**Atmospheric Fluidized Bed Combustion
Advanced System Concepts Applicable to
Small Industrial and Commercial Markets**

Topical Report: Level 2

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EXECUTIVE SUMMARY

As part of an overall strategy to promote FBC coal combustion and to improve the marketability of the eastern coals, the U.S. Department of Energy's Morgantown Energy Research Center (METC) awarded a three level contract to Riley Stoker Corporation (Riley) to develop advanced Multi Solids Fluidized Bed (MSFB) boiler designs. The steam capacities for these boilers fall in the ranges which are primarily served by the oil and gas industry.

The first level of this contract targeted the small package boiler (10,000- 50,000 lb/hr steam) and industrial size boiler (75,000-150,000 lb/hr steam) markets. Two representative sizes, 30,000 lb/hr and 110,000 lb/hr of steam, were selected for the two categories for a detailed technical and economic evaluation. Technically, both the designs showed promise, however, the advanced industrial design was favored on economic considerations. It was thus selected for further study in the second level of the contract.

Results of this Level-2 effort are presented in this report. It consisted of testing the design concept in Riley's 4.4 MBtu/hr pilot MSFB facility located at Riley Research Center in Worcester, Mass. The design and economics of the proof of concept facility developed in Level-1 of the contract was then revised in accordance with the findings of the pilot test program.

A host site for commercial demonstration in Level-3 of the contract was also secured. It was determined that co-firing coal in combination with paper de-inking sludge will broaden the applicability of the design beyond conventional markets. International Paper (IP), the largest paper company in the world, is willing to participate in this part of the program. IP has offered its Hammermill operation at Lockhaven, Pa, site of a future paper de-inking plant, for the proof of concept installation. This plant will go in operation in 1994.

It is recommended that METC proceed to the next level of the program for the commercial demonstration of the design developed. The approach necessary to satisfy the needs of the customer while meeting the objectives of this program is presented along with a recommended plan of action.

MSFB PROCESS

Riley's CFB process is characterized by a high velocity entrained bed of fine ash and sorbent particles superimposed on a fluidized bed of large particles in the lower combustor called the dense bed. The size range of particles fed to the dense bed is between 1/4" and 3/4". The unique characteristics of this lower part of the combustor is the key to the operation and performance of the Riley system. It promotes mixing, provides a stable combustion zone and increases the residence time of ash, fuel, and sorbent particles through increased hold up. This feature not only permits operation on a variety of low grade fuels, but also allows the process to accept a wide range of fuel sizes including the run-of-mine coal.

A typical process schematic is shown in Figure ES 1.

The fuel and sorbent are brought in just above the dense bed. A portion of the total air is introduced as primary air through an air distribution system to the bottom region called the reducing zone due to its prevailing sub-stoichiometric conditions. This section of the combustor is refractory lined to protect it from the abrasive action of the dense bed and circulating solids. Also, the refractory aids in warm start-ups by holding heat when the unit is shut down.

An upper water walled oxidizing zone, where the balance of the total combustion air is introduced, provides the appropriate gas residence time to complete combustion and achieve the necessary sulfur dioxide reduction.

Fine particles leaving the top of the combustor vessel are separated from the flue gas by one or more high efficiency cyclones. These cyclones are refractory lined. The separated solids leave the bottom of the cyclones and pass to the an External Heat Exchanger (EHE), a fluidized bed chamber which has installed heating surface for steam production. The hot flue gases leave the top of the cyclones and are ducted to a conventional convective boiler system to generate a portion of the total steam duty and also to preheat the incoming combustion air.

PILOT PROGRAM

The pilot plant tests have established confidence in the advanced FBC boiler concepts by demonstrating the following major goals of the program:

THE MULTI-SOLID FLUIDIZED BED

Figure ES.1 MSFB Process Schematic

- 99.5 % Combustion efficiency
- 90 % Sulfur capture with a Ca/S mole ratio of 1.5
- NO_x levels lower than 0.1 lb/MBtu

The advanced design offers the benefit of a high boiler efficiency of >90%. Although, the pilot plant is not physically set up to evaluate boiler efficiency, the high carbon efficiency demonstrated during the tests did improve confidence in achieving this goal.

A six week test program was conducted encompassing the process characteristics of the advanced design concept. During this testing, the following variables were investigated to determine conditions best suited for meeting goals for combustion efficiency, sulfur capture and NO_x emissions:

- Temperature
- Stoichiometry
- Excess Air
- Superficial Velocity and Residence Time
- Ca/S Mole Ratio
- Combustor Solids Flux
- Sorbent Sensitivity
- Fly Ash Reinjection Ratio
- Ammonia Injection Ratio

Besides the above variables heat transfer measurements were made for the external heat exchanger (EHE). Table ES-1 lists the range of conditions tested.

Information generated in a pre-Level-2 Riley sponsored pilot test program was extensively used as a guide for achieving the goals of the current program. Earlier results had shown that combustion efficiency increases with temperature. The sulfur capture also showed a similar behavior until an optimum temperature was reached after that a reverse relationship was noted. Also, it was determined that combustion efficiency and sulfur capture could be enhanced with fly ash re-injection and increased residence time. In the current data with no fly ash re-injection, combustion efficiency peaked at ~98.5% and sulfur capture at 75% for 1.5 Ca/S mole ratio. This

Table ES-1

NOMINAL LOAD CONDITIONS AND VARIATIONS FOR PILOT TEST PROGRAM

Load (Million Btu/Hr)	4.4	(3.1)
Dense Bed Stoichiometry	0.8	(0.6, 1.0)
Excess Air (%)	25	(15 - 50)
Oxidizing Zone Temperature (°F)	1640	(1540, 1720)
Oxidizing Zone Gas Velocity (fps)	22	(17)
Oxidizing Zone Residence Time (sec)	2.2	(2.8)
Dense Bed Pressure Drop ("wg)	30	(20, 40)
Combustor Pressure Drop ("wg)	15	(10, 20)
Combustor Solids Flux (lb/hr-sqft)	10,000	(5,000 - 15,000)
Ca/S Mole Ratio	1.5	(1.0 - 3.0)
Flyash Reinjection Ratio	0	(0 - 4.0)
Sorbent	1/8" Limestone	(1/8" Dolomite)
NH3/NO Ammonia Injection Ratio	0	(0 - 8)

corresponded to 1640°F and 3.0 sec residence time. There was a remarkable improvement with the introduction of fly ash for the same temperature and residence time. The combustion efficiency jumped to 99.5%, with a fly ash reinjection (FAR) ratio of 4. The FAR ratio is defined as the ratio of the pounds of fly ash reinjected per pound of ash disposed. Sulfur capture did improve but the goal of 90% capture at Ca/S mole ratio of 1.5 was reached only when dolomite was used instead of high calcium limestone. For limestone, Ca/S mole ratio had to be raised to 3.0 for 90% sulfur removal. With dolomite, however, a sulfur capture of 96% was also demonstrated at Ca/S ratio of 2.3

To demonstrate the NO_x goal of 0.1 lb/MBtu in combination with the sulfur capture and combustion efficiency goals it was necessary to use NH₃ injection. A mole ratio of 2 for NH₃/NO was necessary to reach these NO_x levels. Due to physical constraints in the pilot plant it was not possible to optimize the location for ammonia injection. Also, additional work is necessary to improve the technique for dispersing ammonia into the system efficiently.

With the assistance of University of North Dakota Energy and Environmental Research Center (UNDEERC), additional data was collected on N₂O generation in an FBC environment. The measured quantities were between 50-180 ppm over a temperature range of 1540-1700°F. This was within the range measured at other commercial and pilot facilities. N₂O has a reverse relationship with NO with respect to temperature.

To achieve the boiler efficiency goal of 90% in the advanced concept, an excess air level of 10% had been selected in Level-1. This was done to reduce the dry gas losses, an important component in the determination of boiler efficiency. But pilot tests showed that to reach high levels of combustion and sulfur capture efficiencies an excess air of ~25% was necessary. Nonetheless, the data did establish that it was possible to attain 90% boiler efficiency at these higher excess air levels.

Table ES-2 shows the approach taken for the pilot test program to reach all the demonstrable goals simultaneously

The measured heat transfer coefficient in the vertical tube heat exchanger was 50 Btu/hr-Ft²-°F, which was lower than the earlier estimates of 65-70 Btu/hr-ft²-°F. A detailed circulation study will be performed in Level-3 to determine if the commercial

Table ES-2

GOALS COMPARED WITH MSFB PILOT PERFORMANCE

GOALS COMPARED WITH BEST SINGLE TEST
IN THE ADVANCED PILOT MSFB

	<u>Program Goal</u>	<u>Best Test</u>	<u>Conditions</u>
Combustion Efficiency (%)	99.5	99.2	99.6 Cyclone Efficiency
Ca/S Ratio for 90% Sulfur Capture	1.5	1.5	Fly Ash Reinjection Ratio at 2 Dolomite
NO _x Emission (lb/10 ⁶ Btu)	0.10	0.12	Ammonia Injection at 2 NH ₃ /NO

plant will operate in an all natural circulation mode.

Based on this test program, the recommended design parameters for the proof of concept facility are listed in Table ES-3.

ADVANCED DESIGN ECONOMICS UPDATE

Based on the results of the pilot tests, the advanced concept design was revised. The economic data developed in Level-1 was updated to reflect these changes. The total capital cost of the plant was estimated at \$7,790,000 at 1991 dollars. The evaluation criteria provided by METC was used for comparing costs of similarly equipped coal and oil fired units. It was estimated that the payback period for the advanced system is 2.2 years when compared to a #6 oil fired plant operating at a 80% capacity factor. This number goes down to 1.5 years in comparison to the #2 oil plant due to the higher #2 oil prices. The sensitivity of the payback periods to the capacity factor is shown in Figure ES.2. It should be noted that for these evaluations low calcium dolomite (44% Ca) was used as the sorbent. In comparison limestone had a Ca content of 90%. Thus for the same Ca/S mole ratio dolomite produced larger quantities of waste. But dolomite required a lower Ca/S mole ratio of 1.5, exactly half of what is needed for limestone. This made the amount of ash generated to be about the same, irrespective of the sorbent used. This simplified the sorbent economics. Thus, in the case analyzed it was the cost of the delivered stone that determined its selection. There was a marginal reduction in steam costs if dolomite was used instead of limestone. Figure ES-3 shows the reduction in the cost of steam produced as the capacity factor is increases from 0.5 to 0.8

PROOF OF CONCEPT FACILITY

The contract requires that the findings of the Level-2 work be incorporated into the design of a commercial plant that would be located at an industrial facility.

This involves the obtaining of commitment from an industrial customer to participate in the program. He should be willing to make the unit available for performance testing and for the generation of operational and maintenance data.

Although the original contract required that the proof of concept facility be fired exclusively by coal, during the execution of Level-2 it was felt that a wider market could

Table ES-3
RECOMMENDED ADVANCED DESIGN PARAMETERS

Dense Bed Stoichiometry	0.6 - 0.7
Excess Air (%)	25
Oxidizing Zone Temperature (°F)	1620 - 1640
Oxidizing Zone Velocity (fps)	20
Oxidizing Zone Residence Time (sec)	3.0
Combustor Solids Flux (lb/hr-sqft)	8 - 9,000
Cyclone Efficiency (%)	99.7
Flyash Reinjection Ratio	3.0
Sorbent:	Dolomite
NH ₃ /NO, Ammonia Injection Ratio	2
Vertical Tube EHE Heat Transfer Coefficient (Btu/hr sqft-°F)	50

ADVANCED BOILER ECONOMICS

COAL VS OIL/ SORBENT: DOLOMITE

Figure ES.2 Advanced Boiler Economics: Payback Periods

ADVANCED BOILER ECONOMICS

COAL VS OIL/ SORBENT: DOLOMITE

Figure ES.3 Advanced Boiler Economics: Steam Costs

be served if a co-firing approach was used. This would not prevent coal from being the primary fuel but would combine it with another fuel to find additional applications for the system developed. Many of these other fuels are waste materials produced in various industries that have reasonable heating value but not high enough to sustain combustion on their own. One such material is the sludge generated in a paper de-inking plant which has a heating value of approximately 6000-7000 Btu/lb (dry). A vast market potential exists for utilizing this material as more companies are forced into recycling of paper due to environmental and economic factors. Land fills are getting scarce and the costs for their use are soaring. The American Paper Institute has estimated that more than 45% of the old newsprint and magazines will be recycled by the year 1995. This number will go even higher through year 2000. Many of the paper companies are opening up recycling plants. International Paper (IP) has planned a de-inking facility at Lockhaven, Pa which will go into operation in mid-1994 and will produce approximately 50 tons of non-recyclable waste that is a potential co-fire fuel. A preliminary economic analysis performed by IP with Riley's help on this site has shown that an FBC boiler is economically viable against the landfill option. This will reduce the amount of waste and will add the much needed extra steam at a reasonable price.

The steam conditions for this facility are different from those used for the advanced system. IP needs a cogeneration plant rather than a process steam generator. Table ES-4 shows the comparison between the two conditions.

Before designing the proof of concept facility, however, two important issues have to be addressed related to the co-firing of coal and paper de-inking sludge:

- Understanding the material handling characteristics of the coal-sludge combination.
- Emission characterization of the fuel with respect to dioxins and trace metals.

Riley proposes a short test program, an extension of Level-2, for its pilot plant with a representative sludge sample to investigate these issues. If this test program is successful we will be creating a market that has tremendous potential for growth. Thus, final decision on proceeding to Level-3 can be deferred until the results this testing are available.

Table ES-4
STEAM CONDITIONS FOR LEVEL-3 PROOF OF CONCEPT FACILITY

	<u>Original Level 1</u>	<u>Proposed Level-3 (Cogeneration)</u>
Steam Flow (lb/hr)	110,000	100 - 250,000
Pressure (PSIA)	215	875 - 1500
Temperature (°F)	388	850 - 1000

RESULTS AND RECOMMENDATIONS

The findings of the Level-2 work are as follows:

- All of the performance goals and sub-goals of the advanced concept have been demonstrated during the pilot test program.
- The updated economics of the design based on these results has determined that it is a viable design.
- A potential market exists for using the advanced design by co-firing coal in combination with paper de-inking sludge.
- A host site for the proof of concept facility has been secured.

Based on the above, Riley recommends that METC proceed to the next level of the program. Prior to the commencement of the commercial design, it is recommended that a test program be conducted on Riley's pilot plant to investigate issues related to the co-firing approach suggested for Level-3.

SECTION 1.0

INTRODUCTION

Coal is one of the most abundant sources of energy in the world. The vast distribution network can supply it to the most remote parts of the country, and for that matter the world. It has been long used as a primary source of fuel for steam and power generation. Although all the fossil fuels are accountable for their contribution to the greenhouse effect, coal has been singled out as a key contributor to acid rain as well as the carbon dioxide emissions aggravating the greenhouse effect. Added to this is the challenge from the oil and gas industry on price and convenience. It is possible to counter this by providing economically attractive and environmentally acceptable systems that operate more efficiently by employing new innovative advanced combustion technologies. One such technology is the Fluidized Bed Combustion (FBC) which has been favorably received by the industry. Many of the world's largest companies are participating in its development by allocating financial and technical resources. It certainly has helped in design enhancements that have led to improvements in reliability, efficiency, fuel flexibility and emission control.

1.1 PROGRAM OBJECTIVES

The overall objective of this project is to develop a reasonably priced, socially acceptable, coal-fired atmospheric circulating fluidized bed (CFB) combustion technology. "Reasonably priced" means that the technology is economically competitive in domestic and international markets. "Socially acceptable" means that the technology is as clean, easy to use, and reliable as oil-fired technology of equivalent capacity.

The systems developed target two separate markets. The first one is the small package boiler market in the steam capacities of 10,000 to 50,000 lb/hr. The other is the industrial boiler market where the steam demand is between 75,000 to 150,000 lb/hr.

The intent is to offer an alternative to commercially available oil and gas systems. If the developed designs are found to be economically feasible and environmentally acceptable, large reserves of the high sulfur coal available in the eastern US could be

utilized reducing the country's dependence on foreign imported oil. Conceptual designs were prepared for both the sizes in the Level-1 of this program. An economic evaluation was performed to determine the viability of these designs. At the end of Level-1 it was determined that the small package concept would not be very attractive on economic and operability considerations. The payback periods were calculated to be of more than five years at capacity factors of 60%, a norm for this size. Also the ease of operation of a gas or oil fired installation would be very hard to compete with, considering the technical capabilities of a typical operator of such a small unit. The corresponding FBC unit requires special operator training and involves a fair amount of material handling for fuel, sorbent and ash. On the other hand, the economics of the industrial size boiler was found out to be attractive and being installed at a commercial industrial facility would not pose any problems regarding special operators. The payback periods were found out to be between one to three years. Therefore, at the conclusion of Level-1 work, it was recommended that further development be carried out only for the industrial size boiler called the advanced concept. Besides the favorable economics, the industrial advanced system showed promise to meet high combustion efficiency and stringent sulfur capture and NOx control requirements.

The advanced concept was preferred for meeting the following criteria specified in the contract:

- Quick start ups and shut downs.
- Demand responsive, automatic load following through a wide range of outputs.
- Safety and ease of operation, equivalent to oil-fired package boiler.
- Effective, reliable, and automatic control of particulates, sulfur dioxide and nitrogen oxide emissions at levels under the competing oil-fired package boilers burning No. 6 fuel with less than 2 percent sulfur.
- Operability and serviceability equivalent to the oil-fired systems as measured by the number of and qualifications of the operators.
- Clean fuel, ash, and sorbent handling from delivery to disposal, including on-site conveyance and storage.
- Space requirements for combustor and auxiliary equipment compatible with a typical site considering aesthetics and costs.
- Compatibility with conventional heat and steam distribution and

control technology.

- Cost competitive with oil-fired systems when compared by a \$/lb of steam or Btu delivered. This comparison conducted by considering capital amortization, operation/maintenance costs, as well as less tangible costs such as the cost of any additional space or the need of higher level operational personnel.

Level-2 work presented in this report consists of pilot testing of the concept developed in Level-1 to establish the operational and performance boundaries of the total system and to identify any remaining areas which have to be investigated in order to design and build a proof-of-concept test facility. Also, the contract requires that at the conclusion of Level-2 work, a commercial customer be available who is willing to offer a site to demonstrate the concept. This report describes the approach taken to establish a plan for the proof of concept facility in Level-3.

The next level will consist of operating the proof-of-concept test facility to demonstrate the performance of the system under actual conditions. The operability, reliability, and cost of the system will be determined with a high degree of accuracy and reliability. The general practice for obtaining realistic performances will be to have the system operated by the owner of the facility.

The boiler will be designed to demonstrate the following specific performance criteria:

- Turndown greater than 4 to 1.
- 90% SO₂ capture with more efficient sorbent use (Ca/S=1.5).
- NO_x emissions equivalent to or less than 0.1 lb/MBtu.
- Volumetric heat release approaching 0.5 million Btu/hr ft³ or greater.
- An economically attractive and simplified fuel feed system for as-received coal requiring a minimum of preparation.
- Boiler thermal efficiency greater than 90 percent.
- Alternate fuel flexibility.

1.2 LEVEL-2 EXECUTION APPROACH

The Level-2 program had three distinct parts to it. The first part related to the demonstration of the advanced concept in a pilot plant designed to include its unique process and hardware characteristics. Riley's CFB pilot was constructed prior to being

made available for this contract. It was employed to simulate field installations for evaluating new process and design refinements before their implementation. Thus, a data base existed before the commencement of Level-2 work to help develop a philosophy for testing the advanced concept.

In the second part, results of this test program were used to revise the 110,000 lb/hr preliminary design relating to process parameters and hardware selection. After incorporating these modifications a fresh economic analysis was performed to calculate steam costs and pay back periods against oil fired designs.

The third part of Level-2 consisted of steering the design concept into the market place for commercial demonstration. It involved discussions with a number of potential users of the technology developed in the contract. This effort was directed at those industries that are in need of a technology which facilitated the utilization of poor quality fuels in general. Of course, the impetus was to find a customer interested in using the high sulfur eastern coals. Follow up was done with a limited number of candidates. This led to the commitment of a customer to offer an industrial site for locating the proof of concept facility. The approach taken for carrying out this part of the program was to look beyond the current project, finding long term markets where the technology developed could be meaningfully applied. As discussed in Section 8, it resulted in identifying co-firing of coal in combination with de-inking paper sludge as a value added option in the advanced design.

SECTION 2.0

RILEY MSFB DESIGN

Riley has been offering fluidized bed systems since 1982. Its first units went into operation for General Motors in 1987. Since its very first boiler Riley has been aggressively working on refining its CFB designs. Its latest boiler for the City of Wyandotte encompasses the results of many years of development work conducted by Riley through internally funded programs. A major product of this effort was the construction and operation of a pilot plant at its R&D facilities starting in the spring of 1989. To demonstrate Riley's commitment to further this technology, the plant has been now been assigned to the current DOE contract for developing advanced CFB designs.

2.1 PROCESS

This design concept contains many characteristics, listed as follows, which can be refined and optimized to obtain an advanced second generation atmospheric fluidized bed combustion (AFBC) system.

- High combustion efficiency.
- High sulfur capture at lower Ca/S ratios.
- High NO_x reduction through staged combustion.
- High turndown and the ability to respond to more rapid load changes than conventional AFBC due to the de-coupling of combustion and heat transfer.
- Improved system reliability by avoiding erosion and corrosion of steam generation tubes.
- Proven water tube boiler design for coal-fired applications. A fuel and sorbent feed system requiring few feed points, and the capability to accommodate liquid fuels including coal-water-mixtures.
- An ability to handle run of mine coal up to two inches in size.

An important feature of this system is the de-coupling of the heat transfer and combustion processes. Two thirds of the total heat duty is performed outside of the combustor. This helps in optimizing the process performance with in the combustor.

The heat duty is split between the convective pass, located downstream of the primary cyclone(s) and the EHE. As mentioned above, the EHE consists of steam generating tubes immersed in a conventional low velocity fluidized bed. The desired combustor temperature can be obtained by modulating the split of solids going through the EHE and direct into the combustor. Heat transfer in the EHE can be facilitated by changing its bed level thus exposing selected amounts of heating surface to hot solids. This helps in achieving optimum hydrodynamic and process conditions with in the combustor without compromising heat transfer, an important advantage for turn down considerations. A typical process schematic is shown in Figure 2.1.

Temperature plays a prominent role in achieving good process performance regarding combustion, sulfur capture and NO_x emissions. There is a direct relationship with temperature for combustion efficiency. Sulfur capture efficiency, in general, peaks at around 1650°F, after which it tends to go down. There is a reverse relationship for the two primary constituents comprising the oxides of nitrogen, namely NO and N₂O. While NO goes up with temperature, N₂O shows a downward trend (1-4). On the other hand, the total oxides of nitrogen go down with an increase in temperature and level off at approximately 1750°F. The pilot test program also reflected this relationship as discussed in Section 6. Thus, Riley's process philosophy has been fashioned to operate at a temperature of 1650°F which gives good combustion and sulfur capture performance. The NO_x levels are controlled in the post combustion stage by ammonia or urea injection.

2.2 EVOLUTION OF THE ADVANCED CONCEPT

The first Riley plant went into operation in 1987 at a General Motors plant in Fort Wayne, Indiana. Riley's latest installation, sold to the city of Wyandotte in Michigan is due to go in operation May 1992. It will produce 250,000 lb/hr of steam for cogeneration. A listing of the plants sold to date is shown in Table 2-1.

Over the years the Riley design has gone through many refinements, and the advanced concept developed during this contract is a product of this ongoing process. To pursue international markets for the technology, Riley has licensed it to Mitsui Ship Building Corporation of Japan. This relationship has contributed towards improving design and performance through mutual discussions.

THE MULTI-SOLID FLUIDIZED BED

Figure 2.1 FBC Process Schematic

Table 2-1
FIRST GENERATION FACILITIES

<u>Customer</u>	<u>Location</u>	<u>Industry</u>	<u>Startup</u>	<u>Conditions</u>	<u>Fuels & Sorbent</u>
Conoco	Uvalde, Texas	Enhanced Oil Recovery	12/82 1 Unit	50 MM BTU/HR 2450 PSIG 80% Quality	Coals Pet. Coke Limestone
Kerry Coop	Listowell, Ireland	Dairy Products	4/84 1 Unit	117,000 LBS/HR 350 PSIG 435°F	Coal, Peat Woodchips, Sawdust Anthracite
*Gen. Motors Truck and Bus Group	Fort Wayne, Indiana	Automotive	3/87 2 Units	150,000 LBS/HR 700 PSIG 765°F	Coals, Wastes Paint Sludges Limestone
*Kuraray	Japan	Textiles	9/86 1 Unit	154,000 LBS/HR 1280 PSIG 905°F	Anthracite Delayed Coke Limestone
ICI	Scotland	Chemicals	6/87 1 Unit	85,000 LBS/HR 650 PSIG 850°F	Coals, Peat Heavy Oil Plastic Wastes
*University, of Missouri	Columbia, Missouri	University	10/87 1 Unit	200,000 LBS/HR 950 PSIG 850°F	Coals Limestone
*University of Iowa	Iowa City Iowa	University	9/88 1 Unit	170,000 LBS/HR 475 PSIG 750°F	Coals Limestone
*Idemitsu	Japan	Refinery	6/87 1 Unit	660,000 LBS/HR 1865 PSIG 1004°F	Coals Limestone
*A.E. Staley	Decatur, Illinois	Food Processing	10/88 2 Units	375,000 LBS/HR 1265 PSIG 955°F	Coals Limestone
*City of Wyandotte	Wyandotte, Michigan	Municipal Utility	8/92 1 Unit	250,000 LBS/HR 875 PSIG 900°F	Coals Limestone
Riley Stoker	Research Ctr. Worcester, MA	Pilot Plant for MSFB R&D Work	5/89	7 MMBTU/HR No steam Production	Coals Limestone

*Designed under license with or supplied by Riley Stoker

Additional data has been generated at Riley's pilot plant at Worcester, Massachusetts which started operation in 1989. The information gathered on the pilot plant has been successfully utilized to address many of the process related issues in the Riley design. The experience on the pilot plant has been very valuable in finding solutions to operational problems on Riley's field installations. The same pilot plant has now been modified for this contract to carry the development work into the next generation of CFB designs.

The data generated during the test program under this contract has made significant contribution towards refining the Riley design.

Starting with an all refractory design at General Motors, the Riley design now features full waterwall construction and a state of the art vertical tube external heat exchanger which has the potential to operate in natural circulation mode. This would eliminate the need of EHE circulation pumps leading to reduced horse power requirements, hence higher plant efficiencies.

Following is a discussion of the salient features of this design.

Combustor

A completely water walled combustor was designed for the advanced system. This resulted in the elimination of the expansion joint at the junction of the refractory lined dense bed and the water walled upper combustor in the standard Riley designs. A silicon carbide refractory lining will be located over the waterwall in the lower combustor. Silicon carbide is especially erosion resistant, thus allowing a thinner refractory.

This new arrangement will be both lighter and allow for more heat transfer than the old thick refractory arrangement. It has the potential to reduce the structural steel and foundation requirements of the combustor. Reduced refractory mass may also result in decreased start up times since there will be less mass to be heated. Lower maintenance costs associated with silicon carbide refractory comes as an added benefit.

Air Supply and Distributor

The earlier Riley designs have used air pipes with nozzles to supply primary air to the combustor. There have been numerous operating problems due to differential expansion causing seal failures and erosion in the air nozzles. At some locations, poor combustor fluidization has been noticed due to non-uniformity in the air flow through the fluidizing pipes. This has resulted in hot spots and subsequent bed agglomeration problems.

To overcome such problems a water cooled air plenum and distributor equipped with bubble caps was designed for the advanced design. A special waterwall construction was devised to integrate the combustor waterwall circuits with that of the air plenum/distributor. This arrangement is expected to result in a considerable reduction in expansion related problems.

External Heat Exchanger (EHE)

A unique feature of the advanced design is in its EHE. As discussed above, unlike other EHE designs available in the market, the Riley design uses a vertical tube bundle for evaporative duty. The special header arrangement and roof construction utilizes the conventional boiler bank type heating surface installation. The vertical tubes enable the evaporative circuit to operate in a natural circulation mode, hence reducing the circulation pump horsepower requirements if horizontal tube circuits were to be used.

This arrangement also eliminates any need for a refractory lined vent line leaving the EHE. The integrated water walls provide the opening for the vent line by offsetting some of the tubes. With the absence of a standard vent line, refractory and expansion problems are reduced if not eliminated.

SECTION 3

MSFB PILOT DESIGN AND CONSTRUCTION

A pilot-scale Circulating Fluidized Bed (CFB) test facility is located at Riley Research in Worcester, Massachusetts. It was originally built to examine trade-offs among performance parameters such as combustion efficiency and sulfur capture with changes in Riley's Multi-Solids Fluidized Bed (MSFB) technology. Many things were learned about the MSFB process during the original test program, which lasted from May 1989 to April 1990. It assisted operation of Riley's field units. It helped establish the design parameters for the new 275,000 pounds of steam per hour unit at the City of Wyandotte. Many issues regarding unit stability and operation were clarified. Beginning in January 1991, the pilot facility was refurbished and modified to carry-out Level-2 of the Advanced Systems Program with METC.

3.1 FACILITY DESCRIPTION

3.1.1 Original Hardware

The original pilot facility, shown in Figure 3.1, was made up of 4' flanged sections so that it could be taken apart and reassembled to simulate different physical configurations. It was designed for a maximum 7 Million Btu/hr heat input for bituminous coal although it also was used to burn low heating value anthracite culm. It was equipped with many ports to accommodate different air inlet locations, instrumentation, and probes.

An existing positive displacement blower supplied primary and secondary air. Total air flow was controlled by dampering a vent from the atmosphere. Air preheat was accomplished by electric heaters in the primary and secondary air ducts.

Fuel and sorbent were bunkered in 12 ton bins and gravimetrically fed to the combustor through weigh belt feeders and a rotary valve. Dense bed material was fed through the rotary valve and makeup sand was fed through a port on the EHE as needed. Start-up was accomplished by an overbed natural gas burner. The burner was effective when used in conjunction with hot recycle to bring the entire recycle loop to temperature.

Figure 3.1. Isometric of the Original MSFB Pilot Facility.

A high efficiency, refractory lined cyclone was employed for separating solids and gas exiting the combustor. The high capture efficiency facilitated the operation of the unit by minimizing solids make-up requirements. This was especially true with low ash fuels.

The hot compartment and external heat exchanger were originally incorporated in a single unit separated by a refractory wall. During start-up, the EHE bed level was kept low to maintain a high hot recycle temperature. Then, when combustion temperatures were reached, the hot recycle and the cold recycle temperatures were set by intermixing the compartments. A proper balance between temperatures led to smooth recycle flows and stable unit operation.

Originally, city water was used to cool the pilot facility once through. Horizontal tubes were used in the EHE to cool the solids. These tubes were equipped with separate flow meters, flow detectors, and thermocouples to characterize heat transfer in the EHE. Likewise, the flue gas cooler at the outlet of the cyclone was served by a bank of metered circuits.

A flue gas analysis train was located downstream of the flue gas cooler to continuously monitor SO_2 , NO_x , CO, CO_2 and O_2 . Ports for particulate sampling were located at several places for measuring recycle flow rate and evaluating cyclone efficiency. Flue gas particulate cleanup was accomplished by an existing baghouse. A caustic scrubber was installed after the baghouse to meet local SO_2 emission levels in the event sorbent feed was interrupted. The combustor draft was controlled by an existing induced draft fan. The combustor draft was modulated by dampering an inlet to the fan from the atmosphere.

The control room was equipped with a control panel and two computers. The unit was monitored and manually controlled by the control panel. Combustor temperature and total recycle flow were automatically controlled by one computer. This computer also gathered all process data. The other computer was used to display temperature trends over 8 hour periods for assessing unit stability and equilibrium.

3.1.2 Original Test Program

Except for a month long shutdown in December 1989, the pilot facility was operated

essentially 24 hours a day, seven days a week in the original test program. Several tests were run specifically to simulate the performance of field units. These tests demonstrated that combustion efficiency, sulfur capture, NO_x emissions, and CO emissions measured in the pilot facility matched the same measurements in the field. Recommendations were made regarding the field units based on the original pilot test results. The City of Wyandotte MSFB design, which was in the planning stage, was also based on these results.

3.2 PILOT FACILITY REDESIGN

Due to extreme use in the original test program, the pilot facility deteriorated such that many of the components had to be refurbished or replaced. There was much refractory wear in the combustor and the cyclone. The external heat exchanger suffered from water leaks, and the L-valves showed extensive erosion.

To prepare for Level-2 of the Advanced Systems Program, a large effort was internally funded to refurbish the pilot facility. The effort was directed toward two targets. One target was to modify the MSFB process to achieve the performance goals of the program. The process changes required to achieve the performance were established by extrapolating the original test program results. The other target was to make a key hardware changes. The cyclone was modified to maintain the high capture efficiency with changes in the process. Also, the original horizontal tube external heat exchanger was replaced with a vertical tube design. The vertical tube design required changes in the physical arrangement of the pilot MSFB.

3.2.1 Process Changes

The major performance goals of the Level-2 Advanced Systems Program are:

- 99.5% Combustion Efficiency
- 90% Sulfur Capture at 1.5 Ca/S
- 0.10 lbs NO_x per million Btu

These goals are discussed in detail in Section 4. In redesigning the pilot facility, process changes were made aimed at achieving these goals. The intent of the original test program was to test nominal performance levels for the field units. The advanced system levels were not approached. Nevertheless, the data from those tests were

extrapolated to determine the necessary process changes.

Figure 3.2 shows the relationship of combustion efficiency and sulfur capture with gas residence time in the MSFB combustor extrapolated to 3 seconds residence time based on the original test program results. This residence time was the maximum that could be achieved because of constraints in the height and gas velocity in the pilot facility. As can be seen in the figure, 98.5% combustion efficiency and 90% sulfur capture at Ca/S ratio of 2.5 was predicted at 3 seconds residence time. Figure 3.3 shows the extrapolated sulfur capture relationship with Ca/S ratio at 3 seconds residence time. As can be seen, 75% sulfur capture was expected at 1.5 Ca/S mole ratio.

The original test program showed that combustion efficiency and sulfur capture was improved by employing fly ash reinjection. Fly ash reinjection was tested at various fly ash reinjection ratios. This ratio is defined as the amount of fly ash reinjected divided by the net amount of fly ash collected by the baghouse. As shown in Figure 3.4, the combustion efficiency goal could be approached from the predicted 98.5% level by reinjecting fly ash at a ratio greater than 1.5. Similarly, it was possible to improve sulfur capture from 75% to 90% at 1.5 Ca/S by employing fly ash reinjection.

The original test program also established good operating combustor temperature, dense bed stoichiometry, dense bed pressure drop, combustor pressure drop, and excess air level. Alumina beads were found to be the ideal dense bed material since little dense bed make-up was required. The alumina beads also gave stable unit operation.

In the original test program, the lowest load achieved in the pilot facility was 4 Million Btu per hour. This load was operated with great confidence. Indeed, it gave the best performance. In the Level-2 program, a nominal operating condition was defined as the parameters that gave the best performance in the original test program. Table 3-1 summarizes this condition. As can be seen, a residence time of 2.2 seconds was possible at 4 Million Btu per hour by assembling the pilot facility to its greatest height. All new equipment was designed based on this nominal condition. Mass and heat balances were completed to determine the required process flows. Figure 3.5 shows the process flow diagram for the pilot facility at nominal load.

Figure 3.2. Predicted relationship of combustion efficiency and sulfur capture with gas residence time

Figure 3.3. Predicted relationship of sulfur capture with Ca/S mole ratio at 3 seconds residence time.

Figure 3.4. Predicted relationship of combustion efficiency and sulfur capture with fly ash re-injection ratio at 3 seconds residence time.

Table 3-1

PROCESS CONDITIONS IN THE PILOT MSFB
ADVANCED CFB CONFIGURATION

Nominal Load: 4,400,000 Btu/hr

τ_{ox} , Residence Time	2.2 seconds
U_{ox} , Oxidizing Zone Gas Velocity	20 fps
FAR, Flyash Reinjection Ratio	>1
SR_{db} , Dense Bed Stoichiometry	0.8
XSA, Excess Air	>20%
T_{ox} , Oxidizing Zone Temperature	1650°F
G_s , Combustor Solids Flux	10,000 lb/sq ft/hr

Low Load: 3,000,000 Btu/hr

τ_{ox} , Residence Time	3.2 seconds
U_{ox} , Oxidizing Zone Gas Velocity	15 fps
FAR, Flyash Reinjection	>1
SR_{db} , Dense Bed Stoichiometry	0.8
XSA, Excess Air	>20%
T_{ox} , Oxidizing Zone Temperature	1650°F
G_s , Combustor Solids Flux	10,000 lb/sq ft/hr

DESIGN VARIABLES

Oxi. Zone Velocity	<u>20 fps</u>
Oxi. Zone Res. Time	<u>2.2 sec</u>
D.B. Stoichiometry	<u>0.8</u>
Excess Air	<u>20 %</u>
Oxi. Zone Temp.	<u>1650 F</u>
D.B. Temperature	<u>1550 F</u>
Cold Recycle Temp.	<u>1100 F</u>
Hot Recycle Temp.	<u>1600 F</u>
Oxi. Zone Solids Flux	<u>10,000 lb/ft²</u>
Cyclone Efficiency	<u>99.7 %</u>
Fly Ash Reinj. Ratio	<u>1.0</u>
System Heat Loss	<u>15 %</u>

TEMPERATURES (F)

1) Main Header	<u>100</u>
2) Primary Air	<u>500</u>
3) D.B. Temp.	<u>1600</u>
4) Oxi. Zone Temp.	<u>1650</u>
5) Cyclone Inlet	<u>1650</u>
6) Cyclone Dipleg	<u>1650</u>
7) Hot Recycle	<u>1600</u>
8) Cold Recycle	<u>1100</u>
9) EHE Vent	<u>1000</u>
10) Secondary Air	<u>500</u>
11) Cyclone Outlet	<u>1600</u>
12) F.G. Cooler Out	<u>300</u>
13) Baghouse Outlet	<u>300</u>
14) ID Fan	<u>250</u>

PRESSURES

1) Main Header	<u>5 psig</u>
2) Primary Air	<u>2 psig</u>
3) Hot Comp.	<u>2 psig</u>
4) EHE	<u>2 psig</u>
5) Comb. Draft	<u>0 °WC</u>
6) Cyclone Inlet	<u>-15 °WC</u>
7) EHE Vent	<u>-2 °WC</u>
8) Cyclone Out	<u>-17 °WC</u>
9) Baghouse	<u>-21 °WC</u>
10) ID Fan	<u>-25 °WC</u>

GAS FLOWS (pph)

1) Burner Air	<u>0</u>
2) Primary Air	<u>2560</u>
3) Second. Air	<u>120</u>
4) Hot Comp.	<u>300</u>
5) EHE	<u>760</u>
6) SA Vent	<u>0</u>
7) F.G. Flow	<u>4160</u>
8) LD. Fan	<u>4160</u>

SOLIDS FLOWS (pph)

1) Fuel	<u>340</u>
2) Limestone	<u>76</u>
3) Comb. Solids	<u>30625</u>
4) Hot Recycle	<u>15550</u>
5) Cold Recycle	<u>15075</u>
6) Sand Make-up	<u>-55</u>
7) Cyclone Out	<u>92</u>
8) Baghouse	<u>92</u>
9) Net Baghouse	<u>46</u>
10) Fly Ash Reinj.	<u>46</u>

CONCENTRATION

1) Oxygen	<u>3.5 %</u>
2) Carbon Dioxide	<u>15.5 %</u>
3) Sulfur Dioxide	<u>344 @3%O₂</u>
4) Nitrogen Oxide	<u>74 @3%O₂</u>

Figure 3.5. Pilot MSFB process flow diagram for the Advanced MSFB Program.

To achieve the required 3 seconds residence time, a new low load condition was defined for the pilot facility. This condition is also listed in Table 3-1. This load had not been previously operated in the pilot facility. In this condition, the gas velocity approaches 15 feet per second. It was not known how well the unit would operate at this velocity.

3.2.2 Hardware Changes

The downrating of the load to 4.4 million Btu/hr also required a change in an important piece of hardware: the cyclone. At the end of the original test program, the cyclone had to be refurbished due to extensive wearing. It was not returned to its original dimensions, however. The cyclone inlet dimensions were reduced to maintain the same inlet velocity as at the original 7 million Btu/hr heat rate. The inlet area was reduced 30% to handle the reduced load. The expected cyclone efficiency was 99.7%.

Another required change was the incorporation of the vertical tube external heat exchanger. As shown in Figure 3.6, it was decided to separate the hot compartment and EHE into two compartments and wrap the EHE around the combustor to best accommodate the new design. The drawback was that the wrap around design was new. In the side by side design, the solids fell over a weir from the hot compartment to the EHE. As mentioned before, mixing of the two compartment was possible by raising the EHE bed level. This allowed the recycle temperatures to be modulated. In the wrap around design, an overflow chute was required to transfer solids from the hot compartment to the EHE. No solids inter-mixing between the EHE and hot compartment was possible. There was some concern about solids movement in the wrap around design since solids had to move laterally rather than vertically. Ultimately, the wrap around design was chosen because of the accumulated clearance problems incorporating a vertical-tube EHE side by side with the hot compartment.

Since solids mixing was not possible with the new design, the tube bank had to be carefully designed so that the cold recycle temperature fell within an operating window. It was recognized, however, that a vertical tube design did allow for a degree of flexibility. This was due to the fact that heat transfer is mostly controlled by how much the tubes are immersed in the bed. This can be seen in the heat transfer model for a fluidized bed:

$$Q_{EHE} = U_0 A_0 (T_{EHE} - T_{tubes})$$

where Q_{EHE} = the EHE heat duty (Btu/hr)

U_0 = Overall heat transfer coefficient (Btu/hr ft² °F)

A_0 = Outside tube surface area covered by solids (ft²)

T_{EHE} = EHE fluidized bed temperature (°F)

T_{tubes} = Cooling tube temperature (°F)

This model assumes the bed is isothermal. Once the EHE was designed with a fixed number of tubes, a range of operating loads was handled by maintaining a high bed level so that the covered surface area was high at high loads and decreasing the bed level to decrease the covered surface area at low loads. In that way, the EHE heat duty was varied with load and the cold recycle temperature maintained within the operating window.

Sensitivity studies were done using the heat transfer model to predict the effect of pilot load on EHE bed temperature for a fixed heat transfer coefficient. Since one of the objectives of the test program was to measure the overall heat transfer coefficient, this coefficient was not known in the design phase. Therefore, further sensitivity studies were done to predict the effect of heat transfer coefficient on the EHE bed temperature for a fixed load. The EHE height and the required number of cooling tubes were determined based on these sensitivity studies. The required EHE plan area was then determined after the tubes were arranged. The resulting EHE design is shown in Figure 3.7.

The vertical tube EHE was designed so that it could be disassembled to access the cooling tubes. This could be done by removing the outer refractory shell of the EHE upper module. The tubes could then be replaced if needed. Provisions were made to add or remove tubes in case the actual heat transfer coefficient was found to lower or higher than design.

3.3 NEW AND REFURBISHED PILOT FACILITY COMPONENTS

Figure 3.8 shows the arrangement of the pilot MSFB configured for the Advanced Systems Program. The new, refurbished and existing equipment are indicated. As can be seen, all of the cyclone and many of the combustor pieces were refurbished with new refractory. The lower combustor section was reworked. A new secondary air port was installed in this section to maximize the oxidizing zone height. The lower

Figure 3.7. The Vertical Tube External Heat Exchanger - Pilot MSFB Facility.

Figure 3-8. The pilot MSFB at Riley Research configured for Level-2 of the advanced system program.

combustor required a new, smaller plan area for the downgrade in load.

A new hot compartment, EHE, hot recycle L-valve and cold recycle L-valve were designed and installed. The only items reused were the original 310 stainless steel fluidizing tubes. These were found to be in good shape after the original test program.

In general, the refractory was made up of three components to maintain low skin temperatures. The low skin temperatures were required to minimize expansion and system heat losses, and for safety. The inner material was a hardface refractory to resist wear from the hot solids and gases. Different type of hardfaces were used depending upon the degree of wear expected. For instance, a hard alumina-chromic oxide refractory was used in the dense bed area of the lower combustor section. The middle material was a castable refractory of varying thickness depending upon the inner surface temperature. The outer material was an insulating board.

The EHE vent cooler and secondary air system was also replaced. To change the bed level in the EHE to accommodate load changes and for heat transfer tests, a new sand make-up feeder and EHE ash cooler were installed. New water plumbing was required for the vertical tube EHE, and new flow meters and flow detectors were installed. As part of a capital improvement project, the once through cooling water system was replaced with a circulating water system with a cooling tower.

3.4 INSTRUMENTATION AND DATA ACQUISITION

The data acquisition computer was a COMPAQ-286 PC. A commercially available software program called LT/Control supplied by Labtech Corporation was used to display and record the data and to control the recycle L-valves. Figure 3.9 shows the display screens that the operator used to help operate the pilot facility. Another COMPAQ-286 was used to display temperature trends as shown in Figure 3.10. This computer was programmed using another Labtech product called LT/Notebook. The temperature trends were very useful for determining the stability and equilibrium of the pilot facility.

The data acquisition computer recorded the data every 10 seconds and stored it on hard-disk in binary format. After a test, this data was plotted to determine the most representative time period. Average values of the data were calculated and tabulated

Figure 3.9 Pilot MSFB Display Screens

Figure 3.10 Pilot MSFB Temperature Trends Display Screen

for the test period. A typical test result is shown in Table 3-2.

Figure 3.11 is an instrumentation diagram showing the location of the thermocouples and pressure transducers on the pilot facility. In addition to being recorded by computer, the temperatures were read on digital displays on the control panel and the pressures were read on local magnahelics. Fuel and sorbent flows were controlled and monitored by feeder controller located on the control panel. Fly ash reinjection and sand make-up were measured by calibrated screw feeders. Scales were set-up to measure the various ash flows from the pilot facility. Flue gas instruments were closely monitored to be sure that reliable emissions data were being taken.

Table 3-2

MSFB PILOT FACILITY
COMPUTER RECORDED DATA

TEST : Test 18
START : 07/30/91 02:00
FINISH: 07/30/91 07:00

		AVERAGE	STDDEV
1 Fuel	pph	338	8
2 Limestone	pph	52	1
3 Burner Gas	acfh	0	0
4 PA	pph	2549	54
5 Burner Air	pph	0	0
6 SA	pph	1019	68
7 SA MakeUp Air	pph	175	45
8 SA Vent Air	pph	5	18
9 Hot EHE Air	pph	283	11
10 Cold EHE Air	pph	772	11
11 T Low	F	410	63
12 T Red 1	F	1823	714
13 T Red 2	F	1599	8
14 T Red 3	F	1600	7
15 T Lower Comb	F	1602	8
16 T Oxi 1	F	1622	7
17 T Oxi 2	F	1642	4
18 T Oxi 3	F	1610	2
19 T Upper Comb	F	1643	3
20 T Cyclone In	F	1634	4
21 T Cyclone Out	F	1054	2
22 T Dip Leg	F	1505	6
23 T Hot EHE	F	1656	0
24 T Cold EHE	F	1277	4
25 T EHE Vent	F	830	3
26 T VentClr Out	F	589	2
27 T SA	F	336	5
28 T Hot Recy	F	1607	6
29 T Cold Recy	F	1292	11
30 T FGC Out	F	311	1
31 T Baghouse	F	255	0
32 T Scrubber Out	F	114	0
33 dP DB	inWC	40.6	3.1
34 Comb Draft	inWC	-1.0	1.1
35 dP Comb	inWC	15.0	.7
36 dP Cyclone	inWC	3.9	.3
37 dP Hot EHE	inWC	81.1	2.9
38 dP Cold EHE	inWC	27.8	1.0
39 P EHE Vent	inWC	.2	.0
40 P Baghouse	inWC	-25.7	.4
41 P ID Fan Inlet	inWC	-27.7	.0
42 SO2	ppm	805	85
43 NOx	ppm	145	5
44 N2O	ppm	21	11
45 CO	ppm	9	2
46 CO2	%	13.8	.2
47 O2	%	5.4	.7
48 Uo DB	fps	9.1	.7
49 Uo Lower Comb	fps	23.2	.5
50 Uo Upper Comb	fps	20.2	.4
51 Uo Hot EHE	fps	.8	.0
52 Uo Cold EHE	fps	.6	.0
53 SR DB		.80	.02
54 SR Lower Comb		.80	.02
55 SR Upper Comb		1.18	.03
56 Hot Recy Ctrl	% Open	49	8
57 Cold Recy Ctrl	% Open	41	7

Figure 3-11. Instrumentation diagram for the Advanced MSFB pilot facility.

SECTION 4

THE TEST APPROACH

4.1 LEVEL-2 GOALS

Level-2 of the Advanced Systems Program encompassed pilot-scale testing of the Advanced MSFB concept. Pilot testing targeted many of the overall goals of the program. These goals are listed in Table 4-1. Some of the overall goals could not be directly addressed using the pilot facility, however. For instance, a boiler efficiency was not directly measured since no steam was produced. To address the overall goals, specific Level-2 goals were set for performance variables that could be measured in the pilot facility.

4.1.1 Process Related Goals

One of the program's overall goals calls for 90% boiler efficiency. Boiler efficiency depends primarily upon the excess air, stack temperature, and carbon loss. The original heat loss calculation at the proposal phase of the program showed that 10% excess air, 200°F stack temperature, and 1/2% carbon loss gave a boiler efficiency of 91.1% without margin. However, later in-house testing on the pilot MSFB proved that a 10% excess air level was detrimental to sulfur capture and carbon loss. So the excess air level goal was dropped before Level-2 testing began. The 200°F stack temperature goal required the use of a large corrosion resistant air heater. This had economic implications that will be discussed in Section 7. A subsequent heat loss calculation at 25% excess air, 200°F stack temperature, and 1/2% carbon loss gave a 90.7% boiler efficiency. These conditions, which still meet the goal, were felt to be more feasible for the advanced MSFB concept.

Table 4-2 compares the boiler efficiency and related goals with the norm for industry. As can be seen, the carbon loss goal well exceeds the industry norm. The carbon loss can be directly measured in the pilot facility. Therefore, a carbon loss of 1/2% (or equivalently, a combustion efficiency of 99.5%) was made a Level-2 goal. Air preheat was accomplished by electric heaters in the pilot facility rather than an air-flue gas heat exchanger due to its small size. Therefore, the 200°F stack temperature goal was not addressed in the pilot testing.

Table 4-1
ORIGINAL ADVANCED SYSTEMS PROGRAM OVERALL GOALS

Process

Boiler Efficiency	90%
Excess Air	10%
Stack Temperature	200°F
Carbon Loss	< 0.5%
Ca/S Mole Ratio	1.5 for 90% Sulfur Capture
NO _x	0.10 lb/MilBtu
Particulate	0.03 lb/MilBtu
Turndown	4:1

Hardware

Vertical Tube External Heat Exchanger
 Water Cooled Primary air Distributor

Table 4-2
ADVANCED BOILER EFFICIENCY GOALS AND INDUSTRY NORMS

	<u>Industry Norm</u>	<u>Advanced Design</u>
Boiler Efficiency	87 - 88%	90% or Higher
Excess Air	20 - 25%	25%
Carbon Loss	2%	< 0.5%
Stack Temperature	250°F - 275°F	200°F

Sulfur capture and NO_x emission can also be directly measured in the pilot facility. In Table 4-3, the sulfur capture and NO_x emission goals are compared to industry norms. These too exceed what is normally seen in industry.

The stack particulate emission goal depends mainly on the performance of a baghouse. The existing baghouse on the pilot facility was not scalable to any field conditions. Therefore, particulate emissions were outside the scope of the pilot scale test program.

The turndown ratio was an issue in the advanced concept because the process changes discussed in Section 3 called for decreasing the design gas velocity to 15 feet per second. At a gas velocity of 30 feet per second, a turndown ratio of 4:1 was possible. At 15 feet per second gas velocity, 4:1 turndown is possible only by using flue gas recirculation. While it was possible to install a flue gas recirculation fan in the pilot facility, it was decided that pursuing flue gas recirculation may have interfered with the primary goals of combustion efficiency, sulfur capture, and NO_x emissions.

The combustion efficiency, sulfur capture, and NO_x emission goals were targeted as the major goals of the pilot test program. As will be discussed, other sub-goals were added to Level-2 in the course of the test program. Economic issues were temporarily put aside in developing technical approaches to achieve the goals. The technical issues are contrasted with the economic issues in Section 7.

4.1.2 Hardware Related Goals

An implied goal of the test program was to characterize heat transfer in the vertical tube EHE. Specifically, a heat transfer coefficient was needed for the vertical tube design. Riley's field units employ a horizontal tube EHE where the coefficient is well characterized. However, the coefficient was expected to be different in the vertical tube EHE. Determining the coefficient was even more important since natural circulation was proposed for the commercial vertical tube design. It was needed to assess the feasibility of natural water circulation in the EHE. As will be discussed, the approach to measuring the heat transfer coefficient in the test program was simple and direct.

A water cooled primary air distributor was successfully used in the original, in-house test program. It was determined to be perfectly feasible for the advanced design. It

Table 4-3
SULFUR CAPTURE AND NO_x EMISSION
COMPARED TO INDUSTRY NORM

	<u>Industry Norm</u>	<u>Advanced Design</u>
Ca/S Mole Ratio for 90% Sulfur Capture	2.5 - 3.0	1.5
NO _x Emissions (lb/10 ⁶ Btu)	0.40	0.10

was used in the Level-2 test program without modification.

4.2 TECHNICAL APPROACH

4.2.1 Combustion Efficiency and Sulfur Capture

Table 4-4 illustrates the technical approaches that were used to obtain the process goals in this program. As discussed in Section 3, the pilot MSFB cyclone was redesigned to give a high capture efficiency. Improving cyclone efficiency improves combustion efficiency by retaining carbon in the solids recirculation loop. Similarly, sorbent inventory is maintained in the system improving sulfur capture. Therefore, both the combustion efficiency and the sulfur capture goals were targeted by improving the cyclone efficiency.

The pilot facility was configured to give the maximum residence time possible. Increased residence time improves carbon burnout and provides more time for the sorbent to capture sulfur. Therefore, increasing residence time was aimed at both combustion efficiency and sulfur capture.

As shown in Section 3, after maximizing combustion efficiency and sulfur capture with improved cyclone efficiency and residence time, it was possible to achieve the program goals for these performance variables by applying fly ash reinjection. Reinjecting fly ash into the combustor provided a straightforward means to improve these performance variables since the fly ash is high in carbon and calcium content. Therefore, fly ash reinjection was directed toward both combustion efficiency and sulfur capture.

In the course of testing, a sub-goal to the sulfur capture goal was added. This sub-goal was to demonstrate 96% sulfur capture and to determine the required Ca/S ratio. This was done to test the limits of sulfur capture in the advanced MSFB concept.

4.2.2 Oxides of Nitrogen

The technical approaches for achieving combustion efficiency and sulfur capture were not expected to effect a change in NO_x emissions. A method for reducing NO_x that was successfully applied in combustion was ammonia injection ^(5,6). Injecting ammonia drives NO_x species to N₂ as shown by the following reactions:

Table 4-4

TECHNICAL APPROACH - PROCESS

It was planned to achieve the NO_x emission goal by injecting ammonia after the combustion efficiency and sulfur capture goals were achieved. An unacceptable effect of ammonia injection is unreacted ammonia emitting with the flue gas. It was necessary, therefore, to measure this ammonia slip when injecting ammonia. This was done by a wet chemical sampling technique (7). When injecting ammonia, the quantity was precisely determined to prevent unreacted ammonia contaminating the flue gas.

In addition, it was necessary to protect the flue gas instruments from ammonia. Boric acid was added to the water bubbler in the flue gas sampling system to neutralize ammonia during the ammonia injection tests. The boric acid also captured sulfur dioxide. Therefore, it was not possible to monitor SO₂ emissions during the short-term ammonia injection tests.

When the pilot MSFB was reassembled for the advanced concept test program, provisions were made for injecting the ammonia in the combustor and cyclone outlet tube. The combustor operated near the optimum temperature for NO_x reduction with ammonia injection. However, there was concern that ammonia may not penetrate the solids laden combustor or some unknown ash chemistry would take place. The cyclone outlet was the best location for ammonia injection since in the MSFB process the cyclone also operated at the optimum temperature. It was also relatively lean in solids there. Unfortunately, unlike the field units, the flue gas cooler was closely coupled with the cyclone in the pilot MSFB. The gases dropped in temperature rapidly after exiting the cyclone. This did not provide enough residence time for the ammonia reaction to occur. It was not possible to separate the flue gas cooler from the cyclone due to constraints in height. However, it was thought that ammonia could be injected into the base of the cyclone outlet tube. It was long enough to provide the required residence time. Therefore, an ammonia injector was installed there.

Although currently there are no regulations for nitrous oxide (N₂O) emissions, recent measurements on various fluidized bed installations has generated tremendous interest. It is possible that in the future new regulations will be instituted for N₂O. Anticipating this development, short term monitoring of N₂O was added as a subgoal of the NO_x emission goal. An engineer from UNDEERC was brought in to measure N₂O during one week of the test program. There was concern that ammonia injection

would promote the formation of N_2O . Therefore, much of the nitrous oxide emission measurements were planned along with ammonia injection tests.

4.2.3 EHE Heat Transfer Coefficient

Table 4-5 outlines the technical approach for determining the overall EHE heat transfer coefficient. As shown in the following equation, the overall heat transfer coefficient can be expressed as follows:

$$1/U_O = 1/h_S + \Delta x/k + 1/h_i$$

where k = tube thermoconductivity (Btu/hr·ft·°F)

Δx = tube thickness (ft)

h_i = inside heat transfer coefficient (Btu/hr·sqft·°F)

h_S = solids to tube heat transfer coefficient (Btu/hr·sqft·°F)

In the case of the vertical tube EHE, the only unknown coefficient was the solids to tube heat transfer coefficient for a vertical tube immersed in a fluidized bed. Several literature sources were examined where this coefficient was measured (8). These sources were useful for bracketing the heat transfer coefficient in the sensitivity studies mentioned in Section 3. However, these literature sources were confined to bench scale experiments rather than larger pilot scale tests.

One proposed method for determining the solids side heat transfer coefficient was to immerse a probe into the pilot EHE bed. A large developmental effort would have been required to apply this probe. Another method was to outfit a single cooling tube with tube side thermocouples and in-bed thermocouples to measure the temperature field around a tube. The question of how long these thermocouples would last and the uncertainty of where to put the thermocouples seemed to lead away from this approach.

The vertical tube EHE installed in the pilot facility was a partial full scale model of a prototype EHE with similar material and tube spacings. Of course, the only difference was that fewer tubes were required for the pilot facility. Given this, the simple approach was to measure the overall heat transfer coefficient directly for the pilot EHE. This was done by using the heat transfer model discussed in Section 3 and solving for the overall heat transfer coefficient.

Table 4-5
TECHNICAL APPROACH - HARDWARE

- EHE Overall Heat Transfer Coefficient
 - Install vertical tube EHE
 - Obtain system heat balance closure
 - EHE heat duty
 - EHE temperature distribution and pressure profile

$$U_O = Q_{EHE}/(A_O (T_{EHE} - T_{tubes}))$$

First it was necessary to determine Q_{EHE} , the EHE heat duty, by measuring the cooling water flow rate and temperature rise. In actual practice, these measurements were made for each individual boiler tube. Then, the amount of boiler tube surface area covered by solids, A_O , was determined. To do this, the EHE bed level was measured. A technique successfully applied on cold flow fluidized bed models and in large scale hot fluidized beds in the field was to measure the vertical static pressure profile. In these cases, a pressure probe was immersed into the bed and the static pressure profile recorded as the probe was pulled out. The top of the bed was discerned when the static pressure equalled the freeboard pressure. Indeed, the splash zone above a fluidized bed was seen by the fluctuations in the pressure measurements.

Another measurement necessary was T_{EHE} , the EHE bed temperature. The outside tube temperature, T_{tube} , was assumed to be equal to the water exit temperature. As shown in Section 3, the vertical EHE wrapped around the combustor in the pilot facility. This meant that solids had to move laterally through the EHE. This was unlike the original side by side design where solids fall through the cooling tubes. It was felt that, in this case, a temperature gradient may have arisen in the EHE bed instead of an isothermal bed. Therefore, a provision was made to measure the gradient by installing 11 thermocouple probes in the EHE as shown in Figure 4.1. If a significant gradient was measured, then another heat transfer model could be developed to calculate the overall heat transfer coefficient.

4.3 THE TEST PLAN

The test plan, shown in Table 4-6, was made final in April of 1991. It was developed to accomplish the following objectives

- Demonstrate the Test Program Goals and Subgoals
- Characterize EHE Heat Transfer
- Examine Performance Trade-offs

The test program was not strictly followed in chronological order. Rather, it was used as a check-list of required tests for achieving the objectives. The program was split into three series.

THERMOCOUPLE PROBE

Figure 4.1 Thermocouple probes for measuring the temperature distribution in the vertical tube external heat exchanger.

Table 4-6

TEST PLAN OF APRIL, 1991 FOR LEVEL-2
OF THE ADVANCED SYSTEMS PROGRAM

<u>TEST SERIES</u>	<u>TEST DAY DESCRIPTION</u>	<u>LOAD (10⁶ Btu/h)</u>
Baseline	1 Nominal load w/o flyash reinjection.	4.4
	2 Nominal load w/ flyash reinjection.	4.4
	3 EHE heat trans. coefficient.	4.4
Demonstrate Program Goals	4 Low gas velocity/high res. time.	3.0
	5 Unstaged (1.0 d.b. stoichiometry).	*
	6 Staged (0.6 d.b. stoichiometry).	*
	7 High excess air.	*
	8 Underbed flyash reinjection.	*
	9, 10, 11 Other idea (e.g. Dolomite).	*
Parametric Tests	12 Short term ammonia inj. tests.	*
	13 & 14 Effect of excess air	*
	15 & 16 Effect of flyash reinjection ratio.	*
	17 & 18 Effect of temperature.	*
	19 & 20 Effect of dense bed press. drop.	*
	21 & 22 Effect of solids flux.	*
Combustor Height	23 Effect of Ca/S on sulfur capture.	*
	24 & 25 Remove combustor sections.	4.4
	26 High d.b. stoichiometry.	6.5
	27 & 28 Selected parameter at reduced height.	*

* denotes optimum load

A baseline series was specified where the pilot facility was operated at the previously defined nominal load condition. This condition gave the best performance in the original in-house test program concluded one year before. This was done to determine if the original performance could be duplicated considering the extensive refurbishment work and the installation of new equipment. Included in the baseline series were some short term tests to measure the EHE heat transfer coefficient. It was necessary to characterize the EHE heat transfer to assess if the number of cooling tubes was adequate to give a good cold recycle temperature. It was expected that these short term EHE heat transfer tests would be done throughout the test program.

The next series involved applying the technical approaches described above to achieve the program goals. This series was highly flexible. Tests were even allotted for unplanned ideas that may have been conceived during the test program. For example, replacing limestone with dolomite as the sorbent was considered a possibility in planning the test program.

The first approach to demonstrate the goals was to maximize the residence time by dropping below nominal load conditions. This reduced the gas velocity to 15 feet per second. This had not been previously tested in the pilot facility. It was expected that this was the minimum gas velocity possible before the solids recycle dropped out of suspension. This velocity increased the residence time to a maximum of 3 seconds.

The second approach was to apply fly ash reinjection. Extensive use of fly ash reinjection was planned for the majority of the tests. Provisions were made to inject the fly ash over the dense bed and into the dense bed by injecting into the hot recycle line. Other ideas including testing unstaged (1.0 dense bed stoichiometry) and testing at high excess air were allotted tests. However, these tests were expected to impact the performance less.

In the third series, an extensive parametric study was planned after the program goals were demonstrated to examine performance trade-offs. In this study, one parameter was varied while all other parameters were kept constant. In this way, the effect of the parameter on combustion efficiency, sulfur capture, and NO_x emission was gauged. At least three tests were planned for each variable. Short term tests of varying the Ca/S ratio to characterize SO₂ reduction was included in this series.

An additional fourth series was tentatively planned for the pilot facility. If there were time, oxidizing zone sections of the combustor would be removed to shorten the residence time. This would have required at least a three day shut down for a hired crane to remove the pieces. The purpose of this test series was to separate the effect of residence time from gas velocity on performance. In the first approach to demonstrate the program goals, the residence time can be changed by changing the gas velocity. However, it is not known whether a resulting change in performance is due to the change in residence time or the change in velocity or a combination of both. By removing combustor sections, the effect of residence time on combustion efficiency, sulfur capture, and NO_x emissions could be singled out.

4.4 THE TEST PROCEDURE

The great thing to know about operating the MSFB pilot facility is that things happen slowly. The high solids inventory in the pilot facility acts as a sink for changes in materials and flows. Based on the original in-house test program and the field units, it was known that at least 12 hours were required after a change was made before the unit reached equilibrium and that at least 4 more hours were required to establish a steady state condition.

Given this, three shifts were set-up so that the pilot facility could be operated 24 hours a day. The shifts consisted of a pilot facility operator and a technician. The people who operated the facility were the same people who worked the original in-house test program. So they knew how the pilot facility operated. Still, a training session was held two days before round the clock operation began. At this session, the program goals, test plan, and time frame were presented. The needed information and laboratory analysis were explained. The next day, the pilot facility was walked down to discuss new equipment and review existing equipment. This gave the project a sense of direction.

To determine when steady state was achieved in the pilot facility, several key variables were monitored continuously during the test program. The flue gas instruments were continuously read and recorded by computer. The SO₂ emission was particularly examined to monitor sulfur capture. However, the prime variable for determining steady state was the carbon content in the baghouse ash. Ash samples were taken every hour and sixteen hour laboratory coverage was set up to measure the carbon

content of the samples. When the carbon in the ash stopped changing, the unit was assumed to have reached steady state. Since the total ash flow was necessary for determining the combustion efficiency, it was weighed hourly and monitored. Sometimes ash was continuously removed from the EHE. Ash samples were taken once a shift from this source, and the ash flow was monitored hourly.

Fly ash reinjection was tested at various fly ash reinjection ratios. This ratio is defined as the mass of ash reinjected per mass of ash ultimately collected by the baghouse. In practice, a baseline test was established without fly ash reinjection. After the baseline, fly ash was reinjected keeping all other conditions constant. Since a portion of the baghouse ash was being reinjected, the net baghouse ash collected decreased to a new equilibrium. Therefore, the fly ash reinjection ratio was determined only after the test when the new equilibrium was established. Even though a certain fly ash reinjection ratio was targeted, the ultimate ratio was usually different than the target depending on how the mass flows shifted.

Previewing some pilot test results, a typical progression of tests is illustrated in Figure 4.2. Here, at the end of test 14, the oxidizing zone temperature was changed from 1635°F to 1715°F to start test 15. This was done by incrementally changing the setpoint in the computer that automatically controlled the L-valves. The computer decreased the cold recycle flow by decreasing L-valve control air and held cold recycle flow to maintain the new setpoint. The excess air was also changed by adjusting the air flows. Note the time scale in the figure. After establishing the new setpoint at 20:00, it wasn't until 8 AM the next morning that the carbon content in the baghouse ash and the emissions reached steady state. The test was continued four hours. This time period was important so that a composite sample of the hourly ash samples could be made. A comprehensive analysis of the composite sample was done so that a material balance could be calculated for the test. After this time period, test 16 was started by returning the temperature back to 1635°F. Note that both the change and the response were quicker for test 16.

The operator of the pilot facility had to be skilled to keep it on-line for the duration of a test. For instance, early in test 15, there was a temporary loss of temperature. The SO₂ and the CO emission spiked. This was caused by the fuel rotary valve tripping. After recovering fuel flow, the test was continued across three shifts.

Figure 4. 2. Ash Carbon Content and Emission Trends for Three Tests.

Each shift was responsible for taking a fuel and a sorbent sample. Later, samples that were closest to the 4 hour steady state period were chosen for comprehensive analysis. An ultimate analysis was done for the fuel sample and the calcium content was done for the sorbent to characterize the Ca/S ratio. Even though the computer continuously monitored the temperatures, pressures, emissions and flows, each shift was also required to take manual readings twice in a shift. This was done to ensure that the instrumentation was operating and to back-up the computer recorded data. Temperatures were recorded from digital readouts, and pressures were read from local magnahelics. Flue gas emissions instrumentation was particularly monitored since frequent blowback of the sample line was required especially during fly ash reinjection.

Periodically when the pilot facility was operating steady state, special measurements were taken. For instance, an isokinetic solids sample was pulled from the cyclone inlet for several tests to measure the solids recycle flow rate. This was done to determine the effect of solids recycle flux on performance and to measure the cyclone capture efficiency. These special measurements also included pressure probing the EHE to determine the bed level and temperature probing to measure the temperature distribution as outlined above. Oxygen concentration profiles were measured in the upper combustor to account for the air flows into the combustor.

Table 4-7 summarizes the test procedure used in the test program as outlined above. A list spelling out this procedure was the second most important document in the test program and was posted in the control room. The most important document, posted prominently in the control room, was a sheet spelling out the operating conditions and purpose of a test. A typical example is shown in Figure 4.3. This sheet conveyed critical information across the three shifts. An hour overlap between the shifts was built in so that the operator and technician of a closing shift could discuss problems and operating details with an opening shift. Communication was essential for the resulting smooth operation of the pilot facility.

Table 4-7
TEST PROCEDURE

- 24 hour/day operation
- Fuel and sorbent sample taken every 8 Hours
- Ash weighed hourly
- Carbon in the ash analyzed hourly
- Temperature, pressure, and flow continuously monitored
- Emissions continuously monitored
- Composite sample analyzed for calcium products
- Special Measurements

MSFB PILOT OPERATING CONDITIONS

TEST NO. 12

AS OF: 7/17/91

Purpose: Decrease dense bed stoichiometry from 0.8 to 0.6

*Trim secondary air make up to maintain 4.4% oxygen.

Settings:

Fuel Flow (lbs/hr)	<u>342</u>
Limestone Flow (lbs/hr)	<u>83</u>
Primary Air Flow (lgs/hr)	<u>1920</u>
Primary Air Temp. (°F)	<u>350</u>
Hot Comp. Air (lbs/hr)	<u>300</u>
EHE Air (lbs/hr)	<u>760</u>
Secondary Air Makeup (lbs/hr)	<u>860*</u>
Secondary Air Vent (lbs/hr)	<u>0</u>
*Secondary Air Flow (lbs/hr)	<u>1920</u>
Avg. Red. Zone Temp. (°F)	<u>1600</u>
Avg. Oxi. Zone Temp. (°F)	<u>1650</u>
D.B. Pressure Drop ("WC)	<u>30-40</u>
Comb. Pressure Drop ("WC)	<u>15</u>

Expected:

D.B. Velocity (fps)	<u>18</u>
Oxi. Zone Velocity (fps)	<u>20</u>
D.B. Stoichiometry	<u>0.6</u>
Oxi. Zone Stoichiometry	<u>1.2</u>
Hot Comp. Air Vel. (fps)	<u>1.0</u>
EHE Fluid Air Vel. (fps)	<u>0.4</u>

*SECONDARY AIR FLOW = HOT COMP. AIR
+ EHE AIR + SA MAKEUP - SA VENT

*TRIM SECONDARY AIR FLOW TO GIVE
OXYGEN READING OF: 4.4%

Other Instructions:

- 1) Maintain flyash reinjection at 160 pph.
- 2) Maintain EHE $\Delta P \sim 36$ "WC. Use EHE ash drain periodically.
- 3) Weigh EHE ash collection hourly
- 4) Each shift get EHE drain sample.

Figure 4.3 Typical Pilot MSFB Operating Instruction Sheet

SECTION 5

TEST PREPARATIONS AND OVERVIEW

5.1 DESIGN AND CONSTRUCTION

The design and construction phase of the Level-2 program shortly followed the original, in-house test program. In this phase, the pilot facility was disassembled and refurbished under considerable cost sharing by Riley. The following components were modified or refurbished:

- The lower combustor section
- The combustor top and several upper combustor sections
- Several cyclone sections
- Hot compartment and EHE fluidizing tubes
- The secondary air piping
- The EHE ash cooler
- The ash disposal system
- Instrumentation and the data acquisition system

Refurbishment included relining the combustor and cyclone components with new abrasion resistant refractory and regasketing all the flanged connections to reduce air leakage.

The following new components were designed and fabricated:

- The cyclone discharge into the hot compartment
- The hot compartment
- The EHE including water piping and instrumentation
- The EHE vent cooler
- The hot L-valve
- The cold L-valve

- Additional support steel and platforms

The new components were fabricated to satisfy the needs of the Level-2 program and to replace components beyond repair.

5.2 SHAKEDOWN

A fairly extensive shakedown was performed on the facility with its own set of goals.

Prior to twenty-four hour operation, it was necessary to conduct hot air testing to check out the instrumentation and calibrate the air flow meters. It was necessary to calibrate the pressure drops for the hot compartment, EHE, and dense bed air distributors with the measured air flows. This was done so that the pressure drops for each of the fluidized beds in these components could be continuously monitored by computer. The fluidized bed pressure drops indicated the solids inventory in the pilot facility. The total pressure drop through each air distributor and bed was measured by pressure transducer. The computer subtracted the air distributor pressure drop given the air flow rate and the air distributor calibration and displayed and recorded each bed pressure drop.

On Tuesday evening, June 11th, twenty-four operation commenced firing natural gas through the start-up burner. The primary purpose was for final bake-out of all new refractory. The bake-out procedure is illustrated in Figure 5.1. Temperatures were slowly raised in the combustor to bake-out the new refractory there. To bake-out the refractory in the new hot compartment, EHE, and L-valves, it was necessary to slowly bring up solids recirculation. This required great skill since the solids were hard to control at low temperatures.

During the natural gas bake-out, the oxygen concentration in the flue gas was continuously monitored. This was done to calibrate the measured air and natural gas flows with the oxygen concentration to determine the air in-leakage to the pilot facility. This calibration was incrementally done at increasing combustor pressure drop and temperature since the air leakage was known to increase with load from previous in-house test experience. The start-up burner was increased to the nominal load of 4 million Btu per hour. It was found that the air in-leakage was about 10% of the total air flow at this load. This was found to be manageable as long as it was monitored. When firing coal, this air in-leakage was accounted in each test after the carbon content in

Figure A

Run primary air heater to give a combustor temperature of 250°F for 5-6 hours. (Figure A)

Start gas burner. Raise combustor temperature 200°F/hour until 1000°F is reached.

Figure B

Bring on hot recycle. Maintain hot recycle temperatures 1000°F for 5-6 hours. (Figure B)

Figure C

Increase gas burner to raise combustor temperature at 200°F/hour to 1600°F. Increase hot recycle to bring hot recycle temperature to 1000°F. Start cold recycle. Maintain cold recycle temperatures at 200-400°F for 5-6 hours. (Figure C)

Figure 5.1 Refractory Bakeout Procedure

the ash samples were measured to determine the amount of carbon burned.

Another goal of shakedown was to measure the cyclone efficiency. It was necessary to pull an isokinetic solids sample at the cyclone inlet to get a total recycle flow rate and simultaneously measure the amount of ash collected by the baghouse to determine the amount that got through the cyclone. There were start-up problems to contend with, however. So this was not done during shakedown. However, it was immediately apparent that the cyclone was doing a good job retaining solids because little sand make-up was required. The cyclone efficiency was measured several times during the test program and the results are discussed in Section 6.

5.3 TEST OVERVIEW

On Monday, June 17th, coal burning commenced in the pilot facility burning a low sulfur, pulverized coal available on site. Full load conditions were achieved within half a day, but performance was bad as indicated by the high carbon content in the baghouse ash. This was due to the fine, pulverized coal bypassing the cyclone. On the evening of June 18th, the pulverized coal ran out, and the design, crushed coal at 3% sulfur started feeding into the combustor. Ash carbon content decreased as sulfur dioxide levels increased as expected.

The test program continued for six weeks. There was a shutdown for maintenance during the week of the Fourth of July. Otherwise, the pilot facility ran twenty-four hours a day, 5 days a week without any unplanned shutdowns. During this time, the major objectives of the test plan were addressed.

Figure 5.2 illustrates a typical operating week for the MSFB pilot facility. At midnight, Sunday night, the morning shift came in and started the pilot facility warming it up on natural gas. At 1100°F, coal and sorbent were started and gradually increased. Usually, at 6:00 am, the coal was at design flow and the gas burner shut off. The day shift took over at 8:00 am. The test conditions were established during the day. The flue gas instruments were calibrated. The ash containers were emptied and tare weights recorded. A test was considered underway when the hourly ash sample collection and the recording of ash weights started.

On Monday evening, the evening shift maintained conditions and monitored

SHIFT	HOURS	SUNDAY	MONDAY	TUESDAY	WEDNESDAY	THURSDAY	FRIDAY	SATURDAY
MORNING	23:00 – 08:00	No coverage.	Warm-up.	Maintain conditions.	Maintain conditions.	Maintain conditions.	Maintain conditions.	No coverage.
DAY	07:00 – 16:00	No coverage.	Establish test conditions. Calibrate equipment.	TEST	TEST	TEST	TEST	No coverage.
EVENING	15:00 – 24:00	No coverage.	Maintain conditions.	Establish new conditions.	Establish new conditions.	Establish new conditions.	Shutdown. Prepare for next test.	No coverage.

Figure 5.2 Typical Operating Week.

instrumentation. The morning shift did the same on Tuesday morning. At 8:00 am, the laboratory analyzed the samples collected during the morning shift in reverse of when they were collected. It was clear if the pilot reached steady state or not from these analyses. Special tests such as cyclone inlet sampling or EHE probing was conducted during the day. Typically at 16:00 on Tuesday, the first test was complete. Then the evening shift established conditions for the next test. If the next test called for fly ash reinjection, it was started.

As can be seen in Figure 5.2, this procedure continued such that four tests were accomplished by Friday evening. Then, the unit was shut-down and cooled off. The facility was readied for the next test week. The evening crew secured the facility at 12 midnight.

Table 5-1 is a check list showing what was accomplished compared to the test plan. Twenty-four tests were completed in the test program. The majority of the tests were directed toward achieving the program goals. Among the tests directed toward the program goals, some parametric tests were completed to examine performance trade-offs. To keep the overall program within budget, the test program was stopped Saturday morning, August 3rd. Changing the combustor height as outlined in the final phase of the test plan was not done.

Data analysis was performed during the test period to keep track of the progress toward the performance goals. After the test program, however, there was a two month period where extensive lab analysis was done. Outstanding fuel and sorbent samples were analyzed and composite ash analysis completed. Final data reduction of the combustion efficiency, sulfur capture, NO_x emissions was completed in this period. Heat and mass balances were computed. The end of the data analysis period was marked by a presentation at the Morgantown Energy Technology Center on October 3rd. The results of the test program are discussed in Section 6.

Table 5-1

CHECK LIST OF ACCOMPLISHED TESTS COMPARED TO TEST PLAN

<u>TEST SERIES</u>	<u>DESCRIPTION</u>	<u>TEST NO.</u>
Baseline	Nominal load w/o flyash reinjection	1,4
	Nominal load w/flyash reinjection	5
	EHE heat trans. coefficient	2,4,6,7,11,14,20
Demonstrate Program Goals	Low gas velocity/high res. time	6,7,8
	Unstaged (1.0 d.b. stoichiometry)	13
	Staged (0.6 d.b. stoichiometry)	12
	High excess air	16
	Underbed flyash reinjection	9
	Other idea (e.g. Dolomite)	19-24
	Short term ammonia inj. tests	17,18,23
Parametric Tests	Effect of excess air	1-4
	Effect of flyash reinjection ratio	18 and other tests
	Effect of temperature	10b,10c,15
	Effect of dense bed press. drop	14
	Effect of solids flux	10,10a,10d
	Effect of Ca/S on sulfur capture	9a-d,11,17 & other tests
Combustor Height	Remove combustor sections	--
	High d.b. stoic. & low res. time	--
	Selected parameter at reduced height	--

SECTION 6

TEST RESULTS

The results of testing under Level-2 of the Advanced Systems Program are tabulated in Appendix A. The tests that were performed in the MSFB pilot facility were primarily directed toward the performance goals listed in Table 6-1. All the performance goals were individually demonstrated during the six week test program. What is important, however, is the best that was demonstrated simultaneously in one test. The last day of testing was devoted to measure the best performance. Everything that was learned from the previous tests were applied to that day. The results of this test will be explained at the end of the section.

6.1 FUEL AND SORBENT ANALYSIS

Table 6-2 lists the average ultimate analysis of the fuel samples collected for each test. The fuel was a bituminous coal mined in Crooksville, Ohio. It ranged in size from 2-1/2" to 0". There was a variation in the sulfur content in the fuel as shown by the standard deviation. A definite week to week change in the sulfur was observed in the fuel analysis. This can be explained by the weekly delivery of coal to the facility. Therefore, a weekly average of the fuel analyses was used for calculating the performance of the pilot MSFB.

Table 6-3 lists the average sorbent analysis of the limestone that was used in the majority of the tests. The analysis was very consistent during the test program. As can be seen, the limestone was very calcitic. The nominal size of the limestone was 1/8".

6.2 COMBUSTION EFFICIENCY AND SULFUR CAPTURE

The first approach to achieving the combustion efficiency and sulfur capture performance goals was to reduce the load of the pilot facility below its nominal rating. This increased the residence time in the combustor by decreasing the flue gas velocity approaching a minimum of 15 feet per second. Figure 6.1 shows the effect of increased residence time on combustion efficiency and sulfur capture. These results are compared with the program goals and the predicted relationship from the original, in-house test program. As can be seen, the improvement in combustion efficiency and

Table 6-1
ADVANCED CFB PILOT-SCALE GOALS AND SUBGOALS

	<u>Goal</u>	<u>Subgoal</u>
Boiler Efficiency	90% (99.5% Combustion Efficiency)	--
Sulfur Capture	90% at 1.5 Ca/S	96%
Nitrogen Oxides	0.10 lb/10 ⁶ Btu NO _x Emission	Measure N ₂ O Emissions
Vertical tube EHE	Measure heat transfer coefficient	--

Table 6-2
AVERAGE FUEL ANALYSIS

Size: 1-1/2" x 0

	<u>Average</u>	<u>Std. Dev.</u>	<u>Range</u>
HHV, Higher Heating Value (Btu/lb)	12,284	464	11,220 - 12,930
Moisture (%)	6.1	1.0	4.5 - 7.7
Carbon (%)	68.4	2.4	63.0 - 72.1
Hydrogen (%)	5.1	0.2	4.8 - 5.4
Nitrogen (%)	1.2	0.1	0.9 - 1.3
Sulfur (%)	2.8	0.4	2.2 - 3.7
Ash (%)	6.9	1.8	4.3 - 11.5
Oxygen (%)	9.5	0.4	8.8 - 10.2

	100.0		

Table 6-3
AVERAGE SORBENT ANALYSIS

Size:	1/8" x O"
Calcium Carbonate	90.25%
Magnesium Carbonate	1.50%
Ash	3.50%
Moisture (by difference)	4.75%

Figure 6.1. Pilot MSFB combustion efficiency and sulfur capture using limestone without fly ash reinjection compared with the predicted performance and the program goals.

sulfur capture that was predicted matched fairly well with the results.

The second approach was to reinject fly ash into the combustor as collected from the baghouse of the pilot MSFB. Figure 6.2 shows the effect of the fly ash reinjection ratio on combustion efficiency and sulfur capture. As can be seen, much of the variation in combustion efficiency can be explained by the fly ash reinjection ratio. However, the predicted effect was optimistic compared to the actual. Note that the combustion efficiency goal of 99.5% was achieved as long as a fly ash reinjection ratio of 4 was maintained.

Fly ash reinjection affected the process in two ways to achieve the combustion efficiency goal. First, fly ash is relatively rich in unburned carbon since carbon can penetrate the cyclone better than inert ash. This is due to its low specific gravity compared with inert ash. Reinjecting this carbon is a straightforward means of improving the combustion efficiency.

Another reason was the shift in the mass flows that was apparent during testing. With no fly ash reinjection, all ash generated in the pilot MSFB was discarded from the baghouse. When fly ash reinjection was applied, a portion of the ash had to be discarded from the external heat exchanger (EHE). This shift was prompted by the high efficiency cyclone that was designed for the pilot MSFB. When reinjecting fly ash, a portion of it is captured by the cyclone. This increased the bed level in the EHE. To maintain a constant bed level, the EHE was drained periodically through the EHE ash cooler. Table 6-4 lists a typical analysis of the particle size distribution and the carbon content of the baghouse and EHE ash. As can be seen, the EHE ash is much lower in carbon content than the baghouse ash. On the average, the EHE ash was about 0.25% carbon as compared to 3% carbon in the baghouse ash with fly ash reinjection. Increasing fly ash reinjection increased the amount of low carbon ash discarded from the EHE resulting in improved combustion efficiency.

The cyclone efficiency actually measured in the pilot MSFB averaged at 99.6%. This efficiency was close but short of the 99.7% design value. This may explain why more fly ash reinjection was required to meet the combustion efficiency goal than expected.

As can be seen in Figure 6.2, sulfur capture also improved with fly ash reinjection. However, while the combustion efficiency goal was achieved, the sulfur capture goal

Figure 6-2. Pilot MSFB combustion efficiency and sulfur capture for limestone as a function of fly ash reInjection ratio compared with the predicted performance and the program goals. Note that the combustion efficiency goal was met at a fly ash reInjection ratio of 4.

Table 6-4

TYPICAL PARTICLE SIZE ANALYSIS OF THE EHE ASH
AND THE BAGHOUSE ASH

Size Range (microns)	E H E		A s h		Baghouse Ash	
	%	% Carbon	%	% Carbon	%	% Carbon
above 1,180	1.5	2.22				
1,180 - 600	2.3	0.28				
600 - 300	12.5	0				
300 - 150	57.5	0	5.3	0.16		
150 - 75	23.7	0	11.0	1.70		
75 - 45	2.5	0.24	20.4	3.95		
passing 45	0	0	63.3	4.89		
Total Carbon (%)		0.15		4.39		

still proved to be elusive with the limestone sorbent. It was decided, therefore, to procure a different sorbent to try and achieve the sulfur capture goal. Dolomite and a pulverized limestone were purchased for testing. Due to budget constraints, it was possible to test only the dolomite in the last week of the test program.

Dolomite is well known for its sulfur capture properties. It contains half the calcium as limestone, but is enriched in magnesium. It is believed that the magnesium in dolomite increases the porosity making dolomite more reactive. However, two times more dolomite is required than limestone to capture sulfur at the same Ca/S ratio because it contains half the calcium as limestone. Three dolomites were evaluated with respect to their performance at one of Riley's commercial MSFB installations. The best performing stone was selected for testing in the pilot MSFB. The analysis of the selected dolomite is shown in Table 6-5.

Figure 6.3 compares the performance of dolomite with limestone without fly ash reinjection. As can be seen, at 1.5 Ca/S ratio, sulfur capture improves from 60% to 80%. There is no degradation in combustion efficiency. As shown in Figure 6.4, the sulfur capture goal was reached when dolomite was used in conjunction with fly ash reinjection. The sulfur capture goal was met as long as a fly ash reinjection ratio greater than 2 was maintained. It was noted during testing that the fly ash was finer with dolomite than with limestone. The load to the baghouse was greater. Therefore, maintaining a high fly ash reinjection ratio was more difficult. This was expected because dolomite is softer and easier to break than limestone.

While many of the tests were done at a Ca/S ratio of 1.5 to judge progress toward the program goal, a range of Ca/S ratios was also tested as shown in Figure 6.5. Several results can be summarized by examining this figure. Sulfur capture was around 60% at 1.5 Ca/S ratio with the 1/8" limestone without fly ash reinjection. Decreasing the load brought the sulfur capture to 70%. Reinjecting fly ash at a ratio better than 2 at the base load also brought sulfur capture to 70%. 90% sulfur capture was achieved at 3.0 Ca/S at this condition. When using dolomite at the base load, sulfur capture close to 80% can be achieved at 1.5 Ca/S without fly ash reinjection. Dolomite in combination with fly ash reinjection brought sulfur capture to the primary goal of 90% at Ca/S, and the subgoal of 96% sulfur capture was achieved at 2.5 Ca/S at this condition.

Figure 6.5 shows an important consideration in the economics of using dolomite as a

Table 6-5
AVERAGE SORBENT ANALYSIS

Size:	1/8" x 0"	
	<u>Limestone</u>	<u>Dolomite</u>
Calcium Carbonate	90.25%	6.0%
Magnesium Carbonate	1.50%	44.0%
Ash	3.50%	37.0%
Moisture (by difference)	4.75%	13.0%
	<hr/>	<hr/>
	100.00%	100.0%

Figure 6.3. Combustion efficiency and sulfur capture for dolomite and limestone without fly ash reinjection compared to program goals.

Figure 6.4. Combustion efficiency and sulfur capture for dolomite as a function of fly ash reinjection ratio compared with limestone and the program goals. Note that the sulfur capture goal was met using dolomite in conjunction with fly ash reinjection.

Figure 6-5. Sulfur capture as a function of Ca/S ratio measured in the advanced MSFB pilot facility.

sorbent as compared to limestone. Note that 90% sulfur capture can be achieved at 1.5 Ca/S with dolomite as compared to 3.0 Ca/S with limestone. Since dolomite has half the calcium as limestone, the same amount of dolomite is needed as limestone to give the same sulfur capture. Therefore, the sorbent economics are simplified. The ash disposal costs are the same. Only the cost of the sorbent and the transportation cost need to be considered in evaluating costs.

Figure 6.6 shows the calcium utilization for these tests against Ca/S ratio. Calcium utilization is the measured sulfur capture divided by the required Ca/S ratio. This figure shows the measured calcium utilization versus the maximum possible utilization, which is 100% at 1.0 Ca/S. The sulfur capture goal calls for 60% utilization at 1.5 Ca/S as compared to the maximum possible 67% calcium utilization. Dolomite in combination with fly ash reinjection met the goal.

To summarize, Table 6-6 shows the best combustion efficiency and sulfur capture that was individually achieved in the test program. The conditions that were applied to achieve the goals are also listed. During several of the tests discussed so far, short term ammonia injection tests were completed. Nitrous oxide measurements were also made. These results are discussed next.

6.3 OXIDES OF NITROGEN

In the fifth week of testing, nitrous oxide (N_2O) measurements were made by a gas chromatograph as operated by an engineer from the University of North Dakota Energy and Environmental Research Center. Data from the gas chromatograph was independently acquired by the pilot facility data acquisition computer. Figure 6.7 shows NO and N_2O emissions versus combustor temperature in the pilot MSFB. As expected, NO emissions increased and N_2O emissions decreased as combustor temperature was increased. Note that N_2O is similar in magnitude as NO. Also note how the sum of NO and N_2O emissions decreased with combustor temperature. That means that temperature should be maximized to minimize all oxides of nitrogen. Of course, since sulfur capture decreases with temperature, the trade-off between nitrogen oxides and sulfur capture with temperature must be considered.

Much of the nitrous oxide emission measurements were made along with ammonia injection testing. The first couple of times ammonia injection was tried, there was no

Figure 6.6. Calcium utilization versus Ca/S ratio compared with the maximum utilization possible.

Table 6-6
BEST INDIVIDUAL PERFORMANCE

	<u>Best Value</u>	<u>Test No.</u>	<u>Conditions</u>
Combustion Efficiency	99.5%	5	99.6% Cyclone Efficiency Fly Ash Reinjection Ratio = 5
Sulfur Capture	96% at 2.3 Ca/S	23	Dolomite Fly Ash Reinjection Ratio = 2

Figure 6.7. Nitric oxide (NO) and nitrous oxide (N₂O) emissions from the pilot MSFB.

reduction in NO injecting either into the combustor or the cyclone outlet. Nitrous oxide went from 85 to 93-99 PPM and some ammonia gas was detected in the flue gas sample. At the time, undiluted ammonia was injected through a single nozzle discharger. It was surmised that the ammonia was decomposing in the nozzle tip following a path to N_2 . So a distributor was installed through a port in the combustor and the ammonia was injected with compressed air. Typically 15 SCFM of compressed carrier air was used to inject the ammonia. When ammonia injection was tried with the new distributor, NO decreased substantially. Unfortunately, there was no means to install a similar distributor in the cyclone outlet tube without a long term shutdown of the pilot facility. Once an effective means of injecting the ammonia was found, short-term ammonia injection testing was done into the combustor at several elevations and several operating conditions. The ammonia to NO ratio was varied for these tests and the ammonia slip was measured.

Figure 6.8 and 6.9 show the reduction of NO by ammonia injection at two excess air levels. The measurement of ammonia slip is also shown. At 35% excess air, N_2O increased about 10% at a better than 4.5 NH_3/NO mole ratio. This is contrasted with the approximately 50% reduction in NO. At 25% excess air, 75% NO reduction was achieved at 4.5 NH_3/NO ratio. Unfortunately, this test did not take place when nitrous oxide was being measured. Note that at 25% excess air, the goal of 0.10 pounds of NO per million Btu was achieved at a NH_3/NO ratio of 3. The ammonia slip was approaching 30 PPM at this condition.

Figure 6.10 shows the effect of ammonia injection at three temperatures on a percent reduction scale. NO reduction was found to be more effective at the two higher temperatures. These two temperatures fall into the temperature window for NO reduction as determined in other studies ^(5,6). Ammonia injection was tried at 24 feet and 8 feet below the cyclone inlet, but there was little difference in the results. There was also little difference with or without fly ash reinjection.

A chemical balance shows that only 30% of the injected ammonia can be accounted as being converted to NO, N_2O , and NH_3 slip. The rest may be reacting with solids, dropping out as ammonia sulfate $[NH_4]_2SO_4$, or decomposing to N_2 in the ammonia injection system. It was felt that injecting ammonia at the cyclone outlet was a more ideal location. Because of the inability to inject ammonia after the cyclone, the NO reductions measured were probably not the best that can be achieved in the advanced

Figure 6.8. Ammonia injection tests in the pilot MSFB at 25% excess air.

Figure 6.9. Ammonia injection tests in the pilot MSFB at 35% excess air.

Figure 6-10. Percent reduction of NO versus NH₃/NO mole ratio as a function of temperature.

MSFB system.

Appendix B summarizes the nitrous oxide measurements and the results of injecting ammonia in the pilot MSFB.

6.4 PERFORMANCE TRADE-OFFS

An extensive parametric test matrix was planned following demonstrating the program goals. These tests involved changing one process parameter while holding other parameters constant. In this way, the effect of the parameter on performance could be gauged, and the trade-offs among the performance parameters could be evaluated. The program goal effort required more tests than planned. The parametric test matrix was not carried out due to budget constraints. However, during the program goal effort, some tests were done comparing one process parameter with another. Other parameters, however, were not necessarily held constant in these tests.

Figure 6.11, for instance, shows the effect of excess air on combustion efficiency, sulfur capture, NO_x and CO emissions for tests without fly ash reinjection and operating at the same temperature and Ca/S mole ratio. This figure shows how keeping excess air above 25% can help sulfur capture, but at the expense of NO_x . The effect on combustion efficiency and CO is small. This result is consistent with what was found in the original test program and the 25% excess air recommendation for the City of Wyandotte design.

Figure 6.12 shows the effect of temperature on the performance. This figure illustrates the trade-off that must be made between combustion efficiency and CO versus sulfur capture and NO_x when recommending a combustor temperature. It should be noted that the excess air was 20% for the test at 1620°F and 30% for the test at 1720°F. The change in excess air had little effect on combustion efficiency based on the excess air comparisons. However, it is surmized that excess air had a large effect on sulfur capture and NO_x . The slope of the sulfur capture line would have been greater and the slope of the NO_x line would have been less if excess air were held constant.

In Table 6-7, the baseline process parameters in the pilot MSFB are compared with the variations that were tested during the program goal effort. These variations include the dense bed stoichiometry, dense bed pressure drop, and combustor pressure drop.

Figure 6.11. Parametric Comparisons - The Effect of Excess Air

Figure 6.12. Parametric Comparisons - The Effect of Temperature

Table 6-7
NOMINAL LOAD CONDITIONS AND VARIATIONS

	<u>Nominal Conditions</u>	<u>Variations</u>
Load, (Million Btu/hr)	4.4	(3.1, 4.4)
SR _{db} , Dense Bed Stoichiometry	0.8	(0.6, 0.8, 1.0)
XSA, Excess Air (%)	25	(15 - 50)
T _{ox} , Oxidizing Zone Temperature (°F)	1640	(1540, 1640, 1720)
V _{ox} , Oxidizing Zone Gas Velocity (fps)	22	(17, 22)
τ_{ox} , Oxidizing Zone Residence Time (sec)	2.2	(2.2, 2.8)
ΔP_{db} , Dense Bed Pressure Drop ("WC)	30	(20, 30, 40)
ΔP_{cmb} , Combustor Pressure Drop ("WC)	15	(10, 15, 20)
G _S , Combustor Solids Flux (lb/sq ft/hr)	10,000	(5,000 - 15,000)
Ca/S Ratio	1.5	(1.0 - 3.0)
FAR, Fly Ash Reinjection Ratio	0	(0 - 4.0)
Sorbent	Limestone	(Limestone, Dolomite)
NH ₃ /NO, Ammonia Injection Ratio	0	(0 - 8)

These variations, however, showed little effect on the combustion efficiency, sulfur capture, and emissions within the ranges tested. Overbed fly ash reinjection was compared to in-bed fly ash reinjection, but the results were not much different. These comparisons coupled with what is known from the original in-house development program helped in the design of the commercial, proof-of-concept system.

6.5 EHE HEAT TRANSFER

While the pilot facility was operating at a steady state condition, pressure and temperature measurements were periodically made in the vertical tube external heat exchanger. As discussed previously, these measurements were necessary for determining the EHE heat transfer coefficient. In turn, a heat transfer coefficient for a vertical tube EHE was necessary for designing the proof-of-concept system.

Figures 6.13 and 6.14 show the EHE temperature and pressure measurements along with the heat transfer calculations for two loads. These two figures illustrate how the EHE was used to handle changes in load. When testing at high load, the EHE bed level was raised by filling the EHE with sand at the start of the test. The surface area covered by the bed was high and therefore, the heat duty of the EHE was high. When going to low loads, the EHE bed level was lowered by draining solids through an ash cooler. The covered tube area decreased. This decreased the EHE heat duty to handle the lower load.

As can be seen in the temperature distribution, there was a small temperature gradient from the EHE inlet to the L-valve drain. In calculating the EHE heat transfer coefficient, however, it was assumed that the bed was isothermal since the standard deviation among the temperature measurements was within 150°F. Note that a heat balance around the pilot MSFB was closed to within 99%. This gives high confidence that the heat transfer of the EHE was successfully characterized. Table 6-8 summarizes the heat transfer coefficient for each test where it was measured. The tests span several loads and operating conditions, but the heat transfer coefficient did not vary much. The average coefficient for the tests was 50 Btu/hr sqft °F.

6.6 MASS, HEAT AND SULFUR BALANCES

Heat and material balances were made to assure the integrity of the results. The material balance was very important since the combustion efficiency calculation

$$\text{LOAD} = 4.4 \times 10^6 \text{ Btu/hr}$$

$$\text{HEAT BAL CLOSURE} = 99 \%$$

$$Q_{E-E} = 1.50 \times 10^6 \text{ Btu/hr}$$

$$A_O = 6126 \text{ sq. inches}$$

$$\text{Avg. } T_{E-E} = 940 \text{ } ^\circ\text{F}$$

$$T_W = 150 \text{ } ^\circ\text{F}$$

$$U_O = \frac{Q_{E-E}}{A_O (T_{E-E} - T_W)}$$

$$= 45 \text{ Btu/hr sqft } ^\circ\text{F}$$

Figure 6.13. Pressure and temperature profiles and the calculation of the overall heat transfer coefficient at nominal load.

$$\text{LOAD} = 3.2 \times 10^6 \text{ Btu/hr}$$

$$\text{HEAT BAL. CLOSURE} = 99 \%$$

$$Q_{E-E} = 1.26 \times 10^6 \text{ Btu/hr}$$

$$A_o = 3864 \text{ sq. inches}$$

$$\text{Avg. } T_{E-E} = 980^\circ \text{F}$$

$$T_w = 140^\circ \text{F}$$

$$U_o = \frac{Q_{E-E}}{A_o (T_{E-E} - T_w)}$$

$$= 55 \text{ Btu/hr sqft } ^\circ \text{F}$$

Figure 6.14. Pressure and temperature profiles and the calculation of the overall heat transfer coefficient at low load.

Table 6-8
SUMMARY OF HEAT TRANSFER COEFFICIENTS

<u>Test No.</u>	<u>U_o (Btu/hr/sq ft/°F)</u>
2	45
4	56
6	41
7	55
11	52
14	47
20	48

Average = 50 Btu/hr/sq ft/°F

strongly depended upon the ash outflow. Heat balances were made for several tests between heat generated by the fuel and input by the air heaters and what was picked up by the water circuits and lost from the system. As mentioned before, these heat balances were closed to within 99%.

Two methods were used to balance the material flows. In the mass balance method, the ash generated from burning the fuel and utilizing the sorbent was compared with what was collected from the baghouse and drained from the EHE. In the calcium balance method, the measured concentration of calcium in the baghouse ash and the EHE drain ash was compared with what was expected from the amount of calcium fed with the sorbent.

The mass balance assumes that the solids inventory in the system was constant during a test. Solids inventory was maintained by monitoring the pressure drop in the EHE and the dense bed. Ash was drained or make-up sand added to the EHE to maintain pressure drops. Dense bed make-up was not required since alumina beads were used. As compared to the commercial, prototype unit, the solids inventory in the pilot MSFB was large with respect to the mass flows. This was due to the fact that while the mass flows were approximately 1/35th scale of the prototype, the refractory thickness and the tube sizes were full scale making the hot compartment and EHE larger in geometric scale. Therefore, it was difficult to demonstrate a mass balance because a small change in solids inventory gave a significant difference between the calculated ash generated and the measured outflows.

In the calcium balance method, the ash flows were determined by leveraging the smaller sorbent flow by the calcium concentration in the larger ash flow. Imprecise calcium concentrations can give errors in the estimated flow. When the ash is removed from both the baghouse and the EHE, then a ash flow split was required. Usually, the measured ash flow split between the baghouse and the EHE was used in this case to complete the calcium balance.

The mass balance and the calcium balance calculations were compared to the measured flows. In tests 11, 14, 21, and 23, the measured ash flow did not compare at all because the measured flow was not well established. These tests were short term or the baghouse dumped ash inconsistently. In these cases, the mass balance ash flow was used to calculate combustion efficiency.

Figure 6.15 compares the calculated total ash flow by mass balance with the measured total ash flow. As can be seen, the calculated mass flow averages 20% less than the measured ash flow. As discussed before, the solids inventory was probably not constant during the tests. The initial solids charge to the EHE was 5000 pounds. A calculated deficit of 20 lbs/hr from a measured 100 lbs/hr amounts to a 0.4% change in inventory per hour. This change in inventory could only be detected across an extremely long period.

Figure 6.16 compares the ash flow calculated from a calcium balance with the measured ash flow. The calculated flows agreed with the measured flows for about half the tests. The calcium concentrations measured for the tests where there was greater than 20% disagreement may not have been representative.

After considering the mass and calcium balances, it was decided that the measured ash flows be used for calculating combustion efficiency except for those cases where the measured ash was not well established. However, these balances were also considered in calculating the sulfur capture based on the ash analysis. The sulfur capture based on ash analysis was compared to what was calculated from SO₂ emissions measured by the flue gas instrument. Figure 6.17 shows the comparison. As can be seen, half the tests showed agreement within 20%. This is fairly good agreement considering that the ash analysis method is accounting for a large majority of the sulfur (90% if the sulfur capture goal is realized). It is easier to measure the smaller amount of sulfur exiting with the flue gas. In the end, the sulfur capture results were based on the flue gas analysis rather than the ash analysis. The derivation of sulfur capture from the ash analysis was considered only to judge the quality of individual tests.

6.7 PILOT FACILITY RESPONSE AND OPERATION

Some interesting notes on how the pilot facility responded to certain changes were made during the test program.

There was a marked difference between nominal load (4.4 million Btu/hr) operation and low load (3.1 million Btu/hr) operation. The unit performance changed very slowly with a change in a process parameter at low load. For instance, SO₂ emissions took at least 4 hours to decrease when Ca/S ratio was increased. Changes were seen within

Figure 6 .15. The total ash generated by the Advanced Pilot MSFB as calculated by mass balance versus the measured ash. Mass balance based on weekly fuel analysis. Numbers refer to test numbers.

Figure 6.16. Calculated total ash flow by calcium balance versus measured ash flow.
Calcium balance based on weekly fuel analysis. Numbers refer to test numbers.

Figure 6 .17. Calculated sulfur capture from ash analysis versus sulfur capture based on flue gas analysis. Numbers refer to test numbers.

an hour for the same change at nominal load operation. The sluggish response at low load will not translate to the prototype because of the difference in scale between the solids inventory and mass flows in the pilot facility versus the prototype. Also it was hard to control excess air at low load due to increased uncertainty in air flow measurements. The pilot MSFB was operated mostly at nominal load during the test program for these reasons.

Fly ash was first injected above the dense bed. When fly ash was reinjected at this location, the dense bed temperature decreased about 300°F. Since the cold recycle L-valve was tuned to control the upper combustor temperature in the pilot MSFB, this response was caused by an increase in cold recycle flow to control the upper combustor temperature. This meant that there was increased combustion in the upper combustor, and the dense bed was quenched with the increased cold recycle. In an attempt to change this, the fly ash was injected into the hot L-valve to accomplish in-bed fly ash reinjection. However, the dense bed quenching did not change. Apparently, the fly ash did not mix well with the dense bed. As mentioned previously, there was no difference in combustion efficiency or sulfur capture between overbed or in-bed fly ash reinjection. The combustion efficiency and sulfur capture was much improved with fly ash reinjection than without. However, there was not time to explore other injection methods such as multiple injection ports that may have minimized dense bed quenching and further improved performance.

As was mentioned before, dolomite gave a much finer ash when it was used as a sorbent than with limestone. This was attributed to the increased breakage of the dolomite. This in turn improved its reactivity. Draining from the EHE was not required using dolomite with fly ash reinjection. Combustion efficiency could have been improved if a higher fly ash reinjection ratio could be maintained to require draining low carbon ash from the EHE. This again suggests the use of multiple fly ash reinjectors to increase fly ash reinjection capacity.

6.8 THE BEST SINGLE TEST

The last day of operation was devoted to demonstrating the best performance in the pilot MSFB. The shift schedule was extended beyond a normal shutdown to maintain the condition. Previous to the test, three tests (tests 21, 22, and 23) were completed using dolomite in conjunction with fly ash reinjection. In these tests the Ca/S ratio was

varied from 1.0, 1.5, to 2.3 respectively. In test 22, the sulfur capture goal of 90% at 1.5 Ca/S ratio was closely matched. In the final test, the Ca/S ratio was returned to 1.5, the fly ash reinjection was maximized, and ammonia injection was applied to attempt to achieve all the goals at once.

Table 6-9 lists the results of this test compared to the program goals. As can be seen, the results almost matched the performance goals. Also shown in Table 6-9 are the conditions which gave the best performance in the pilot MSFB. Equally important to the combustion efficiency was the 99.6% efficient cyclone and the use of fly ash reinjection. The sulfur capture goal was achieved with fly ash reinjection in combination with using dolomite as a sorbent. As can be seen, ammonia injection at 2 NH₃/NO ratio closely matched the NO_x emission goal. This goal could easily be exceeded by increasing the NH₃/NO ratio, but the ammonia slip may be excessive.

Table 6-10 lists the design parameters recommended for the advanced proof-of-concept facility. These recommendations are based on the current test results, the results of the original test program, and what is known about the performance of the field units. It should be noted that while the nominal load gas velocity is recommended for the advanced design, the higher residence time that was approached during low load testing is recommended to provide some margin in the performance.

The fact that the Level-2 goals were closely demonstrated suggests that the technical approaches were sound. It was strongly felt that the combustion efficiency goal could have been met in the final test if the fly ash reinjection ratio was higher. Since the fly ash was finer using dolomite, a higher reinjection flow rate was required to maintain the same fly ash reinjection ratio as achieved using limestone. Extensive wear was found in the eductor used to transport fly ash when the test program was finished. This wear, caused by extensive use of fly ash reinjection system, also limited the reinjection flow rate. Using multiple injectors to increase the reinjection capacity and to better mix the fly ash into the dense bed has a strong potential for improving performance even beyond what was obtained in the test program.

Table 6-9

GOALS COMPARED WITH BEST SINGLE TEST
IN THE ADVANCED PILOT MSFB

	<u>Program Goal</u>	<u>Best Test</u>	<u>Conditions</u>
Combustion Efficiency (%)	99.5	99.2	99.6 Cyclone Efficiency
Ca/S Ratio for 90% Sulfur Capture	1.5	1.5	Fly Ash Reinjection Ratio at 2 Dolomite
NO _x Emission (lb/10 ⁶ Btu)	0.10	0.12	Ammonia Injection at 2 NH ₃ /NO

Vertical Tube EHE Heat Transfer Coefficient = 50 Btu/hr·sqft·°F

Table 6-10
RECOMMENDED ADVANCED MSFB DESIGN

SR_{db} , Dense Bed Stoichiometry	0.7 - 0.8
XSA, Excess Air (%)	25 - 30
T_{ox} , Oxidizing Zone Temperature (°F)	1620 - 1640
V_{ox} , Oxidizing Zone Velocity (fps)	20
τ_{ox} , Oxidizing Zone Residence Time (sec)	3.0
G_S , Combustor Solids Flux (lb/sq ft/hr)	8,000 - 9,000
η_{cyc} , Cyclone Efficiency (%)	99.7
FAR, Fly Ash Reinjection Ratio	3
Sorbent	Dolomite
NH_3/NO , Ammonia Injection Ratio	2
Vertical Tube EHE Heat Transfer Coefficient (Btu/hr/sq ft/°F)	50

SECTION 7

ADVANCED DESIGN AND ECONOMICS UPDATE

During the design and engineering phase of Level-2, some hardware changes were incorporated into the original design developed earlier in Level-1. This was intended to reduce any uncertainties introduced in the design due to an unproven equipment. It was particularly true of the loop seal design for solids recirculation which was under development in collaboration with Jenike and Johanson. A conventional design was employed in its place for the pilot plant as well as the proof of concept design. It should be noted, however, that the loop seal concept has not been abandoned. It has just been taken out to enhance the current design reliability. It is possible that, after further evaluations a loop seal can be added to enhance future MSFB designs.

The design developed for the proof of concept facility in Level-1 was re-evaluated to determine the impact of the Level-2 test program findings on the design. This evaluation resulted in the revision of some process design parameters as discussed in Section 6 and listed in Table 7-1. These process and hardware modifications in the design are expected to improve system performance and reliability.

Revised economic data on the steam costs and pay back periods was recalculated to reflect these changes.

It should be noted, however, that during this review the major goals of the program related to combustion and emission performance were kept in focus all along.

7.1 SOLIDS RECYCLE SYSTEM

The original proof of concept facility design proposed in the Level-1 topical report (9) employed a loop seal arrangement for solids recycle. This was done to alleviate hot solids flow problems that may be tied to the L-valve design. Riley's current CFB facilities are equipped with L-valves for solids recirculation. All these problems cannot be solely attributed to the recycle flow system design but to a combination of factors with both process and mechanical origins, listed as follows:

- Poor combustion in the combustor

Table 7-1
ADVANCED CONCEPT PARAMETERS

	<u>Recommended</u>	<u>Original</u>
Dense Bed Stoichiometry	0.6 - 0.7	0.8
Excess Air (%)	25-30	10
Oxidizing Zone Temperature (°F)	1620-1640	1600
Oxidizing Zone Velocity (fps)	20	20 - 25
Oxidizing Zone Residence Time (sec)	3.0	1.5 - 2.0
Combustor Solids Flux (lb/hr-sqft)	8-9,000	10 - 12,000
Cyclone Efficiency (%)	99.7	99.6
Flyash Reinjection Ratio	3.0	-
Sorbent:	Dolomite	Limestone
NH ₃ /NO, Ammonia Injection Ratio	2	-
Vertical Tube EHE Heat Transfer Coefficient (Btu/hr-sqft-°F)	50	65

- Burning in the L-valves
- Improper mixing
- Temperature influenced solids flow characteristics

These factors have resulted in the formation of agglomerates in the combustor and the L-valves. Modulating the process conditions and hydrodynamics in the combustor has resulted in improved operation.

In the current Riley design all of the cyclone solids are taken to a dual compartment EHE. The first compartment is a holding zone for the solids with an L-valve at the bottom. Based on the overall heat and mass balance a portion of the solids are diverted through this L-valve to the combustor. The balance of the solids cascade over a partition wall to the second compartment equipped with heating surface. The solids are cooled and flow through another L-valve back into the combustor. The split of solids between the two compartments is modulated to achieve proper process conditions in the combustor.

The operating experience on this design has identified a number of areas which have not performed satisfactorily. One of these is the partition wall refractory which has failed at many locations affecting the plant reliability. An alternative is to separate the hot and cold compartments eliminating the need of a partition wall. This approach led us to a loop seal arrangement concept. A preliminary design of this loop seal was presented in the Level-1 topical report ⁽⁹⁾.

Under this new arrangement, solids leaving the cyclone are taken to a loop seal where they are split into two streams, one going direct to the combustor and the other to the EHE for cooling. This would require development of a loop seal with the capability to split, seal and control the solids flow. It should also be able to operate smoothly under a variety of conditions. In collaboration with Jenike and Johanson, a company well known for its expertise in solids flow technology, a short program was initiated by Riley during the engineering phase of Level-2.

In this test program, flow characteristics of typical FBC solids under representative environment were tested. A full scale loop seal model for Riley's pilot plant was constructed and operated by Jenike and Johanson at its test facilities located in San Luis Obispo, California. The tests were carried out under cold and hot conditions

representative of the temperatures and pressure differentials in a CFB system. Based on these tests, a design was proposed for the Level-2 pilot program. It is shown in Figure 7.1. As seen in this figure the plant is equipped with two different loop seals. The first one is a splitter loop seal which diverts solids in controlled quantities to the combustor and the EHE. The second loop seal is located under the EHE which is used to transport the cooled solids from the EHE to the combustor. The control of the solids stream is achieved by injecting air at different strategic locations.

The test report from Jenike and Johanson on this program is presented in Appendix B.

Although the results of this testing were promising, it was decided to proceed with an alternate design for Level-2. The primary reason for this was to keep focus on the goals of Level-2 which placed more emphasis on demonstrating the process features of the advanced concept. If a decision was made to go with the new loop seal design, it would be mandatory to first have the recycle system operate efficiently. This would mean that hardware development was taking precedence over process demonstration, a deviation from the original goal of the program. And so, a modified arrangement of EHE and L-valves was selected for the solids recycle as described in Section 3. Because this design operated satisfactorily on the pilot plant it was also adopted for the proof of concept facility.

7.2 PROCESS DESIGN PARAMETERS

The Level-2 test program provided the background for the selection of design parameters for the proof of concept facility. Table 7-1 shows the original and recommended process parameters. In the original design, an excess air of only 10% was suggested. The pilot data showed that to achieve proper combustion and sulfur capture goals it was necessary to go higher levels of excess air, preferably 25%. For the commercial plant, this certainly will increase the air and flue gas handled by the FD and ID fans. It also will affect the overall boiler efficiency. But there are trade offs that need to be examined. A smooth operation with full emissions compliance will definitely be better than trying to attain high efficiency goals with unreliable operation. Customers can tolerate a lower boiler efficiency but an unreliable plant has financial ramifications that go beyond the boiler itself. Non-availability of a plant at critical time periods can be devastating for the customer.

Another process parameter related to the boiler efficiency is the stack temperature. In

Figure 7.1 Proposed Loop Seal Arrangement

the original design presented in Level-1, a stack temperature of 200°F was recommended. This results in the selection of expensive air heater materials due to dew point considerations. This adds to the capital costs of the plant. The value in going to this low a stack temperature is questionable. It certainly improves the stack gas losses but again a customer is open to compromise on the efficiency. Improved operation is what he demands. There are areas that can be improved that would interest him more like reduced NO_x levels and better sorbent utilization. These items have already been addressed in the discussion of test results in Section 6. The proposed design parameters have incorporated these findings. The revised efficiency calculations performed for this report, however, are still based on a stack temperature of 200°F. It is recommended that a higher temperature of 250°F be considered at the time of the final design of the proof of concept facility. It will result in the lowering of the boiler efficiency by ~0.5% and may eliminate the need of special materials of construction in the air heaters.

Other modifications/changes have been included into the advanced design as a result of the Level-2 test program. One of them is the fly ash reinjection system. Recycling a portion of the baghouse solids back into the combustor results in significant improvement in combustion efficiency and sulfur capture. It is recommended that four pounds of ash be recirculated for every pound of disposed ash. As a result of the larger percentage of fines in the system, the bag house may grow in size. Again, there is trade off between the capital costs and performance. One can selectively choose the equipment which makes the overall plant performance economically attractive.

In the design developed for the proof of concept facility, emphasis has been put on achieving high combustion and sulfur capture efficiencies in the combustor. The recommended oxidizing zone temperature in the combustor is 1640°F at 25% excess air. It should be noted, however, that temperature and oxygen levels that encourage high carbon burn up and efficient sulfur capture may lead to poor NO_x control. To counter this, provision has been made to have a post NO_x control by NH₃ injection. The pilot test results have determined that a mole ratio of two moles NH₃ per mole of NO is a reasonable amount in order to attain acceptable NO_x levels.

An important goal of this program was to reach 90% sulfur capture with a Ca/S mole ratio of 1.5. As discussed in Section 6, it was possible only with dolomite (44% Ca), which demonstrated better Ca utilization than limestone (90%Ca). Similar sulfur

capture was achieved for limestone at a Ca/S mole ratio of 3.0 which is double the mole ratio needed for the dolomite. But, the smaller calcium content in dolomite negates its advantage in the lower Ca/S mole ratio requirement. In the case of limestone, on the other hand, the poorer calcium utilization and higher Ca/S mole ratio is compensated by its higher calcium concentration. Therefore, for 90% sulfur capture, the total sorbent needed is about the same whether we use limestone or dolomite. There is not much difference in spent ash quantities produced for the two sorbents either. As a result, the selection of the sorbent will be dependent on its delivered cost. For example, an efficient dolomite, delivered inexpensively can be more attractive than a costly high calcium limestone. These site specific factors contribute towards establishing the economics of operating this plant.

7.3 HEAT & MASS BALANCE

Based on the above discussion and using the original steam conditions, shown in Table 7-2, a heat and mass balance has been prepared. The coal and dolomite analyses used for these calculations are shown in Table 7-3. The combustion calculations are summarized in Table 7-4 which shows the estimated boiler losses. As can be seen from this table, even with the revised design parameters, the efficiency remains above the targeted 90% level. Figure 7.2 and Table 7-5 summarize the new heat and mass balance.

The physical design proposed for Level-1 has also been revised to reflect the new numbers. The combustor sizing, solids recirculation system design and the heating surfaces have been recalculated. These changes have been incorporated into a revised arrangement drawing. It is shown in Figures 7.3 and 7.4. The new heat distribution and the physical dimensions of the pressure parts are presented in Table 7-6.

The combustor dimensions in the oxidizing zone are 12'1"x 8'8". To have 3.0 seconds of residence time with 20 ft/sec gas velocity, the height will be 56'. Approximately 0.5 second of residence time is available in the reducing zone with a velocity of 25 ft/sec. The total height of the combustor is 66'. The boiler bank reduces the temperature of the gases from 1640°F to 954°F. This permits construction of the economizer surface out of carbon steel. The temperature leaving the economizer is 281°F. The reduction of the gas temperature from 281° to ~200°F is where special material requirements

Table 7-2
ADVANCED CONCEPT STEAM CONDITIONS

Main Steam Flow	pph	110,000
Steam Outlet Pressure	psia	214.7
Drum Pressure	psia	214.7
Feedwater Pressure	psia	245.7
Feedwater Temperature	°F	100
Sat Temp @ Drum Press	°F	393
Outlet Steam Temperature	°F	393
Feedwater Enthalpy	Btu/lb	68.6
Sat Steam Enthalpy	Btu/lb	1200.1
Sat Water Enthalpy	Btu/lb	368.1
Heat of Evap	MBtu/hr	124.5
Total Output	MBtu/hr	124.5

Table 7-3

FUEL AND SORBENT ANALYSES

COAL ANALYSIS

	<u>Wt. %</u> (as fired)
Carbon	61.10
Hydrogen	4.60
Oxygen	9.40
Sulfur	3.30
Nitrogen	1.10
Moisture	9.40
Ash	11.10
Higher Heating Value: Btu/lb	11,075

DOLOMITE ANALYSIS

	<u>Wt. %</u> (as fired)
CaCO ₃	44.0
MgCO ₃	37.0
Inerts	13.0
Moisture	6.0

Table 7-4
ESTIMATED BOILER LOSSES

Dry Gas Loss	2.65 %
Moisture in Fuel Loss	0.93
Moisture from Combustion of Hydrogen Loss	4.06
Moisture in Air Loss	0.07
Sorbent Moisture Loss	0.07
Calcination Loss	1.65
Sensible Heat Loss From Solids	0.11
Unburnt Carbon Loss	0.40
Radiation Loss	2.00
Sulfation gain	-1.80
Fan/Pump Horse Power Gain	-0.73
 Total Losses	 9.40
 Boiler Efficiency	 90.60 %

Figure 7.2 Advanced System Process Flow Schematic

Table 7-5
ADVANCED CONCEPT STREAM IDENTIFICATION

Stream #	Description	Temp. (°F)	Press. (in WG)	Gas Flow (lb/hr)	Solids Flow (lb/hr)
1.	Coal	80			12,500
2.	Dolomite	80			4,396
3.	FD Fan Discharge	90	18		131,200
4.	Air Leaving Air Heater #2	133	17.5		131,200
5.	Air Leaving Air Heater #1	185	17		131,200
6.	EHE/Hot Compartment Air	200	68		3,140
7.	Primary Air	200	68		83,970
8.	Secondary Air	185	16.5		44,090
9.	EHE/Hot Compartment Vent Air	1100	0		3,140
10.	Flue Gases Leaving Combustor	1640	-20		144,000
11.	Solids Leaving Cyclone Bottoms	1600			940,940
12.	Flue Gases Leaving Cyclone	1640	-24	144,000	4,060
13.	Flue Gases Leaving Boiler Bank	954	-26	144,000	
14.	Flue Gas Leaving Economizer	281	-27	144,000	
15.	Flue Gases Leaving Baghouse	281	-28	144,000	
16.	Solids Leaving Baghouse	281			4,060
17.	Flue Gases Leaving Air Heater #1	238	-30	144,000	
18.	Flue Gases Leaving Air Heater #2	202	+3	144,000	
19.	Flue Gases to Stack	200	+1	144,000	
20.	Flyash Reinjection	150			3,248
21.	Flyash Discharge	150			812

SECTIONAL R.H. SIDE VIEW

Figure 7.3 Advanced System Arrangement - Side View

Figure 7.4 Advanced System Arrangement - Front View

Table 7-6

ADVANCED CONCEPT PHYSICAL DATA

HEAT DISTRIBUTION

Combustor	43.20x10 ⁶ Btu/hr
Boiler Bank	25.02x10 ⁶
EHE	28.72x10 ⁶
Economizer	27.56x10 ⁶
Total	124.5x10 ⁶ Btu/hr

BOILER BANK

Drum Centers	15'
Upper Drum	48" Dia
Lower Drum	48" Dia
Tube Dia	2"
Transverse Spacing	4"
Lateral Spacing	4"
Total Heating Surface	3,141 ft. ²

ECONOMIZER

Tube OD	2.5"
Transverse Spacing	4"
Lateral Spacing	4"
Total Heating Surface	8,346 ft ²

EHE HEATING SURFACE

Tube OD	2.5"
Transverse Spacing	4"
Lateral Spacing	4"
Total Heating Surface	754 ft ²

COMBUSTOR WALLS

Tube Dia	2.5"
Combustor Cross Section	12'1 x 8'8"
Total Height	66 ft.
Water Wall Surface (exposed)	2300 ft ²

are introduced. Those sections of the air heater where the temperature is lower than 250°F are glass lined for protection against acid condensation.

To keep erosion due to fine solids entrained in the flue gas to a minimum, the convective section geometry is selected to keep the superficial gas velocity under 50 ft/sec.

7.4 REVISED ECONOMICS

The economics of the advanced system presented earlier in Level-1 has been affected by the changes recommended as a result of the Level-2 test program. A new economic analysis thus was performed to incorporate findings of Level-2 work. The items considered for the revised economics are as follows:

- Combustor Size
- Solids Recycle System
- Fly Ash Reinjection System
- Heating Surfaces
- NH₃ Injection System
- Sorbent Selection
- Escalation

After revising these costs, new payback periods and steam generation costs were calculated.

Economic Analysis

There are three major cost considerations in an economic analysis

- Capital cost
- Fuel cost
- Operational cost

The capital cost is the price paid for a unit of specified design output steam (flow at pressure). The capital cost differential is the amount of excess price paid for a coal fired CFB boiler to substitute for an oil fired system of the same rating. The fuel cost is

that paid for oil and is to be compared to that paid for coal and limestone/dolomite. The fuel cost differential will favor coal fired CFB over oil at present market fuel prices. Future fuel prices are expected to rise more quickly for oil (3.0% annual increase) than for coal (0.07% annual increase), thus enhancing this advantage. Operational costs should include day-to-day normal plant cost of manpower and utilities for running the plant, maintenance, and repair regardless of the type of unit or fuel fired. When the three major cost considerations are compared, the payback of excess capital investment for coal fired CFB boilers should be of the order of three to five years to make it look attractive in comparison to conventional systems.

A revised cost analysis for the 110,000 lb/hr advanced MSFB boiler system is presented in this section. Sensitivity studies showing the effects of the capacity factor on the total steam cost are also included.

Riley Stoker's cost data base for recent boiler offerings and updated vender quotations have been used to predict capital costs in this study. Therefore, the information presented is consistent with today's market conditions.

Table 7-7 is a breakdown by major component of the total installed cost (TIC) of an advanced concept FBC steam plant. Each component cost includes the basic equipment cost, installation cost on its share of the foundations and structural steel, wiring and piping. The plant itself is considered a new addition to an existing industry, or a portion of a totally new industrial site. To this extent it is virtually grass-roots and shares a minimum of existing equipment. The availability of an operating crew is assumed.

The total capital investment costs for an advanced concept CFB plant are estimated to be \$7,790,027 as tabulated in Table 7-7. This cost was used in comparing the advanced concept CFB to a #6 and a #2 oil fired unit. Capital costs for the oil fired units are \$3,482,325 and \$3,594,360 respectively. Using an 50% capacity factor, the payback period for the coal fired system using dolomite as the sorbent is 4.7 years when compared to a unit fueled by #6 oil and 2.8 years when compared to one firing #2 oil. Corresponding numbers for limestone are 5.4 and 3.1 years. Sensitivity studies investigating the effect of capacity factor showed that for dolomite these payback periods can be reduced to 2.2 and 1.5 years respectively as shown in Table 7-8 at an 80% capacity factor. For limestone, the pay back periods are calculated as 2.4 and

Table 7-7

ADVANCED CONCEPT PLANT - COMPONENT INSTALLED COSTS

	<u>1991 \$ (X 1000)</u>
Combustor, Heat Exchanger, Convective Boiler Bank	\$1,940
Cyclones	237
Fans, Flowlers	396
Solids Handling and Storage	1,264
Flyash Reinjection	541
L-Valves	200
Back End Conditioning and Cleanup, Including Stack	1,644
Controls, Installed and Operational	1,021
Site Preparation	168
Superstructures, Lighting and Paint	199
Balance of Plant, Auxiliaries	180
Total Installed Cost	\$7,790

**PAYBACK PERIOD (years)
FOR A 110,000 LB/HR ADVANCED BOILER
COMPARED WITH OIL-FIRED PACKAGE BOILERS
SORBENT: DOLOMITE**

Capital Cost – Coal FBC	7,790,027			
Capital Cost – #6 Oil	3,482,325			
Inc. Capital Cost – Coal	4,307,702			
CAPACITY FACTOR	0.50	0.60	0.70	0.80
ANNUAL FUEL COSTS				
Coal	1,486,740	1,784,088	2,081,436	2,378,783
Fuel oil (#6)	3,525,855	4,313,182	5,032,045	5,750,909
Fuel Cost Savings	2,039,115	2,529,094	2,950,610	3,372,125
ANNUAL OPERATING COSTS (= O&M – FUEL COST)				
Coal	2,294,691	2,357,861	2,421,032	2,484,203
Fuel Oil (#6)	1,163,334	1,093,401	1,091,930	1,090,459
Increased Op Cost of Coal	1,131,356	1,264,461	1,329,102	1,393,743
Coal Annual Savings	907,759	1,264,633	1,621,508	1,978,382
Payback Period (yrs)	4.7	3.4	2.7	2.2
Capital Cost – Coal FBC	7,790,027			
Capital Cost – #2 Oil	3,594,360			
Inc. Capital Cost – Coal	4,195,667			
CAPACITY FACTOR	0.50	0.60	0.70	0.80
ANNUAL FUEL COSTS				
Coal	1,486,740	1,741,638	2,031,911	2,322,184
Fuel Oil (#2)	4,085,176	4,759,428	5,552,666	6,345,904
Fuel Cost Savings	2,598,436	3,017,790	3,520,755	4,023,720
ANNUAL OPERATING COSTS (= O&M COST – FUEL COST)				
Coal	2,294,691	2,400,311	2,470,557	2,540,802
Fuel Oil (#2)	1,171,737	1,326,742	1,362,761	1,398,780
Increased Op Cost of Coal	1,122,954	1,073,569	1,107,796	1,142,022
Coal Annual Savings	1,475,482	1,944,221	2,412,959	2,881,698
Payback Period (yrs)	2.8	2.2	1.7	1.5

Table 7-8 Payback Period Comparison: Dolomite

1.6 years at 80% capacity factors. This data is presented in Table 7-9.

Other costs, such as labor for both the coal-fired and oil-fired plants, are derived assuming completely new operating staffs. The selection of the Riley RX boiler as the basis of our oil-fired cost comparison with an advanced coal-fired fluidized bed concept lies in the fact that its construction and mode of installation are similar. Both units share the feature of modularity, and therefore derive the benefit of having a large proportion of the boiler being shop fabricated. Shop construction is approximately 40% lower than field construction. The leverage of lower shop rates and higher shop efficiency is evident in the oil fuel system TIC cost of just above \$34/lb of installed steam capacity as shown in Tables 7-10 and 7-11. This cost falls between published costs for packaged shop assembled and field erected units for complete waterwall, oil-fired steam plants. Minimizing field construction labor, therefore is the key element in the cost competitiveness of the oil-fired boiler market. The corresponding costs for the coal fired systems using dolomite and limestone as sorbents are presented in Tables 7-12 and 7-13.

Assumptions used in the analysis are:

- 24 hour operator and assistant attendance
- Percentages for maintenance of coal fired, No. 6 oil fired, and No. 2 oil fired plants are 7%, 4.5% and 4% of Total Installed Costs, respectively.
- Three day fuel storage assumed for both oil and coal.

The costs expressed in terms of 1000 lb steam generated are plotted in Figures 7.5 and 7.6.

With the assumptions specified by DOE for this analysis, the cost competitiveness of this technology with oil firing is evident when operating and maintenance, as well as capital costs are considered. Using dolomite, at 50% annual capacity factor, the total cost for the advanced CFB is \$11.16/1000 lb steam. This is comparable with \$11.54 for the No. 6 oil-fired system, and less than the \$12.83 for the No. 2 oil-fired plant. Figures 7.5 and 7.6 also show that as the capacity factor increases, the advanced CFB costs go down to \$8.45 for dolomite and \$8.70 for limestone in comparison to \$10.16 for #6 oil and \$11.42 for #2 oil. Sensitivity of the type of sorbent selected on the overall economics is marginal. But it is very evident from the economic analysis that

**PAYBACK PERIOD (years)
FOR A 110,000 LB/HR ADVANCED BOILER
COMPARED WITH OIL – FIRED PACKAGE BOILERS
SORBENT: LIMESTONE**

Capital Cost – Coal FBC	7,790,027
Capital Cost – #6 Oil	3,482,325
Inc. Capital Cost – Coal	4,307,702

CAPACITY FACTOR	0.50	0.60	0.70	0.80
ANNUAL FUEL COSTS				
Coal	1,486,740	1,784,088	2,081,436	2,378,783
Fuel oil (#6)	3,525,855	4,313,182	5,032,045	5,750,909
Fuel Cost Savings	2,039,115	2,529,094	2,950,610	3,372,125
ANNUAL OPERATING COSTS (= O&M – FUEL COST)				
Coal	2,407,323	2,493,020	2,578,717	2,664,414
Fuel Oil (#6)	1,163,334	1,093,401	1,091,930	1,090,459
Increased Op Cost of Coal	1,243,989	1,399,619	1,486,787	1,573,955
Coal Annual Savings	795,127	1,129,475	1,463,823	1,798,171
Payback Period (yrs)	5.4	3.8	2.9	2.4

Capital Cost – Coal FBC	7,790,027
Capital Cost – #2 Oil	3,594,360
Inc. Capital Cost – Coal	4,195,667

ANNUAL FUEL COSTS				
Coal	1,486,740	1,741,638	2,031,911	2,322,184
Fuel Oil (#2)	4,085,176	4,759,428	5,552,666	6,345,904
Fuel Cost Savings	2,598,436	3,017,790	3,520,755	4,023,720
CAPACITY FACTOR	0.50	0.60	0.70	0.80
ANNUAL OPERATING COSTS (= O&M COST – FUEL COST)				
Coal	2,407,323	2,535,470	2,628,242	2,721,013
Fuel Oil (#2)	1,171,737	1,326,742	1,362,761	1,398,780
Increased Op Cost of Coal	1,235,586	1,208,728	1,265,481	1,322,233
Coal Annual Savings	1,362,850	1,809,062	2,255,274	2,701,487
Payback Period (yrs)	3.1	2.3	1.9	1.6

Table 7-9 Payback Period comparison: Limestone

Table 7-10

COMPARATIVE PLANT COSTS: #6 OIL

COMPARATIVE PLANT COSTS
FOR A 110,000 LB/HR ADVANCED BOILER
FUEL - #6 OIL

TOTAL INSTALLED COSTS	3,482,325	3,482,325	3,482,325	3,482,325
Capacity Factor (%)	0.50	0.60	0.70	0.80
MAINTENANCE COST SUMMARY				
electricity	56,280	67,536	78,792	90,048
fuel	3,525,855	4,231,026	4,936,197	5,641,368
water & chemicals	10,080	11,046	12,012	12,978
sorbent	0	0	0	0
disposal & sand	0	0	0	0
labor	835,800	835,800	835,800	835,800
maintenance	156,705	156,705	156,705	156,705
taxes & insurance	104,470	104,470	104,470	104,470
total annual O&M	4,689,189	5,406,582	6,123,975	6,841,368
REQUIRED CAPITAL COST SUMMARY				
total installed costs	3,482,325	3,482,325	3,482,325	3,482,325
start-up costs	390,766	450,549	510,331	570,114
construction interest	417,879	417,879	417,879	417,879
working capital	1,172,297	1,351,646	1,530,994	1,710,342
total capital requirement	5,463,267	5,702,398	5,941,529	6,180,660
ANNUALIZED COST SUMMARY				
annual capital recovery	872,812	911,015	949,219	987,422
federal income tax	0	0	0	0
total debt service	872,812	911,015	949,219	987,422
total annual O&M	4,689,189	5,406,582	6,123,975	6,841,368
total annual cost	5,562,001	6,317,597	7,073,194	7,828,790
STEAM COST SUMMARY (\$/1000 LB STEAM)				
annual capital steam cost	\$1.81	\$1.58	\$1.41	\$1.28
annual O&M steam cost	\$9.73	\$9.35	\$9.08	\$8.87
total steam cost	\$11.54	\$10.93	\$10.49	\$10.16

Table 7-11

COMPARATIVE PLANT COSTS: #2 OIL

COMPARATIVE PLANT COSTS
FOR A 110,000 LB/HR ADVANCED BOILER
FUEL - #2 OIL

TOTAL INSTALLED COSTS	3,594,360	3,594,360	3,594,360	3,594,360
Capacity Factor (%)	0.50	0.60	0.70	0.80
MAINTENANCE COST SUMMARY				
electricity	56,280	67,536	78,792	90,048
fuel	4,085,176	4,902,211	5,719,246	6,536,281
water & chemicals	10,080	11,046	12,012	12,978
sorbent	0	0	0	0
disposal & sand	0	0	0	0
labor	835,800	835,800	835,800	835,800
maintenance	161,746	161,746	161,746	161,746
taxes & insurance	107,831	107,831	107,831	107,831
total annual O&M	5,256,913	6,086,170	6,915,427	7,744,684
REQUIRED CAPITAL COST SUMMARY				
total installed costs	3,594,360	3,594,360	3,594,360	3,594,360
start-up costs	438,076	507,181	576,286	645,390
construction interest	431,323	431,323	431,323	431,323
working capital	1,314,228	1,521,542	1,728,857	1,936,171
total capital requirement	5,777,987	6,054,406	6,330,826	6,607,245
ANNUALIZED COST SUMMARY				
annual capital recovery	923,091	967,252	1,011,413	1,055,573
federal income tax	0	0	0	0
total debt service	923,091	967,252	1,011,413	1,055,573
total annual O&M	5,256,913	6,086,170	6,915,427	7,744,684
total annual cost	6,180,004	7,053,422	7,926,840	8,800,258
STEAM COST SUMMARY (\$/1000 LB STEAM)				
annual capital steam cost	\$1.92	\$1.67	\$1.50	\$1.37
annual O&M steam cost	\$10.91	\$10.53	\$10.25	\$10.05
total steam cost	\$12.83	\$12.20	\$11.75	\$11.42

**COMPARATIVE PLANT COSTS
FOR A 110,000 LB/HR ADVANCED BOILER
COAL FIRED FLUIDIZED BED
SORBENT: DOLOMITE**

TOTAL INSTALLED COSTS	7,790,027	7,790,027	7,790,027	7,790,027
Capacity Factor	0.50	0.60	0.70	0.80

MAINTENANCE COST SUMMARY

Electricity	73,710	88,452	103,194	117,936
Fuel	1,486,740	1,784,088	2,081,436	2,378,783
Water & chemicals	10,080	11,046	12,012	12,978
Sorbent (dolomite)	168,958	202,750	236,541	270,333
Disposal & sand	68,355	82,026	95,697	109,368
Labor	1,194,585	1,194,585	1,194,585	1,194,585
Maintenance	545,302	545,302	545,302	545,302
Taxes & insurance	233,701	233,701	233,701	233,701
Total annual O&M	3,781,430	4,141,949	4,502,468	4,862,986

REQUIRED CAPITAL COST SUMMARY

Total installed costs	7,790,027	7,790,027	7,790,027	7,790,027
Start-up costs	315,119	345,162	375,206	405,249
Construction interest	934,803	934,803	934,803	934,803
Working capital	945,358	1,035,487	1,125,617	1,215,747
Total capital requirement	9,985,307	10,105,480	10,225,653	10,345,826

ANNUALIZED COST SUMMARY

Annual capital recovery	1,595,253	1,614,451	1,633,650	1,652,849
Federal income tax	0	0	0	0
Total debt service	1,595,253	1,614,451	1,633,650	1,652,849
Total annual O&M	3,781,430	4,141,949	4,502,468	4,862,986
Total annual cost	5,376,683	5,756,400	6,136,118	6,515,835

**STEAM COST SUMMARY
(\$/1000 LB STEAM)**

Annual capital steam cost	\$3.31	\$2.79	\$2.42	\$2.14
Annual O&M steam cost	\$7.85	\$7.16	\$6.68	\$6.31
Total steam cost	\$11.16	\$9.96	\$9.10	\$8.45

Table 7-12 Comparative Plant Costs: Coal/Dolomite

COMPARATIVE PLANT COSTS
FOR A 110,000 LB/HR ADVANCED BOILER
COAL-FIRED FLUIDIZED BED
SORBENT : LIMESTONE

TOTAL INSTALLED COSTS	7,790,027	7,790,027	7,790,027	7,790,027
Capacity Factor (%)	0.50	0.60	0.70	0.80
MAINTENANCE COST SUMMARY				
electricity	73,710	88,452	103,194	117,936
fuel	1,486,740	1,784,088	2,081,436	2,378,783
water & chemicals	10,080	11,046	12,012	12,978
sorbent	281,590	337,908	394,226	450,544
disposal & sand	68,355	82,026	95,697	109,368
labor	1,194,585	1,194,585	1,194,585	1,194,585
maintenance	545,302	545,302	545,302	545,302
taxes & insurance	233,701	233,701	233,701	233,701
total annual O&M	3,894,063	4,277,108	4,660,153	5,043,197
REQUIRED CAPITAL COST SUMMARY				
total installed costs	7,790,027	7,790,027	7,790,027	7,790,027
start-up costs	324,505	356,426	388,346	420,266
construction interest	934,803	934,803	934,803	934,803
working capital	973,516	1,069,277	1,165,038	1,260,799
total capital requirement	10,022,851	10,150,533	10,278,214	10,405,896
ANNUALIZED COST SUMMARY				
annual capital recovery	1,601,251	1,621,649	1,642,048	1,662,446
federal income tax	0	0	0	0
total debt service	1,601,251	1,621,649	1,642,048	1,662,446
total annual O&M	3,894,063	4,277,108	4,660,153	5,043,197
total annual cost	5,495,313	5,898,757	6,302,200	6,705,643
STEAM COST SUMMARY (\$/1000 LB STEAM)				
annual capital steam cost	\$3.32	\$2.80	\$2.43	\$2.16
annual O&M steam cost	\$8.08	\$7.40	\$6.91	\$6.54
total steam cost	\$11.41	\$10.20	\$9.34	\$8.70

Table 7-13 Comparative Plant Costs: Coal/Limestone

ADVANCED BOILER ECONOMICS

Coal vs. Oil / Sorbent: Dolomite

CAPACITY FACTOR

Figure 7.5 Advanced Boiler Economics - Steam Costs: Dolomite

ADVANCED BOILER ECONOMICS

Coal vs. Oil / Sorbent: Limestone

CAPACITY FACTOR

Figure 7.6 Advanced Boiler Economics - Steam Costs: Limestone

the coal fired CFB is strongly favored against the oil fired systems. Figures 7.7 and 7.8 pictorially compare the payback periods of coal fired systems with #2 and #6 oil for different capacity factors and sorbents.

ADVANCED BOILER ECONOMICS

Coal vs. Oil / Sorbent: Dolomite

CAPACITY FACTOR

Figure 7.7 Advanced Boiler Economics - Payback Period: Dolomite

ADVANCED BOILER ECONOMICS

Coal vs. Oil / Sorbent: Limestone

Figure 7.8 Advanced Boiler Economics - Payback Period: Limestone

SECTION 8

PROOF OF CONCEPT FACILITY

Success of a development project lies in how effectively the generated information is translated into a commercial product. So, the design development and economic analysis of the coal fired CFB system carried out under this contract would not be complete until a commercial demonstration plant is designed and operated using the results of the program.

In order to achieve this goal Riley has devised an approach to locate potential users of the FBC technology who would find value in the performance and attractive economics of the advanced concept. Working as part of a team, a motivated customer can make useful contributions to the design resulting in the product's acceptance in the targeted markets. Riley has followed this philosophy during the selection of a candidate for the proof of concept facility and has secured a site that will mutually benefit both Riley and the selected customer.

8.1 SELECTION CRITERIA

DOE contract with METC has specific guidelines for Riley to follow in making recommendations for proceeding to the next level of the contract. These guidelines evaluate the results of the work carried out during the execution of Level-2 of the contract. Only after it is determined that this criteria has been fully satisfied will DOE consider proceeding to Level-3. Following are the three major components of this criteria:

- The integration of the subsystem has resulted in an AFBC system that provides competitive performance at a reasonable price.
- The system has demonstrated that it is competitive with similarly sized gas and oil fired systems when evaluated with the cost assumptions provided by DOE.
- An industrial user is committed to sponsoring an AFBC demonstration by offering a site and operating a new AFBC system concept while cost and performance data is collected.

As discussed in Section 6 the pilot plant operated according to expectations and major

performance goals were demonstrated during the six week test run of the pilot plant. The economics of the design as presented in Section 7 also looks attractive, in comparison to the oil fired systems for the advanced concept, sizes ranging between 75,000 to 150,000 lb/hr. The contract did not ask for the economic analysis against gas fired designs.

After having succeeded in the demonstrating the performance and economic viability of the design, Riley's effort moved toward the third important criteria for proceeding to the next level of the contract. This consists of securing a host site for the demonstration of the concept. At the conclusion of the Level-2 work Riley was able to obtain conditional support from one company and strong support from another, pending the findings of a detailed economic evaluation of the design by the prospective customers.

Discussions are continuing with the most promising customer to generate additional performance and cost data. The approach taken for securing the host site is described in the following section.

8.2 TARGETING THE PAPER INDUSTRY

A number of candidates were considered for the proof of concept facility for Level-3 of the contract. The original plan was to construct this plant for the small institutional user. Hence the earlier contacts were with those companies and institutions that needed small package size boilers with 10,000 to 50,000 lb/hr of steam production, mostly for space heating. Although the design had technical merit, it was up against commercially available products that would be very difficult to beat on many fronts. The existing designs in the market are very compact and easy to operate. A coal fired design, on the other hand, needs more space and special operator skills beyond the ones required for the current designs would be needed, at least during the initial stages of operation. During Level-1 it was determined that the small coal fired FBC boiler design was not a viable economic alternative to the conventional oil fired units. The economics was even less attractive when compared with the gas fired plants.

However, the advanced industrial size design (75,000-150,000 lb/hr) was found to be economically feasible against the corresponding oil fired boilers. As noted in Section 7, in one instance, this design has the potential to pay back in slightly over one year.

Therefore, at the end of Level-1, it was recommended that it would be meaningful to proceed to Level-2, only with the larger capacity design.

This meant additional work in contacting companies, primarily industrial manufacturers, who would fall under this larger size category. Although the scope of Level-3 work did not change, there was a definite change in direction for the proof of concept facility. With a different market being targeted, the boiler specifications were revised. It should be noted that many of the present day industrial users prefer to have cogeneration capability in their new boilers. Steam conditions for such a system are different from those selected for this contract. Table 8-1 compares the steam conditions for a typical cogeneration system with those for this contract. Riley believes that making such a compromise is very essential because it is in accordance with the market needs and any new product being offered should be able to conform to customer needs. The long term success of this project is also dependent on how effectively we are able to utilize the developed information for commercial applications. That justifies the expense of sponsoring collaborative developmental projects between DOE and industry.

Riley's sales and marketing force was again called upon to assist in identifying the industries that would benefit from this development. During this evaluation process an industry that showed tremendous promise was recognized. It is the Paper and Pulp industry which is in dire need of a system that could help reduce its burden of a variety of wood derived wastes generated during paper manufacturing and recycling.

Taking lead from this vast market potential, Riley contacted all the major paper manufacturers in the country. The response was very positive. After these preliminary contacts, only the five companies listed in Table 8-2 were selected for further discussions. These companies represented a total of seven potential sites for the proof of concept facility.

Riley believes that the paper and pulp industry is a potential beneficiary of the development work carried out on this contract. Although the original contract was conceived with the intent of firing the FBC steam generators exclusively with coal, it was decided to explore the possibility of using combination fuels. Due to the limited scope of this contract, it is rather difficult to address every aspect of the paper and pulp industry where FBC technology could find application. But an area which has been

Table 8-1
STEAM CONDITIONS FOR LEVEL-3 PROOF OF CONCEPT FACILITY

	<u>Original Level 1</u>	<u>Proposed Level-3 (Cogeneration)</u>
Steam Flow (lb/hr)	110,000	100 - 250,000
Pressure (PSIA)	215	875 - 1500
Temperature (°F)	388	850 - 1000

Table 8-2

HOST SITE SELECTION ACTIVITY

<u>COMPANY</u>	<u>PRODUCT</u>	<u>STEAM REQUIREMENT</u>	<u>FUEL</u>
<u>International Paper</u>			
A) Lockhaven, PA	Copying Paper	150,000 lb/hr	Coal, Chips, Sludge
B) Towanda, PA	Hard Board	100,000 lb/hr	Chips, Sludge, Coal
<u>Stone Container</u>			
A) Ontonagan, MI	Packing Container Paper	100,000 lb/hr	Coal, Sludge, Chips
<u>Container Corporation</u>			
A) Fernandina Beach, FL	Packing Container Paper	250,000 lb/hr	Coal, Sludge, Chips
<u>James River</u>			
A) Wauna Mill, OR	Paper	100, 00 lb/hr	Bark, Sludge, Tires
B) Berlin, NH	Paper	Not determined	Coal, Sludge, Chips, Waste Paper
<u>Kimberly Clark</u>			
A) Coosa Pines, AL	Pine Pulp and Newsprint	120,000 lb/hr	Coal, Sludge, Paper Waste

receiving much attention in this industry lately, paper recycling, has shown very good promise. Discussions with many of the paper manufacturing companies and contacts with organizations related to the Paper and Pulp Industry has identified a waste problem that is just beginning to surface in the recycling of paper.

To avoid the project from being overwhelmed by the problems of this particular industry, it was decided that only one specific area, namely, waste material generated in a paper de-inking process be addressed and pursued. The U.S. paper industry continues to demonstrate its commitment to using increased amounts of recycled fiber by the abundance of de-inking projects planned or proposed to start up in the next couple of years. The amount of non-recyclable effluent produced in a de-inking plant amounts to approximately 10-15 % by weight of the recycled paper. On a dry basis it has a heating value of 6000-7000 Btu/lb. Unless efficient means of combusting this material are found it will be disposed of in a land fill. Riley proposes to burn it in combination with coal in an FBC boiler. By doing this we are able to reduce the land fill costs and generate steam for the processing plant. For the customer, utilization of this material as fuel is a value added option.

Although many paper companies have cut back on capital spending, they have continued to invest in secondary fiber projects. This is largely due to legislation pressures requiring increased recycled fiber content in newsprint. Also, it is becoming economically viable to use secondary fiber in fine as well as package grade paper manufacturing. Tables 8-3 through 8-5 list projects planned through 1992 (10). According to the American Paper Institute (API), there is a steeper rise in the paper and paper board capacity relative to that of wood pulp capacity. This reflects the industry's preference for the recovered waste paper to furnish the paper making fibre rather than fresh wood pulp (11). According to the studies carried out by API, 45% of all the news papers and magazines will be recycled in the country by the year 1995. This will be in the range of 40 million tons, a 50% rise in the over 1988 levels. Targets for year 2000 are even more ambitious (12).

8.2.1 The De-inking Process

In a de-inking system, waste newspaper and magazines are first turned into a pulp to defiber the paper. Caustic is added in the pulper to help separate the ink from fibers and a small amount of hydrogen peroxide is added to prevent yellowing.

Table 8-3

PAPER DE-INKING MARKET ACTIVITIES IN NORTH AMERICA
NEWSPRINT

<u>Company/Mill/Site</u>	<u>Startup</u>	<u>De-Inking capacity (mtpy)</u>
Abitibi-Price, Thunder Bay, Ont.	May '92	55,000
Alabama River Newsprint Claiborne, Ala	Spring '92	100,000
Atlantic Newsprint, Whitby, Ont.	Jan. '91	146,000
Augusta Newsprint, Augusta, Ga.	Mar. '91	86,750
Bear Island, Ashland, Va.	Late '92	43,500
Boise Cascade, W. Tacoma, Wash.	1992	70,000
Bowater, Calhoun, Tenn.	Sept. '91	109,500
Canadian Pacific Forest Products Gatineau, Que.	Late '91	182,500
Canadian Pacific Forest Products Thunder Bay, Ont.	Late '91	100,375
Cascades/Donohue/Maclaren Cap de la Madeleine, Que.	Early '92	90,000
Chammpion Internantional Sheldon, Texas	1992	138,800
Daishowa Forest, Quebec City, Que.	Feb. '92	110,000
Daishowa Forest, Port Angeles, Wash.	Spring '92	65,000
Inland Empire, Millwood, Wash.	Late '91	37,000
Kruger, Bromptonville, Que.	1992	41,640
Newstech Recycling, Vancouver, B.C.	Nov. '91	122,300
North Pacific Paper, Longview, Wash.	Apr. '91	156,000
Smurfit Newsprint N.Y. State (site not chosen)	1994	n.a.
Stone-Consolidated, Shawinigan, Que. (under study)	n.a.	70,000
Alberta Newsprint, Whitecourt, Alta.	n.a.	n.a.
Fletchers Challenge, Elk Falls and Crofton, B.C.	n.a.	n.a.
Georgia-Pacific, E. Millinocket, Maine	n.a.	86,750
Manistique (Kruger) Manistique, Mich.	Proposed	34,700
Kenaf Paper, Raymondville, Texas	n.a.	n.a.

Table 8-4

PAPER DEINKING MARKET ACTIVITIES IN NORTH AMERICA
TISSUE

<u>Company/Mill/Site</u>	<u>Startup</u>	<u>De-inking Capacity (mtpy)</u>
Scott, Owensboro, Ky.	Proposed	n.a.
Perkins Papers, Candiac, Que.	Aug. '91	48,500
Atlas Paper, Hialeah, Fla.	June '91	21,000
Putney Paper, Putney, Vt.	Feb. '91	15,000
Marcal Paper, Elmwood, N.J.	n.a.	n.a.
Bay West (Mosinee), Middletown, Ohio	1991	79,810
Fort Howard, Rincon, Ga.	1991	n.a.
American Power, Bala Cynwyd, Pa.	1993	138,800
Orford Recycling, Drummondville, Que.	n.a.	120,000
Fox River Fiber, De Pere, Wis.	July '91	69,400
Mississippi River Corp. Natchez, Miss.	Jan. '91	n.a.
Caithness King, Pejepscot, Maine	1993	190,850
Caithness King, Midland, Mich.	1992	190,850
Ponderosa, New York, N.Y.	n.a.	n.a.
Ponderosa, Augusta, Ga.	1992	34,700
Ponderosa, Oshkosh, Wis.	1992	34,700
Ponderosa, Memphis, Tenn.	1992	34,700

Table 8-5
PAPER DE-INKING MARKET ACTIVITIES IN NORTH AMERICA
PRINTING/WRITING

<u>Company/Mill/Site</u>	<u>Startup</u>	<u>De-inking capacity (mtpy)</u>
Noranda Inc. Thorold, Ont.	Proposed	120,000.
Noranda Forest Recycled Papers, Thorold, Ont.	1991	32,965
Patriot Paper, Boston, Mass.	Aug. '91	34,700
Riverside Paper, DePere, Wis.	n.a.	n.a.

From the pulper, the pulp is taken to a screen to remove coarse material and debris followed by fine screening to remove the smaller particles of debris.

The clean pulp then goes through a flotation step to remove ink in a highly concentrated form. The pulp is then washed in a gravity drum washer. High speed belt washers can also be employed for this washing operation. This is followed by bleaching with hydrogen peroxide or sodium hydrosulphite. After a final thickening step the product is taken to the paper machine. Figure 8.1, shows all the operations performed during the de-inking process.

An important part of any de-inking system is the water clarification and the treatment and disposal of sludge from various points in the system. The major sources of waste or rejects from the system are as follows:

- Pulper detrashing
- Final stage of coarse and fine screening
- Flotation
- Cleaner
- Thickener white water
- Liquid cyclones

The solid material from the pulper detrasher and liquid cyclones is disposed of directly into a reject handling system. The coarse and fine screen rejects are combined with flotation cell rejects and cleaner rejects and sent to a sludge chest and then to a sludge press for thickening and disposal. White water from the washer, press and cleaners is sent to a clarifier to remove suspended solids such as ink, clay and fines, which are then sent to a sludge press. The clarified water is reused in the system.

The non-recyclable rejects sludge generated at various locations in the process is the material that is proposed to be combined with coal for combustion in an FBC system.

8.2.2 Facility Location

With the assistance of the National Council on Air and Stream Improvement (NCASI), Riley was able to contact more than seventy members of NCASI who are potential

PROCESS SCHEMATIC PAPER RECYCLING AND DE-INKING

Figure 8.1 Process Schematic Paper Recycling and De-inking

candidates for this project. Follow up was done with six companies and two of these, International Paper (IP) and Stone Container showed consistent interest in the project.

The most promising among these is IP's Hammermill operations at Lockhaven, PA, which has been selected as the site for a paper recycling and de-inking plant designed to process newspaper and magazines. This plant will be constructed by IP under a German license to turn the recycled paper into high quality printing and writing paper. Approximately 400 tons/day of recyclable paper will be brought to Lockhaven from the neighboring towns. The program is co-sponsored by the local communities and the state of Pennsylvania. The recycling plant will generate 30-40 tons/day of waste sludge, which, according to the current plans will be disposed off in a landfill already approved by the IP board of directors. It has 45-50% moisture and has a heating value of ~ 6000-7000 Btu/lb (dry).

8.2.3 Special Issues

Riley believes that by burning this sludge in combination with coal in the advanced FBC boiler, developed under this contract, a new market would be tapped for this technology. But, success of this effort will depend entirely on how well we are able to identify uncertainties related to the concept presented and preparing ahead for the potential problems that may come about during the development of the design. These uncertainties must be resolved before a design is finalized for submission to the customer. The following are among the more important items that need to be examined before an efficient design can be established:

- Understanding of the material handling characteristics of the sludge which has 45-50% total moisture content.
- Emission characterization of this fuel. Besides the continuous monitoring of SO₂, NO_x, CO, CO₂, it will be advisable that special measurements be made to determine if any trace metals and/or dioxins are released as a result of combustion of this fuel. This is important because the sludge is produced from de-inking of colored and chlorine bleached paper which may have environmental ramifications.

Being the largest paper manufacturing company in the world, there is a strong interest in this project at IP. The incentives to recycle waste paper are continually increasing due to greater world demand for paper, decreased landfill space in urban areas and an increasing quantity of solid waste to be disposed or processed. The Hammermill

location is only one of the plants that IP has considered for this application.

Results of the preliminary economic evaluations conducted on this location by IP have been very promising. These evaluations are based on the data furnished by Riley on the Advanced Concept design. Moreover, data expected from the proposed test program will provide additional information to assess the emissions and fuel handling aspects of this material.

Based on these results, another site specific detailed economic analysis will be performed in collaboration with International Paper incorporating the following items:

1. The cost savings associated with the land fill costs.
2. Utilization of inexpensive and locally available coal and limestone.
3. Extra Steam availability.
4. Payback on the cost of capital.

To protect the economic interests of DOE, Riley and the host site owner, International Paper, this information will be generated prior to the commencement of Level-3 work.

SECTION 9.0

LEVEL 3 PROGRESSION CRITERIA

The progression criteria for proceeding to Level 3 are:

- The integration of the subsystems has resulted in an AFBC system that provides competitive performance at a reasonable price.
- The system has demonstrated that it will be competitive with similarly sized gas- and oil-fired systems when evaluated with the cost assumptions provided by DOE.
- An industrial user is committed to sponsoring an AFBC demonstration by offering a site and operating a new AFBC system concept while cost and performance data are collected.

SECTION 10.0

CONCLUSIONS AND RECOMMENDATIONS

Riley has successfully executed the Level-2 work of the contract and has satisfied the requirements for proceeding to the next level for the construction and operation of the proof of concept facility. It has been determined that a commercial need exists for the industrial size FBC systems offered by this project. The advanced concept design has met all the performance goals set in the contract. The economics of the design is favored against the competing oil systems. A potential host site has been secured for the demonstration of this concept in an industrial environment.

Riley recommends that DOE continue with its efforts toward the development of an affordable and efficient FBC technology for the industrial customers. We also propose that this commercial demonstration in Level-3 be carried out with the co-firing of coal in combination with paper de-inking sludge. This suggestion is made after examining the potential needs of the market.

In Level-2, Riley has gone beyond just testing of the concept for its process performance. It has also investigated market opportunities in American industry that could benefit from this development. Contacts and discussions with various paper and pulp companies have identified a critical need for finding a solution to its non-recyclable waste fibre sludge problem that demands immediate attention. A collaborative effort by the boiler manufacturers and the paper industry would come to the mutual benefit of both. This would also mean that two fuels, namely, high sulfur coal and waste sludges that have been the target of environmentalists will be able to contribute towards a cleaner environment. The offer of participation from International Paper, the largest paper company in the world, should be seen as a very significant message coming from this industry in general. According to Riley's market analysis this interest is bound to grow as more companies start to recycle paper.

The boiler industry must be adequately prepared and positioned to take on this task. To achieve this goal, Riley recommends a step approach as follows:

- Conduct a short test program on the pilot plant to evaluate material handling and emission characteristics of the coal-paper sludge

combination and the disposed ash.

- Provide funding for the site specific engineering and economics analysis
- Based on the results of the above two items, decide to proceed to the construction and operation of the proof of concept facility.

This program will enable us to focus on the specific issues faced by this industry prior to the design of the proof of concept facility.

SECTION 11.0

REFERENCES

1. Bramer, E.A., et.al., "Nitrous Oxide and Nitric Oxide Emissions by Fluidized Bed Combustion", 11th International Conference on Fluidized Bed Combustion, Montreal, Canada, April 21-24, 1991.
2. Brown, R., Muzio, L., "N₂O Emissions from Fluidized Bed Combustion", 11th International Conference on Fluidized Bed Combustion, Montreal, Canada, April 21-24, 1991.
3. Jahkola, A., et.a., "Emission and Reduction of NO_x and N₂O in PFB - Combustion of Peat and Coal", 11th International Conference on Fluidized Bed Combustion, Montreal, Canada, April 21-24, 1991.
4. Moritomi, H., et.a., "NO_x Formation Mechanism of Circulating Fluidized Bed Combustion", 11th International Conference on Fluidized Bed Combustion, Montreal, Canada, April 21-24, 1991.
5. Lyon, R.K., Longwell, J.P., "Selective, Non-Catalytic Reduction of NO_x by NH₃", Proc. of the NO_x Control Technology Seminar, San Francisco, CA., Feb. 5-6, 1976, Special Report 39.
6. Muzio, L.J., Arand, J.K., Teixeira, D.P., "Gas Phase Decomposition of Nitric Oxide in Combustion Products", Ibid.
7. EPRI Report CS-4386.
8. Kunii and Levenspiel FLUIDIZATION ENGINEERING.
9. "Atmospheric Fluidized Bed Combustion Advanced System Concepts Applicable to Small Industrial and Commercial Markets", Topical Report, METC Contract DE-AC21-87MC23295.
10. "Pulp & Paper", May 1991, P52-129.
11. "Paper Recycling: The Views to 1995", A report prepared by Franklin Associates for American Paper Institute, February 1990.
12. "Publically Announced Expansion Plans for Increased Use of Recoverd Paper", American Paper Institute Report, November 1991.

APPENDIX A

DOE Advanced CFB
Performance Database
As of: 10/02/91

Week	Test	DATE	SUR FUR	HRV	LOAD	SORBENT	Vox	tau	Tox	Stach	XSA	CPdb	CPdb	FAR	CE	Ca/S	SC	CU	MOx	CO	ES	Mo
1	1	06/19	3.0	12,313	4.1 Limestone	22	2.2	1650	0.8	0.8	35	21	20	0.0	97.9	1.6	50	30	0.27	11	9	7140
	2	06/20	3.0	12,313	4.4 Limestone	21	2.3	1673	0.8	0.8	17	20	14	0.0	97.7	1.6	50	30	0.27	11	9	7140
	3	06/20	3.0	12,313	4.2 Limestone	22	2.2	1656	0.8	0.8	25	19	11	0.0	MA	1.7	62	34	0.21	9	7	56
2	4	06/25	3.0	12,267	4.0 Limestone	22	2.2	1645	0.9	0.9	39	30	12	0.0	98.2	1.8	76	42	0.30	MA	MA	4710
	5	06/27	3.0	12,267	4.0 Limestone	22	2.2	1650	0.8	0.8	35	25	25	0.0	98.4	1.7	74	44	0.30	4	4	97.56
	6	06/28	3.0	12,267	3.1 Limestone	16	2.9	1640	0.9	0.9	35	28	11	0.0	98.1	1.4	72	52	0.31	23	23	4710
3	7	07/08	3.2	12,050	3.1 Limestone	17	2.8	1637	0.8	0.8	42	33	10	0.0	98.9	1.4	74	53	0.37	19	19	4710
	8	07/10	3.2	12,050	3.1 Limestone	18	2.6	1635	0.8	0.8	49	27	10	1.5	98.8	1.4	75	53	0.34	15	15	4710
	9	07/11	3.2	12,050	3.1 Limestone	17	2.7	1632	0.8	0.8	43	27	10	1.4	98.6	1.7	76	45	0.32	10	10	4710
4	9a	07/11	3.2	12,050	3.1 Limestone	17	2.8	1630	0.8	0.8	43	27	10	NA	MA	1.7	79	38	0.38	8	8	4710
	9b	07/11	3.2	12,050	3.1 Limestone	17	2.8	1630	0.8	0.8	46	27	10	NA	MA	2.1	84	36	0.32	10	10	4710
	9c	07/12	3.2	12,050	3.1 Limestone	17	2.8	1630	0.8	0.8	43	27	10	2.0	99.3	2.3	89	32	0.32	11	11	4710
5	9d	07/12	3.2	12,050	3.1 Limestone	17	2.8	1630	0.8	0.8	43	27	10	NA	MA	2.8	89	38	0.22	11	11	4710
	10	07/15	2.8	12,406	4.2 Limestone	21	2.3	1650	0.8	0.8	25	25	15	1.3	98.5	1.5	65	43	0.15	14	14	4710
	10a	07/16	2.8	12,406	4.2 Limestone	21	2.3	1640	0.8	0.8	23	30	15	NA	MA	1.5	66	43	0.14	56	56	4710
6	10b	07/16	2.8	12,406	4.2 Limestone	20	2.4	1570	0.8	0.8	23	30	15	NA	MA	1.5	65	43	0.14	56	56	4710
	10c	07/16	2.8	12,406	4.2 Limestone	20	2.4	1540	0.8	0.8	23	30	15	NA	MA	1.5	65	43	0.14	56	56	4710
	10d	07/16	2.8	12,406	4.2 Limestone	21	2.3	1635	0.8	0.8	23	30	20	NA	MA	1.5	68	35	0.23	17	17	4710
7	11	07/17	2.8	12,406	4.2 Limestone	21	2.3	1631	0.6	0.6	35	27	15	2.4	99.2	2.5	84	34	0.24	15	15	4710
	12	07/18	2.8	12,406	4.2 Limestone	23	2.1	1631	0.6	0.6	35	32	15	3.6	99.3	2.5	84	34	0.24	15	15	4710
	13	07/18	2.8	12,406	4.2 Limestone	23	2.0	1644	1.0	1.0	40	25	15	1.4	98.8	2.5	76	30	0.26	11	11	4710
8	14	07/23	2.3	12,847	4.4 Limestone	21	2.3	1617	0.8	0.8	21	39	15	3.4	99.1	3.0	75	25	0.16	45	45	4710
	15	07/24	2.3	12,847	4.4 Limestone	23	2.1	1713	0.8	0.8	29	36	15	3.5	99.6	3.0	62	21	0.29	6	6	4710
	16	07/24	2.3	12,847	4.4 Limestone	23	2.1	1631	0.8	0.8	32	37	15	1.9	98.9	3.0	91	30	0.24	16	16	4710
9	17	07/26	2.3	12,847	4.4 Limestone	23	2.1	1633	0.8	0.8	32	37	15	2.8	99.3	2.3	81	35	0.23	17	17	4710
	18	07/26	2.3	12,847	4.4 Limestone	21	2.3	1625	0.8	0.8	31	41	15	0.0	97.6	1.7	60	36	0.23	10	10	4710
	19	07/30	2.7	12,025	4.4 Dolomite	22	2.2	1624	0.8	0.8	30	42	15	0.0	98.0	1.5	79	52	0.28	13	13	4710
10	20	08/01	2.7	12,025	4.4 Dolomite	22	2.2	1628	0.8	0.8	30	39	15	1.4	98.5	1.5	94	78	0.15	19	19	4710
	21	08/01	2.7	12,025	4.4 Dolomite	21	2.3	1625	0.8	0.8	30	42	15	2.5	99.2	1.0	74	78	0.15	19	19	4710
	22	08/02	2.7	12,025	4.4 Dolomite	21	2.3	1627	0.8	0.8	30	41	15	1.7	98.9	1.5	87	57	0.25	12	12	4710
11	23	08/02	2.7	12,025	4.4 Dolomite	21	2.3	1626	0.8	0.8	30	40	15	NA	MA	2.3	96	42	0.24	25	25	4710
	24	08/03	2.7	12,025	4.4 Dolomite	21	2.3	1626	0.8	0.8	30	40	15	1.9	99.2	1.5	89	58	0.24	21	21	4710

* 0.12 lb/10⁶ lb/hr MOx achieved with ammonia injection at 2 NH3/H2O ratio.

APPENDIX B

641-87805 DOE Advanced Concept MSFB
NITROUS OXIDE AND AMMONIA INJECTION TESTING

As of:08/06/91

Test No.	Date	TOXI (F)	INJ. XSA ELEV. (ft)	NH3 FLOW (std l/m)	NOx (ppm)	N2O (ppm)	O2 (%)	NH3 SLIP (ppm)	NH3 ESTIMATED FLUE GAS FLOW(pph)	NH3/NO Mole Ratio	NOx3XO2 (ppm)	N2O3XO2 (ppm)	NH3 SLIP a3XO2 (ppm)
** 14	07/23	1640	25	0.0	140	85	4.5	0	4400	0.0	153	93	0
** 15	07/24	1720	25	0.0	200	44	4.5	0	4400	0.0	218	48	0
** 17	07/25	1640	35	0.0	155	85	5.6	0	4700	0.0	181	99	30
				20.0	90	97	5.6	26	4700	4.5	105	113	NA
				10.0	106	91	5.6	NA	4700	2.3	124	106	NA
				15.0	112	89	5.6	NA	4700	1.1	131	104	NA
				5.0	90	94	5.6	NA	4700	3.4	105	110	NA
				0.0	143	87	5.6	0	4700	0.0	167	102	35
17	07/26	1640	35	0.0	65	95	5.6	30	4700	4.9	76	111	NA
				20.0	93	94	5.6	NA	4700	2.4	109	110	NA
				10.0	79	97	5.6	NA	4700	3.7	92	113	9
				15.0	105	86	5.6	7.5	4700	1.2	123	101	0
				5.0	87	151	6.5	0	5000	0.0	108	188	NA
17	07/26	1540	35	0.0	42	173	6.5	NA	5000	7.6	52	215	NA
				20.0	56	166	6.5	29	5000	3.8	70	206	36
				10.0	165	NA	5.6	0	4700	0.0	193	NA	0
17	07/26	1710	35	0.0	115	NA	5.6	NA	4700	2.1	135	NA	NA
				10.0	89	NA	5.6	17	4700	6.2	104	NA	20
				20.0	161	NA	5.6	0	4550	0.0	181	NA	0
18	07/30	1640	25	0.0	35	NA	5.0	NA	4550	0.0	39	NA	NA
				20.0	74	NA	5.0	21	4550	0.0	83	NA	24
				10.0	110	NA	5.0	NA	4550	0.0	124	NA	NA
				5.0	160	NA	5.0	0	4550	0.0	180	NA	0
23	08/02	1640	25	0.0	40	NA	5.0	NA	4550	4.5	96	NA	NA
				20.0	85	NA	5.0	23	4550	2.3	146	NA	26
				10.0	130	NA	5.0	NA	4550	1.1	NA	NA	NA

COMMENTS
** Test 14 - Brief ammonia injection test took place using single nozzle injection with no carrier gas. No change in NOx, but N2O went to 93-99 ppm and ammonia slip 30 ppm.

** Test 15 - Ammonia injection tried again with single injection nozzle and no carrier gas. No change in N2O. Ammonia slip at 10 ppm.

APPENDIX C

Permission granted to copy

Loop Seal for Pilot Plant
Report 902288

Jenike & Johanson, inc.

Riley Research

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Report on Loop Seal for Pilot Plant Riley Research

1. Introduction

Riley Research has a contract with the US Department of Energy to investigate an improved design of Riley's Multi-Solid circulating Fluidized Bed (MSFB) at their Worcester, Massachusetts pilot plant. The company would like to test a loop seal designed by Jenike & Johanson, Inc., which promises to be able to control the flow of solid through the system better than the present L valves.

2. Layout

The existing pilot plant is arranged with the cyclone and external heat exchanger (EHE) on the west side of the combustor. Vertical pipes below the EHE with L valves at their lower ends control the flow of solids.

The new layout will incorporate a new EHE with vertical tubes. It appears that the new layout with loop seals can fit into the west side of the combustor as before.

3. Process

The process operates at gas pressures as follows:

Top of cyclone	minus 12 inches water gage
Top of EHE	minus 5 inches water gage
Bottom of EHE	plus 43 inches water gage
Bottom of combustor	plus 35 inches water gage

During startup, all of the solids will bypass the EHE. The rate will be approximately 38,000 lb/hr. During steady state conditions the flows will be approximately 38,000 lb/hr through the EHE and approximately 4500 lb/hr bypassing the EHE (hot recycle).

Three loop seals are proposed. The upper loop seal, between the cyclone and EHE will be required to seal against a pressure differential of 7 inches w.g. Two lower loop seals, a "hot" loop seal to control the flowing solid that bypasses the EHE and a "cold" loop seal to control the flowing solid that passes through the EHE, will be required to seal against a pressure differential of 40 inches w.g.

October 22, 1990
(revised 10/31/90)

**Report on
Loop Seal for Pilot Plant
Riley Research**

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4. Model Tests

4.1 Loop Seal

Full scale models of both the loop seal and the splitter arrangement were made and tested. In order to reproduce the hydrodynamic conditions in the model at room temperature that will exist in the pilot plant at up to 1500°F it is necessary to scale the model using various dimensionless numbers.

Using the paper "Scaling the Hydrodynamics of Fluidized Bed Combustors with Cold Models: Experimental Confirmation" by L.R. Glicksman et al, published in the Proceedings of the 1987 International Conference on Fluidized Bed Combustion as a basis, the following approach was used:

Conditions affecting the operation of both the loop seal and the splitter are at incipient fluidization. The Reynolds number considers inertial forces and gas viscosity, and the Archimedes number considers particle and gas density, both of which are important at incipient fluidization. Scaling each separately is unsatisfactory and therefore, use was made of the Kunii-Levenspiel equation. This is combination of Reynolds and Archimedes numbers and predicts minimum fluidization velocity.

$$\frac{1.75}{\phi(\epsilon_{mf})^3} (Re_{mf})^2 + 150 \left(\frac{1 - \epsilon_{mf}}{\phi^2(\epsilon_{mf})^3} \right) Re_{mf} - Ar = 0$$

$$\text{where } Re_{mf} = \frac{D_p \rho_g U_{mf}}{\mu} = \text{Reynolds number at minimum fluidization velocity}$$

$$Ar = \frac{(D_p)^3 \rho_g (\rho_s - \rho_g) g}{\mu^2} = \text{Archimedes number}$$

ϕ = shape factor

ϵ_{mf} = minimum fluidized-bed voidage

ρ_s = particle density of fluidizing solids, lb/ft³

D_p = average particle diameter, ft

ρ_g = density of fluidizing gas, lb/ft³

U_{mf} = minimum fluidization velocity, ft/s

μ = viscosity of fluidizing gas, lb / (ft · s)

g = acceleration of gravity, 32.2 ft/s²

In order to ensure dynamic similarity it is necessary to have the ratio of solids to gas densities similar in the cold model and pilot plant. Ash and air densities in the pilot plant of 94 pcf and 0.024 pcf respectively were used based on the assumption that the average temperature at the loop seals will be approximately 1200°F. This dictates a solids bulk density of 294 pcf in the cold model. A sample of steel shot was obtained that matched the shape and particle size distribution of ash from Riley's MSFB combustor in operation at A.E. Staley. Using this material in a full scale model of the pilot plant ensures similarity in particle shape, geometric similarity in both the model and the material, and dynamic similarity in solids-to-gas density ratio.

Permeability tests were run on both the ash and the steel shot. The permeability constant (K_0) determined from the tests is a measure of the superficial gas velocity, when the gas pressure gradient acting through a bed of the solid equals the bulk density of the solids.

The ratio of the Kunii-Levenspiel minimum fluidization velocity to the permeability constant for each material was found.

Table A

	Staley Ash 1200°F	Steel Shot 72°F
U_{mf}	0.026 ft/sec	0.199 ft/sec
K_0	0.020 ft/sec	0.161 ft/sec
U_{mf}/K_0	1.28	1.24

This agreement is very good, and is close enough to allow measurements taken with the shot in the cold model to be directly used as if they were taken in the hot pilot plant.

The laboratory model was made of sheet metal in the loop seal part with clear plexiglas in the dipleg. The five groups of nozzles as shown in Figures 4, 5 and 6 were connected to five separate plenum chambers which could be independently controlled.

Experiments were first run without back pressure to determine if the loop seal would operate satisfactorily without air injection in the bottom of the curved part. While it is relatively easy to inject air at this point in the model, it is very difficult to do the same in the pilot plant. The air injection ports of the model were covered with porous plastic pieces to prevent them from becoming clogged with solid particles but in the pilot plant this approach is not available. Therefore, attempts were made to get all of the air injection into surfaces that are nearly vertical. This allows the angle of repose to limit movement of the solid back down the nozzle.

The tests showed that flow could be controlled without air injection in the bottom of the loop. Under these conditions, and especially at high solids flow rates the air injection at the wide part of the riser leg was ineffective in controlling the solids flow rate. However, air injected in the narrow part at the base of the riser section was very effective.

The model was set up so that the loop seal would feed solid into a container whose air pressure could be controlled. Figure 9 shows the arrangement.

Tests were run at conditions representing gas pressure differentials between the cyclone and the EHE (7 inches water gage) and between the top of the EHE and the combustor (40 inches water gage). Table B shows the air flow rates and equilibrium conditions found in the tests.

Table B

Pressure Differential Across Loop Seal (inches water gage)	Air Flow Through Nozzle Groups (SCFH)				
	Set 1	Set 2	Set 3	Set 4	Set 5
7	40	60	120	200	200
40	80	120	160	270	270

At the required solids flow rate of ash of approximately 40,000 lb/hr (125,000 lb/hr of steel shot) the air flow rates given in Table B allowed the model to run at steady state with a level of solid in the dipleg. For the 7 inch w.g. pressure differential the level was 2 ft above the discharge of the loop seal and for the 40 inch w.g. pressure differential, 4 ft. When solids flow into the top of the dipleg was stopped, solids continued to flow from the loop seal, but the rate decreased as the solids head differential decreased. The flow of solid eventually stopped when the difference in gas pressure balanced the difference in head of solid.

If flow of solid continued into the top of the dipleg but air flows to nozzle sets 1 and 2 were reduced, the flow of solid through the loop seal was reduced and the level of solid built up in the dipleg. If air flow to sets 1 and 2 was turned off, solids flow out of the loop seal stopped.

If the level of solid in the dipleg was above the equilibrium point, and air flows to nozzle sets 1 and 2 were set at the rates given in Table B, the flow rate of solid through the seal would increase. The solids flow rate was higher than approximately 40,000 lb/hr (of ash) and slowly decreased until the equilibrium position was reached.

4.2 Splitter

In order to split the stream of flowing solids to either enter, or bypass, the EHE a splitter design using the angle of repose of the ash as a mechanism was built and tested. The angle of repose of a bulk solid is a function of, among other things, its degree of fluidization. As the bulk solid becomes more fluidized its angle of repose decreases.

The design shown in Figures 7, 8 and 10 allows the splitter to control the flow streams to the two legs. It is necessary to have independent control of the nozzle at the center of each terrace. If all three nozzles on each terrace are attached to a common plenum chamber, the air takes the path of least resistance through the two side nozzles where the depth of solid is lower.

The air flow rates through the nozzles as shown in Table C were established with steel shot falling onto the upper portion of the splitter. The air flow rates required in the pilot plant may be lower than these rates for two reasons: Firstly, the solid exiting the loop seal will be more aerated and secondly, the viscosity of air at 1200°F is higher than room temperature air. The ratio of densities of the air at the different conditions was accounted for in the model, but not viscosity.

Table C

Solids Flow Rate lb/hr	Air Flow to Splitter (each step)	
	Air Flow to Both Outer Nozzles (SCFH)	Air Flow to Central Nozzle (SCFH)
4,500	40 (20 each)	30
38,000	80 (40 each)	65

It was relatively easy to control the flow rate of bulk through the splitter model by controlling air flow rates to the nozzles.

The air flow to the lower nozzles had more control over the flow rate of solid than the air to the upper nozzles. For this reason we feel that the viscosity effect will be easy to control, if it is a factor that appears in the pilot plant. If the higher viscosity of high temperature air in the pilot plant makes the ash stay fluidized longer than the steel shot in the cold model, it may be necessary to reduce the amount of air flowing to the upper nozzles. Lower air flows to the upper nozzles will not reduce the effectiveness of the splitter and solids flow will continue to be controlled by the air flow directed to the lowest nozzles.

It is expected that actual air flow rates required to control the solids will be fixed by some experimentation in the field.

Figure 1
Loop Seal for Pilot Plant

Note:
Design of Loop Seal is
proprietary to Jenike
& Johanson, Inc.
Patent pending.

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Design of Loop Seal is
proprietary to Jenike
& Johanson, Inc.
Patent pending.

Figure 3
South Elevation

Figure 4
Loop Seal for Pilot Plant

Note:
Design of Loop Seal is
proprietary to Jenike
& Johanson, Inc.
Patent pending.

Front View

Figure 5
Loop Seal for Pilot Plant

Note:
Design of Loop Seal is
proprietary to Jenike
& Johanson, Inc.
Patent pending.

Back View

Figure 6
Loop Seal for Pilot Plant

Figure 7
Details of Splitter

Nozzles in each step have two rotometers to control air flow connected as shown.

Figure 8
Details of Plumbing to Splitter

APPENDIX D

LARGE PARTICLE FLUIDIZATION

Technical Report

Prepared for

Riley Research Center

45, Mckeen Road

Worcester, MA 01610

Reference No: P.O. W-236600-87805 dated

October 5, 1988

V.K. Mathur and J.Talukdar

Department of Chemical Engineering

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Durham, NH

September 15, 1991

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1 INTRODUCTION

The circulating fluidized bed technology has made inroads into the combustion industry since the late 1970s, and appears to have excellent prospects in the future. The success of the circulating fluidized bed (CFB) boiler is due to the many advantages it offers over conventional bubbling fluidized beds and pulverized combustors. Among them are its superior carbon burnout efficiency, a lowered temperature operation resulting in less nitrogen oxide emissions, efficient sulfur capture, excellent turn-down capabilities, and the ability to burn a variety of fuels. However, there is a lack of understanding of fundamental hydrodynamics and heat transfer phenomena. This weakness in knowledge manifests itself in scale-up from pilot plant operation to a full scale combustor. The CFBs are relatively new and their behavior and operation need further study. Prediction of regimes of flow, of the mixing of gas and solids and of heat transfer coefficients thus becomes considerably more difficult.

The Battelle technology utilizes a multi-solids fluidized bed (MSFB) design with a dense bed of bubbling inert materials superimposed on a circulating bed of fine particles. This dense bed acts as a resistance to the fast moving circulating coal particles, slowing them down by the repeated collisions and friction. This allows for increased residence time of the fines resulting in more complete combustion. The Battelle MSFB is a special case of the CFB systems. Understanding the hydrodynamics of this system was the primary objective of this study with particular emphasis on the bubbling(dense) bed region of the combustor.

The hydrodynamic behavior of any fluidized bed may be viewed and analyzed from a gross(or macroscopic) viewpoint and a fine(or microscopic) scale. The gross behavior of a fluidized bed is characterized by the pressure drop across the bed, the average bed voidage, and the bed expansion. The fine behavior of the bed is typically an analysis of the movement of gas and solids in a microscopic sense. Foremost in such a description is the bubble behavior, characterized by the three bubble properties - bubble rise velocity, bubble diameter(or length), and bubble frequency.

This study has been undertaken with the objective of characterizing the bubbling dense bed section of a MSFB. The primary focus has been towards understanding the microscopic behavior of the circulating fine particles as they passed through the dense bed of coarse particles. The significance of this study is twofold. From a purely academic point of view, there are a few data/model available to analyze the behavior of coarse particle systems, particularly those systems which have fine solids passing through. From an industrial point of view, the Battelle design utilizes a coarse particle system with fast circulating fines passing through. Thus an enhancement in the knowledge of the hydrodynamic behavior of a coarse particle system with fines passing through, would represent important progress, both in academia as well as in industry.

2 LITERATURE REVIEW

Since the Battelle MSFB system consists of both a bubbling fluidized bed and a pneumatic transport bed, a literature search has been conducted with the objective of reviewing the existing hydrodynamic information relating to both the areas. Literature pertinent to solids recirculating devices is also reviewed, as part of this study is devoted towards characterizing such a device. Detailed literature review is available elsewhere [1].

3 THEORY AND PROPOSED MODEL DEVELOPMENT

Of all the current CFBs in operation, the multi-solids fluidized bed is a unique process developed by the Battelle Memorial Institute, Columbus, OH. As shown in Fig. 1, it consists primarily of a dense bed of inert large particles superimposed on a circulating fluidized bed of fine particles. The coarse particles remain at the bottom of the riser section and serve a number of purposes. They tend to grind the larger lumps of freshly fed coal and also act as a zone of intense gas-solids mixing for efficient combustion. The superficial gas velocity is maintained such that it far exceeds the terminal velocity of the fines; it is, however, greater than the minimum fluidization velocity of the large inert particles but less than their terminal velocity. Thus, the fines entrain once they pass through the dense bed. The coarse particles remain in a bubbling mode at the lower end of the riser section.

The primary objective of this investigation is to understand the mechanics of flow behavior of the fine particles as they pass through the bubbling bed of coarse solids. Flow behavior is characterized by bubble properties of the dense bed. Solids residence times are measured and a mechanistic model developed based on the gas-fine particle-coarse particle interactions.

3.1 Models for Bubble Properties in the Dense Bed

Till the present date, there is no available model which adequately describes bubble characteristics of a coarse particle bubbling bed with fines circulating through it. There are a few empirical models that predict bubble diameter, bubble rise velocity, and bubble frequency in coarse particle bubbling beds. These models have been reviewed and their performance has been numerically evaluated with the data from this study [2]. It has been found that these empirical models deviate from experimentally measured values by as much as 400 % in their prediction of bubble length and bubble velocity, when applied to a system of coarse particles with fines passing through them. As a result of the current investigation there are substantial data available on bubble parameters. These bubble characteristics portray the microscopic behavior of the bed and are indicative of the quality of fluidization.

3.2 Models for Residence Time of the Fine Particles in the Dense Bed

An exhaustive literature search shows only one available model for the prediction of residence time of fine particles through a coarse particle bubbling bed [3]. An earlier work by the same researcher resulted in the basic formulation of the momentum balance on the fines upon which the later model is based.

3.3 Rationale for the choice of Fluidizing Materials - Scaling Law Hypothesis

Three fine particle-coarse particle systems have been chosen for this study. The fluidizing medium in each case is air at ambient temperature and pressure. The systems are sand circulating through rocks, sand through spherical alumina balls, and copper dust through steel shots. The first two systems are materials used in an actual combustor, operating at elevated temperatures. The last system (copper-steel) is a result of the scaling law hypothesis as proposed and experimentally verified by Nicastró et al. [4]. The laws are derived from the non-dimensional equations of conservation of mass and momentum, of both the particulate and the gas phase. These non-dimensional continuity and Navier-Stokes equations when coupled with Ergun's equation of flow through a porous medium, result in a set of dimensionless 'scaling groups'. The following groups are important for scaling of gas-solids fluidized bed : Re_p , the particle Reynolds number, $\frac{\rho_s}{\rho}$, the ratio of solid density to gas density, Ar , the Archimedes number, Fr , the Froude number, $\frac{H}{d_p}$, and $\frac{D}{d_p}$. H and D are the column height and diameter, respectively and d_p is the mean particle diameter. In the scaling method, the cold flow model is so designed that the model values of these scaling groups are similar to the values for the full scale hot combustor. In this way the gas and solids inertia, buoyancy, and viscous drag forces are balanced between the cold model and the full scale combustor [5].

3.4 Development of a Proposed Mathematical Model to Predict Residence Times of Fine Particles in the Dense Bed Section of a MSFB

Residence time data for fines circulating through the dense bed are experimentally obtained at the UNH unit, using a defluidization technique. These data did not check with the Fan et al. expression. The single most important reason for this anomaly is the Fan model contains no explicit dependance of circulating solids mass flux in the velocity expression (momentum balance). The experimental finding of this study clearly shows such a dependance, hence the need to incorporate this in the model.

Thus, a mathematical model is proposed to account for the behavior of fine particles circulating through the dense bed. The model is based on a force balance approach and

Figure 1: Flow of Fines in a Bed of Coarse Particles

follows a development similar to that of Fan et al. [3,6].

3.4.1 Model Development

Consider the flow of gas and fine particles through a bubbling (dense) bed of coarse solids as shown in Figure 1. A force balance on the fines across the differential length dL may be represented as:

$$dm_2 \frac{dU_{p2}}{dt} = dF_{drag} - dF_{gravity} - dF_{friction} - dF_{collision} \quad (3.1)$$

Here, dm is the effective weight of the fine particles in the section dL . Subsequently, subscripts 1 and 2 will refer to the coarse and fine particles, respectively. Thus, U_{p2} refers to the velocity of fine particles as they move upwards through the dense bed.

It has been noted by Fan et al.[4] that fine particles quickly achieve their steady state velocity within a fraction of a millimeter from being introduced into the dense bed. Thus the term $dm_2 \frac{dU_{p2}}{dt}$ is equal to 0. Therefore equation (3.1) becomes a balance between the upward drag force and the downward gravitational, frictional and collision forces.

$$dF_{drag} - dF_{gravity} - dF_{friction} - dF_{collision} = 0 \quad (3.2)$$

Considering the individual terms in equation(3.2) in more detail, dF_{drag} is the drag force, acting on all the fine particles by the fluidizing medium(air). This is also equal to the sum of the drag forces acting on one single particle. This may be represented as:

$$dF_{drag} = \frac{\frac{3}{4} C_{DM} \rho_a (U_a - U_{p2})^2}{\rho_{p2} d_{p2}} \Delta m_2 \quad (3.3)$$

where, $U_a = \frac{U_a}{\epsilon}$, and C_{DM} is the multi-particle drag coefficient.

Yang [7] suggested a relationship between C_{DM} and C_{DS} , the single-particle drag coefficient as follows:

$$C_{DM} = \epsilon^{-4.7} C_{DS} \quad (3.4)$$

In this model, the void fraction used is ϵ_f , which is the void fraction in the phase containing primarily fine particles. Thus,

$$C_{DM} = \epsilon_f^{-4.7} C_{DS} \quad (3.5)$$

where,

$$\begin{aligned} C_{DS} &= \frac{24}{Re_{p2}}, Re_{p2} \leq 2 \\ &= \frac{18.5}{Re_{p2}^{0.6}}, 2 \leq Re_{p2} \leq 500 \\ &= 0.44, Re_{p2} \geq 500 \end{aligned}$$

The gravitational force term is

$$dF_{gravity} = g(1 - \frac{\rho_a}{\rho_{p2}}) dm_2 \quad (3.6)$$

The frictional term accounts for the friction between small particles and large particles, between small particles themselves, and between small particles and the wall;

$$dF_{fric} = dF_{fric,1-2} + dF_{fric,2-2} + dF_{fric,2-wall} \quad (3.7)$$

Breault [8] developed an expression to account for these effects. However, his expression has been modified here to account for the different physical system. The mass-flux dependence of velocity has been included in the friction term, which gives it a certain degree of empiricity. The above individual frictional effects have been combined, resulting in the following expression for the frictional force term:

$$dF_{fric} = 10 \left(\frac{1-\epsilon}{\epsilon} \right)^n U_{p2} D \frac{1}{\phi d_{p2}} \left(\frac{W_g}{W_{p2}} \right)^p dm_2 \quad (3.8)$$

where $\frac{W_g}{W_{p2}}$ is the ratio of mass flux of gas to small particles, p is a fitting parameter, ϕ is the sphericity, and ϵ is the void fraction of the dense bed, and n is a system-dependent

parameter that can be evaluated based on system properties. A discussion of n and p follows later on, in the Results and Discussion chapter of this thesis. Note, ϵ is not equal to ϵ_f .

An expression for the collision force has been derived by Marble [9], from first principles, considering elastic collisions between small and large rigid spherical bodies. The collision force term is a major addition to the ideas presented in earlier models, and enhances the understanding of the fine particle flow behavior through the dense bed. This term implies momentum changes to the fine particles as a result of elastic collisions between the small and large particles. The original expression is modified here to include the effect of mass flux:

$$dF_{\text{collision}} = \frac{6(U_{p1} - U_{p2})^2(1 - \epsilon)}{\phi d_{p1}} \left(\frac{W_g}{W_{p2}} \right)^p dm_2 \quad (3.9)$$

Now, collecting all the terms and substituting in equation(3.2), and noting that dm_2 cancels out, we have:

$$\begin{aligned} \frac{\frac{3}{4} C_{DS} \epsilon_f^{-4.7} \rho_a (U_a - U_{p2})^2}{\rho_{p2} d_{p2}} - g \left(1 - \frac{\rho_a}{\rho_{p2}} \right) - 10 \left(\frac{1 - \epsilon}{\epsilon} \right)^n U_{p2} D \frac{1}{\phi d_{p1}} \left(\frac{W_g}{W_{p2}} \right)^p \\ - \frac{6(U_{p1} - U_{p2})^2(1 - \epsilon)}{\phi_1 d_{p1}} \left(\frac{W_g}{W_{p2}} \right)^p = 0 \end{aligned}$$

or,

$$\begin{aligned} \frac{\frac{3}{4} C_{DS} \epsilon_f^{-4.7} \rho_a (U_a - U_{p2})^2}{\rho_{p2} d_{p2}} &= g \left(1 - \frac{\rho_a}{\rho_{p2}} \right) \\ &+ 10 \left(\frac{1 - \epsilon}{\epsilon} \right)^n U_{p2} D \frac{1}{\phi d_{p1}} \left(\frac{W_g}{W_{p2}} \right)^p \\ &+ \frac{6(U_{p1} - U_{p2})^2(1 - \epsilon)}{\phi_1 d_{p1}} \left(\frac{W_g}{W_{p2}} \right)^p \end{aligned}$$

For any specified small particle-large particle-gas system, the system-dependent parameters n and p may be evaluated. The above equation can then be solved for the small-particle velocity, U_{p2} . Once U_{p2} is determined, the residence time in seconds per meter of the dense bed can be determined from

$$\tau_r = \frac{1}{U_{p2}} \quad (3.10)$$

if U_{p2} is in m/sec. The model described above has been put in the form of a computer code, for the evaluation of residence times with varying independent parameters. The computer code is presented in reference [1].

4 EXPERIMENTAL APPARATUS AND PROCEDURE

The study was conducted in several phases. The first stage consisted of building a bench scale two-dimensional model and visually observing the hydrodynamic behavior of bubbling coarse particles. This was followed by the design, fabrication and operation of 0.038m (1.5") and 0.102m (4") cold circulating fluidized bed units with the objective of studying bubble dynamics. The fluctuating pressure signal technique was used to obtain bubble characteristics. An on-line data acquisition system with associated software was utilized for the purpose. The next phase consisted of obtaining data from a pilot unit which had a very similar data acquisition system. This consisted of a 0.229m x 0.229m (9" x 9") square cross section and was used to obtain bubble characteristics data. As a final step in trying to understand the dynamics of the dense bed region, an in-depth study was conducted to obtain fine particle residence time data. A defluidization technique was used in a 0.102m unit to get residence time values of fine particles in the dense bed. This is of critical importance for the MSFB boiler, as it is a measure of how long burning coal particles stay in the dense bed section of the combustor.

4.1 The 0.102m (4") UNH Circulating Fluidized Bed Experimental Unit

A cold circulating fluidized bed unit was designed, fabricated and operated at the UNH facility (Figure 2). This experimental unit was constructed primarily of cast plexiglass tubing and was designed to operate at atmospheric pressure and temperature. The system consisted of an Ingersoll-Rand air compressor as the primary source of fluidizing air. This was a reciprocating machine equipped with an 80 gallon tank capable of supplying $0.019 \text{ Sm}^3/\text{s}$ (2400 SCFH) of air at 446.1 kPa (50 psig). The main air line from the compressor passed through a pressure regulator, whose outlet pressure was adjusted to about 274 kPa. This system resulted in minimal air pressure fluctuations. Air was supplied to the bottom of the riser section, below the distributor, DI. Air was also supplied to the L-valve in the form of motive and assist air. The L-valve is a non-mechanical device to recirculate solids. It can also be used to measure solids flow rate; in this phase, however, it was used primarily as a solids recirculating device. Details of the L-valve aeration are provided in the results and discussion section. Additional air was sometimes needed for effective fluidization of a large bed inventory of steel shots. This was provided by a Rotron manufactured cyclonair positive-displacement blower, capable of delivering $0.033 \text{ Sm}^3/\text{s}$ (4200 SCFH) at 12.16 kPa pressure. The solids loop consisted the riser section, R, a gas-solids disengaging device in the form of a cyclone, CY, a downcomer, DO and a solids return device (L-valve).

The principal region under investigation was the dense bed, DB. This section, was

Figure 2: The UNE Circulating Fluidized Bed Unit with Data Acquisition System

located at the lower end of the riser. The riser diameter could be changed from 0.038 m to 0.102 m ; columns of both diameters were tested using a variety of solid materials. The pressure probes consisted of $3.175 \times 10^{-3} \text{ m}$ (1/8") diameter steel tubes, that were 0.152 m (6") long and 'snubbed' with glass wool at the active end. Four such probes were inserted into the dense bed such that their tips lay on the central axis of the riser. A rib attached to the wall of the plexiglass column by glue avoided direct attachment of the probes onto the bed wall, and helped maintain their rigidity. The other ends of the probes were connected by $6.35 \times 10^{-3} \text{ m}$ (1/4") flexible tygon tubing to high speed pressure transducers. A battery of 4 transducers was mounted close to the pressure probes to minimize pressure signal lag time. Details of other units and hardwares are available elsewhere [1].

5 RESULTS AND DISCUSSIONS

A summary of the results pertaining to the bubble properties and residence time of fine particles in the dense bed of a CFB is presented here. The readers are requested to refer to reference [1] for details.

5.1 Bubble Properties in the Dense Bed of a CFB

The passage of gas bubbles causes pressure fluctuations in a fluidized bed. This property has been utilized to characterize the bubbles with respect to size, velocity and frequency. Size of a bubble has been defined as the 'pierced length of bubble', velocity as the 'rise velocity of the bubble' and frequency as the number of bubbles passing a given point in the bed per second. Characteristics of the gas bubbles control the performance of the fluidized bed. Size, velocity and frequency control solids mixing, convection and diffusion of gas components (reactants and products), elutriation of fine particles and attrition of large particles. Hence, it is of utmost importance to measure bubble properties and establish their dependence on the major operating parameters. Large particles, when fluidized, typically give rise to slow moving bubbles; the passage of circulating fine particles through coarse solids however changes bubble characteristics considerably. The objective of this part of the study was to quantify such changes in bubble properties with independent variables such as solids mass flux.

Data were collected from three experimental units - a 0.038m (1.5") diameter column at the UNH facility, a 0.102m (4") diameter column also built and operated at UNH, and a 0.229m (9") square fluidizing column at the Riley Stoker Corporation in Worcester, Mass. All three equipment were 'cold' circulating fluidized bed units. Two air-solids systems were primarily used in the three set-ups : copper dust circulating through steel pellets and sand circulating through rocks. The first system was based on a 'scaling law' analysis and

simulated the conditions of a hot combustor when fluidized with air at room temperature. The second system consisted of materials actually used in a hot CFB, but operated at room temperature. Thus, data from the copper-steel system can be viewed as that resulting from the operation of a *HOT* unit, and the sand-rock system data as that from a *COLD* unit using materials used in the combustor. Nicastro et al. [4] have shown using experiments and a statistical analysis technique that the 'scaling law materials' yield the pertinent data; this operation using scaling law materials in a cold unit is similar to the operation of a hot unit. If the hot bed materials are used in a cold flow unit, erroneous conclusions are inferred. Thus the copper-steel data of bubble characteristics and residence time of fines in the dense bed truly represent the behavior of an actual hot combustor. Additional rationale for the choice of these systems is described elsewhere [1].

5.2 Bubble Characteristics Data from Experimental Units

The bubble properties in the dense bed were measured using the pressure fluctuation technique, as described earlier. Bubble length in this study was estimated as 'the pierced length of bubble'; subsequently, pierced length of bubble will be simply referred to as 'bubble length' or 'length of bubble'. Bubble rise velocity was calculated from the pressure signals as the time for the pressure signal from the bubble to traverse unit distance. Bubble frequency was the number of bubbles passing the pressure probe per unit time. Experiments were conducted for the following cases to obtain bubble parameters.

Table 1: Bubble Characteristics from Experimental Units

Column <i>m</i>	Solids Systems —	Air Velocity <i>m/s</i>	Air Mass Flux <i>kg/s * 10³</i>	Bubble Parameters —
0.038 (<i>UNH</i>)	<i>Copper — steel</i> <i>Sand — rock</i>	3.62-4.48 3.19	4.13-5.1 4.41	<i>Length, Velocity, Frequency</i> <i>Length, Velocity, Frequency</i>
0.102 (<i>UNH</i>)	<i>Copper — steel</i> <i>Sand — rock</i>	3.62-4.48 3.19	4.13-5.1 4.41	<i>Length, Velocity, Frequency</i> <i>Length, Velocity, Frequency</i>
0.229 (<i>RileyStoker</i>)	<i>Copper — steel</i> <i>Sand — rock</i>	3.62-4.48 3.19	4.13-5.1 4.41	<i>Length, Velocity, Frequency</i> <i>Length, Velocity, Frequency</i>

Figures 3 through 6 are comparative plots of bubble length and bubble rise velocities for both solids systems from all three units. Figures 3 and 4 show bubble lengths and bubble velocities versus solids mass flux, respectively, for the copper-steel system. The bubble characteristics data for the 0.038m *UNH* unit appear considerably higher, as compared to

Figure 3: Bubble/slug Pierced Length, Copper-steel System : All Units

Figure 4: Bubble/slug Rise Velocity, Copper-steel System: All Units

Figure 5: Bubble/slug Pierced Length, Sand-rock System : All Units

Figure 6: Bubble/slug Rise Velocity, Sand-rock System: All Units

the other units. Significant effects of the wall in terms of increased slug lengths (growth of the bubble in the radial direction was severely restricted because of the narrow diameter) and increased velocities are considered the cause for this deviation. Figures 5 and 6 show that the 0.229m data are completely out of any banded zone of data fluctuations. The values of both bubble length and bubble rise velocity are an order of magnitude higher than those from the 0.038m and 0.102m units. The possible explanation for this can be that the data for the sand-rock system in the 0.229m unit was taken at an extremely low solids mass flux range. Sand mass flux values ranged from 0.18 to 3.58 kg/sq.m-s. Visual observation showed a freely bubbling regime of fluidization, quite different from the other modes of operation. The escape mechanism for the gas was primarily via bubbles and not slugs. This could be a faster process, since the time it took for the bubbles to coalesce and form the slug over the entire diameter of the column, then expand and rise, was eliminated.

There was an increase in bubble length as solids mass flux was increased. This can be explained if one analyzed the escape mechanism of air and fine particles from the dense bed. As the solids mass flux was increased the dense bed region expanded. More fluidizing air accumulated between regions of high concentration of solids. Ultimately this region of increased fluidizing air (constituting a larger bubble) escaped out of the bed. Thus as solids mass flux increased the bubble length increased.

The bubble rise velocity increased as solids mass flux increased. This occurred because solids moved faster as solids mass flux was increased (solids continuity). The regions of faster rising solids had accompanying regions of voids interspaced, that also moved with greater velocity. Bubble frequency remained fairly constant with increasing solids mass flux at a value in the range of 1.75 to 2. These characteristics of the bubble appeared to be set by the mode of fluidization, which was fixed by the nature of the solids and the fluidizing gas.

All the pressure traces and detailed figures for 0.038m, 0.102m, and 0.229m units for both copper-steel and sand-rock systems are available elsewhere [1].

5.3 Residence Time of Fine Particles in the Dense Bed and Proposed Model

Solids residence time and solids holdup were two parameters that were experimentally measured with varying dense bed void fraction and mass flux of circulating fine particles. Attempts are made to explain the findings based on physical phenomena. A mechanistic model to reflect the flow behavior of the fine particles is proposed and verified against the experimental data. The residence time data resulting from the copper-steel system reflected the performance of the dense bed section of an actual hot combustor in its ability to retain circulating fine particles. The sand-rock and sand-alumina systems are the materials used in the hot unit; however, when they are used in a cold fluidized bed set-up the resulting

data do not simulate a hot operating unit.

5.3.1 Experimental Data Acquisition

Residence time data were obtained from the bench-scale circulating fluidized bed, using the defluidization technique. Three different systems were selected; copper fines circulating through steel balls, sand through rocks and sand through alumina. In all systems the fluidizing medium was air. The mass flux of circulating solids were varied by air inputs to the L-valve. Measurements of this mass flux were made by timing the descent of particles in the downcomer. Dense bed void fraction was varied by changing the bed inventory of coarse solids and fluidizing them with a constant amount of air.

5.3.2 Experimental Residence Time Values

Figure 7 shows the variation in residence time with expanded dense bed voidage for the copper-steel system, for three mass fluxes namely, 44.65, 63.78 and 89.56 kg/sq.m-s. In all three cases the residence time decreased from about 13.12 to 1.64 sec/m of expanded bed, as the void fraction in the dense bed was increased from about 0.68 to 0.82. This decrease in residence time with increasing dense bed voidage is similar in all three systems.

5.4 Development and Verification of the Model

The model as discussed earlier describes a force balance on the fine particles as they pass through a bubbling bed of coarse solids. The drag force is the upward accelerating force, and gravity, friction and collision are the downward retarding forces. The solution of this steady-state force balance yields fine particle velocities. A certain degree of empiricity is introduced in this mathematical model. The friction term includes an empirical constant 'n' which has been evaluated for the three systems using a regression analysis; data used in the determination of 'n' was obtained experimentally as previously discussed. The index 'p' to which the ratio of particle to air mass flux is raised, in both the friction and the collision terms, has also been determined using regression analysis of experimental data. It is found to be 0.25. As a result of this analysis the values of 'n' for the copper-steel, sand-alumina and rock-sand system are 2.9, 0.8 and 2.5, respectively. For this set of 'n' values and the mass flux ratio index of 0.25, the values of AAPD are calculated. The Absolute Average Percentage Deviation (AAPD) is defined as:

$$AAPD = \frac{1}{n} \sum_{i=1}^n \left(\frac{|\tau_{r,exp} - \tau_{r,cal}|}{\tau_{r,exp}} \right) 100\%$$

Here, τ_r is the solids residence time in sec/m of dense bed. The AAPD values for the various systems are listed in Table 2.

Figure 7: Dense Bed Solids Residence Time, 0.102m UNH Unit, Copper-steel System

Table 2: AAPD Values for Copper-steel, Sand-alumina and Sand-rock Systems

Serial No	System	Solids Mass Flux $\frac{kg}{m^2 \cdot s}$	Expanded Bed Voidage	AAPD %
1	Cu-steel	21.71	0.72-0.80	20.0
2	Cu-steel	44.65	0.68-0.81	29.2
3	Cu-steel	63.78	0.68-0.81	24.4
4	Cu-steel	89.56	0.68-0.81	21.5
5	Alu-sand	11.63	0.73-0.94	24.8
6	Alu-sand	16.28	0.73-0.93	22.7
7	Alu-sand	22.20	0.73-0.93	85.2
8	Roc-sand	10.67	0.67	2.59
9	Roc-sand	24.04	0.67	27.5
10	Roc-sand	10.67	0.72	9.48
11	Roc-sand	15.58	0.72	9.45
12	Roc-sand	17.44	0.72	17.39
13	Roc-sand	20.94	0.72	97.03
14	Roc-sand	15.27	0.77	9.59
15	Roc-sand	26.6	0.77	8.47

The model generated curves for the copper-steel and the alumina-rock system are shown in Figures 8 and 9, respectively. In Figure 8, the model traces for three mass fluxes are shown, as the solid lines, with increasing dense bed voidage. There is a reasonable agreement between experimental data and model generated values with AAPD values ranging from 20 to 29.2. Figure 9 shows a greater deviation between experimental and model values with AAPD values ranging from 24.8 to 85.2. This information is also presented in Table 2. The AAPD values for the copper-steel system show less variation as compared to the sand-alumina system. The sand-rock data have been used as an independent check for the model. The AAPD values for this system range from 2.59 to 97.03. This appears to validate the mathematical model, since these data are not used in the model development. The high value of 97.03 may be an experimental error.

5.5 Development of Empirical Equation for Predicting 'n' Values

The empirical constant 'n' of the model previously described together with the other variables such as sphericity, density of particles, and particle size, are summarized in Table 3.

Table 3: Variables Used in the Model and Resulting 'n' Values

Variable	Copper-steel	Sand-alumina	Sand-rock
ϕ_1	1.0	0.95	0.6
ρ_1 [kg/m ³]	7900	3400	2800
ρ_2 [kg/m ³]	8900	2600	2600
d_{p1} [m]	3.175×10^{-3}	6.35×10^{-3}	9.525×10^{-3}
d_{p2} [m]	80×10^{-6}	300×10^{-6}	300×10^{-6}
n	2.9	0.8	2.5
p	0.25	0.25	0.25

The overall mathematical model thus becomes:

$$\frac{\frac{3}{4} C_{D, \epsilon}^{-4.7} \rho_a (U_a - U_{p2})^2}{\rho_{p2} d_{p2}} = g \left(1 - \frac{\rho_a}{\rho_{p2}} \right) + 10 \left(\frac{1 - \epsilon}{\epsilon} \right)^n \frac{U_{p2}}{D} \frac{1}{\phi_1 d_{p1}} \left(\frac{W_g}{W_{p2}} \right)^{0.25} + \frac{6 (U_{p1} - U_{p2})^2 (1 - \epsilon)}{\phi_1 d_{p1}} \left(\frac{W_g}{W_{p2}} \right)^{0.25}$$

The usefulness of such a model is enhanced if the model can be applied to any independent system operating under different conditions and parameters than those used in this

Figure 8: Model Traces of Residence Time Values, UNII 0.102m Unit, Copper-steel System

Figure 9: Model Traces of Residence Time Values, UNII 0.102m Unit, Sand-alumina System

study. An attempt has been made to correlate the empirical constant 'n' with such system-dependent parameters like particle diameters, particle densities etc. Some of these relevant parameters that influence the empirical constant 'n' have been collected in dimensionless groups and a regression analysis has been performed. The results of analysis are tabulated in Table 4. A linear variation of 'n' with the diameter-to-density ratio of fine particle/large

Table 4: Characteristics Groups and 'n' Values

System	Fine Particle Characteristic Group $\frac{\phi_2 d_{p2}}{\rho_2}$	Coarse Particle Characteristic Group $\frac{\phi_1 d_{p1}}{\rho_1}$	Dimensionless Group Ratio $\frac{\frac{\phi_2 d_{p2}}{\rho_2}}{\frac{\phi_1 d_{p1}}{\rho_1}}$	n Values
—	—	—	—	—
<i>Copper - steel</i>	$8.988 * 10^{-9}$	$4.0189 * 10^{-7}$	0.0224	2.9
<i>Sand - alumina</i>	$1.154 * 10^{-7}$	$1.774 * 10^{-6}$	0.065	0.8
<i>Sand - rock</i>	$1.154 * 10^{-7}$	$2.041 * 10^{-6}$	0.056	2.5

particle is observed. The resulting expression for 'n' in terms of the fine and coarse particle characteristics is found to be:

$$n = -32 \frac{\frac{\phi_2 d_{p2}}{\rho_2}}{\frac{\phi_1 d_{p1}}{\rho_1}} + 3.6 \quad (5.11)$$

This linearity will facilitate the use of the model. If the system parameters are known, 'n' can be determined from the correlation. Hence, fine particle residence time can be calculated using the model.

6 CONCLUSIONS

The major findings of this study are:

- (1) Two bench scale CFB systems were built and operated at the UNH facility, and a pilot scale model was operated at the Riley Stoker facility. All units were operated in 'cold'. The three units were 0.038m, 0.102m, and 0.229m in effective diameter, respectively. These three units were successfully used to collect data of bubble length, bubble rise velocity and bubble frequency.
- (2) An on-line data acquisition system was successfully used to obtain pressure-time signals (traces). The pressure-fluctuation technique was used to analyze these signals. The technique essentially consisted of estimating bubble characteristics (length, velocity and frequency) from the separation distance of the peaks and troughs of the

sinusoidal signals. It was found to be a robust technique and accurately describe the bubble dynamics of the dense bed, that could be used for a wide range of solid loadings and a variety of materials (including conducting powders).

- (3) It was found that bubble properties, were dependent on the column diameter, with the smallest typically giving rise to the longest and fastest bubbles. Bubble frequency remained almost independent of column diameter. The bubble data from the 0.038m UNH unit suffered from 'wall effects'.
- (4) The materials chosen for study were based on the scaling law hypothesis. As a comparison, the actual materials used in the combustor were also used. This resulted in the use of copper-steel, sand-rock, and sand-alumina systems. These systems were used for the bubble characteristic and/or the residence time studies. An analysis of bubble data showed a striking difference between the values obtained from the different systems. Since the copper-steel system actually simulated the materials in a hot combustor, those data may be used to design a coal-burning MSFB.
- (5) The solids residence times were determined by a defluidization technique for various gas-solids systems in the 0.102m UNH unit.
- (6) A mathematical model to predict the residence time of the fines in the dense bed was developed from first principles, based on a momentum balance approach. The friction and collision terms appeared separately. The effect of solids mass flux was incorporated, improving upon existing model(s). The following model has been proposed for predicting the residence time of fine particles flowing through a bubbling fluidized bed of coarse solids:

$$\frac{\frac{3}{4} C_D \epsilon_f^{-4.7} \rho_a (U_a - U_{p2})^2}{\rho_{p2} d_{p2}} = g \left(1 - \frac{\rho_a}{\rho_{p2}}\right) + 10 \left(\frac{1 - \epsilon}{\epsilon}\right)^n U_{p2} D \frac{1}{\phi d_{p1}} \left(\frac{W_g}{W_{p2}}\right)^{0.25} + \frac{6(U_{p1} - U_{p2})^2}{\phi d_{p1}} \left(\frac{W_g}{W_{p2}}\right)^{0.25}$$

The model developed here employs a semi-empirical constant, n . An empirical equation was developed relating n and system dependent parameters such as fine and coarse particle properties. Thus, once a physical system was chosen the n may be evaluated, and the model may be used to predict residence time values of the fine particles in the dense bed. The empirical equation for estimating n is :

$$n = -32 \frac{(\phi d_p / \rho)_2}{(\phi d_p / \rho)_1} + 3.6$$

- (7) Residence time values as predicted by the model were compared with experimental data. The AAPD values range from 20 to 29.2 for the copper-steel system, from 22.7 to 85.2 for the sand-alumina system, and from 2.59 to 97.03 for the sand-rock system.

7 RECOMMENDATIONS

The following recommendations are made:

- (1) A better picture may emerge if more sophisticated high-speed photography is employed in a two-dimensional setup. The images of bubbles erupting, of the fines short circuiting through the slow-rising bubbles, and the nature of gas-solids mixing, may be recorded and analyzed to give further insight into the dense bed dynamics.
- (2) The pressure fluctuation technique is a robust method for obtaining bubble dynamics in a fluidized bed. The data collected in this study were from cold units using scaling law materials. Efforts should be made to acquire bubble data from a hot operating MSFB using the pressure fluctuating technique, and the signals should be compared to those obtained from a cold flow unit using both scaling law materials and actual materials used in the hot combustor. The comparisons may be made based on statistical properties of the signal. This would also verify the validity of scaling law materials in a MSFB.
- (3) As of this date, no model exists that predicts bubble behavior in a coarse particle system with fines passing through it. Models should be developed to predict bubble length, bubble velocity and bubble frequency in such a system. Data from the present study may be used to validate the models.
- (4) Efforts should be made to obtain residence time data from a hot pilot-plant scale combustor, and compare them with data from the present study, to further prove the validity of the scaling laws. Data from a hot unit may also be used to further validate the mathematical model developed in this study.
- (5) The mechanism of gas-mixing and solids-mixing in a coarse particle system is not well understood. Efforts should be made to gain a better understanding of such processes, both by designing experiments and by mathematical modelling of the physical phenomena.
- (6) There is a certain degree of empiricity in the model developed here for the prediction of residence time of fine particles. Efforts should be made to understand and incorporate all the effects in the dynamics of the dense bed; this would include electrostatic effects,

attrition, agglomeration of particles, etc. If these effects are included the empiricity may be avoided.

- (7) As industrial units are currently in operation using a CFB technology, efforts should be made to obtain operating data and use them for further verification of the currently available residence time models.
- (8) Only two sections of the MSFB process are studied in detail in this investigation - the dense bed section and the solids recirculating device. The dynamics of solids flow in the freeboard region above the dense bed and the gas-solids separation device (cyclone) need to be studied; a knowledge of hydrodynamics of the dense bed, the freeboard region, and the cyclone would enhance the understanding of the gas-solids flow behavior in the entire loop.
- (9) The pressure drop studies of the dense bed as part of its gross characterization should be extended to include more data. Such data are available as a result of the current study, but need to be analyzed to get a more complete picture of the dependency of pressure drop on various independent parameters.
- (10) As a natural extension of any hydrodynamic study, the associated heat transfer aspects need to be investigated. Combustion is a phenomenon in which the hydrodynamics and heat transfer are coupled. Once the hydrodynamics are thoroughly understood, efforts should be made to understand the heat transfer process. Such an understanding can be attained both by designing experiments and by modelling the physical phenomena.

8 NOMENCLATURE

C_D	Drag Coefficient for a Single Sphere
C_{DM}	Multi-Particle Drag Coefficient
d_p	Particle Diameter, m
d_{p1}	Coarse Particle Diameter, m
d_{p2}	Fine Particle Diameter, m
D	Bed Diameter, m
g	Acceleration due to Gravity, $\frac{m}{s^2}$
m	Mass of Particles, kg
n	System Dependent Parameter used in Model
p	Fitting Parameter of Model
Re_p	Particle Reynolds Number
U_o	Superficial Gas Velocity, $\frac{m}{s}$
U_a	Actual Gas(Air) Velocity in Bed, $\frac{m}{s}$
U_{p1}	Coarse Particle Velocity, $\frac{m}{s}$
U_{p2}	Fine Particle Velocity, $\frac{m}{s}$
W_g	Mass Flux of Gas (Fluidizing Medium), $\frac{kg}{m^2 \cdot s}$
W_{p2}	Solids Mass Flux of Fine Particles, $\frac{kg}{m^2 \cdot s}$

Greek Symbols

dm	Infinitesimal Amount of Fine Particles in Section ΔL of the Dense Bed, kg
ϵ	Average Bed Void Fraction
ϵ_f	Voidage of the Fine Particle Phase
ϵ_p	Average Fine Particle Holdup
ϕ	Sphericity of Coarse Particles
ρ_a	Density of Air (Fluidizing Medium), $\frac{kg}{m^3}$
ρ_p	Density of Fine Particles, $\frac{kg}{m^3}$
τ_r	Average Residence Time of Fine Particles in the Dense Bed, $\frac{s}{m}$

Subscripts

- 1 Referring to the Coarse Particles
- 2 Referring to the Fine Particles

9 REFERENCES

- 1) Talukdar, J., Hydrodynamic Studies of the Flow of Fine Particles through a Fluidized Bed of Coarse Solids, Ph.D. Dissertation, University of New Hampshire, 1991.
- 2) Mathur, V.K., Technical Report to Riley Stoker Corporation, Worcester, MA, 1989.
- 3) Fan, L.S., Toda, M. and S. Satija, Hold-up of Fine Particles in the Packed Dense Bed of the Multisolid Pneumatic Transport Bed, Powder Technology, Vol. 36, 1983, 107.
- 4) Nicastro, M.T. and L.R. Glicksman, Experimental Verification of Scaling Relationships for Fluidized Beds, Chem. Eng. Sci., Vol. 39, No. 9, 1984, 1381.
- 5) Glicksman, L.R., Scaling Relationships for Fluidized Beds, Chem. Eng. Sci., Vol. 39, No. 9, 1373.
- 6) Kitano, K., Wisecarver, K.D., Satija, S. and L.S. Fan, Holdup of Fine Particles in the Fluidized Dense Bed of the Multisolid Pneumatic Transport Bed, Ind. Eng. Chem. Res., Vol. 27, 1988, 125.
- 7) Yang, W.C., Mathematical Definition of Choking Phenomenon and Mathematical Model for Predicting Choking Velocity and Choking Voidage, AIChE Journal, Vol. 21, No. 5, 1975, 1013.
- 8) Breault, R.W., Hydrodynamic Characteristics and Coal Combustion Modeling of a High Velocity Fluidized Bed, Ph.D. Dissertation, University of New Hampshire, 1985.
- 9) Marble, F.E., Mechanism of Particle Collision in the One- Dimensional Dynamics of Gas-Particle Mixtures, The Physics of Fluids, Vol. 7, No. 8, 1964, 1270

APPENDIX E

Volume Two

1989 INTERNATIONAL CONFERENCE ON FLUIDIZED BED COMBUSTION

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MULTI-SOLIDS FLUIDIZED BED (MSFB) COMBUSTOR: MODELING AND ANALYSIS

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ABSTRACT

The development of Circulating atmospheric Fluidized Bed Combustors (CFBC) has grown rapidly during the past few years. CFBC's are complex chemical reactors in which coal combustion, sulfur capture, and nitrogen oxide formation and destruction takes place. Advanced coal combustion in CFBC is believed to be able to compete with oil fired boilers and to meet more strict emission limits. This paper discusses part of the current research activities stemming from a U.S. D.O.E. contract to develop an advanced concept atmospheric fluidized bed combustor (AFBC) to achieve +90% boiler efficiency, high turndown, +90% Sulfur removal at Ca/S molar ratio less than 1.5 and NO_x emissions less than 0.1 lb per MMBtu.

The Multi-solids Fluidized Bed (MSFB) Combustor is a unique high velocity process which is the base for our advanced design and consists of an entrained bed of fine ash and limestone particles superimposed on a fluidized bed of large particles. The low dense bed enhances mixing and increases combustion residence time of entrained particles. Coal devolatilization, volatile combustion, initiation of char combustion, and limestone calcination are accomplished in the lower dense bed, followed by char burnout and desulfurization which are completed in the upper lean phase oxidizing zone of the MSFB, which operate at superficial gas velocities of 7.5 - 9.1 m/s (25 - 30 ft/s).

A major advantage of the MSFB steam generator is the capability of utilizing low quality, high sulfur coals, and meeting the clean air act limitation of 0.52 kg sulfur dioxide per GJ (1.2 lb per MMBtu). In addition, the MSFB process decouples heat transfer and combustion processes. This allows greater flexibility to optimize the reactor design.

To characterize the performance of the dense bed and to interpret plant data, a model is developed for the lower fuel-rich fluidized bed part of the MSFB to predict pressure drop, particle size range, moisture evaporation, coal devolatilization, volatile combustion and char burnout. Solids flow is approximated by a series of reactors with mass interchange between each two consecutive reactors, and entrainment out of the bed. This model is applicable to the lower zone of any CFBC. However a model of heat transfer to the bubbling bed tube bundle would be a necessary addition for bubbling bed applications. Model results are presented and discussed.

Keywords: Fluidization, Circulating Fluidized Beds, Atmospheric Fluidized Bed Combustion, Two-Phase Flow, Multisolids.

INTRODUCTION

This paper describes the modeling of the dense bed of a Circulating Fluidized Bed Combustor (CFBC). Circulating fluidized bed combustion processes have evolved from bubbling fluidized bed combustor (Crouse and Nazemi, 1981, Farag et.al, 1987) to eliminate shortcomings such as poor carbon burnout and poor turndown characteristics, while preserving the heat transfer characteristics and the solid mixing capability associated with the bubbling bed. Excellent gas-solid contacting efficiency has been demonstrated in chemical processes where circulating beds have been utilized, e.g., Lanneau, (1960), Massimilla (1973), and Petersen (1980). Coal combustion tests in Circulating Fluidized Bed (CFB) have shown excellent carbon burnout and superb desulfurization efficiency at low Ca/S ratio, e.g., Engstrom (1980), Petersen (1980), Breault (1987),

Tang (1987). Staging the combustion in CFB results in reduced NO_x emissions, (Breault, 1987).

The design of this CFB systems involves the physical separation of the energy release and heat recovery sections, e.g., Reh (1979), Engstrom (1980), Keairns (1978), and Yerushalmi (1980). This results in more efficient start-up and simplified fuel feeding equipment. The Multi-Solids Fluidized Bed (MSFB) Combustor is a circulating fluidized bed combustor (CFBC) which has commercially shown the minimization of turndown problems, and better carbon conversion than bubbling bed units.

The MSFB process (figure 1) consists of an entrained bed of fine ash and limestone particles superimposed on a fluidized dense bed of large particles, (Nack et. al., 1987, Breault, 1987). The lower dense bed of large particles which do not leave the combustor, runs at fuel rich conditions, and serves to promote mixing, increase the residence time of the entrained ash, fuel, and limestone particles within the combustor. The combustor operates at superficial gas velocities of 7.5 to 9.1 m/s (25 to 30 ft/sec).

An important feature of the MSFBC is the essentially decoupled heat transfer and combustion processes. Most heat recovery occurs in an external heat exchanger (EHE) and a convective boiler. The EHE consists of heat exchanger tubes in a conventional fluidized bed operated at low velocities of 0.3 to 0.7 m/s (1 to 2 ft/sec). The combustor temperature is controlled to 1100 to 1200 K (1500 to 1700 F) by the recycling of entrained solids through the EHE.

A primary hot cyclone and non-mechanical "L" valves are used to separate and transport these solids. Flue gases leaving the cyclone separator are cooled in a conventional convective boiler.

NO_x control is achieved through staged combustion by operating the dense bed of the combustor under fuel rich conditions, and introducing secondary air in the entrained-bed region, or equivalent freeboard, to complete combustion. Since there are no heat exchange tubes immersed in the dense bed of the combustor there is no danger of corrosion, or erosion. There are waterwalls in the upper combustor which remove approximately five to ten percent of the total energy.

Objectives

It is critical to the design of the MSFB to be able to accurately project the fuel load to the dense bed. This determines heat release rates, heat transfer and bed depth requirements to insure appropriate temperatures, adequate fuel burnout and effective desulfurization. The fuel load distribution depends upon the extent to which the feed coal devolatilizes and forms char and volatiles, and to the extent to which these products are entrained in the off gas from the dense bed region. It also depends on the carbon flowing in from the lower hot and cold recycle streams.

The overall goal of this work is the development of advanced analytical and computer models to simulate the steady operation of the lower fuel-rich dense bed. The output of this model will be used as the input to the upper oxidizing zone transport bed

model developed by Breault (1985, 1987). The upper oxidizing entrained bed model is a one-dimensional solution to the differential solids and gas continuity equations, the mixture momentum balance, the solid and gas energy balances including combustion, desulfurization and NO_x formation and destruction chemical processes.

The combined model will be important in the analysis of pilot plant data, in the refinement of fluidized bed design, and in establishing the technical and economic feasibility for advanced second generation MSFB boiler modules in the 75,000 to 150,000 lb steam/hr size range.

The dense bed model discussed in this paper includes fluid mechanics, moisture evaporation, flow parameters, pressure drop, and devolatilization rate and chemistry of the lower bed of the MSFB process. Evaporation and devolatilization time calculations, and chemical reactions between the devolatilized species have also been included. These are discussed in the background.

BACKGROUND

This paper chronicles the results of model development and analysis done for the dense bed region of the Multi-Solids Circulating Fluidized Bed (MSFB) Combustor. Analyses, in this case, means the combined basic and engineering approach to yield information needed by the engineer. The engineering approach to furnace behavior, as distinguished from purely basic approach, is reasonably well discussed in the literature, e.g., Beer and Sidall (1972), and Thekdi et.al. (1969).

A generalized and rigorous solution for the dense bed of an MSFB is not yet available. Realizing that in economic life the designer/manufacturer can not justify non-delivery of hardware on the basis of awaiting research results, it was important to develop a combined theoretical/engineering approach to obtain needed results.

The fuel, coal, a non-conventional solid, is assumed to be described by its ultimate proximate and sulfur analyses. The prediction of MSFB performance is being solved by the "slicing technique" in conjunction with the following component parts:

1. Overall material balance.
2. Hydrodynamic parameters and pressure drop computation.
3. Moisture evaporation.
4. Coal devolatilization rate, and char composition.
5. Devolatilization chemistry.
6. Residence time of solids.
7. Elemental balances.
8. Flow model for gases and solids.
9. Reasonably rapid computer solutions.

The different feeds to the MSFB dense bed are

- 1- Coal (fuel) feed: with a specified rate, ultimate, proximate, and sulfur analyses. If the higher heating value of the coal is not supplied, it will be estimated using available correlations (IGT, 1978)

- 2- Sorbent (limestone) feed, with a specified rate and composition.
- 3- Recycled solids feeds. These are assumed to enter the bed at 1100F and 1600F with mass rates many times that of the fresh fuel rate. The composition of these recycled solids is specified.
- 4- Primary air, which may be specified either as a rate or as a percent of stoichiometric air.

All material is assumed to leave the dense bed in the gas stream.

Overall Material Balance: Based on the input streams the total amount of gas leaving, and its elemental composition are determined. The hourly and volumetric firing rates are calculated.

Hydrodynamic Parameters: From the dense bed mass, density and geometry data, the void fraction is calculated. Based on the effective particle diameter the minimum fluidization and terminal particle velocities and Reynolds number are computed. The minimum fluidization Froude number is computed to determine the fluidization mode. Pressure drop across the bed is computed using Ergun (1952) equation, based on the effective particle size, average gas and particle properties, and minimum fluidization conditions. Distributor pressure drop is taken to be 30% of the bed drop, or 15 inches of water, whichever is larger.

Based on the superficial air velocity, and the coal properties, the minimum (to prevent entrainment) and maximum (to be insure fluidization) particle diameter size is determined. The incoming air velocity is tested to insure that it is bracketed between terminal and minimum fluidization velocity for both with and without recycle contributions.

Recycle contributions are calculated from Fan et.al., (1983, 1985), who studied the effect of air and fine particles on the fluidization and entrainment of a bed of coarse particles. Their measurements showed that the minimum fluidization velocity increases initially as the flow rate of fines is increased. Further increase in the fines rate resulted in noticeable decrease in the minimum fluidization velocity. The study showed that due to drag reduction, a fluidizing mixture of air and fine particles may have a lower minimum fluidization velocity than in the absence of fines. Consequently a correlation was presented to relate the minimum fluidization Reynolds number to the mass flow rate of fines, the Archimedes number, as well as other fluidized bed properties. Similarly the terminal velocity of the coarse particles is lower with recycle than without recycle.

Moisture Evaporation Time: Jung (1980) and Jung and Stanmore (1980) estimated a 10 second drying period for wet brown coal, prior to devolatilization. For anthracite the evaporation time is about 0. Accordingly, a linear relation has been assumed up to a time of 10 sec and a moisture contents of 20%. Above 20% the evaporation time is kept at 10 sec.

Devolatilization Time: The volatile fraction of coals normally have a significant contribution to the total amount of heat released during combustion.

The rate of heat release from the combustion of volatile depends on the rate of volatile escape from the coal particle and combination with oxygen in the gas phase. It is therefore important to include a volatile release model in simulating MSFBC dense zone. Considerable work has been done on devolatilization of coal particles, e.g., Howard (1981), Baron et.al (1977), LaNauze (1982), Morris and Keairns (1979), and Salam (1982). In the present work the char diffusion-limited Model of LaNauze and further developed by Salam (1982) is adopted. In this model, the devolatilization time is assumed to follow a power-law relationship of the form

$$t_v = k (d_c)^a / [O_2]^b$$

which relates the devolatilization time t_v , in seconds, to the coal particle diameter d_c , in millimeters, and the oxygen concentration $[O_2]$ in volume %. The values of the constants are $k=6.53$; $a=1.19$ and $b=0.17$ (LaNauze, 1985, and Pillai, 1981, and 1985). The value of b indicates that as the oxygen concentration decreases t_v will increase. This has been confirmed by several investigators and can be explained as follows. Coal particle heating is principally due to radiation, and with a fully developed flame the heating rate would be very high. At low oxygen concentrations the flame would not be fully developed, reducing the heating rate to the devolatilizing coal particle.

The extent of volatile release, or coal weight loss upon heating, depends on the final temperature, as well as the coal composition. The correlations developed by the Institute of Gas Technology (IGT, 1978) for the percent weight loss of coal as a function of the ultimate and proximate analyses and of the temperature, are used to verify the maximum volatile matter of the coal.

Volatile Composition: The devolatilization from the dense bed is modeled as follows: Coal breaks into volatile matter (VM), water vapor from moisture, char and ash. Investigations have shown that all hydrogen in coal is released in the volatile matter and non remains in char, (Salam, 1982). Consequently, it has been assumed that char mass fractions of all elements (on a moisture ash-free basis) are in the same proportions as in the parent coal except that all hydrogen in the parent coal appears in the volatiles. The pyritic sulfur content of the coal is assumed to follow a different path. Half of the pyritic sulfur is taken to appear in ash, and the other half is distributed in the volatiles and char as follows: 62% goes to volatiles and 38% goes to char. Thus the ash in char will consist of the ash in the parent coal and 50% of the pyritic sulfur.

Once the volatiles are released the following chemistry is imposed, in an ascending order:

- 1- Convert volatile sulfur to hydrogen sulfide
 $S + 2H = H_2S$
- 2- Convert volatile nitrogen to ammonia
 $N + 3H = NH_3$
- 3- Convert volatile oxygen either to carbon monoxide or water vapor

- $O + C = CO$
 $O + 2H = H_2O$
 4- Convert volatile carbon and remaining volatile hydrogen to hydrocarbon
 $yC + zH = C_y H_z$
 y and z depend on the coal.

Once these reactions are completed the oxidation reactions take place. Oxygen in primary air converts ammonia to nitrogen oxide

$$2 NH_3 + 5O = 2NO + 3H_2O$$

Hydrogen Sulfide is oxidized by primary air to sulfur dioxide

$$H_2S + 3O = SO_2 + H_2O$$

Hydrocarbon is oxidized by primary air to carbon monoxide and water vapor

$$C_y H_z + (y+z/2)O = y CO + (z/2) H_2O$$

NO_x Assumptions

Many processes contribute to the formation of NO in fluidized bed coal combustors, so that models to describe its formation are complex and only partially predictive. In estimating the emissions of NO_x from the MSFB, experimental results obtained under conditions pertinent to those encountered in the various stages of the process are used in addition to the model developed by Breault (1987).

The first step is the determination of the distribution of coal nitrogen between the char and volatiles. For a bed temperature of 1117 K, and taking into account the particle temperature overshoot of 150 to 200 K, it has been estimated that approximately 50 percent of the nitrogen will be evolved with the volatiles. It is estimated that 25% of the char nitrogen gets converted to NO, and that practically all NO comes from nitrogen in coal. Based on the above the following approximations have been made.

- No thermal NO_x is formed
- All NO_x comes from fuel-bound nitrogen
- Volatile Fuel-Bound Nitrogen gets released at a rate proportional to the volatiles release rate. Nitrogen is released as ammonia, which is then oxidized by primary air to NO + H₂O.

Flow Model: Circulating fluidized beds are characterized by fast fluidization, extensive solids backmixing, uniform gas temperatures and high gas velocity resulting in high processing capacity with small vessel diameters. Excellent gas-solid contacting has been reported by several investigators. This has led to the assumption of a well mixed model for the gas in the different regions (or zones) of the bed. While axial mixing is high in the shallow dense bed, radial (or lateral) mixing may not be as efficient. Consequently, the dense bed is analyzed as a series of CSTR's through which the gas flows in parallel arrangement and the solids flow in series. Thus the cross section of the bubbling bed is divided into NZONE horizontal zones, each with a uniform temperature and composition (Figure 2). Solids may transport across the interface of zones. Coal enters just above the bed, at zone NZONE, while the

recycle streams enter into zone number 1. Solids may be interchanged between consecutive zones. The mass rate of solids interchanged between two adjacent zones is specified as a dimensionless exchange factor, representing the total mass moving from one zone into the other zone divided by the input mass rate of the coal fuel and limestone sorbent.

Range of Flow Rates: Interest is in the 75,000 lb steam/hr and larger units that would utilize as much as 30,000 lb coal/hr (produce about 300,000 lb steam/hr). For larger units the boiler is usually equipped with parallel fuel input locations and parallel recycle. Such a boiler can be treated as parallel sections. The main interest here is in one section in which steam generation is of the order of 75,000 to 100,000 lb steam per hour.

GENERAL COMPUTER MODEL DESCRIPTION

A comprehensive computer model has been developed to simulate the operation of the dense bed of an MSFB burning coal. The model includes data base of gas and solid component properties. It simulates a set of well-stirred fluidized beds with gas flow in parallel and solid flow in series, operating at steady state, with fresh coal, sorbent and air feeds, and recycle streams. The model is modular in structure and consists of a number of submodels dealing with the phenomena of fluidization, pressure drop, char and volatiles formation and reactions, etc.

The computer code for the model is written FORTRAN 77 and is comprised of several subprograms (subroutines and functions). It is initiated by an executive main. The program begins by reading in the input parameters followed by calls to different subroutines to perform hydrodynamic calculations, pressure drop estimation, moisture evaporation, coal devolatilization, volatiles reactions, material and elemental balances.

Model Input: The input to the package include:

- title
- bed information; e.g., height, width and depth, mass, density, and effective particle diameter,
- fuel information; e.g., rate, proximate, ultimate, and sulfur analyses, and higher heating value,
- sorbent information; e.g., flow rate and composition,
- recycle information
- air composition and its stoichiometric fraction,
- number of zones in the bed and the interchange variables
- the starting temperature of each zone
- number of carbon atoms in the hydrocarbon released from volatiles

Computation Sequence: After data input and initialization the program:

- verifies the consistency of input
- calculates bed parameters and void fraction
- calculates average bed temperature, based on input zonal temperatures, and computes the

- weight loss in Carbon due to devolatilization
- calculates char and volatiles mass fractions, and mass rates
- determines the primary air feed rate, and the stoichiometric air
- estimates the coal heating value, if not supplied, and calculates the firing rate, and the volumetric firing flux
- calculates input flow rates of each element based on input streams
- calculates char and volatiles composition
- calculates moisture evaporation time, using Jung's results
- calculates devolatilization time, using Salam's equation
- performs Volatile chemistry calculations
- determines total amount of gas leaving
- computes the minimum fluidization and terminal particle velocities and Reynolds number.
- determines the allowable range of particle sizes.
- test that the air velocity is bracketed between terminal and minimum fluidization velocity
- calculates average solid residence time in bed, and the average zone residence time
- performs mass balances on zones to determine the mass flow rate of gas leaving each zone, and check the consistency of data by verifying total mass balance
- performs overall elemental balances on C, Ca, H, N, O, S
- performs zonal calculations of moisture evaporation. It compares the solid residence time in each zone to the moisture evaporation time and iterates for the water evaporated in each zone. An overall mass balance is done to insure consistency of calculations.
- performs zonal calculations of volatiles release. It compares the post-evaporation solid residence time in each zone to the estimated devolatilization time and iterates for the devolatilization rate in each zone. Overall and elemental balances are done to insure consistency of calculations.

MODEL PREDICTION

The present model has been used to simulate an MSFB dense bed of a combustor generating 75,000 lbs steam/hr, and the results has been compared to available data.

The bed is 0.9 m high and 1.3 m in diameter with an effective particle size of 12.5 mm. Coal is fed at the rate of 0.88 kg/s (3.5 tons/hour) with the following proximate analysis: 10% ash, 45% fixed carbon, 10% moisture and 35% volatiles, and ultimate analysis of: 80% carbon, 5.9% hydrogen, 1% nitrogen, 10.1% oxygen and 3% sulfur (1% organic and 2% pyritic). Limestone (95% CaCO_3 , 2.5% of CaSO_4 and of CaSO_3) is fed at the rate of 0.227 kg/s (0.9 ton/hr). 60% of the recycle is CaO . The primary air input is 45% of the stoichiometric air, and the dense bed was divided into three equal volume zones. The hydrocarbon released was assumed to C_3H_y .

Devolatilization: Predicted devolatilization yields have been found to compare well with published carbonization assays of some coals (Selvig and Ode, 1957). For a lignite coal the model predicts a yield of 0.57 kg char and 0.43 kg volatiles per kg coal. The reported carbonization yields at 550 C are 0.65 and 0.35, respectively. This is a good agreement since the 550 C carbonization yields less volatiles than the ASTM procedure at 950 C. The volatile hydrocarbon released is predicted to be C_3H_7 , at the rate of 0.24 kg per kg coal. This, too, compares well with the carbonization yield of 0.23 kg per kg coal. Based on the IGT (1978) correlation the effect of temperature on the weight percent loss of coal due to devolatilization has been calculated. The results are plotted in Figure 3.

Pressure Drop: For a one meter (40 inches) high bed the computed pressure drop is 44.5 inch of water. This compares well with the rule of thumb of about 1 inch of water for every inch of bed.

Evaporation: Moisture evaporation calculations gave the distribution of water release in the bed, while volatile release calculation, combined with solid zoning flow result in the profile of volatile release rate (Figure 4) and the distribution of volatiles gaseous species across the dense bed, e.g., hydrocarbon, ammonia, hydrogen sulfide, sulfur dioxide, etc. Figure 4 shows the distribution of moisture evaporation and devolatilization in the combustor. At low recycle rate complete evaporation occurs in zone 3 (where coal is fed) and devolatilization is completed in zones 3 and 2. Further computer experimentations indicate that at high recycle rate evaporation is completed in zones 3, 2 and 1 (Figure 5). In this case no devolatilization takes place in zones 3 or 2 and the devolatilization of coal is not completed in the dense bed. Figure 5 indicates that only 9% of the coal is devolatilized in zone 1. The results also indicate some limitations on the interzone solid exchange rate to accomplish the desired water evaporation and/or volatile release.

The present simulative package will be combined with a model of the upper transport bed regime, COMSUL, which describes the combustion and desulfurization in the transport (lean) bed [Breault, 1987] to provide an integrated simulative system to be used in characterizing the performance of the MSFBC process.

SUMMARY

A model has been developed to characterize the dense bed region of the Multi-Solids Fluidized Bed (MSBF) Combustor. The model takes into account coal particle moisture evaporation, volatiles release, pressure drop and bed hydrodynamics. It incorporates a multi-zone solid flows pattern allowing for solid interchange between adjacent zones. The MSFBC is described by a set of continuous stirred reactors with parallel gas flow and series solid flow. The predictions of the model provide the input needed to the simulation of the upper transport (lean) bed. Combination of the two model will allow accurate prediction of coal

combustion, sulfur emission and capture and nitrogen oxide emissions.

REFERENCES

1. Baron, R.E., Beer, J.M., Borhi, G., Hodges, J.L. and Sarofim, A.F., 1977, Proc. Fifth International Conference Fluid Combustion, Washington, Vol II, p 437.
2. Beer, J.M., and Sidall, R.G., 1972, ASME paper 72-WA/HT, ASME Winter Meeting, NY, NY.
3. Borgvardt, R.H., 1970, Environmental Science and Technology, Vol. 4, No. 1,.
4. Breault, R.W., 1985, Ph.D. Thesis, University of New Hampshire, Durham, NH, 1985.
5. Breault, R.W., 1987, 9th Conference on Fluidized Bed Combustion, Boston, MA, May 3-7.
6. Breault, R.W., 1987, AIChE National Spring Meeting March 29, April 2, Houston TX.
7. Breault, R.W., 1987, ASME Winter Annual Meeting, December 14-18, Boston, MA.
8. Crouse, F.W., and Nazemi, A.H., 1981, "Characterization of Second Generation Fluidized Bed Combustion Concepts", Annual AIChE meeting, New Orleans, LA, Nov 8-12.,
9. Elliott, M.A., ed., 1981, "Chemistry of Coal Utilization Second Supplementary Volume", John Wiley, NY.
10. Engstrom, F., 1980, Proceeding of 6th International conference on FBC, 2, 616.
11. Ergun, S., 1952, Chemical Engineering Progress, 48, 89
12. Fan, L.S., Satija, S., 1985, "Terminal Velocity of Dense Particles in the Multi-Solid Pneumatic Transport Bed", Chem Eng. Sci, Vol. 40, No. 2, pp 259-267.
13. Fan, L.S., Toda, M. and Satija, S., 1983, "Fundamental Characteristic of a Multi-solid Pneumatic Transport Bed: Minimum fluidization Velocity of the Dense Bed", Fourth Fluidization Conference. (Eds., Kunii and Toei), Engineering Foundation, N.Y.
14. Farag, I.H., Breault, R., Karri, S.B.R., and Tsai, K.Y., 1987, Heat and Technology, Vol 5, No 1-2, pp 1-18
15. Howard, J.B., 1981 in "Chemistry of Coal Utilization", 2nd suppl. vol. (Ed. M.A. Elliott), John Wiley and Sons, N.Y., p 665.
16. IGT, 1978, Coal Conversion Systems Technical Data Book, prepared for DOE, Contract No Ex-76-C-01-2286.
17. Jung, K.M. 1980, Eng. Sci. Thesis, Department of Chemical Engineering, University of Melbourne, Australia.
18. Jung, K.M. and Stanmore, B.R. 1980, Fuel, 59, 74
19. Keairns, D.L. et.al., 1978, Inter Soc. Energy Conf, San Diego, CA
20. LaNauze, R.D., 1982, Fuel, 61, 771
21. LaNauze, R.D.; 1985, in "Fluidization", Eds. J.F. Davidson, R. Clift and D. Harrison, Academic Press, 2nd Edn., pp 113-124
22. Lanneau, K.P., 1960, Trans. of Inst. of Chemical Engineering, 38, 125
23. Massimilla, L., 1973, AIChE Symposium Ser., 129, 69, 11
24. Morris, J.P., and Keairns, D.L., 1979, Fuel, 58, 465
25. Nack, H. et al., 1977, Proceedings of the 5th International Conference on FBC, 3, 223
26. Petersen, V., 1980, Proceedings of 6th International Conference on FBC, 2, 212
27. Pillai, K.K., 1981, J. Inst. Energy, September, 142
28. Pillai, K.K., 1985, J. Inst. Energy, March, 3
29. Salam, F.F., Shen X.L. and Bibbs, B.M., 1988, Fuel, 67, 414
30. Selvig, W.A. and Ode, W.H., 1957, Bulletin 571, Bureau of Mines, Washington, D.C.
31. Tang, J., 1987, "Assessment of the Ahlstrom Pyroflow Circulation in Conventional Fluidized Bed Combustion System", Proceeding of 9th Int. Conf. on FBC, Boston, MA, May 3-7.
32. Thekdi, A.C., et. al., 1969, Technical report No. FS/CGSSC-12/69-4, Dept. of Fuel Science, The Pennsylvania State University.
33. Yerushalmi, T., 1980, paper at AIChE mtg, Philadelphia, PA

FIGURE 1: SCHEMATIC OF MULTISOLIDS FLUIDIZED BED (MSFB) COMBUSTOR PROCESS

FIGURE 3: TEMPERATURE EFFECT ON COAL DEVOLATILIZATION

FIGURE 4: EVAPORATION AND COAL DEVOLATILIZATION PROFILES

FIGURE 2: SCHEMATIC OF DENSE BED SUBDIVIDED INTO CSTR ZONES

FIGURE 5: EVAPORATION AND COAL DEVOLATILIZATION PROFILES

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