certain investments may be deferred, the future cost of money. Fortunately, in the field of steam power generation an extensive statistical base exists on which reasonable projections of future costs may be made. The various alternatives are then evaluated on the present-worth basis. The best apparent choice is that alternative which has the lowest present worth. Computerized evaluations make possible sensitivity checks, i.e., the effect of an incremental change in each cost ingredient may be evaluated so as to determine which are the most significant.

Unfortunately, when air pollution control is added to the list of plant requirements and this requirement also includes control of gaseous emissions, the statistical base becomes very limited. In addition, even current capital and operating costs involved in pollution control techniques, other than the selection of a low sulfur fuel, are based on a limited number of "paper" evaluations.

This section of the report is not a complete investment appraisal; instead it is intended to indicate how limestone addition to a fluidized-bed boiler may affect costs. These data might then be utilized in a complete investment appraisal.

9.1.2 Treatment of Incremental Costs. The costs which are included in the evaluation are those which are directly attributable to limestone injection. It must be assumed that in the selection of a fluidized-bed boiler, limestone injection is not treated as an afterthought. The boilers, the plant, the auxiliaries and the pollution control systems are designed with maximum degree of integration. Examples of such integration include: a single receiving point for coal and limestone; a single bulk conveyor system; a single, but properly, partitioned storage silo; the use of the preheated combustion air or possibly flue gas to dry the limestone before pulverizing; and the use of the boiler's induced draft fan to provide any suction required on the limestone system. The boiler itself receives all dust vented from the limestone handling and storage system, etc.

When a single system serves two functions, it is reasonable to attribute only an incremental cost to the function being evaluated. Therefore, limestone addition to a new, properly designed fluidized-bed boiler plant is far less costly than limestone addition to an existing plant or to a new plant in which pollution control is an afterthought.

The cost estimates were based on the assumption that if a new plant were being built it would include two boilers. Costs were therefore estimated for the 500,000 lb/hr plant and then divided by two to indicate costs attributable to a single boiler. This approach was taken in order that this report be consistent with earlier analyses. For readers who wish to determine capital costs for a single boiler installation or for more than two boilers, the well-known six-tenths factor has been found to apply to equipment and construction of this type.

9.2 BASIS OF PERFORMANCE ESTIMATES

The analysis of limestone use has been based on the experimental performance data obtained with the single full-scale fluidized-bed module. This module has many features in common, especially dimensions, with a 250,000 lb/hr shop-assembled boiler. The key dimensions which are similar are bed height and cell width.

9.2.1 Limestone vs Dolomite. Performance data were obtained for both dolomite and high calcium limestone. These data indicated that dolomite (53% $CaCO_3$) was superior in controlling emissions when the measure of superiority was stoichiometric addition rate based on the calcium content only. However, limestone and dolomite are both sold on a weight basis with little regard to chemical composition. So, although the calcium in limestone is less effective than that in dolomite, a much lesser total weight of limestone is required for a given SO_2 reduction. For this reason the economic analysis is based on the data obtained with the high calcium limestone (97% $CaCO_3$).

9.2.2 Raw Stone vs Hydrate. Hydrates of limestone and dolomite were also evaluated as an additive and these were found to be slightly more effective than the raw stones. However, as in the case of dolomite vs limestone noted above, the cost of a ton of calcium delivered as the hydrate is higher than the calcium delivered in the raw stone. Since the slightly higher utilization of the hydrate does not compensate for its much greater cost, only the raw stone has been considered in the evaluation.

9.2.3 Particle Size. Increased utilization of the raw stone is found with decreasing particle size. This is illustrated in Figure 27 where SO_2 reduction is plotted against particle size for constant additive rate. Since the smallest particle size used, -325 mesh, gave the best results, this size has been assumed for the economic evaluation.

9.3 PERFORMANCE DATA

The reduction of sulfur dioxide emission from the full scale module using 1359 limestone at a bed temperature of 1600°F was noted earlier to be about the same as that achieved in the pilot scale unit. It was also noted that if percent reduction is plotted against stoichiometric ratio similar values are found for both the 2.6% and 4.5% sulfur coals. This plot is given as Figure 42, and is an average of the 1359 lines in Figures 25 and 26.

When the stoichiometric ratio is converted to a weight basis, pounds of limestone per 100 pounds of coal, separate curves are generated for each coal. These are given in Figures 43 and 44. For additional clarity, the ordinate in these figures was converted to the ratio--sulfur in emissions/sulfur input.

POPE EVANS AND ROBBINS

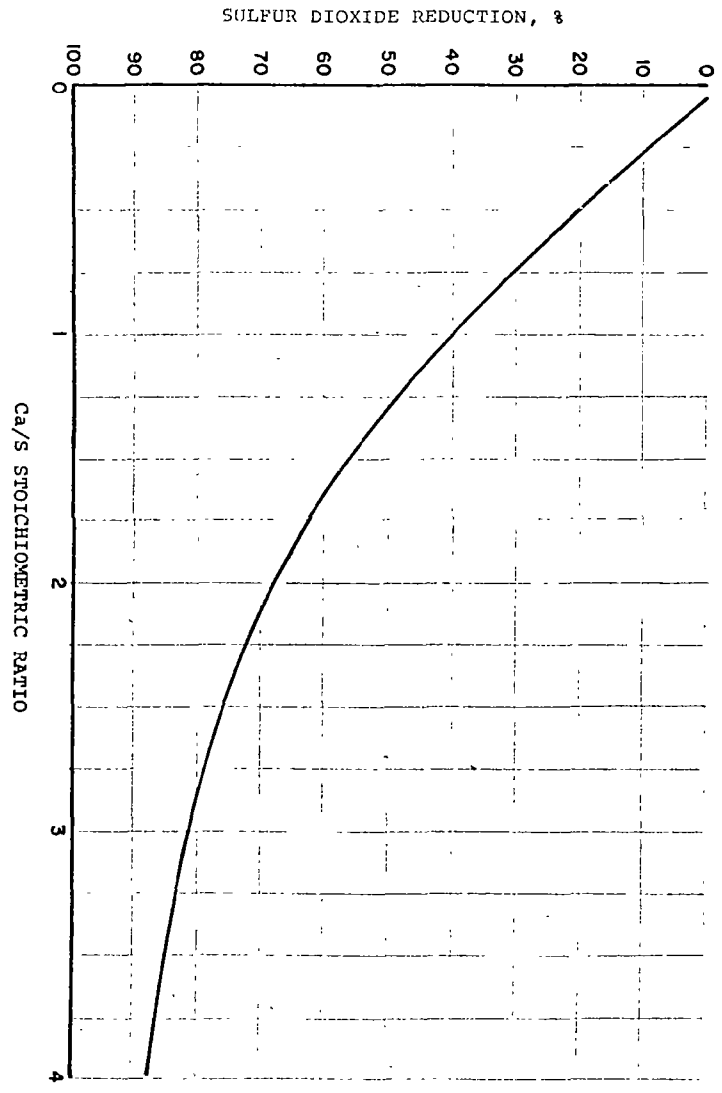


FIGURE 42. ESTIMATED REDUCTION IN SULFUR DIOXIDE EMISSION BY ADDITION OF 1359 LIMESTONE AT VARIOUS STOICHIOMETRIC RATIOS

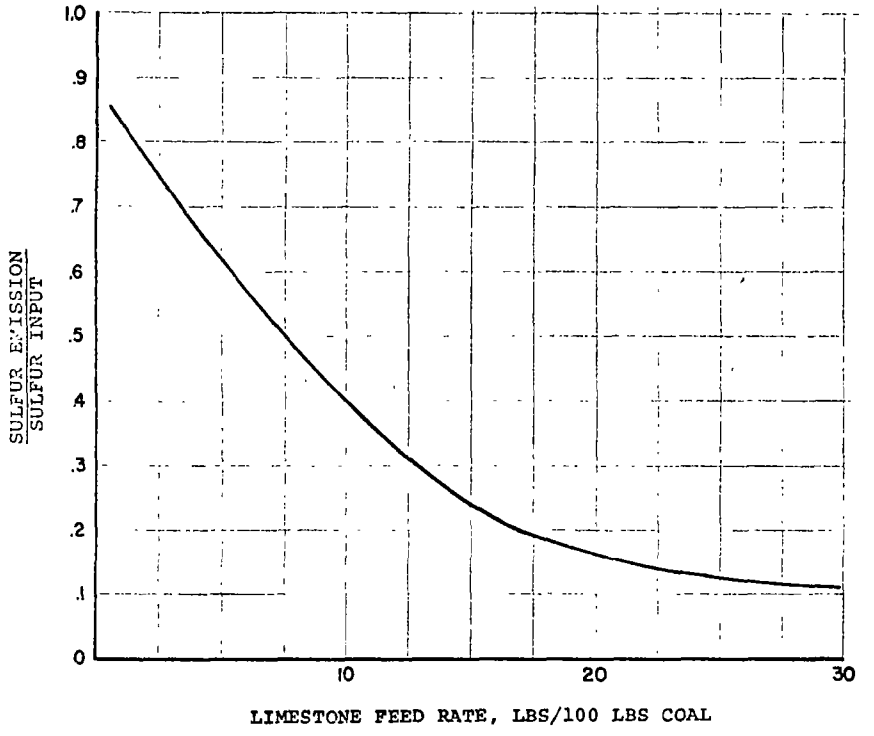


FIGURE 44. RATIO OF SULFUR EMISSION TO SULFUR INPUT VS LIMESTONE FEED RATE FOR THE 2.6% S COAL

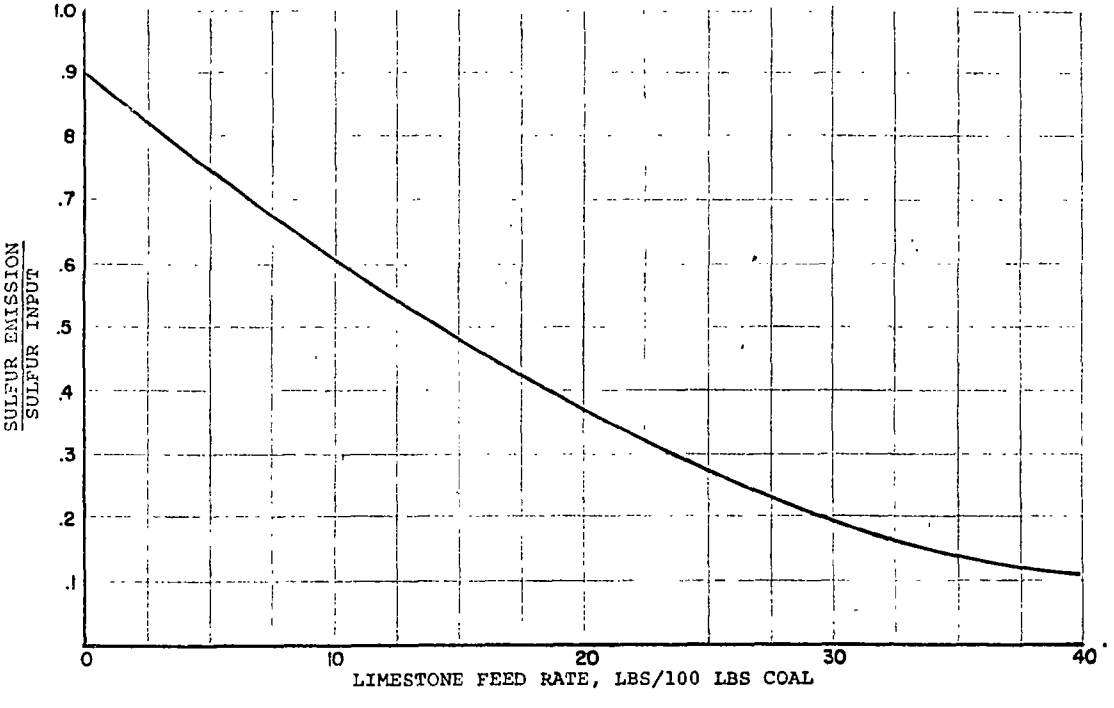


FIGURE 43. RATIO OF SULFUR EMISSION TO SULFUR INPUT VS LIMESTONE FLOW RATE FOR THE 4.5% S COAL

It was found that even without additive not all the sulfur would appear as SO_2 in the flue gas. Typically, 10% of the input sulfur was found in the ash. Therefore, Figures 43 and 44 show the ordinate at 90% with a zero additive rate. These curves then form the basis for the operating cost analysis in that they relate sulfur emissions to the required weight of additive.

9.4 CAPITAL REQUIREMENTS FOR EQUIPMENT

9.4.1 Description of Raw Stone Feed System. Limestone must be received, stored, prepared and injected, captured and disposed of. As noted earlier, the major portion of the system has been integrated with the coal handling system, and therefore the size of the system is relatively independent of the sulfur content of the fuel. The limestone injection system is charged with a storage silo increment, dust collector increment, etc. (See Table XII). In some instances, pneumatic ash conveyors for example, the smallest system commercially available would be used with or without limestone addition.

A block outline of the combined coal/limestone/ash handling system is given as Figure 45. The system shown and the costs tabulated below are assumed constant regardless of the sulfur content of the coal and the degree of emission control required. Although some capital cost reduction would be achieved for a precisely sized system, it would be poor judgment for the plant designers not to provide for use of high sulfur coal even though use of a lower sulfur coal is planned, and for maximum emission control since doing so would not affect the capital significantly.

9.4.2 Description of Dust Collector System. The reacted limestone is carried out of the boiler, along with the carbon rich fly ash in the flue gas. Collected in a cyclone, the spent stone and fly ash are pneumatically injected into the Carbon-Burnup Cell*. Here the carbon content of the fly ash is burned in an oxygen rich, high temperature environment. Carried out once again by flue gas, the spent limestone and fly ash are collected in the secondary mechanical collector for disposal. Depending on local regulations regarding

*The Carbon Burnup Cell is an integral component of a fluidized-bed boiler in which the relatively unreactive carbon remaining in fly ash can be burned so as to improve the boiler's efficiency. It is fully described in U.S. Patent 3,508,506.

TABLE XII. SUMMARY OF CAPITAL COST COMPONENTS FOR LIMESTONE ADDITION PER BOILER. 500,000 LB STEAM/HR PLANT CONSISTING OF TWO 250,000 LB/HR COAL-FIRED, FLUIDIZED-BED BOILERS

Line		
1	Incremental site improvements	$ 1,000
2	Incremental unloading hopper, storage silos, transfer belt and bucket elevator	7,000
3	Surge hopper	2,000
4	Dryer, pulverizer and classifier	38,000
5	Storage hopper	2,000
6	Incremental mechanical handling and injection systems	10,000
7	Incremental dust collector costs	8,000
8	Incremental ash handling and storage	5,000
9	Controls and instruments	10,000
10	Miscellaneous steel	5,000
11	Incremental electrical, mechanical, utilities, etc.	10,000
	Subtotal (Lines 1 through 11)	$ 98,000
13	Contingency @ 10% of Line 12	9,800
14	Total (Line 12 + Line 13)	$107,800

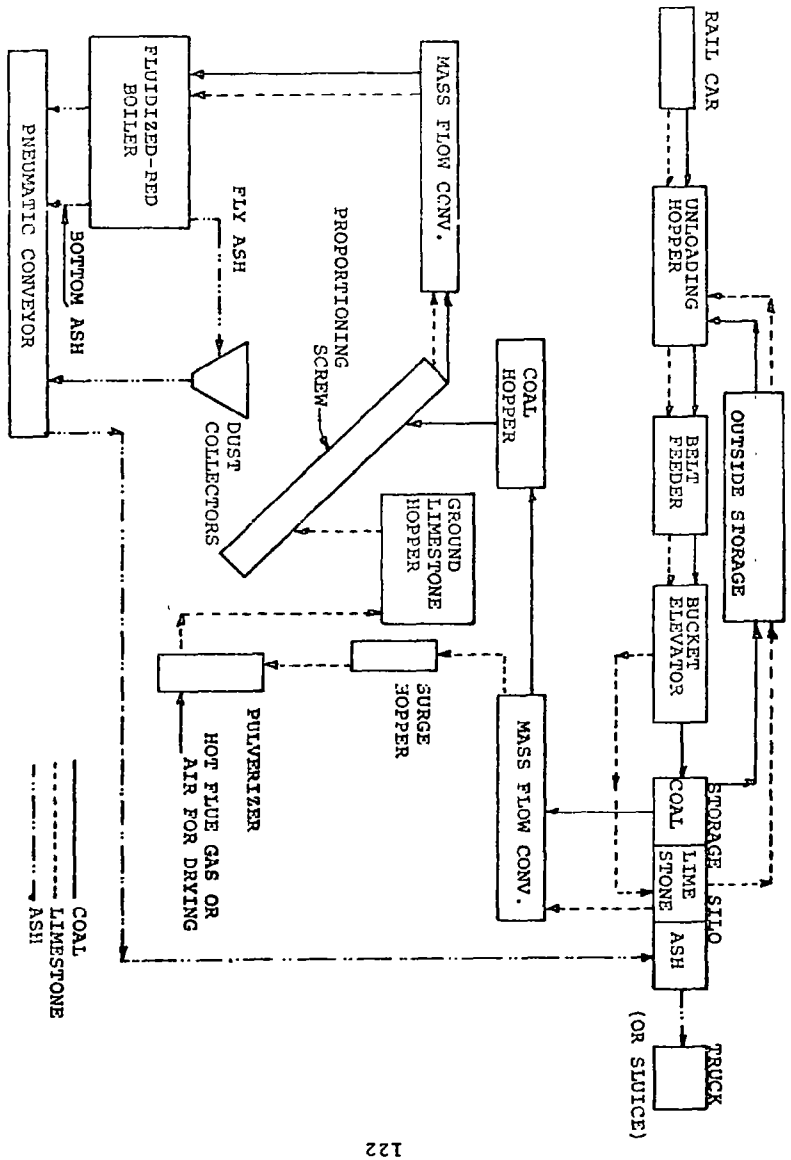


FIGURE 45. SCHEMATIC ARRANGEMENT OF THE SORBENT FEED SYSTEM FOR A FULL SCALE FLUIDIZED-BED BOILER

particulate emissions, an electrostatic precipitator* or wet scrubber may also be required. It should be assumed that near urban areas regulations will require particulate emissions on the order of 0.2 lbs per 10^6 Btu input.

The ash content of the coal is assumed to be 10% and 7% for the 4.5% S and 2.6% S coals respectively. The total particulate matter emanating from the combustion of each coal is shown in Figures 46 and 47 as a function of additive feed ratio. Curves which show the probable variation in precipitator load were added. The curves assume all ash goes overhead, 40% utilization of CaO, 85% efficiency on the mechanical collector and 10% carbon in the fly ash. Omitted is bed material attrition which may add to the particulate load.

Discussions with precipitator manufacturers failed to provide a basis on which to estimate the costs of additional capacity requirements due to limestone addition. It appears that the resistivity of fly ash increases when SO_3 is not present. However, carbon in the fly ash may compensate so that precipitator efficiency will not be seriously impaired.*

The preferred method of defining precipitator requirements is to use one of the portable or pilot precipitators owned by precipitator manufacturers.

Cyclone collector costs are relatively independent of dust loading except that an increment has been provided for heavy duty construction, increased hopper capacities and increased unloading capacities.

The size distribution and composition of the fly ash emanating from a fluidized-bed boiler is now under study. Some preliminary work has indicated that about 95% of the particles leaving the combustor are collectable in a low efficiency mechanical collector. Of the particles bypassing the collector, 99.9% were under 20 microns.

Ash is moved to the ash section of the common silo via a pneumatic conveyor. Except for the increased silo capacity requirement due to the added limestone essentially no capital cost increase is required for ash disposal.

*See Appendix A, Enclosure 45 for proposed design arrangement for minimizing electrostatic precipitator costs.

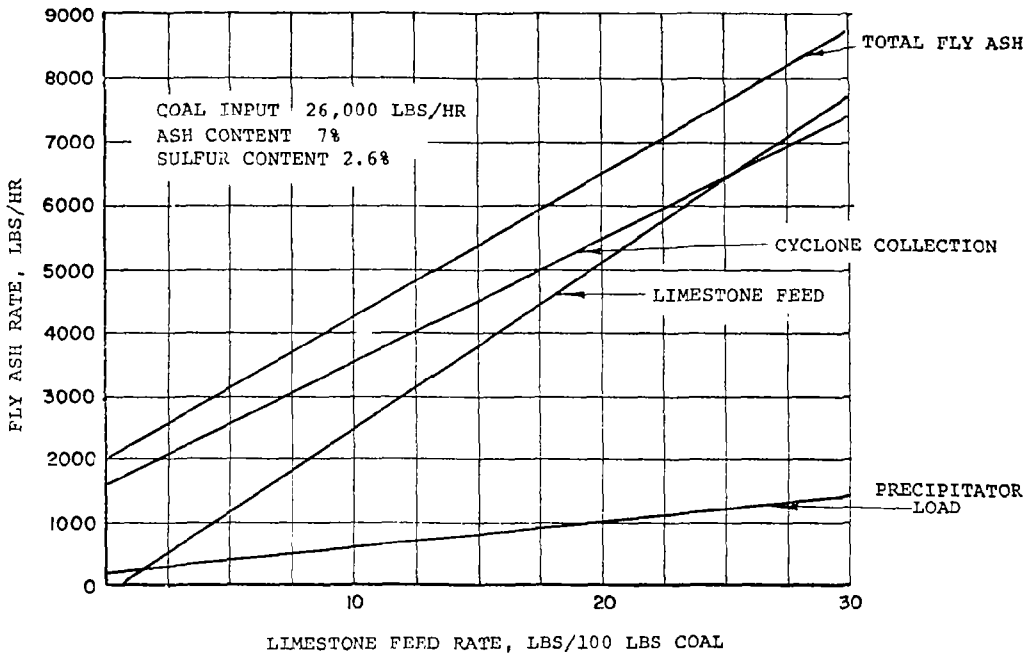


FIGURE 47. TOTAL FLY ASH RATES FOR FULL LOAD OPERATION OF A 250,000 LB PER HR. FLUIDIZED-BED BOILER WITH LIMESTONE ADDITION (2.6% S COAL)

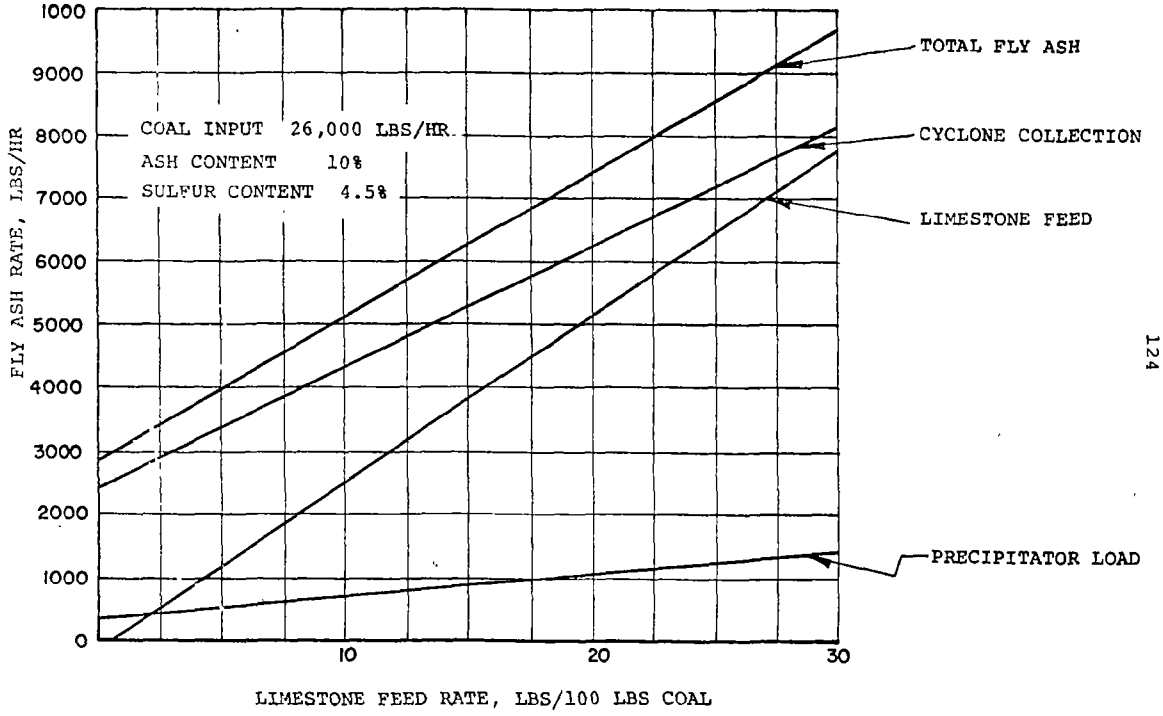


FIGURE 46. TOTAL FLY ASH RATES FOR FULL LOAD OPERATION OF A 250,000 LB PER HR. FLUIDIZED-BED BOILER WITH LIMESTONE ADDITION (4.5% S COAL)

9.4.3 Summary of Capital Costs. The cost data which appears in Table XII was based on manufacturer's information where applicable and on published estimates for components and systems. Outdated information was adjusted using the well-known Marshall and Stevens Equipment Cost Index. As noted earlier, these costs represent one-half the incremental cost of including limestone addition in the design and construction of a new plant containing two 250,000 lb/hr fluidized-bed boilers.

9.5 ANNUAL OPERATING COSTS

The major element of operating cost is the delivered cost of the raw limestone. Other components of operating costs are incremental labor costs, incremental maintenance costs, increased disposal costs, power costs for pulverizing, recovery of capital, taxes and insurance and a small cost for the thermal effect of limestone additions.

9.5.1 Delivered Cost of Limestone. Delivered costs of limestone are variable, as are the costs of coal, and dependent on plant location, rate of consumption, mode of transportation, and market conditions. The most definitive evaluation of limestone economics, by TVA, assumed a cost of $2.05 per ton for crushed limestone. Studies by Esso Research and Engineering and A. M. Kinney, Inc. also used this limestone cost. This cost, as in the TVA study, is a $1.35 per ton vendor's cost and a $0.70 per ton shipping cost.

The same value, based on $2.05 per ton, will be assumed in this evaluation, although costs above or below this value may be found to be more appropriate in an actual investment analysis.

9.5.2 Incremental Labor Cost (Plant Handling). Two men are employed in the 500,000 lb/hr steam plant as coal and ash handlers. They both work during the day shift, five days per week. During other periods, materials are drawn from live storage. No increase in staffing requirements is anticipated as a result of the decision to use limestone injection. To account for occasional overtime, however, a cost of $0.15 per ton of limestone has been allocated for plant handling.

A watch supervisor and a watch fireman monitor the operation of the plant's two boilers and auxiliaries. No extra watch positions are required due to the addition of limestone.

9.5.3 Incremental Power Costs (Pulverizing). The only significant power requirement because of limestone addition is that due to the pulverizer. An evaluation of limestone grinding by A.M. Kinney, Inc. indicated less than 35 kwh per ton of stone, while TVA's evaluation indicates that ~43 kwh per ton of limestone is required to grind to 99%, -325 mesh. At a conservative $0.009/kwh the power cost would be on the order of $0.32 to $0.39 per ton of stone. A cost of $0.40 per ton will be used in this analysis.

9.5.4 Thermal Effect. When limestone utilization approaches about 40%, it is possible to realize a net thermal gain from limestone injection. Depending on the cost of coal, the method of drying, the exit gas temperature and the precise degree of utilization, costs of from 1¢ to 5¢ per ton of raw stone might be used for this factor. This analysis will use 5¢ per ton of raw limestone for thermal effect.

9.5.5 Incremental Maintenance Costs. In many economic analyses, annual maintenance is simply assumed at 2 - 5% of capital. In this study, the incremental maintenance costs are assumed to be made up of a fixed portion, and a value dependent upon throughput. For the fixed portion, 2½% of the incremental investment will be used. For the tonnage dependent portion, a value of $0.20 per ton will be used to account for pulverizer wear. The sum of these two factors will exceed 5% of capital for several of the cases analyzed below.

9.5.6 Disposal Costs. Fly-ash disposal costs are the most variable ingredient in any industrial coal-fired boiler cost analysis. Costs may vary between $0.00 per ton to $1.00 per ton depending on local market conditions for fly ash or the distance to a landfill.

Ash disposal is often by sluice to a fill area. In this case, ash disposal costs are more properly expressed as a capital cost (see TVA's treatment, for example). For this evaluation, a cost of $0.25 per ton of raw stone will be assumed to be borne by the steam plant.

9.5.7 Summary of Annual Operating Cost Ingredients. Shown below is a summary of the cost ingredients for the annual operating cost analysis. The costs are divided into two categories--fixed costs, independent of the degree of emission control, and variable costs which are proportional to the degree of control.

TABLE XIII. OPERATING COST INGREDIENTS SUMMARY FOR FINE LIMESTONE INJECTION IN A 500,000 LB/HR FLUIDIZED-BED BOILER PLANT*

A. Fixed Costs

1. Interest, depreciation, taxes and insurance @ 14% of $107,800 = $15,100/annum
2. Maintenance @ 2-1/2% of $107,800 = 2,700
   Total $17,800/annum

Fixed cost per ton of coal,
13 tph x 6,000 hrs/yr. → $17,800/(13 x 6,000) = $0.23/ton of coal

B. Variable Costs

	$/Ton of Limestone
1. Limestone, 1/4" x 0, vender's price	1.35
2. Shipping	.70
3. Power for pulverizing	.40
4. Thermal effect	.05
5. Incremental maintenance	.20
6. Disposal	.25
Total	2.95

*Made up of two 250,000 lb/hr boilers

9.5.8 Annual Operating Costs. Applying the cost data of the previous section and the limestone requirements for the two coals (4.5% S and 2.6% S) from Figures 43 and 44, the annual operating costs are shown below.

The two measures of performance (emission equivalent, % S, and SO_2 removed, %) are two ways of expressing the same thing. Emission equivalent, % S, is related to % SO_2 removed by the equation:

$$\text{Emission equivalent, \% S} = \frac{(100 - SO_2 \text{ removed, \%}) \times (\text{Actual \% S in coal})}{100}$$

TABLE XIV. ANNUAL OPERATING COST DATA FOR FINE LIMESTONE INJECTION

Case 1. For the 4.5% Sulfur Coal						
1.	Emission equivalent, % S	3.5	2.5	1.5	1.0	0.6
2.	SO_2 removed, %	22	45	67	78	87
3.	Additive rate, Tons of limestone/ton to coal	.037	.12	.21	.28	.37
4.	Fixed cost, $/ton of coal	.23	.23	.23	.23	.23
5.	Variable cost, $/ton of coal	.11	.35	.62	.83	1.09
6.	Total Cost*, $/ton of coal (4 + 5)	.34	.58	.85	1.06	1.32
Case 2. For the 2.6% Sulfur Coal						
1.	Emission equivalent, % S	2.0	1.5	1.0	0.6	
2.	SO_2 removed, %	23	42	62	77	
3.	Additive rate, Tons of limestone/ton to coal	.028	.065	.105	.155	
4.	Fixed cost, $/ton of coal	.23	.23	.23	.23	
5.	Variable cost, $/ton of coal	.08	.19	.31	.46	
6.	Total Cost*, $/ton of coal (4 + 5)	.31	.42	.54	.69	

*These results are plotted in Figures 48 and 49.

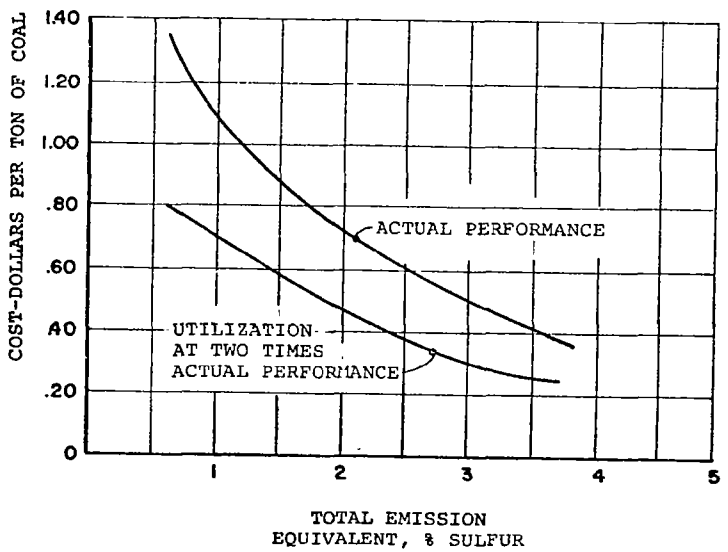


FIGURE 48. ESTIMATED TOTAL COST OF CONVERTING A HIGH SULFUR (4.5%) COAL TO A LOWER SULFUR COAL EQUIVALENT BY LIMESTONE ADDITION TO A 250,000 LB/HR FLUIDIZED-BED BOILER

POPE, EVANS AND ROBBINS

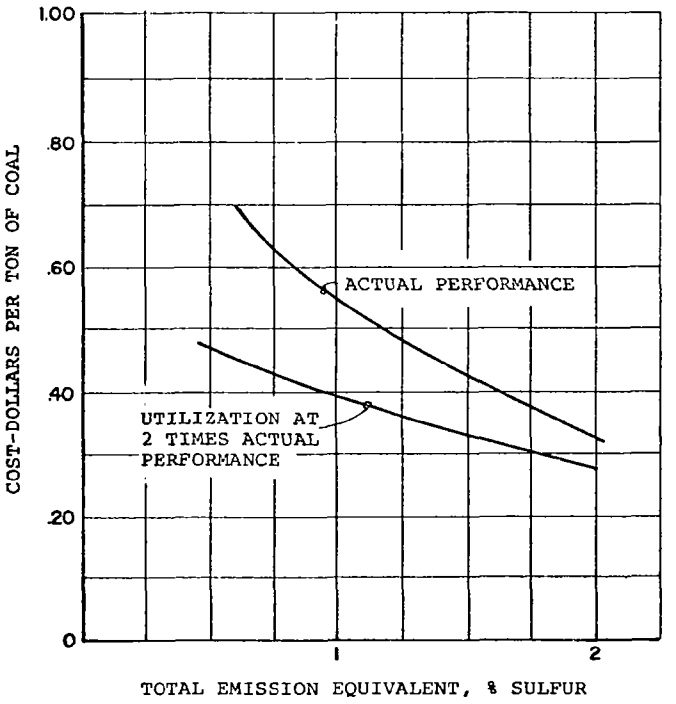


FIGURE 49. ESTIMATED TOTAL COST OF CONVERTING A MEDIUM SULFUR (2.6%) COAL TO A LOWER SULFUR COAL EQUIVALENT BY LIMESTONE ADDITION TO A 250,000 LB/HR FLUIDIZED-BED BOILER

POPE, EVANS AND ROBBINS

The results of one additional case is also plotted on these figures. A lower curve shows the costs for a utilization rate twice as good as that actually found experimentally in the fluidized-bed boiler. This indicates a hypothetical lower limit to costs in a fluidized-bed boiler if additional research reveals methods of achieving a utilization on the order of 80%. Results on this order have been reported by British experimenters.

9.6 COMPARISON WITH COSTS FOR ALTERNATIVE METHODS

9.6.1 Use of Low Sulfur Coal. The use of low sulfur coal may be an economical alternative to limestone addition where low sulfur coal is locally available.

Where this coal is not available locally, it must be shipped and this may markedly increase its cost. This is the case for Chicago, as an example, where the low sulfur coal might come from West Virginia.* Table XV presents the costs for burning three "local" coals in a fluidized-bed boiler with limestone addition and costs for burning an "imported" low sulfur coal in the same boiler without limestone. The costs for limestone addition are derived from the test program performance curves.

It is clear from this comparison that, for the case estimated, the cost of energy is less with limestone injection than with the low sulfur coal.

9.6.2 Limestone Injection into Conventional Boilers. Dry limestone injection into conventional boilers may be somewhat less effective than injection into a fluidized-bed boiler. Until test results from operation, full-scale units of both designs are available, no economic comparisons are meaningful.

9.6.3 Other Flue Gas Control Processes. A number of survey articles have been published reviewing the costs of the alternative stack gas cleaning processes.

Almost all of this work pertains to large utility boilers, not industrial boilers, and is therefore not truly comparable to the data presented above. In every case, the capital costs are significantly higher than for limestone injection and would be more unfavorable when reduced in scale. Some process developers claim a profit on operations when markets exist for the sulfur form produced and other factors are favorable.

*Low sulfur coal from Wyoming is presently being brought into Chicago by Commonwealth Edison and with shipping costs alone exceeding $8.00 per ton.

TABLE XV. COST OF REDUCING A HIGH SULFUR COAL TO EQUIVALENT 0.7% SULFUR COAL IN A FLUIDIZED-BED BOILER, COMPARED WITH COST OF PURCHASE OF LOW SULFUR COAL IN THE CHICAGO AREA

Coal Source				Coal Cost,		Sulfur Reduction Requirements, %	Limestone Injection Cost ¢/MBtu	Total* ¢/MBtu
State	County	Heating Value, Btu/lb	Sulfur Content, %	$/ton	¢/MBtu			
W. Va.	Mingo	13,800	0.7	11.32	41.02	Base	Base	41.02
W. Ky.	Hopkins	12,500	3.2	8.45	33.88	78	4.40	38.28
Ill.	Franklin	12,300	2.0	8.20	33.33	65	3.76	37.09
Ind.	Sullivan	11,700	3.1	7.87	33.63	77	4.40	37.03

*Total cost to produce emissions equivalent to the 0.7% sulfur coal by use of low sulfur coal or limestone injection

This could never be the case for pulverized limestone injection. Continuing research on flue-gas cleaning processes may provide a process applicable to industrial boilers of either the conventional or fluidized-bed design.

9.7 CONCLUSIONS

Review of the results outlined above lead to the following observations:

1. Limestone injection to a fluidized-bed boiler could be used at a reasonably low capital cost ($3/kw) when the limestone system is treated as an integral part of the steam supply system.

2. Operating costs for limestone injection to a fluidized-bed boiler will be a small multiple of the raw stone cost (~1.5) when the plant design is such that increased labor requirements are avoided.

3. In those areas where coal enjoys a natural cost advantage over natural gas, a fluidized-bed boiler with limestone injection may provide the plant owner with an economically feasible method of providing steam and complying with local air quality regulations. Conventional boilers may not, in many cases, be able to provide such a feasible alternative.

4. One final conclusion is warranted, in part by the results discussed above and in part by information recently published by the Federal Power Commission on declining gas reserves: when the investment appraisal techniques utilized by a potential boiler plant owner provide for a sophisticated treatment of cost trends, coal-fired, fluidized-bed boilers utilizing limestone injection may appear favorable even when coal does not currently enjoy a natural cost advantage.

APPENDIX A

ENCLOSURES

POPE, EVANS AND ROBBINS