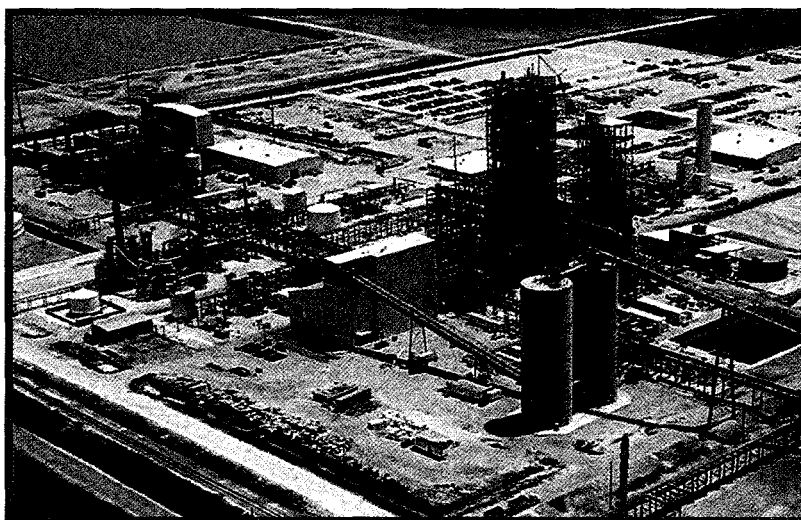


# Fifth Annual Clean Coal Technology Conference

## TECHNICAL PAPERS

*Powering the Next Millennium*



*Hyatt Regency Westshore*  
*January 7-10, 1997, Tampa, Florida*

# MASTER



Photograph

Tampa Electric Company's Integrated  
Gasification Combined-Cycle Plant  
(Polk Power Station)

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# **Fifth Annual Clean Coal Technology Conference**

## **Technical Papers**

**January 7-10, 1997  
Tampa, Florida**

## **Powering the Next Millennium**

**T**he Fifth Annual Clean Coal Technology Conference will focus on presenting strategies and approaches that will enable clean coal technologies to resolve the competing, interrelated demands for power, economic viability, and environmental constraints associated with the use of coal in the post-2000 era. The program will address the dynamic changes that will result from utility competition and industry restructuring, and to the evolution of markets abroad. Current projections for electricity highlight the preferential role that electric power will have in accomplishing the long-range goals of most nations. Increased demands can be met by utilizing coal in technologies that achieve environmental goals while keeping the cost-per-unit of energy competitive. Results from the projects in the DOE Clean Coal Technology Demonstration Program confirm that technology is the pathway to achieving these goals.

The industry/government partnership, cemented over the past 10 years, is focussed on moving the clean coal technologies into the domestic and international marketplaces. The Fifth Annual Clean Coal Technology Conference will provide a forum to discuss these benchmark issues and the essential role and need for these technologies in the post-2000 era.

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# **Technical Session I**

## **Advanced Coal Process Systems**

## **Fuel and Power Coproduction**

### **The Liquid Phase Methanol (LPMEOH™) Process Demonstration at Kingsport**

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**Fifth Annual DOE Clean Coal Technology Conference  
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#### **Abstract**

The Liquid Phase Methanol (LPMEOH™) process uses a slurry bubble column reactor to convert syngas (primarily a mixture of carbon monoxide and hydrogen) to methanol. Because of its superior heat management, the process is able to be designed to directly handle the carbon monoxide (CO) - rich syngas characteristic of the gasification of coal, petroleum coke, residual oil, wastes, or of other hydrocarbon feedstocks. When added to an integrated gasification combined cycle (IGCC) power plant, the LPMEOH™ process converts a portion of the CO-rich syngas produced by the gasifier to methanol, and the remainder of the unconverted gas is used to fuel the gas turbine combined-cycle power plant. The LPMEOH™ process has the flexibility to operate in a daily electricity demand load-following manner. Coproduction of power and methanol via IGCC and the LPMEOH™ process provides opportunities for energy storage for electrical demand peak shaving, clean fuel for export, and/or chemical methanol sales.

## Introduction

The LPMEOH<sup>TM</sup> technology was developed during the 1980's, with the financial support of the U. S. Department of Energy (DOE). The concept was proven in over 7,400 hours of test operation in a DOE-owned, 3,200 gallons (U.S.) of methanol per day process development unit located at LaPorte, Texas. (Ref. a). The commercial-scale demonstration plant for the technology has been constructed and is now being commissioned at Eastman Chemical Company's coal gasification facility in Kingsport, Tennessee under the DOE's Clean Coal Technology Program. The LPMEOH<sup>TM</sup> plant will demonstrate the production of at least 80,000 gallons of methanol per day, and will simulate operation for the IGCC coproduction of power and methanol application. Construction began in October of 1995 and was completed in December of 1996. Commissioning will be completed and startup will begin in January of 1997, and will be followed by four years of operation to demonstrate the commercial advantages of the technology.

Air Products and Eastman formed the "Air Products Liquid Phase Conversion Co., L.P." partnership to execute the demonstration project. The partnership owns the LPMEOH<sup>TM</sup> demonstration plant. Air Products manages the demonstration project and provides technology analysis and direction for the demonstration. Air Products also provided the design, procurement, and construction of the LPMEOH<sup>TM</sup> demonstration plant (i.e., a turnkey plant). Eastman provides the host site, performs the permitting and operation of the LPMEOH<sup>TM</sup> unit, and supplies the supporting auxiliaries, the synthesis gas, and takes the product methanol.

The LPMEOH<sup>TM</sup> plant will demonstrate production of at least 80,000 gallons of methanol per day, from a portion of the available clean synthesis gas. Most of the product methanol will be refined to chemical-grade quality (99.85 wt % purity via distillation) and used by Eastman as replacement chemical feedstock in the commercial facility. A portion of the product methanol will be withdrawn prior to purification (about 98 wt % purity) and used in the off-site product-use tests.

This paper gives a review of: I - Commercial Application for the LPMEOH<sup>TM</sup> process technology; II - Demonstration Plant - Test Plans, highlighting the operational and product-use testing plans to confirm the commercial application; and III - Demonstration Plant Design, Construction and Startup - Status, highlighting the design and integration of the demonstration plant at Kingsport, and of the accomplishments during the design and construction phase.

## I - Commercial Application

### Technology Description

The heart of the LPMEOH™ process is the slurry bubble column reactor (Figure 1).

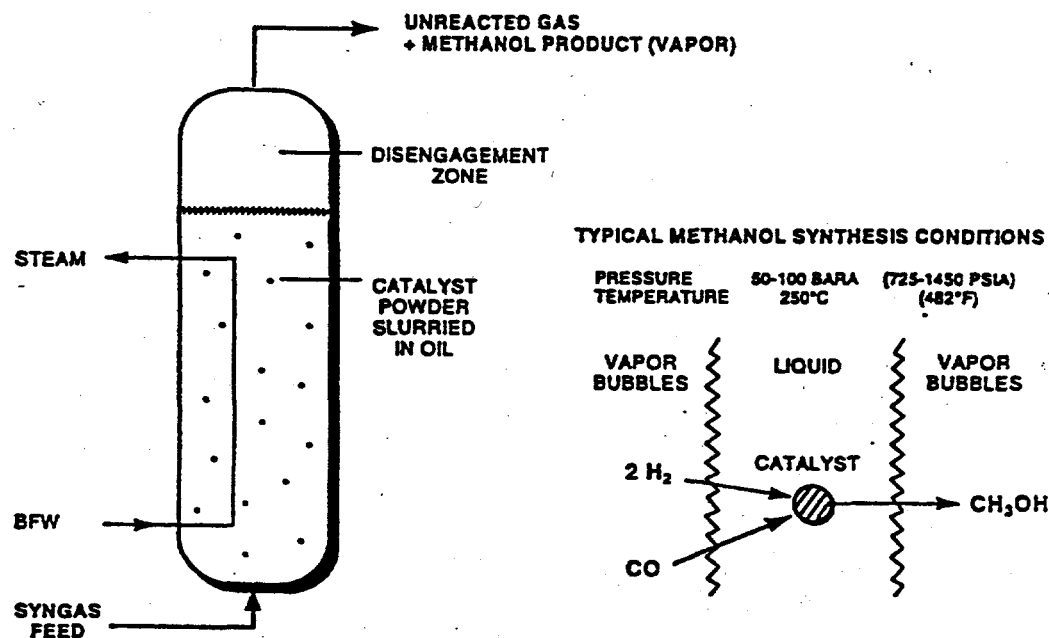


Figure 1. LPMEOH™ Reactor and Reaction Schematics

The liquid medium is the feature that differentiates the LPMEOH™ process from conventional technology. Conventional methanol reactors use fixed beds of catalyst pellets and operate in the gas phase. The LPMEOH™ reactor uses catalyst in powder form, slurried in an inert mineral oil. The mineral oil acts as a temperature moderator and a heat removal medium, transferring the heat of reaction from the catalyst surface via the liquid slurry to boiling water in an internal tubular heat exchanger. Since the heat transfer coefficient on the slurry side of the heat exchanger is relatively large, the heat exchanger occupies only a small fraction of the cross-sectional area of the reactor. The slurry reactor can thus achieve high syngas conversion per pass, due to its capability to remove heat and maintain a constant, highly uniform temperature through the entire length of the reactor.

Because of the LPMEOH<sup>TM</sup> reactor's unique temperature control capabilities, it is able to directly process syngas which is rich in carbon oxides (carbon monoxide and carbon dioxide). Gas phase methanol technology would require such a feedstock to undergo stoichiometry adjustment by the water gas shift reaction (to increase the hydrogen content) and carbon dioxide (CO<sub>2</sub>) removal (to reduce the excess carbon oxides). In a gas phase reactor, temperature moderation is only achieved by recycling large amounts of hydrogen (H<sub>2</sub>)-rich gas, utilizing the higher heat capacity of H<sub>2</sub> gas as compared to carbon monoxide (CO) gas. Typically a gas phase reactor is limited to about 16% CO gas in the inlet to the reactor, in order to limit the conversion per pass to avoid excess heating. In contrast, with the LPMEOH<sup>TM</sup> reactor, CO gas concentrations in excess of 50% have been routinely tested without any adverse effect on the catalyst activity.

A second differentiating feature of the LPMEOH<sup>TM</sup> reactor is its robust character. The slurry reactor is suitable for rapid ramping, idling, and even extreme stop/start actions. The thermal moderation provided by the liquid inventory in the reactor acts to buffer sharp transient operations that would not normally be tolerable in a gas phase methanol synthesis reactor.

A third differentiating feature of the LPMEOH<sup>TM</sup> process is that a high quality methanol product is produced directly from syngas which is rich in carbon-oxides. Gas phase methanol synthesis, which relies on hydrogen-rich syngas, results in a crude methanol product with to 4 to 20% water by weight. The product from the LPMEOH<sup>TM</sup> process typically contains only 1% water by weight. This methanol product, coproduced with IGCC, is therefore suitable for many applications, and at a substantial savings in purification costs. The steam produced in the LPMEOH<sup>TM</sup> reactor is suitable for purification of the methanol product (for upgrading to a higher quality) or for use in the IGCC power generation cycle.

Another unique feature of the LPMEOH<sup>TM</sup> process is the ability to add fresh catalyst on-line. Methanol catalyst deactivates at a slow rate. With the LPMEOH<sup>TM</sup> reactor, spent catalyst slurry may be withdrawn and fresh catalyst slurry added on a periodic batch basis. This allows continuous, uninterrupted operation and also the maintenance of a high productivity level in the reactor. Furthermore, choice of replacement rate permits optimization of productivity versus catalyst replacement cost.

## IGCC Coproduction Options

The LPMEOH™ process is a very effective technology for converting a portion of the H<sub>2</sub> and CO in an IGCC electric power plant's coal-derived syngas to methanol. The process is very flexible in being able to process many variations in syngas composition. The LPMEOH™ process can be used with an IGCC electric power plant (Ref. b), to provide the once-through methanol production as depicted in Figure 2. The process can be designed to operate in a continuous, baseload manner, converting syngas from oversized gasifiers or from a spare gasifier. The process can also be designed to operate only during periods of off-peak electric power demand to consume a portion of the excess syngas and allow the electricity output from the combined-cycle power unit to be turned down. In this latter circumstance, the gasification unit continues to operate at full baseload capacity, so the IGCC facility's major capital asset is fully utilized. In either baseload or cycling operation, partial conversion of between 20% and 33% of the IGCC plant's syngas is optimal, and conversion of up to 50% is feasible.

The design configuration for the LPMEOH™ process depends upon the degree of conversion of syngas (or the quantity of methanol relative to the power plant size). The feed gas pressure is a prime determinant of the degree of syngas conversion, as shown in Figure 3. Reaction pressure for methanol synthesis design is usually 750 psia or higher.

In its simplest configuration, syngas (feed gas) at its maximum available pressure from the IGCC electric power plant is passed once, without recycle through the LPMEOH™ plant (Figure 4), and partially converted to methanol. The unreacted gas is returned to the IGCC power plant's combustion turbines.

Of course, the richer the once-through syngas is in CO, the more the production is limited by the availability of H<sub>2</sub>. Normally the least expensive methanol conversion cost comes from converting as much hydrogen as is practical; without feed gas compression, unreacted gas recycle or further processing of the feed gas. The higher the pressure at which the syngas is available, the greater is the degree of conversion and the lower the conversion cost.

If greater amounts of syngas conversion are required, different once-through plant design options (Figure 4) are available. There is still no need for upstream stoichiometric adjustment of the feed gas by the water-gas shift reaction and CO<sub>2</sub> removal; so the simplicity of once-through CO-rich gas processing is retained. The LPMEOH™ process design options for greater syngas conversion are:

- Once-Through, with Gas Recycle.

One design technique to increase the degree of syngas conversion is to condense out methanol from the reactor effluent and to recycle part of the unreacted feed gas back to the reactor inlet. With the LPMEOH™ process, this simple recycle refers to recycle of CO-rich gas. The recycle ratio required for the LPMEOH™ is moderate, for example, one part unreacted syngas to one part fresh feed gas. This 1 to 1 recycle ratio is usually

quite effective in optimizing the methanol production. At higher recycle ratios, little is gained, since most of the available  $H_2$  has already been converted to methanol.

- Once-Through, with Water Addition

If additional conversion is desired, the LPMEOH<sup>TM</sup> process design can be altered to generate additional  $H_2$ . The inherent shift activity of the methanol catalyst can be utilized to accommodate a modest amount of shift activity within the reactor. This is done by the addition of water, as steam, to the syngas before it passes through the liquid phase methanol reactor. Within the reactor, the additional steam is converted to  $H_2$ , which is, in turn, converted to methanol. In the water addition case, the increase in conversion is accompanied with a modest increase of water in the crude methanol product and of  $CO_2$  in the reactor effluent gas.

- Once-Through, with Feed Gas Compression

When the feed gas pressure from the IGCC electric power plant is low (e.g. below 750 psia), feed gas compression may be added to the LPMEOH<sup>TM</sup> process design, to increase reactor productivity and the overall conversion of syngas to methanol.

### Baseload Coproduction of Methanol and Power

Process design study work for the LPMEOH<sup>TM</sup> process has been directed towards converting a portion of coal-derived syngas produced in an IGCC electric power plant to methanol. A feed gas containing 35%  $H_2$ , 51%  $CO$ , 13%  $CO_2$  and 1% inerts (nitrogen) was used for preparing the baseload methanol coproduction economics.

With a given gasification plant size, the IGCC coproduction plant can be designed to accommodate a range of methanol to power output ratio's. For example (Ref. c, d), a gasification plant, with two gasifiers of 1735 million Btu (HHV) per hour output each, could be sized for baseload power output of 426 megawatts of electricity (MWe) and for baseload methanol coproduction of 152,000 US gallons per day (G/D). Other methanol and power plant size options for this gasification plant size, are shown in Table 1.

Table 1. Methanol Plant to Power Plant Size Ratio

% of Syngas Converted to Methanol (%)	Baseload Power Plant Size (MWe)	Baseload Methanol Plant Size (G/D)	Methanol Plant to Power Plant Size Ratio (G/D per MWe)
0	500	0	0
13.8	426	152,000	357
20.0	394	210,000	533
30.0	342	330,000	965

The IGCC coproduction plant with 426 MWe of power and 152,000 G/D of methanol is used for the baseload production cost estimate for coproduced methanol, shown in Table 2. If the baseload fuel gas value is \$4.00 per million Btu, then 152,000 G/D of methanol can be coproduced from coal for under 50 cents per gallon.

As one would expect, the methanol production cost is lower at larger methanol plant sizes. Figure 5, shows the effect of plant size for once-through methanol coproduction. Methanol production costs for two of the LPMEOH™ plant design options for higher syngas conversion: 1 to 1 gas recycle, and 1 to 1 gas recycle with water addition, are also shown.

Today, new methanol plants are being built where natural gas is inexpensive (Chile, Saudi Arabia). These new world scale plants range in size from 700,000 to 900,000 G/D (2000 to 2700 metric tons per day) in size. The economy of scale savings; in natural gas gathering, syngas manufacturing, and in methanol storage and ocean transport facilities; drive these plants to their large size. Estimates (Ref. e, f) show that an 836,000 G/D off-shore methanol plant (with the same, 20% per year capital charge as in Table 1 and Figure 5), with natural gas at \$0.50 to \$1.00 per million Btu, has a total ex-plant methanol production cost of 46 to 50 cents per gallon. Adding ocean freight, duty and receiving terminal storage typically adds 8 to 10 cents per gallon; giving a total delivered U.S. Gulf Coast methanol cost (Chemical Grade) of 55 to 60 cents per gallon.

Figure 5 is interesting, because it provides an unexpected result. Methanol coproduction with IGCC and the once-through LPMEOH™ process, does not need large methanol plant sizes to achieve good economies of scale. The gasification plant is already at a large economical scale for power generation; so the syngas manufacturing economies are already achieved. Methanol storage and transport economies are also achieved by serving local markets, and achieving freight savings over the competing methanol, which is usually shipped from the U. S. Gulf coast.

The 50 cents per gallon coproduction cost for a 152,000 G/D once-through LPMEOH™ plant size is competitive in local markets with new world scale off-shore methanol plants. Figure 5 shows an additional 3 to 4 cent per gallon saving for a 365,000 G/D LPMEOH™ plant size. These additional savings might be used to off-set higher freight costs to more distant local customers; while still maintaining a freight and cost advantage over the imported methanol from the Gulf Coast.



## **Applications for the Coproduced Methanol Product**

The methanol coproduction process studies show that the LPMEOH™ process can produce a clean high quality methanol product at less than 50 cents per gallon; from an abundant, non-inflationary local fuel source (coal). Serving local markets, the methanol coproduced at central IGCC electric power plants, can be a valuable premium fuel or fuel feedstock for many applications, such as:

1. An economical hydrogen source for small fuel cells, which are being developed for transportation applications. Methanol is a storable, and transportable, liquid fuel which can be reformed under mild conditions to provide an economical source of hydrogen for fuel cells.
2. When reformed under mild conditions, may be an economical hydrogen or carbon monoxide source for industrial applications.
3. A substitute for chemical grade methanol being used for MTBE manufacture. (MTBE is added to gasoline to boost octane and to meet environmental clean air mandates. MTBE is one of the major current markets for methanol.)
4. An environmentally advantaged fuel for dispersed electric power stations. Small packaged power plants (combustion turbine, internal combustion engine, or fuel cell) provide power and heat locally, at the use point; eliminating the need for natural gas pipelines and high voltage power lines.
5. Finally, the coproduced methanol may be used by the utility owning the IGCC facility (see Figure 2). Potential uses are: a) as a backup fuel for the IGCC plant's main gas turbines; b.) as a fuel for a separate, dedicated cycling combined-cycle unit at the same site; c.) as the fuel exported to the utility's distributed power generation system(s); or d) as the transportation fuel for the utility's bus or van pool. Since methanol is an ultra-clean (zero sulfur) fuel which burns with very low (better than natural gas) emissions of nitrogen oxides, the incremental power is very clean. Since the methanol is derived from the coal pile, the IGCC facility can be truly independent and self-sufficient for fuel needs. In addition, should the external prices for methanol command higher value to the IGCC plant's owner, the methanol can be exported for additional revenues.

Many of the applications listed above, are embryo developments. Their ultimate market size potential; for transportation applications, for industrial applications and for distributed power generation; could become large. The methanol product specification for the five applications is not adequately known. Therefore, part of the LPMEOH™ demonstration project's program is to confirm the suitability of the methanol product for these (and other) uses. Product-use tests will be used to develop final methanol product specifications. During the demonstration, in the 1998 to 2000 time-frame, about 400,000 gallons of the "as-produced from CO-rich syngas" methanol will be available for off-site product-use testing. The final off-site product-use test plan is now under development. More details will be provided to interested parties.

## **II - Demonstration Plant - Test Plans**

### **Objective**

The LPMEOH™ Process technology is expected to be commercialized as part of an IGCC electric power generation system. The preceding Commercial Application section highlighted the advantages of the LPMEOH™ process. These commercial advantages must be demonstrated and confirmed during operations. Therefore, the demonstration test plan incorporates, but is not limited to, these commercially important aspects of IGCC integration:

- The coproduction of electric power and of added value liquid transportation fuels and/or chemical feedstocks from coal. This coproduction requires that the partial conversion of syngas to storable liquid products be demonstrated.
- Using an energy load-following operating concept which allows conversion of off-peak energy, at attendant low value, into peak energy commanding a higher value. This load-following concept requires that on/off and syngas load-following capabilities be demonstrated.
- Syngas compositions will vary with the type of gasification process technology and gasification plant feed used in the power generation application. Therefore, operation over a wide variety of syngas compositions will be demonstrated.
- Catalyst life, operating on "real" coal-derived syngas, must be demonstrated over a long period of time. Major parameters include reactor operating temperature, concentration of poisons in the reactor feed gas, and catalyst aging and attrition.
- Reactor volumetric productivity must be optimized for future commercial designs. Parameters include: high inlet superficial velocity of feed gas, high slurry catalyst concentration, maximum gassed slurry level, and removal of the heat of reaction.
- Methanol Product, as produced from by the liquid phase reactor from syngas rich in carbon oxides, must be suitable for its intended uses. Off-site methanol product-use testing will confirm the product specification needed for market acceptability.

### **Methanol Operations - Demonstration Test Plan**

Three key results will be used to judge the success of the LPMEOH™ process demonstration during the four years of operational testing:

- Resolution of technical issues involved with scaleup and first time demonstration for various commercial-scale operations
- Acquisition of sufficient engineering data for commercial designs; and
- Industry acceptance.

The demonstration test plan has been established to provide flexibility in order to meet these success criteria. Annual operating plans, with specific targeted test runs, will be

prepared, and revised as necessary. These plans will be tailored to reflect past performance, as well as commercial needs. User involvement will be sought.

The LPMEOH™ operating test plan outline, by year, is summarized in Table 3. The demonstration test plan encompasses the range of conditions and operating circumstances anticipated for methanol coproduction with electric power in an IGCC power plant. Since Kingsport does not have a combined-cycle power generation unit, the tests will simulate the IGCC application. Test duration will be emphasized in the test program. The minimum period for a test condition, short of the rapid ramping tests, is 2 weeks. Numerous tests will have 3–6 week run periods, some 8–12 weeks, and a few key basic tests of 20 to 30 weeks.

Table 3.	LPMEOH™ Demonstration Test Plan Outline
<u>Year 1</u>	Catalyst Aging Catalyst Life Versus LaPorte process development unit and Lab Autoclaves Process Optimization / Maximum Reactor Productivity Catalyst Slurry Concentration Reactor Slurry Level Catalyst Slurry Addition Frequency Test Establishment of Baseline Condition
<u>Years 2 &amp; 3</u>	Catalyst Slurry Addition and Withdrawal at Baseline Condition Catalyst Attrition/Poisons/Activity/Aging Tests Simulation of IGCC Coproduction for: <ol style="list-style-type: none"> <li>1. Synthesis Gas Composition Studies for Commercial Gasifiers                  Texaco, Shell, Destec, British Gas/Lurgi, Other Gasifiers</li> <li>2. IGCC Electrical Demand Load Following:                  Rapid Ramping, Stop/Start (Hot and Cold Standby).</li> <li>3. Additional Industry User Tests</li> </ol> Maximum Catalyst Slurry Concentration Maximum Throughput/Production Rate
<u>Year 4</u>	Stable, extended Operation at Optimum Conditions 99% Availability Potential Alternative Catalyst Test Additional Industry User Tests

**Table 3 - Demonstration Test Plan Outline**

### **III - Demonstration Plant Design, Construction and Startup - Status.**

#### **Kingsport Site**

Eastman began coal gasification operations at Kingsport in 1983. Figure 6 shows an aerial view of Eastman's Kingsport gasification facility. Texaco gasification is used to convert about 1,000 tons-per-day of high-sulfur, Eastern bituminous coal to synthesis gas for the manufacture of methanol, acetic anhydride, and associated products. Air Products provides the oxygen for gasification by a pipeline from an over-the-fence air separation unit. The crude synthesis gas is quenched, partially shifted, treated for acid gas removal (hydrogen sulfide and carbonyl sulfide, and CO<sub>2</sub>, via Rectisol), and partially processed in a cryogenic separation unit to produce separate H<sub>2</sub> and CO streams. The H<sub>2</sub> stream is combined with clean synthesis gas to produce stoichiometrically balanced feed to a conventional gas phase methanol synthesis unit. Methanol from this unit is reacted with recovered acetic acid to produce methyl acetate. Finally, the methyl acetate is reacted with the CO stream to produce the prime product, acetic anhydride (and acetic acid for recycle).

Because the gasification facility produces individual streams of clean synthesis gas, CO, and H<sub>2</sub>-rich gas, there is the capability to blend gases and mimic the gas compositions of a range of gasifiers. Figure 7 shows the process block flow diagram for the Kingsport gasification facility including the LPMEOH<sup>TM</sup> demonstration plant.

#### **Demonstration Plant Design**

The site at Kingsport borders an existing methyl acetate plant, and was relatively level. Figure 8 is an aerial view of the site prior to the start of construction. Some fill was required to provide a 270 ft. by 180 ft. plot for the demonstration plant and tank truck loading areas. An area next to the site was made available for establishing the construction trailer, fabrication, and laydown areas. The job site was fenced off in order to provide a secure site separate from the operating areas of the Eastman facility.

Air Products' Gas Group Engineering Department was responsible for the engineering design and construction of the project. This included detailed design and procurement. Eastman was responsible for the outside battery limits design and construction, the permitting, and for providing the digital control programming. Eastman reviewed the detailed design of the demonstration plant. Most of the equipment and materials were bid competitively from combined Air Products and Eastman developed bid lists. The construction was subcontracted into ten different packages, awarded on fixed priced bidding.

The need to meet all of the test program objectives provided a design challenge for the Air Products/Eastman design team. Of primary importance was the integration of the LPMEOH<sup>TM</sup> demonstration plant within the Kingsport gasification complex. Since the feed composition to the reactor is to be varied from H<sub>2</sub>-lean to H<sub>2</sub>-rich (25% to 70+%

H<sub>2</sub>) and the flow to the reactor by at least a factor of two, all of the product and byproduct streams within and outside the battery limits were affected. Control valves and instrumentation for the demonstration plant were required to have functionality over and beyond those for a normal commercial facility. Extreme cases of about twenty different heat and material balances were considered for specification of each piece of equipment, flow measurement device, control valve, and safety relief device.

Machinery specification was especially challenged by the requirements of the demonstration operating period. The syngas recycle compressor design uniquely considered all of these varying molecular weight streams in tandem with the varying pressure drop requirements. These extremely atypical operating requirements challenged controls and machinery engineers to develop a robust surge control strategy which could adequately protect the compressor. Slurry pump design also considered a wide range of operating conditions, such as changing temperature and viscosity due to varying slurry concentrations.

The heart of the LPMEOH plant is the reactor. The design and fabrication of the reactor vessel and its internal heat exchanger received careful attention. The reactor size is based on a scale-up of the DOE-owned process development unit at LaPorte, Texas. The reactor is a stainless steel clad carbon steel vessel designed for 1000 psig and 600°F. The detail design and materials delivery for the reactor were not a problem. The reactor has an internal heat exchanger for removal of the heat of reaction. The design of the internal heat exchanger required careful analysis of the headers and support system. The fabrication of the bundle required special procedures for welding the tubes in place prior to insertion within the reactor.

The analysis to be performed during the demonstration period requires the collection of high quality engineering data. The gas analysis system is "research quality" in terms of analysis capabilities (number of components and precision) and is also rugged enough to withstand an industrial operating environment. Duplicate flow measurement devices were frequently required to ensure accurate measurement during some of the off-design cases. Temperature and pressure measurement devices over and above what would be required for a commercial plant were installed. All of this data will be processed and stored using a state of the art data acquisition system, Honeywell Total Plant™, which will be integrated with the facility's distributed control system. This system will also provide engineers with remote access to data.

The ultimate goal of the demonstration period is to reach a stable optimized operating condition, with the best combination of the most aggressive operating parameters. These parameters, such as reactor superficial gas velocity, slurry concentration and reactor level, will allow us to maximize the reactor productivity. It will be a continuing goal during the demonstration period to determine and debottleneck limitations of the demonstration plant.

## **Project Schedule**

The DOE approved the site change of the LPMEOH Demonstration to Eastman's Kingsport, TN site in October of 1993. Air Products and Eastman worked with the DOE to define the size of the plant and develop a Statement of Work for the LPMEOH™ Demonstration at Kingsport. This Project Definition phase including a cost estimate was completed in October of 1994. Preliminary Detail Design work on equipment layouts and development of P&ID's began shortly after this. Full authorization from the DOE for Design and Construction was effective February 1, 1995. The reactor was the first piece of equipment to be placed on order in November of 1994. Equipment deliveries began in November of 1995. The State air permit was received in March of 1996.. The DOE completed its National Environmental Policy Act (NEPA) review and issued a Finding of No Significant Impact (FONSI) in June of 1995. Construction at the site began in October of 1995. Construction manpower peaked at 150 people on site in mid-November of 1996. Construction was essentially completed in December of 1996. The overall schedule from authorization to startup was 23 months.

Instrument Loop Checking began in October 1996. Commissioning began in December of 1996, and startup will begin in late January of 1997. The four-year methanol test operation will begin in February of 1997. The operating test program will end in the year 2001. The off-site fuel use tests will be performed over an 18 to 30 month period, beginning in May of 1998.

## **Conclusion**

The LPMEOH™ process is now being demonstrated at commercial scale, under the DOE Clean Coal Technology Program. The demonstration plant, located at Eastman Chemical Company's Kingsport, Tennessee coal gasification facility site, will produce at least 80,000 gallons-per-day of methanol from coal-derived synthesis gas. Startup begins in late January of 1997, followed by a four-year demonstration test period beginning in late February of 1997.

Successful demonstration of the LPMEOH™ technology will add significant flexibility and dispatch benefits to IGCC electric power plants, which have traditionally been viewed as strictly a baseload power generation technology. Now, central clean coal technology processing plants, making coproducts of electricity and methanol; can meet the needs of local communities for dispersed power and transportation fuel. The LPMEOH™ process provides competitive methanol economics at small methanol plant sizes, and a freight and cost advantage in local markets vis-à-vis large off-shore remote gas methanol. Methanol coproduction studies show that methanol at less than 50 cents per gallon can be provided from an abundant, non-inflationary local fuel source (coal). The coproduced methanol may be an economical hydrogen source for small fuel cells, and an environmentally advantaged fuel for dispersed electric power.

**TABLE 2**

**Production Cost Estimate for Coproduced Methanol**  
 LPMEOH Plant Capacity: 152,000 gallons per day (500 sT/D)  
 Capital Investment: \$29 million

Methanol Plant Operation:	(Hours/year)	<b>Baseload</b> 7884 hr/yr.
Methanol Production	(million Gal./year)	49.9
<b>Methanol Production Cost</b>		<b>cents/gallon</b>
<b>Syngas cost:</b>		
Feed Gas @ fuel value (\$4.00/mmBtu)		98.7
Unreacted (CO-rich) gas @ fuel value (\$4.00/mmBtu)		(68.4)
Sub-total; net cost of syngas converted:		<b>30.3</b>
<b>Operating cost</b>		
Catalyst and chemicals		2.6
Export steam		(2.9)
Utilities		0.9
Other (fixed) costs		4.0
Sub-Total; Operating Costs:		<b>4.6</b>
<b>Capital charge @ 20% of investment per year</b>		<b>11.6</b>
<b>Total Methanol Production Cost:</b>		<b>46.5</b>

**Basis:**

*U.S. Gulf Coast Construction, 4thQ 1996 \$*

*Includes owner costs and 30 days of Product Storage*

*CO-rich feed gas from IGCC electric power plant at 1000 psia., with 5ppm (max.) sulfur.,*

*Once-through LPMEOH process design with 1562 mmBtu/hr in, 1082 mmBtu out (HHV)*

*Excludes License and Royalty fee. Air Product's is the LPMEOH proces technology licensor.*

*Product methanol with 1 wt.% water; 'Chem Grade would add 4 to 5 cents per gallon.*

**Table 2. Production Cost Estimate for Coproduced Methanol.**

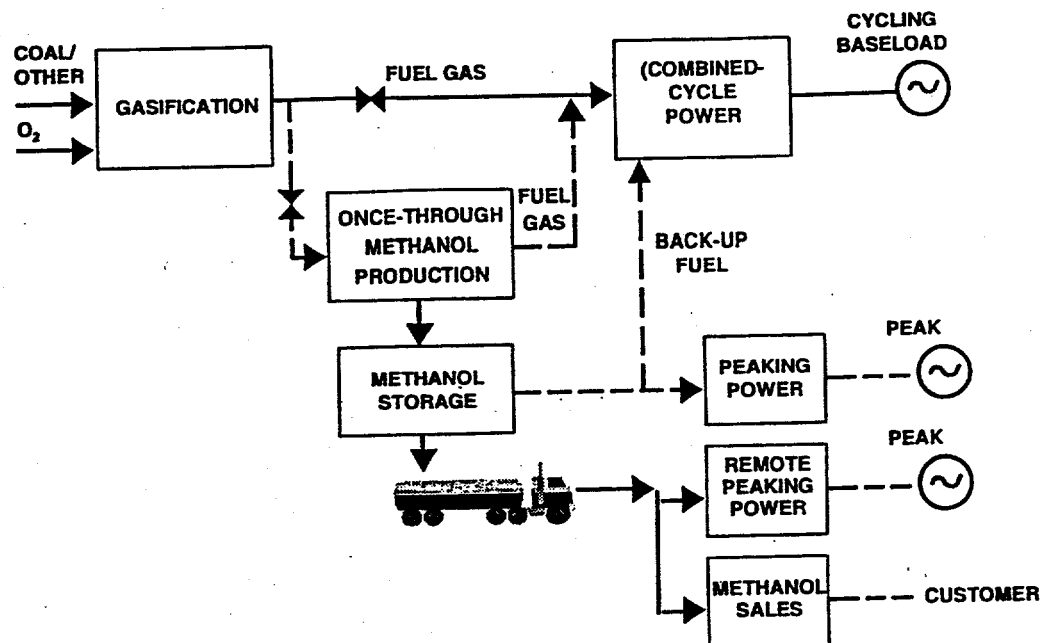


Figure 2. Once-through Methanol Coproduction with IGCC Electric Power

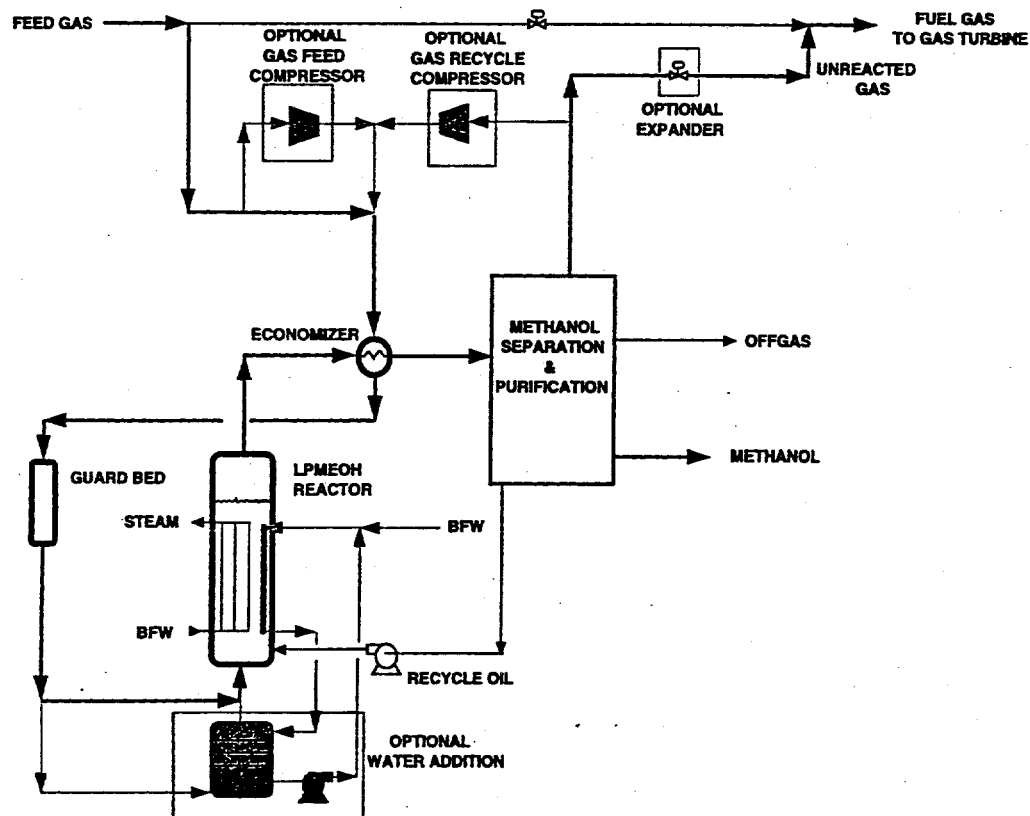


Figure 4. Once-through LPMEOH™ Process Design Options



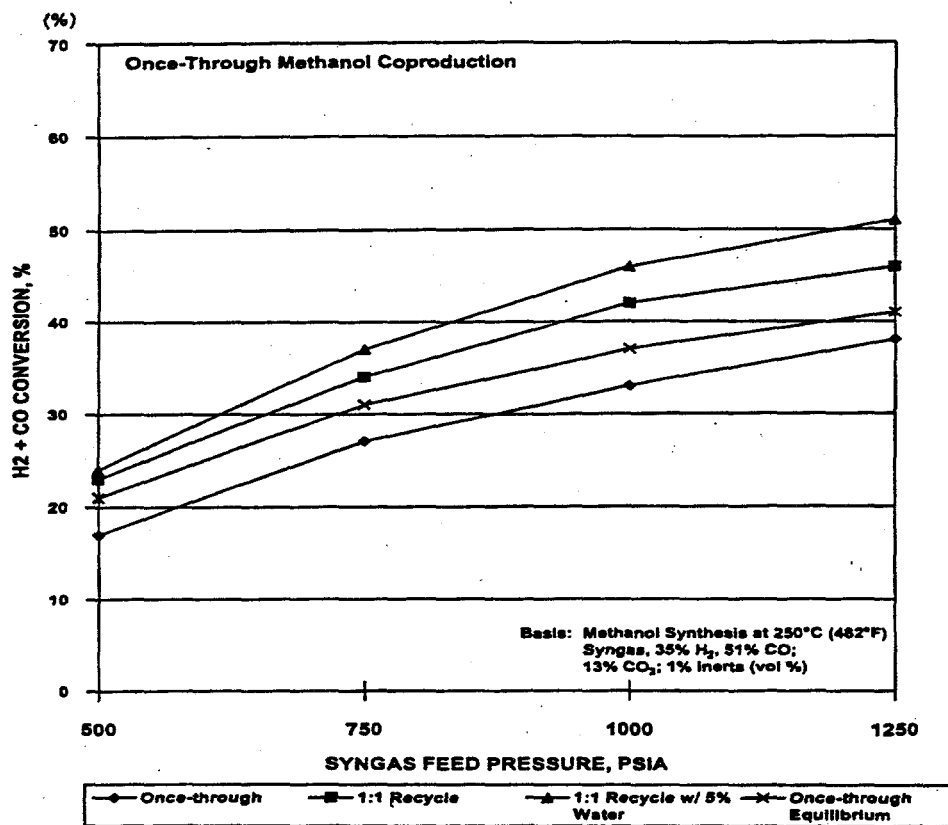


Figure 3. Synthesis Gas Conversion to Methanol

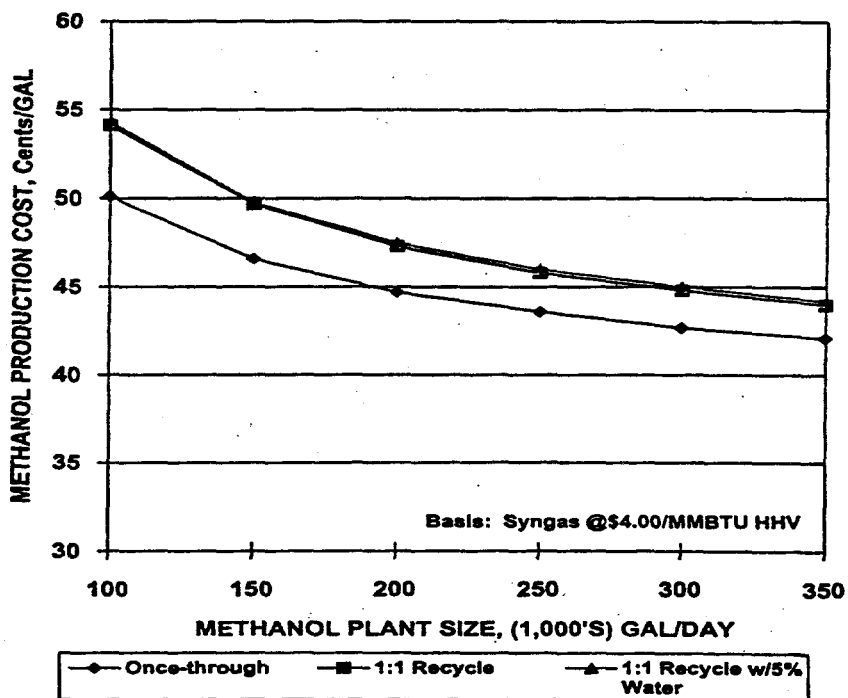


Figure 5. Coproduct Methanol Cost versus Methanol Plant Size.

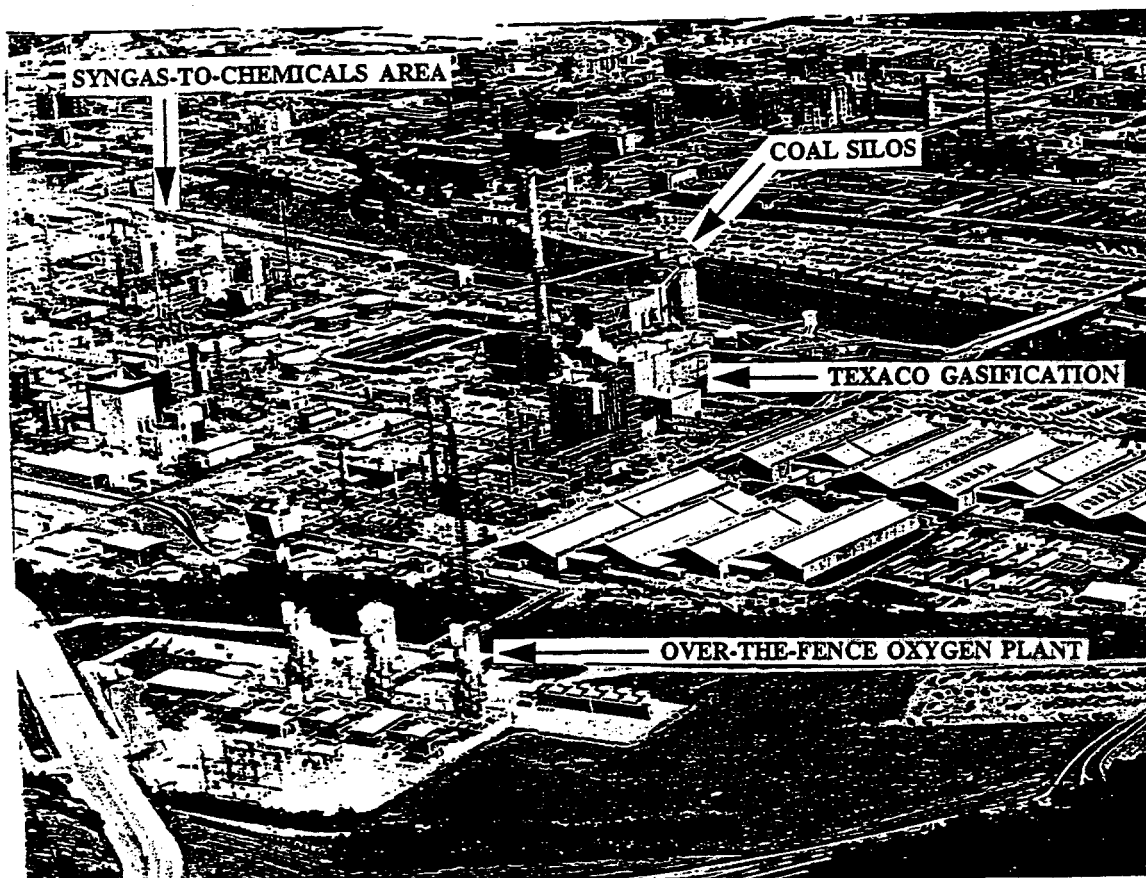


Figure 6. Aerial View of Eastman's Kingsport Complex

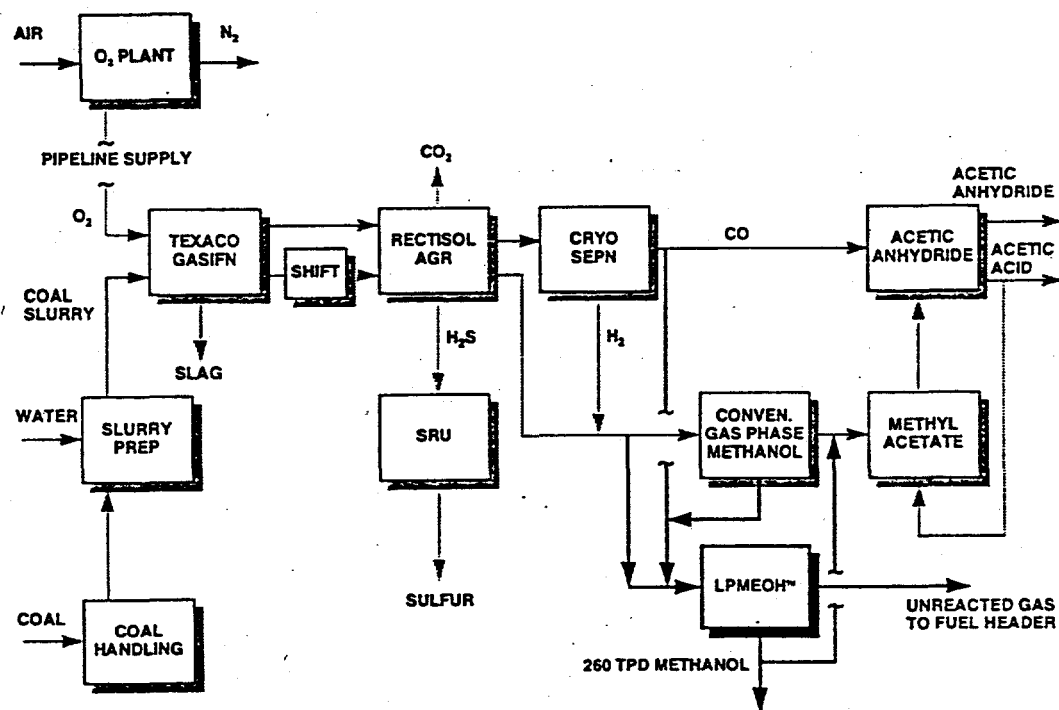


Figure 7. Process Block Flow Diagram of Kingsport Facility Including LPMEOH™ Demonstration Plant.

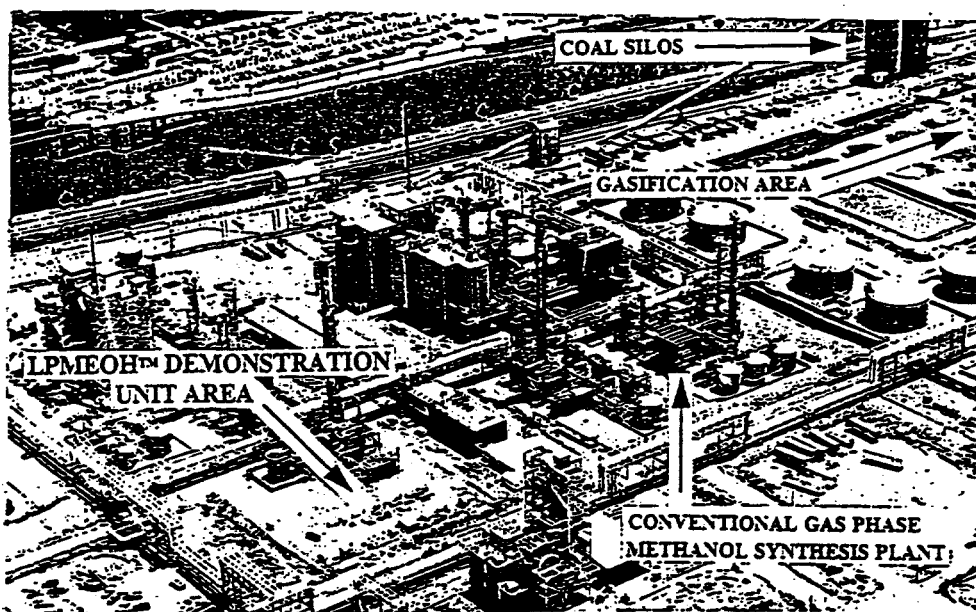


Figure 8. Aerial View of the Site for the LPMEOH™ Demonstration Plant

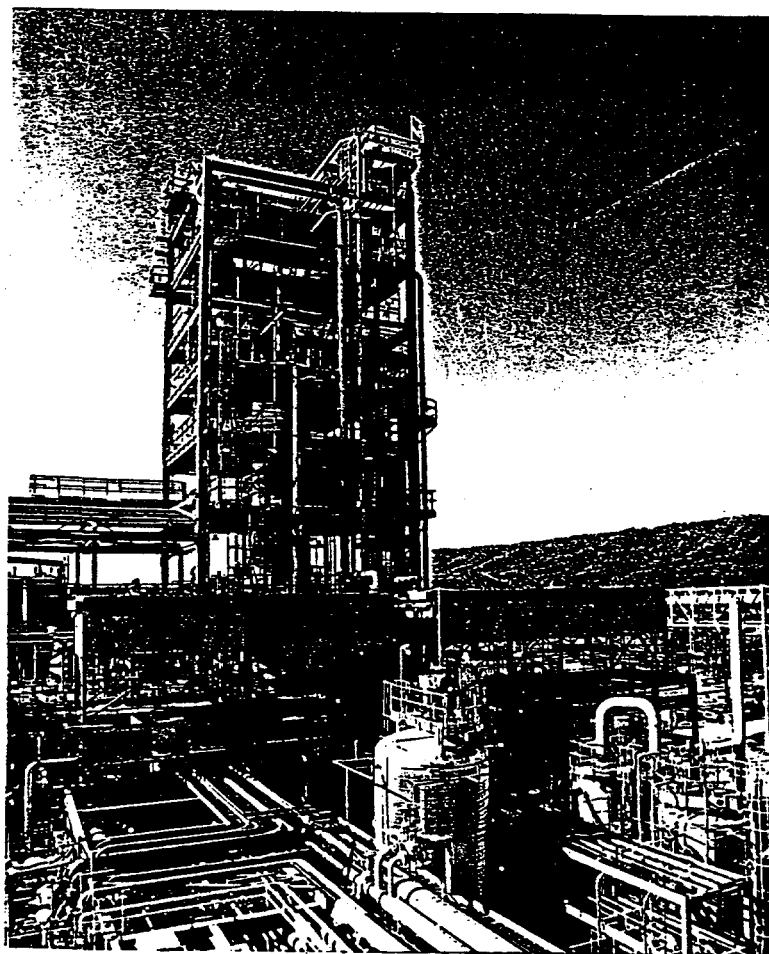


Figure 9. Photograph of the installed LPMEOH™ Demonstration Plant

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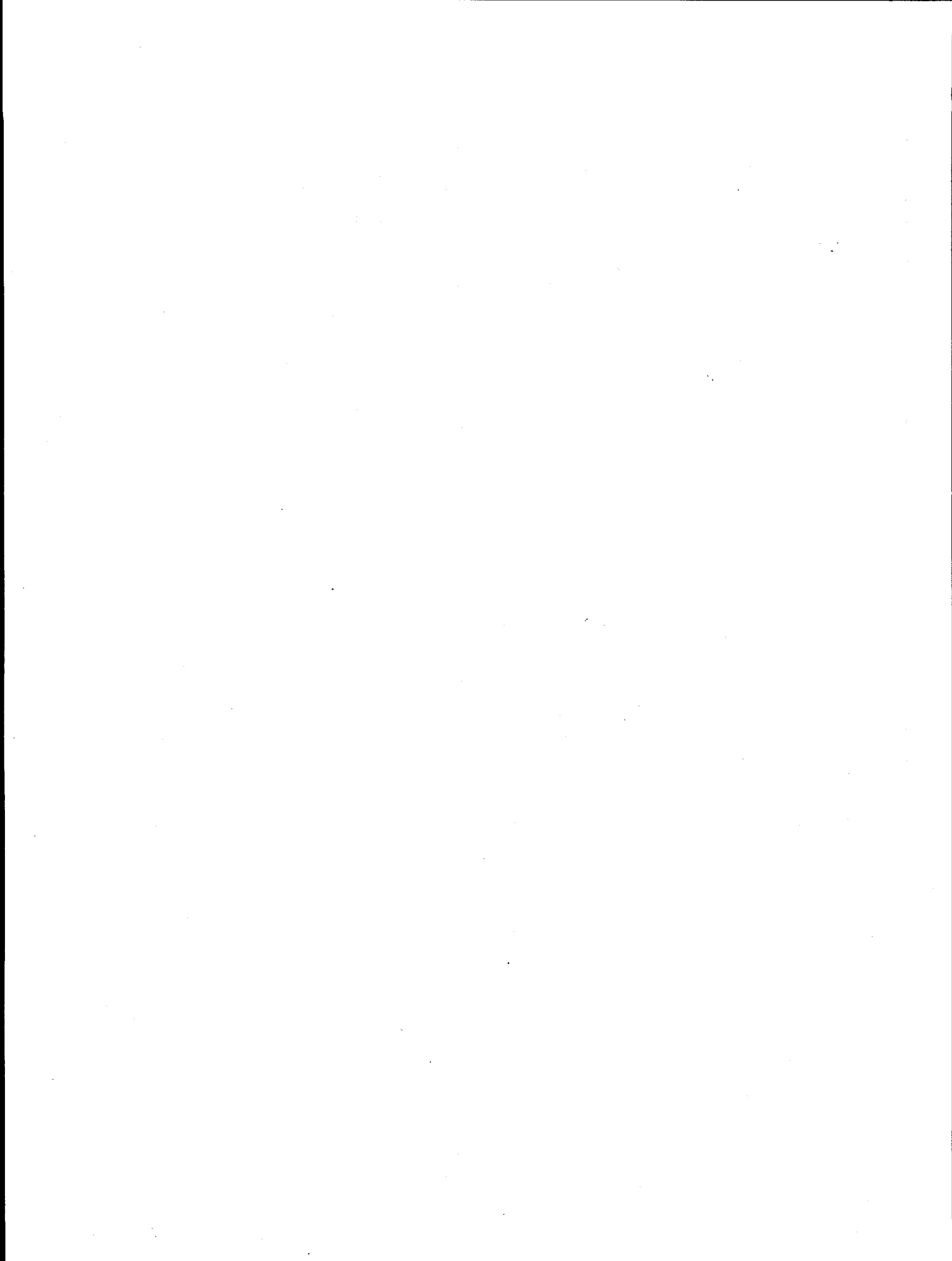
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## **Recent Experience with the CQE™**

**Clark D. Harrison and David B. Kehoe, CQ Inc.  
David C. O'Connor, Electric Power Research Institute  
G. Scott Stallard, Black & Veatch**

Increasing public awareness about the health of the global environment, tightening emissions regulations, growing competition among power producers, and advances in power generation technology are transforming the business of power generation worldwide. This transformation has further complicated fuel purchase decisions that profoundly affect the cost of electricity.

CQE (the Coal Quality Expert) is a software tool that brings a new level of sophistication to fuel decisions by seamlessly integrating the system-wide effects of fuel purchase decisions on power plant performance, emissions, and power generation costs.

The result of a \$21.7 million U.S. Clean Coal Technology project sponsored by the Department of Energy and the Electric Power Research Institute, CQE offers unparalleled advancements in technical capability, flexibility, and integration.

The CQE technology, which addresses fuel quality from the coal mine to the busbar and the stack, is an integration and improvement of predecessor software tools including:

- EPRI's Coal Quality Information System
- EPRI's Coal Cleaning Cost Model
- EPRI's Coal Quality Impact Model
- EPRI and DOE models to predict slagging and fouling

CQE can be used as a stand-alone workstation or as a network application for utilities, coal producers, and equipment manufacturers to perform detailed analyses of the impacts of coal quality, capital improvements, operational changes, and/or environmental compliance alternatives on power plant emissions, performance and production costs. It can be used as a comprehensive, precise and organized methodology for systematically evaluating all such impacts or it may be used in pieces with some default data to perform more strategic or comparative studies.

### **Overview of the Project**

The CQE project was conceived by EPRI to integrate the results and products of several on-going R&D projects into computer software that would become a worldwide standard for addressing fuel-related issues in the power industry. EPRI and DOE sponsored numerous coal quality R&D projects in the late 1970s and early 1980s to carefully examine and document the answers to questions that need to be addressed before a utility can be certain that it is operating its power plants within emissions limitations at the lowest possible cost:

- What are the economics of burning a prospective coal?

- How would the delivered price of coal change if the supplier cleans or blends the coal(s) to produce a product with quality characteristics different than the coal currently delivered to the power station?
- To what degree can the quality of the coal currently delivered to the power station be changed?
- What power plant equipment and systems are most affected or limited by coal quality?
- What are the trade-offs between increased capital spending at the power stations and increased cost of fuel for higher quality?
- How will alternative emissions control strategies affect the production cost of electricity at a specific unit?
- Are the slagging and fouling consequences of burning a prospective coal affordable?

Coal producers and equipment manufacturers must also address these questions from a different perspective to assess the potential value of alternative products and services for utilities. For example, a coal producer contemplating changes to an existing cleaning plant or a manufacturer trying to sell replacement parts for coal pulverizers would both be interested in using a model that could accurately determine pulverizer performance, power consumption and maintenance costs for potential utility customers to provide a fuel that matched plant/unit capabilities and goals. CQE was conceived as the tool to serve the needs of these prospective users as well as the utilities that were already using CQIM and related EPRI and DOE software.

## **Background and History of the Project**

In the mid 1970s, EPRI initiated its effort to understand the linkage between coal quality and power plant performance, emissions, and economics. Initial studies focused on the potential savings in capital cost of new coal-fired power stations that would result from the use of cleaner coal (1). To quantify the costs of producing cleaner coals and to evaluate the potential for physical coal cleaning to improve the quality of U.S. coals for power generation, EPRI initiated a coal cleanability characterization program at the Coal Cleaning Test Facility (CCTF) which it constructed in 1980-81. The facility's mission also included the demonstration of emerging coal cleaning technologies to accelerate their commercial deployment.

In 1982 EPRI started a parallel effort to build a state-of-the-art computer model that would predict power plant performance, production costs, and emissions based on laboratory and bench-scale coal quality measurements. The initial effort was focused on defining the specifications for the model and assembling the proven methodologies for predicting coal quality impacts on various power plant systems and components. A complementary effort to perform laboratory, bench-scale, and pilot-scale coal quality analyses was also initiated by EPRI in the mid 1980s, and since the Coal Cleaning Test Facility became the source for most of the combustion test samples, its name was changed to the Coal Quality Development Center (CQDC).

When the DOE Program Opportunity Notice for the Clean Coal Technology Program was issued on February 17, 1986, Combustion Engineering Inc. on behalf of EPRI prepared a proposal for the development of the Coal Quality Advisor that was later renamed the Coal Quality Expert, or CQE. The project proposed by Combustion Engineering included coal cleanability characterization of selected additional U.S. coals, laboratory, bench-scale, and pilot-scale combustion testing of representative samples of the run-of-mine and clean coal; full-scale power plant testing of those coals to verify coal quality effects; and the development of the software tool that would replace pilot-scale and full-scale demonstrations in the future. The proposal by Combustion Engineering was not selected from the initial awards for Round 1 of the Clean Coal Technology Program, so EPRI proceeded with some aspects of the proposed project in the meantime.

By the time the Combustion Engineering proposal was selected for negotiations in 1988, EPRI had completed the initial version of the Coal Quality Impact Model (CQIM™) and initiated some pilot-scale and commercial power plant testing programs. The result of these efforts and the previous work done by EPRI at the CQDC (and CCTF) were contributed by EPRI to the CQE project and the scope of the project was redefined to incorporate the testing and software development work necessary to complete a rigorous and robust model.

During the course of the project from May 1990 through mid-1996, computer technology and the methodology available to measure and predict coal quality continued to advance, so CQE was developed to incorporate as many of these advancements as possible and to maintain the flexibility to incorporate new features or update existing methodologies economically in the future.

### **Project Organization**

As EPRI's contractor with responsibility for bench-scale and pilot-scale testing to correlate coal quality characteristics to power plant performance, Combustion Engineering (now ABB CE) submitted the proposal for the CQE project to DOE. While the DOE CCT1 project award decisions were being made, EPRI engaged Black & Veatch to develop the original Coal Quality Impact Model software and Electric Power Technologies to conduct full-scale power plant coal quality impact tests. In addition, coal cleanability characterization efforts continued at the CQDC and EPRI developed plans to establish the CQDC as EPRI's wholly-owned subsidiary.

When DOE selected the CQE project for negotiation, EPRI and Combustion Engineering felt that it was appropriate for CQ Inc., EPRI's subsidiary, to integrate and manage the efforts of the project team as shown on the project organization chart, Figure 1-1.

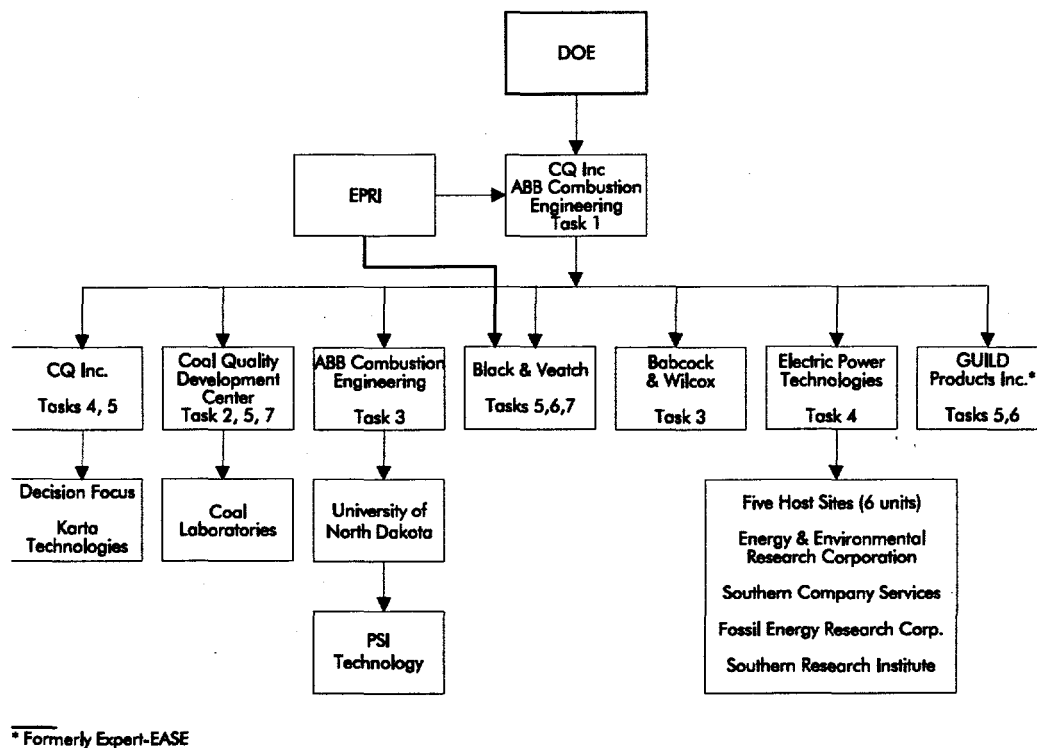
Under this organization, both CQ Inc. and Combustion Engineering executed the Cooperative Agreement with DOE and both contractors became co-prime contractors for the project with project management and administrative duties being delegated to CQ Inc. Consequently, the project was organized so that each participating organization other than EPRI and DOE would be subcontractors to CQ Inc.



As new computer technologies developed during the project and as the definition of CQE became more defined, some logical changes were made in the project organization. All software coding responsibilities were centralized at Black & Veatch. When a decision was made to exclude the Fireside Troubleshooting Guideline from the CQE code, Karta Technologies' role on the project ended, and when CQ Inc. required assistance with the design of the coal cleaning and blending models, Decision Focus was added to the project team as another subcontractor. The roles of the University of North Dakota Energy and Environmental Research Center (UNDEERC) and PSI Technology were also expanded to include the delivery of fouling and slagging prediction methodology to Black & Veatch.

In recognition of the value of CQE to their customers and to continue their support of EPRI's and DOE's coal quality R&D programs, ABB CE willingly reduced its scope and budget on the project to provide funding for more robust slagging and fouling models for CQE. ABB CE led the efforts with UNDEERC and PSI Technology that distinguish CQE from other software tools that rely on empirical indices to indicate potential slagging and fouling problems.

In addition to its role as co-sponsor, EPRI also provided technical leadership to the project for the pilot-scale and full-scale power plant testing programs and directly managed the software development tasks. EPRI's CQIM User's Group provided a sounding board for CQE development ideas and served as a project advisory committee. Moreover, five members of the user's group served as beta test users of the prototype software.



**Figure 1-1**  
**Project Organization Chart**

### Project Description

Although the project mission was to deliver a software tool, the scope of the project included numerous supporting tasks to collect and analyze data to form the basis for CQE algorithms, methodologies and submodels and to verify the accuracy and integrity of the CQE software at the conclusion of the project. These responsibilities are described in Table 1-1.

At the conclusion of each testing program, the responsible contractor prepared a detailed report and data summary for the host utility to use in addressing near-term problems and objectives and to aid the other CQE project contractors in completing their assigned tasks.

**Table 1-1**  
**CQE Organizational Responsibility Assignment**

Test Sites	ABB/CE and PSIT	B&W	B&V	CQ Inc.	UNDEERC	EPT	GUILD
Northeastern	5 DTFS 5 FPTF	NA	need FT/PT/BT data	2 CCC	4 DTFS 5 SEM	3 FT	NA
Watson	2 DTFS 2 FPTF	NA	need FT/PT/BT data	2 CCC	2 DTFS 2 SEM	2 FT	NA
King	NA	2 SBS	need FT/PT/BT data	5 CCC	2 SEM	2 FT	NA
Gaston	1 DTFS 1 FPTF	NA	need FT/PT/BT data	2 CCC	NA	2 FT	NA
Brayton Point	NA	NA	need FT data	NA	NA	2 FT	NA
Brayton Point	NA	NA	need FT data	NA	NA	2 FT	NA
Other CQE Work	commercial applications and slagging models	NA	CQE software developer, CQIM enhancements, ARA	Coal Cleaning Cost Model, CQIS enhancements, select CQE test sites	ash deposition data & fouling models	Fireside Testing Guidelines	develop CQE shell specs

CCC--Coal Cleanability Characterization  
SBS--Small Boiler Simulator (Pilot Test)  
BT--Bench Test  
DTFS--Drop Tube Furnace System

FT--Field Test  
PT--Pilot Test  
FPTF--Fireside Performance Test Facility (Pilot Test)  
SEM--Scanning Electron Microscopy

The highlights of the project are shown in Table 1-2.

The following U.S. electric utilities cofunded the project and participated in the field testing and software development/testing efforts.

Alabama Power Company  
Wilsonville, AL

Northern States Power  
Oak Park, MN

Duquesne Light Company  
Pittsburgh, PA

Public Service Company of Oklahoma  
Oologah, OK

Mississippi Power Company  
Gulfport, MS

Southern Company Services  
Birmingham, AL

New England Power Company  
Somerset, MA

**Table 1-2**  
**Project Accomplishments**

<b>Accomplishment</b>	<b>Date</b>
DOE awarded Cooperative Agreement	5/3/90
First of six field tests started	7/90
Pilot and bench-scale testing started	11/90
CQE specifications completed	2/15/92
Pilot and bench-scale testing completed	6/92
Acid Rain Advisor--first commercial product--released and copy sold	3/93
Completion of all six field tests	4/93
CQ Inc. and B&V signed CQE commercialization agreements	10/13/93
Conceptual design of the general Interactive Output Utility completed	8/94
Partially functional CQE beta version successfully tested	12/94
CQE alpha-version completed	3/31/95
CQE beta version completed and released for testing	6/95
Beta testing complete	11/30/95
CQE revised and issued on CD ROM	12/95
CQE Release 1.1 beta issued	6/7/96
Final Report	8/96
CQE Release 1.0	12/96

CQE builds on existing correlations from worldwide R&D on the impacts of coal quality for specific parts of the total power generation system. CQE features EPRI's CQIM as the calculational foundation for determining the impacts of different coals on plant performance and costs, and EPRI's Coal Quality Information System (CQIS™) provides a national database of coal quality information.

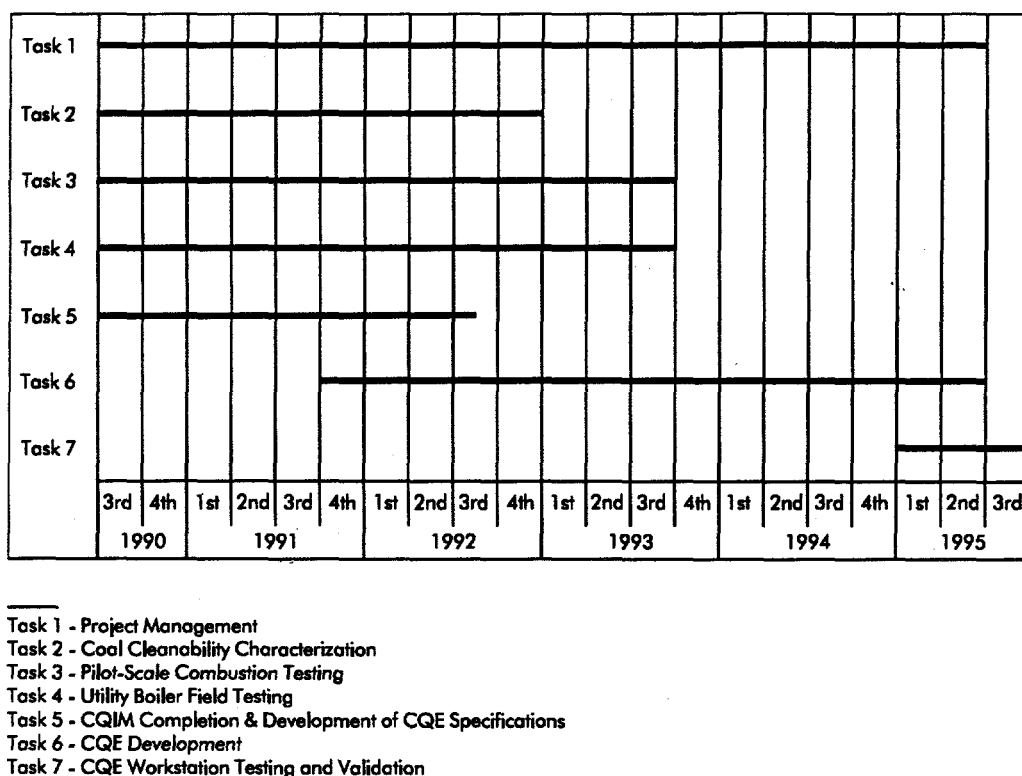
CQE combines the expertise from these established models--or the models themselves--into a single, personal computer-based tool. The electronic consultations that occur transparently between CQE's models let users address all aspects of fuel issues and their corresponding impacts on power generation systems.

This groundwork of established models is complemented by new and enhanced models derived from bench-, pilot-, and full scale test programs. These test programs, which allow coal-related effects to be distinguished from operational or design impacts, are among the most extensive of their kind ever conducted to relate power plant performance and emissions to coal quality.

### **Project Schedule**

The original 42-month project actually spanned 64 months because the required "off-the-shelf" software for OS/2 was late.

The extended duration of the project required increased funding from EPRI and DOE, but it ensured that CQE was adequately planned and that CQE's underlying computer software was adequately proven. The project schedule is given in Figure 1-2.



**Figure 1-2**  
**Project Schedule**

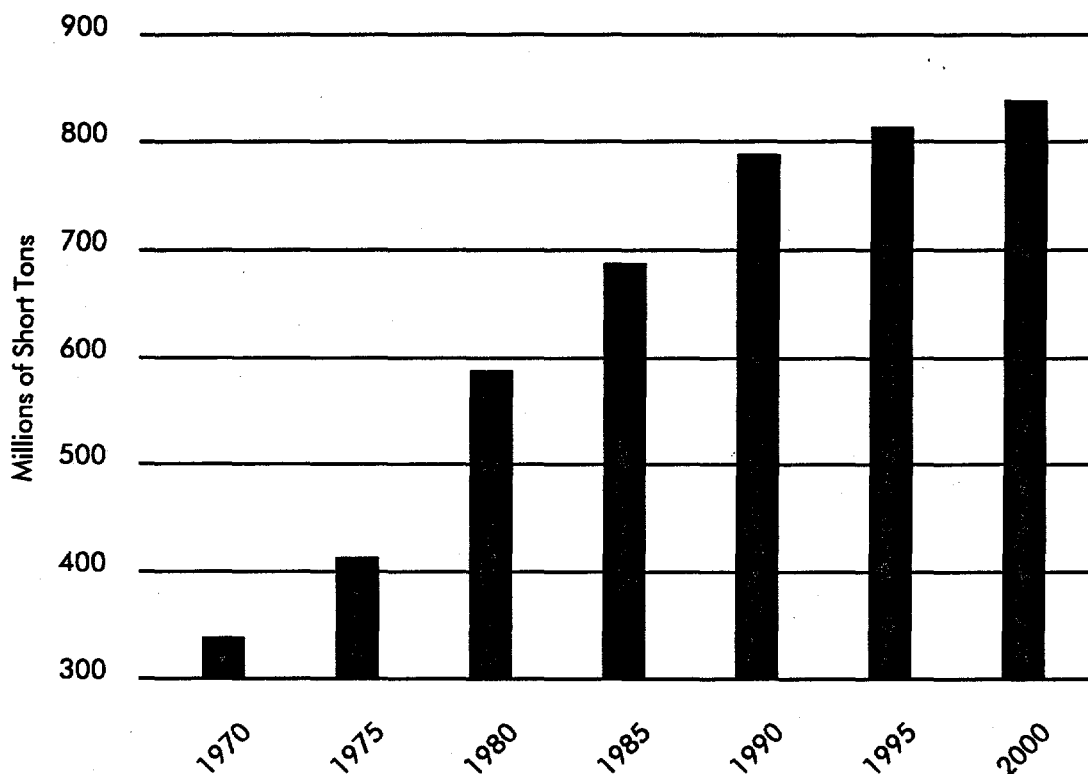
### Objectives of the Project

The work falls under DOE's Clean Coal Technology Program category of "Advanced Coal Cleaning." The 64-month project provides the utility industry with a PC software program to confidently and inexpensively evaluate the potential for coal cleaning, blending, and switching options to reduce emissions while producing the lowest cost electricity. Specifically, this project was designed to:

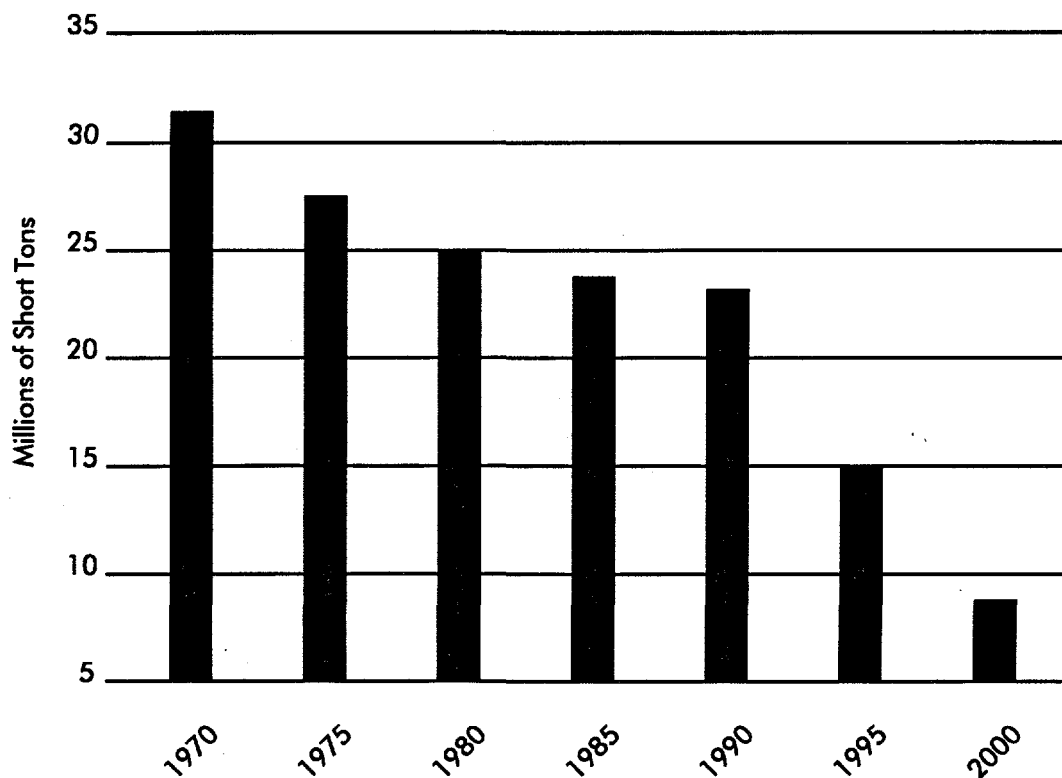
- Enhance the existing Coal Quality Information System (CQIS) database and Coal Quality Impact Model (CQIM) to allow confident assessment of the effects of cleaning on specific boiler cost and performance.
- Develop and validate a methodology, Coal Quality Expert (CQE), which allows accurate and detailed predictions of coal quality impacts on total power plant operating cost and performance.

## Significance of the Project

Originally, coal cleaning technologies were used only to remove ash-forming mineral matter. After passage of the 1970 Clean Air Act, coal cleaning processes were applied to a second purpose--sulfur reduction--accomplished primarily by removing the sulfur-bearing mineral pyrite. A great deal of geochemical information concerning the modes of occurrence of pyrite in coal was gathered and used to develop new methods of sulfur removal and to enhance existing methods. Today, coal cleaning plays a larger role in controlling SO<sub>2</sub> emissions than all post combustion control systems combined. It has led to reduced SO<sub>2</sub> emissions while U.S. coal use by utilities has increased steadily since 1970 (see Figures 1-3 and 1-4).

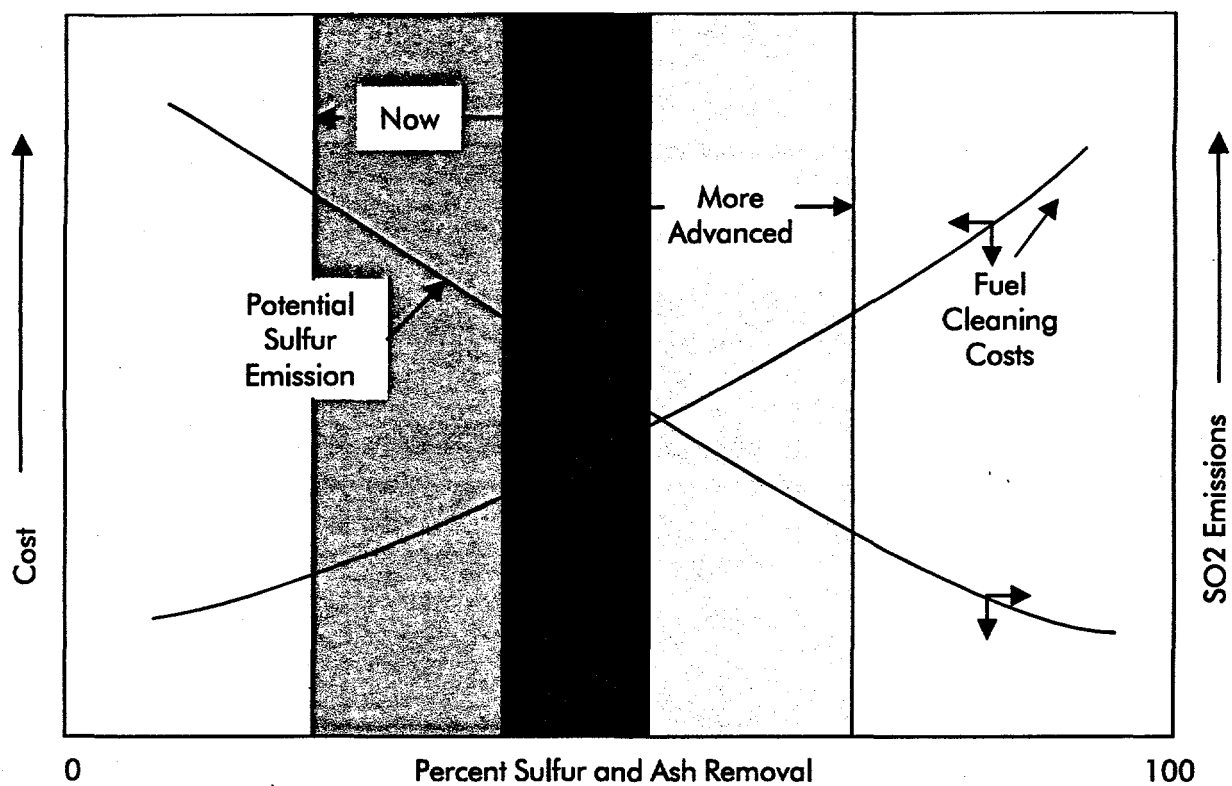


**Figure 1-3**  
**U.S. Utility Coal Use**

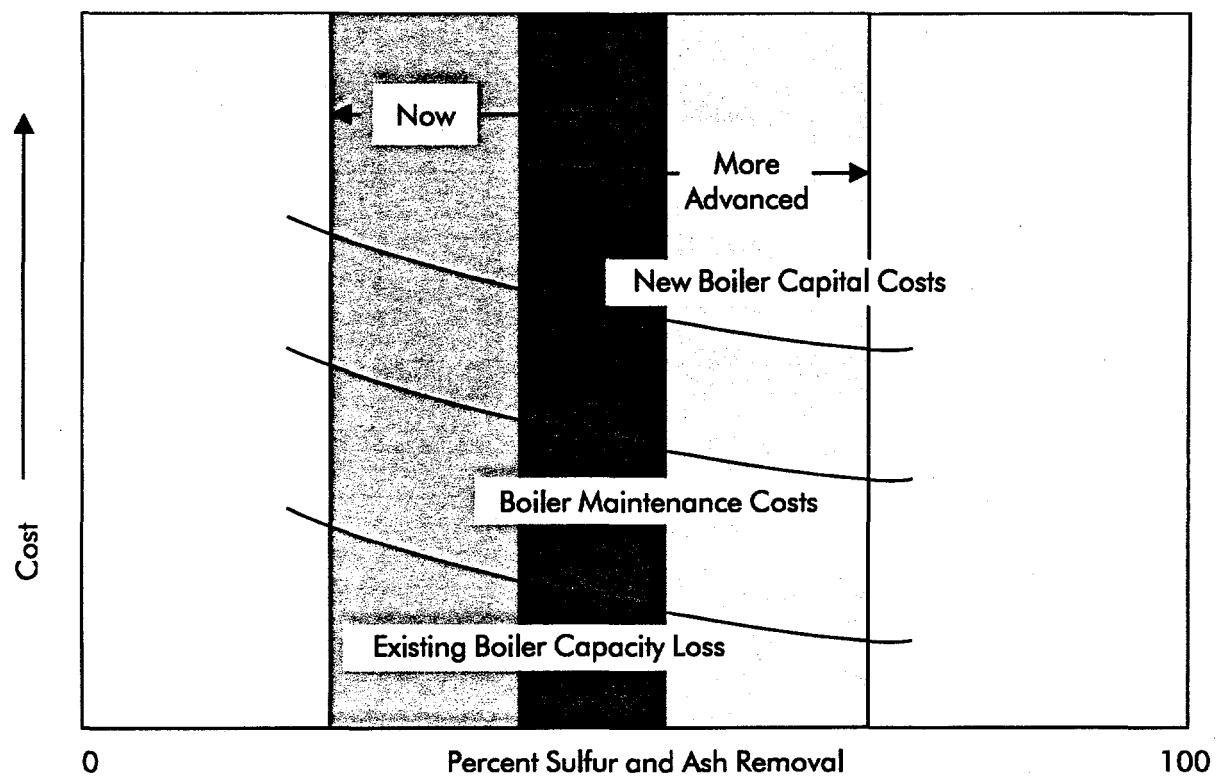


**Figure 1-4**  
**Total U.S. SO<sub>2</sub> Emissions**

Coal cleaning has been commercially demonstrated as a means of reducing sulfur concentrations in some types of coal to levels which allow firing in boilers to conform to environmental standards without using scrubbers. In addition, coal cleaning reduces the concentrations of mineral impurities which may result in significant improvements in boiler performance, reduced maintenance, and increased availability. Figures 1-5 and 1-6 illustrate trade-offs which dictate the feasibility of coal cleaning. Sulfur emissions produced when burning a coal generally decrease with increased levels of cleaning. Fuel costs, however, increase with increased levels of cleaning (Figure 1-5). Another consideration is that performance benefits can increase with increased cleaning for existing units and higher quality fuel reduces new unit capital costs (Figure 1-6).



**Figure 1-5**  
**The Relationship Between Sulfur Emissions and Fuel Costs**

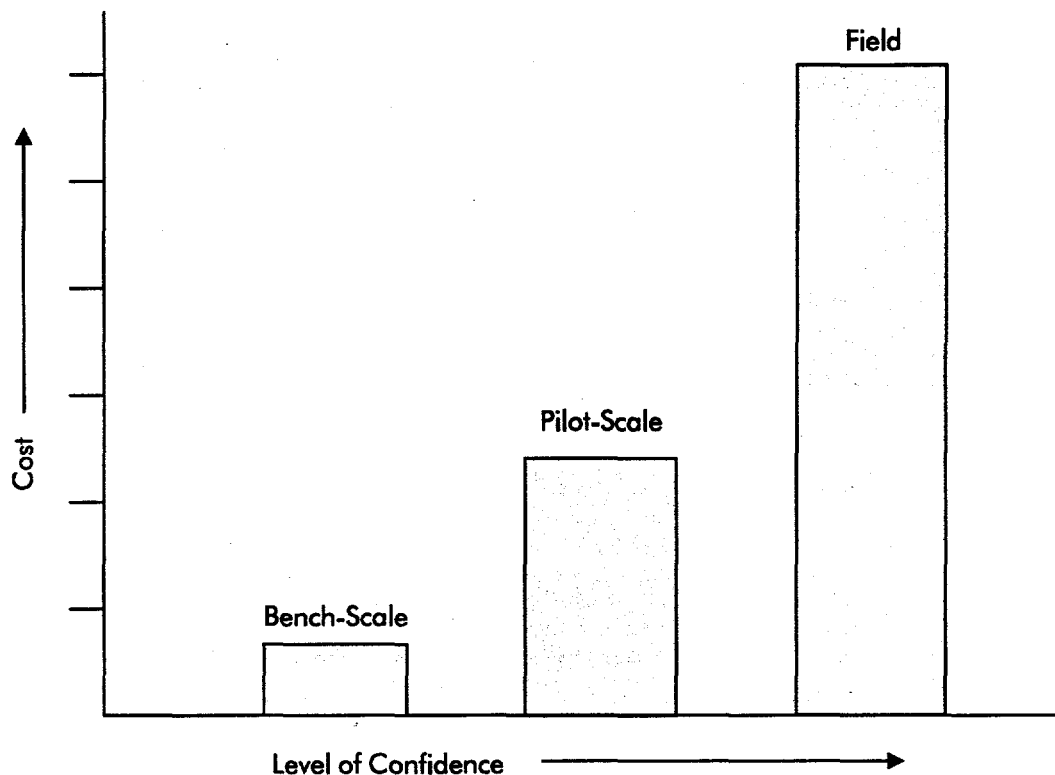


**Figure 1-6**



## Coal Cleaning to Reduce Power Production Cost

Studies have indicated significant economic benefits due to coal cleaning (2). However, to accurately and completely assess the commercial viability of cleaning a particular coal, detailed large-scale combustion testing is necessary. Quantification of performance savings is necessary to compare the economic benefits obtainable through coal cleaning with the costs of other techniques for emission control. Industry currently does not have the capability to reliably predict the performance of cleaned coals without extensive studies. The relationship between level of confidence and testing costs is illustrated in Figure 1-7. Since many of today's bench-scale coal performance indices rely on empirical correlations, extrapolation of these indices to fuels not represented by the specific database used for correlation can be misleading. The need for quick, inexpensive tests that can be reliably used to assess the commercial impacts of coal cleaning is vital to implement clean coal technology. One of the major goals of the program was to develop and demonstrate simple techniques (bench-scale fuel properties and predictive models) to allow industry to confidently assess the overall impacts of coal quality and the economic implications during utilization.



**Figure 1-7**  
**Relationship Between Testing Cost and Confidence Level of Commercial Predictions**

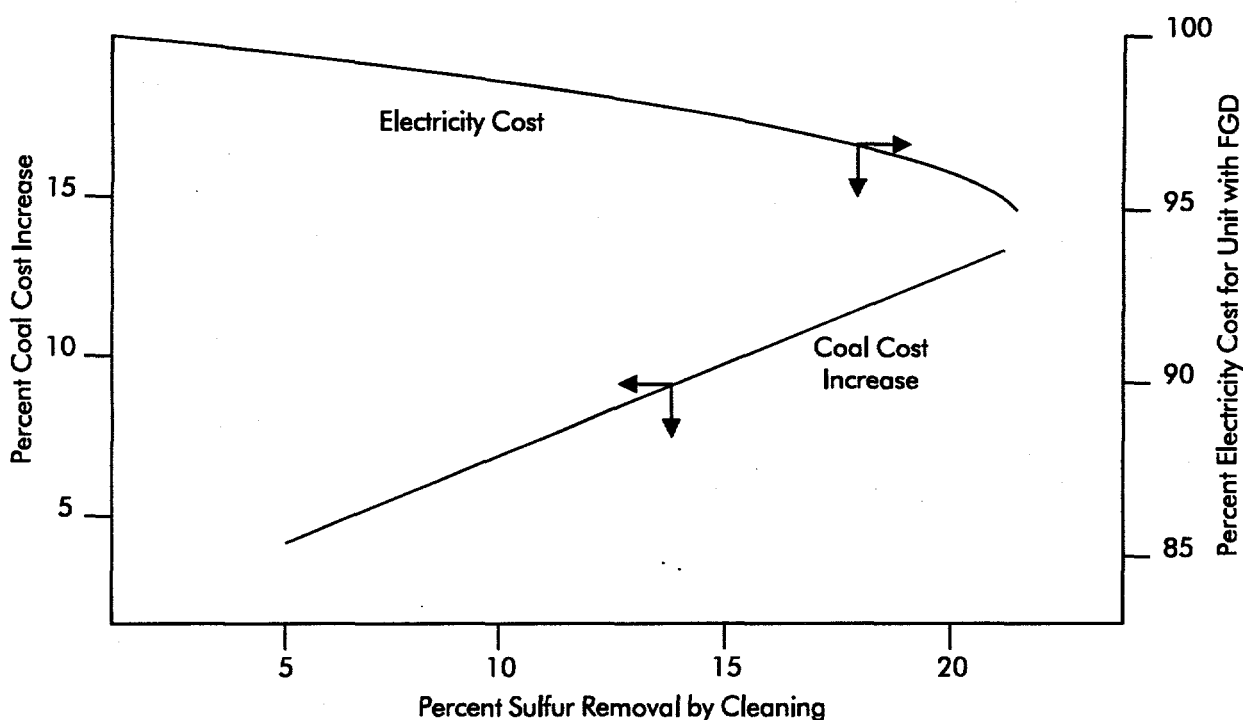
## The Significance of the CQE Tool

Fuel decisions affect nearly every aspect of power generation. Fuel buyers handle transportation issues and coal sourcing; plant engineers evaluate how individual coals behave in a unit; and environmental engineers address compliance and disposal issues. Typically, each expert uses an individual set of assumptions, data, and tools to complete an evaluation, resulting in one-dimensional pictures of fuel-related costs.

CQE integrates these assumptions, data, and tools, creating a unique electronic forum within which experts can efficiently and effectively share their knowledge and results.

The power of the forum is twofold. It not only centralizes all relevant information, it makes that information available to all other experts as appropriate. The end result of integrating a set of previously isolated analyses is a new capability that provides a complete picture of fuel-related impacts and costs.

One new capability, for instance, is CQE's ability to evaluate the economic tradeoffs between coal cleaning and scrubbing (Figure 1-8). Traditionally, utility engineers would combine results from two different models to compare the costs of cleaning and scrubbing. In contrast, a CQE analysis of cleaning versus scrubbing captures and consolidates the results of required analyses to determine the most cost-effective option or combination of options.



**Figure 1-8**  
**Economic Impact of Coal Cleaning**

## **Commercial Potential and Plans**

An analysis of the market for CQE shows that the most likely customers for CQE are power generation organizations, fuel suppliers, environmental organizations, government organizations, and engineering firms. These world-wide organizations can take advantage of CQE's ability to evaluate the impact of fuel quality on entire generating systems.

EPRI owns the software and distributes CQE to EPRI members for their use, and has contracted CQ Inc. as their commercialization agent. CQE is available to others in the form of three types of licenses: use, consultant, and commercialization. The largest market for use licenses with an introductory price of \$90,000 is power generation organizations. Coal producers and equipment manufacturers are also prospective users. Large architect/engineering firms and boiler manufacturers are most likely to purchase consultant licenses or regional or world-wide commercialization licenses.

Black & Veatch executed the first CQE commercialization license with CQ Inc (as agent) and CQ Inc. is also licensed to commercialize CQE. Under the terms of that license, B&V and CQ Inc. are working collaboratively to sell use and consultant's licenses worldwide to provide consultation to organizations with coal quality projects and to continue the development of CQE software enhancements. Copies of CQE's stand-alone Acid Rain Advisor have been licensed to two U.S. users to date.

## **Conclusions and Recommendations**

CQE will benefit owners and operators of coal-fired power plants in their commitments to produce energy economically and with concern for the environment. Utilities now have a tool to evaluate the system-wide consequences of fuel purchase decisions on power plant performance, emissions, and power generation costs. The software can examine potential changes in coal quality, transportation options, pulverizer performance, boiler slagging and fouling, emissions control alternatives and byproduct disposal for pulverized-coal and cyclone-fired power plants.

CQE will warrant further refinement and updating as new predictive models are validated. Future development of CQE should include coal gasification, fluidized bed boilers, European and Asian boiler design, and post combustion SO<sub>2</sub> and NO<sub>x</sub> control technologies that are successfully demonstrated in U.S. Clean Coal Technology projects.

## **References**

1. *Coal Preparation for Combustion and Conversion*, EPRI AF-791, Project 466-1 Final Report, May 1978, Gibbs & Hill Inc., NV.
2. *Impact of Coal Cleaning on the Cost of New Coal-fired Power Generation*, EPRI CS-1622, Project 1180-2 Final Report, May 1981, Bechtel National Inc., San Francisco, CA.

# **SELF-SCRUBBING COAL - AN INTEGRATED APPROACH TO CLEAN AIR**

**K. E. Harrison, P.E.  
Custom Coals Corporation  
Pittsburgh, Pennsylvania**

## **CONTENT**

**INTRODUCTION AND TECHNOLOGY OVERVIEW**

**PROJECT DESCRIPTION**

**PROJECT STATUS**

**STARTUP AND DEBUGGING**

**CIRCUIT OPTIMIZATION**

**EQUIPMENT PERFORMANCE**

**PLANS FOR COMPLETING THE PROJECT**

## **I. INTRODUCTION**

On October 29, 1992, a Cooperative Agreement was executed by the United States Department of Energy (DOE) and Custom Coals Corporation (CCC). This agreement provides for the design, construction and operation of a coal preparation facility to produce Carefree Coal and Self-Scrubbing Coal, two fuels that will provide many United States utilities the opportunity to achieve compliance with the 1990 Clean Air Act Amendments (CAAA) without incurring major expenditures for power plant modifications.

Carefree Coal is coal cleaned in a proprietary dense-media cyclone circuit, using ultrafine magnetite slurries, to remove noncombustible material, including up to 90% of the pyritic sulfur. Deep cleaning alone, however, cannot produce a compliance fuel from coals with high organic sulfur contents. In these cases, Self-Scrubbing Coal will be produced. Self-Scrubbing Coal is produced in the same manner as Carefree Coal except that the finest fraction of product from the cleaning circuit is mixed with limestone-based additives and briquetted. The reduced ash content of the deeply-cleaned coal will permit the addition of relatively large amounts of sorbent without exceeding boiler ash specifications or overloading electrostatic precipitators. This additive reacts with sulfur dioxide ( $\text{SO}_2$ ) during combustion of the coal to remove most of the remaining sulfur. Overall, sulfur reductions in the range of 80-90% are achieved.

After nearly 5 years of research and development of a proprietary coal cleaning technology coupled with pilot-scale validation studies of this technology and pilot-scale combustion testing of Self-Scrubbing Coal, CCC organized a team of experts to prepare a proposal in response to DOE's Round IV Program Opportunity Notice for its Clean Coal Technology Program under Public Law 101-121 and Public Law 101-512. The main objective of the demonstration project is the production of a coal fuel that will result in up to 90% reduction in sulfur emissions from coal-fired boilers at a cost competitive advantage over other technologies designed to accomplish the same sulfur emissions and over naturally occurring low sulfur coals.

## **II. PROJECT DESCRIPTION**

The Demonstration Project, called the Laurel Facility, consists of a 500 TPH state-of-the-art, coal preparation plant and various product and raw coal handling and storage facilities. During the current project operations phase, the advanced coal cleaning cyclone and various ancillary magnetite recovery schemes are being demonstrated as well as the demonstration of combustion of the Carefree Coal and Self-Scrubbing Coal at full size power plant boilers.

### **Goals**

CCC's goal for the project is to successfully commercialize its first plant and use that success to build a merchant coal preparation business. DOE's goal is to ensure the long term availability of a low cost, environmentally friendly fuel for our nation's long term energy needs.

## **Participants**

The Project Team assembled to carry out the demonstration project includes:

- DOE's Project Management Team from PETC
- Custom Coals Corporation (CCC), overall project manager and lessee of patents for the technology
- Affiliated Engineering Technologies, Inc., design contractor
- Riggs Industries, Construction Managers
- Richmond Power & Light, utility host site
- Centerior Energy, utility host site
- Pennsylvania Power & Light, utility host site

## **III. PROJECT STATUS**

- Design and construction of the facilities was completed in early 1996. Start-up began in late December 1995 and the first coal was processed on February 22, 1996. The plant circuits were fed an increasing amount of throughput and various adjustments to water and media flows were made until, in May of 1996, the facility reached its design capacity. Equipment and circuit optimization testing began immediately thereafter and have continued throughout the remainder of the year.
- One of the test burns, the Carefree Coal test at Pennsylvania Power and Light's Martins Creek Station, was conducted in mid-November. Although several of the plant circuits were performing below the expected proficiency because optimization has not been completed, the overall plant product produced for the test was consistent with the current quality of the plant feed coal.
- The later sections will detail the Start-up, the Circuit Optimization and the Equipment Performance work completed to date and provide the team's plans for completing the demonstration program.
- The project, as approved through Budget Period 3, calls for a total cost of \$87,386,102, with DOE providing \$37,994,437 or 43.5% of the funds. The project is expected to be completed in June 1997.



# **ROSEBUD SYNCOAL PARTNERSHIP SYNCOAL® DEMONSTRATION TECHNOLOGY UPDATE**

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Rosebud SynCoal Partnership  
Billings, Montana**

**Fifth Annual Clean Coal Technology Conference  
January 7-10, 1997  
Tampa, Florida**

## **SYNCOAL® PROCESS IMPROVES LOW-RANK COALS**

An Advanced Coal Conversion Process (ACCP) technology being demonstrated in eastern Montana (USA) at the heart of one of the world's largest coal deposits is providing evidence that the molecular structure of low-rank coals can be altered successfully to produce a unique product for a variety of utility and industrial applications.

The product is called SynCoal® and the process has been developed by the Rosebud SynCoal Partnership (RSCP) through the U.S. Department of Energy's multi-million dollar Clean Coal Technology Program. RSCP is a Colorado (USA) general partnership formed for the purpose of conducting the Clean Coal Technology Program demonstration and the commercializing of the ACCP technology.

Western SynCoal Company, a subsidiary of Montana Power Company's Energy Supply Division, is the managing general partner of RSCP. The other general partner is Scoria Inc. a subsidiary of NRG Energy, the nonutility entity of Northern States Power Company of Minnesota (USA).

Montana Power Company's subsidiary, Western Energy Company, initially developed the ACCP technology and signed the original Cooperative Agreement with the Department of Energy (DOE) to build the demonstration facility under the Clean Coal Technology Program (CCT I). Western Energy then formed Western SynCoal Company and joined with Scoria. RSCP's partners own the technology in undivided interests and have exclusively licensed it to the partnership. The RSCP partnership manages the \$105 million demonstration project adjacent to the Rosebud Mine at Colstrip, Montana and all activities related to technology commercialization. (See Demo Plant Location Map) DOE has committed \$43.125 million in funding to the demonstration project. Rosebud SynCoal is responsible for all additional funding and operation of the project.



The patented ACCP process improves the heating quality of low rank coals to produce an upgraded coal produced called SynCoal®, which is a registered trademark owned by RSCP.

## **Process**

The ACCP demonstration process uses low-pressure, superheated gases to process coal in vibrating fluidized beds. Two vibratory fluidized processing stages are used to heat and convert the coal. This is followed by a water spray quench and a vibratory fluidized stage to cool the coal. Pneumatic separators remove the solid impurities from the dried coal.

There are three major steps to the SynCoal® process: (1) thermal treatment of the coal in an inert atmosphere, (2) inert gas cooling of the hot coal, and (3) removal of ash minerals. See Flow Diagram

(1) During the thermal treatment process, raw coal from the stockpile is screened and fed into a two-stage thermal processing system. In the first vibratory fluidized-bed reactor, surface water is removed from the coal by heating it with hot combustion gas. When the coal exits this reactor, its temperature is slightly higher than that required to evaporate water. The coal is further heated to nearly 300° C (5° F) in a second reactor to a temperature sufficient to remove pore water and prompting decarboxylation. Here, particle shrinkage causes fracturing, destroys moisture reaction sites, and separates out the coal ash minerals.

(2) The coal then enters the coal cooler, where it is cooled to less than 150°F by contact with an inert gas (carbon dioxide and nitrogen at less than 100°F) in a vibrating fluidized bed cooler.

(3) In the last stage -- the coal cleaning system -- cooled coal is fed to deep bed stratifiers where air velocity and vibration separate mineral matter from the coal with rough gravity separation. The low specific gravity fractions are sent to a product conveyor while heavier specific gravity fractions go to fluidized bed separators, for additional ash removal. Fines from various parts of the cleaning process are collected in baghouses and cyclones, cooled and made available as an additional product line.

The SynCoal® is a high quality product with less than 5 percent moisture, sulfur content of 0.5 percent, ash content of about 9 percent, and a heating value of about 11,800 Btu per pound.

When operated continuously, the demonstration plant produces over 1,000 tons per day (up to 300,000 tons per year) of SynCoal® with a 2% moisture content, approximately 11,800 Btu/lb and less than 1.0 pound of SO<sub>2</sub> per million Btu. This product is obtained from Rosebud Mine sub-bituminous coal which starts with 25% moisture, 8,600 Btu/lb and approximately 1.6 pounds of SO<sub>2</sub> per million Btu.

Nearly 1.3 million tons of raw coal has been processed and over 850,000 tons of SynCoal® has been produced through October 1996. The plant has consistently operated at over 100% of design capacity and over 75% availability. **See SynCoal Production and Sales History and Monthly Operating Statistics**

### **Utility Applications - Customer Results**

A SynCoal® test-burn was completed at the 160 MW. J.E. Corette plant in Billings, Montana. A total of 204,478 tons of SynCoal® was burned between mid 1992 and April, 1996. The testing involved both handling and combustion of SynCoal® in a variety of blends. These blends ranged from approximately 15% SynCoal® to approximately 85% SynCoal®. Overall the results indicated that a 50% DSE SynCoal®/raw coal blend provided improved results with SO<sub>2</sub> emissions reduced by 21% overall, generation increased at normal operating loads and no noticeable impact on NO<sub>x</sub> emissions. DSE is a treatment to improve SynCoal®'s bulk handling characteristics when using conventional handling techniques. It controls dusting of the product and provides temporary resistance to spontaneous combustion.

Additionally SynCoal® deslagged the boiler at full load eliminating costly ash shedding operations and provided reduced gas flow resistance in the boiler and convection passage, reducing fan horsepower and improving heat transfer in the boiler area, resulting in increased generation by approximately 3 megawatts on a net basis.

Deliveries of SynCoal® are now being sent to Colstrip Project Units 1 & 2 in Colstrip, Montana. Testing has begun on the use of SynCoal® in these twin 320-megawatt pulverized coal fired plants. The results of these tests will provide information on: boiler efficiency, output, and air emissions. A total of 61,339 tons have been consumed to date.

A new SynCoal® delivery system is being designed which, if installed, would provide selectively controlled pneumatic delivery of SynCoal® to pulverizers individual pulverizers in the two units. This system would allow controlled tests in the two units providing valuable test data on emissions, performance and slagging. The use of both units operating at similar loads and with the same raw coal would provide a unique opportunity to perform directly comparative testing.

In May 1993, 190 tons of Center, North Dakota lignite was processed at the ACCP demonstration facility in Colstrip, producing 10,740 Btu/lb product and 47% reduction in sulfur and a 7% percent reduction in ash. In September 1993, a second test was performed processing 532 tons of lignite, producing a 10,567 Btu/lb product with a 48% sulfur reduction and a 27% ash reduction. The Center lignite before beneficiation had 36% moisture, approximately 6,800 Btu/lb at about 3.0 lbs of SO<sub>2</sub> per million Btu.

Approximately 190 tons of these upgraded products produced in September was returned two days later to the Milton R. Young Unit #1 and burned in an initial test showing dramatic improvement in cyclone combustion, improved slag tapping and a 13%

reduction in boiler air flow, reducing the auxiliary power loads on the forced draft and induced draft fans. Additionally the boiler efficiency increased from 82% to in excess of 86% and the total gross heat rate improved by 123 Btu/kWh hour.

The operation of the cyclone units at the Milton R. Young facility are plagued by cyclone barrel slugging which is typically removed by burning additional No. 2 fuel oil. These units also slag and foul in the boiler and convective passes requiring complete shutdown and cold boiler washing between three and four times a year.

In an effort to reduce these detrimental effects, Minnkota Power has tested the use of SynCoal® as a substitute for fuel oil when removing cyclone slag and also as a steady additive to reduce the boiler slugging and convective fouling to reduce the number of cold boiler washings necessary. The fuel oil substitute testing nicknamed "Klinker Killer" has been successfully tested showing the SynCoal® is at least as effective in removing cyclone barrel clinkers on a Btu for Btu basis as fuel oil. The SynCoal® produces a much higher temperature in the cyclone barrel than lignite increasing the cyclone barrel front wall temperature as much as 900° F and more closely matched the design temperature profile which improves the cyclone combustion operation dramatically.

The testing to support the long term objective has indicated that SynCoal® would be effective in this application although the limited duration of these tests has left them less than fully conclusive.

### **Industry Applications - Customer Results**

Several industrial cement and lime plants have been customers of SynCoal® for an extended period of time. A total of 129,056 tons have been delivered to these customers from 1993 through October 1996. In their testing and use of SynCoal® they have found that it improves their production from their direct fired kiln applications. These improvements are both in capacity and product quality as the steady flame produced by SynCoal® appears to allow tighter process control and process optimization in their operations.

A bentonite producer has been using SynCoal® as an additive in their green sand molding product for use in the foundry industry. The bentonite company has used SynCoal® since 1993 and has taken approximately 30,569 tons. SynCoal® has been found to be a very consistent product allowing their customers to reduce the quantity of additives used and improving the quality of the metal casting produced.

### **Commercialization**

Western SynCoal Company has moved closer to building a \$37.5 million commercial SynCoal plant at Minnkota's Milton R. Young Power Station near Center, North Dakota.

Minnkota is a generation and transmission cooperative supplying wholesale electricity to 12 rural electric cooperatives in eastern North Dakota and northwestern Minnesota.

Minnkota owns and operates the 250-megawatt Unit 1 at the Young Station, and operates the 438-mw Unit 2 which is owned by Square Butte Electric Cooperative of Grand Forks. This power station is already one of the lowest cost electric generating plants in the nation; however, with the use of SynCoal® the operations of the plant could further improve.

The SynCoal plant would produce an estimated 403,000 tons of finished product annually, which would be blended with the lignite. The reduced slagging and fouling improves generating plant maintenance and allows potentially longer runs between downtimes to ultimately produce more electricity. The process is anticipated to boost the lignite heating value by 60 percent and could lower its sulfur content by 50 percent with an anticipated second phase of the project.

### **International**

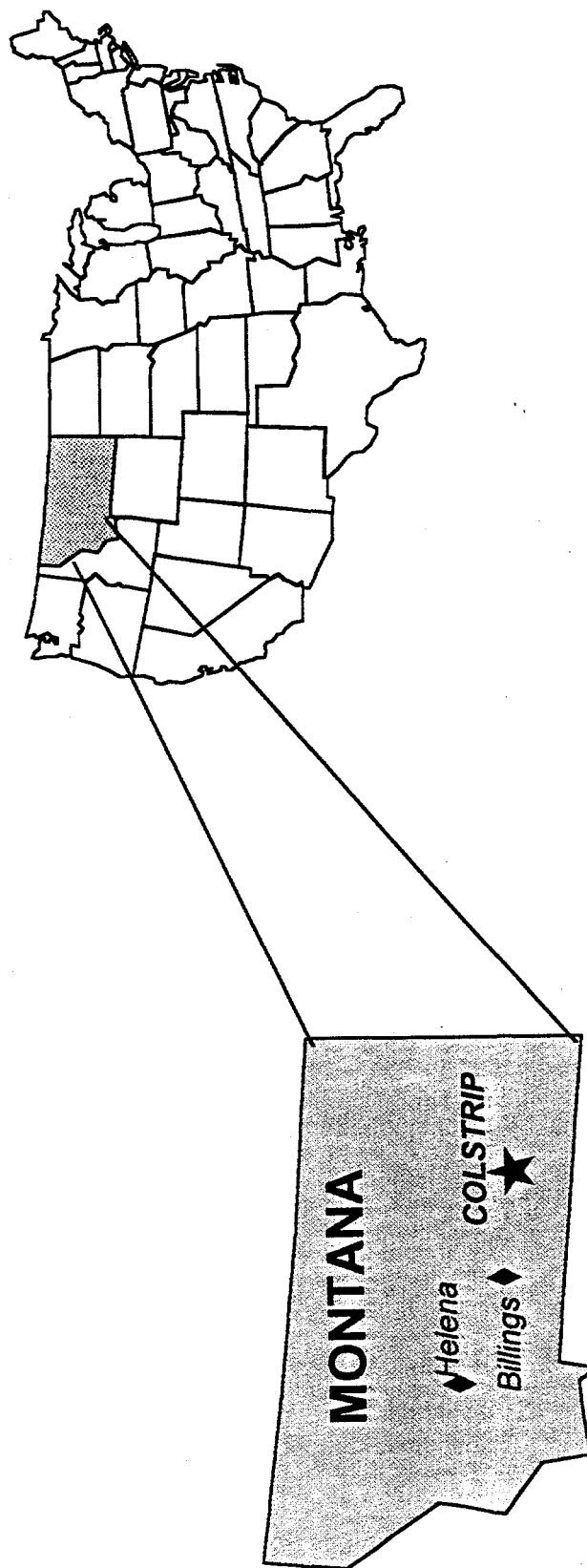
RSCP has been actively marketing and promoting the SynCoal® technology world-wide. RSCP has been working closely with a Japanese equipment and technology company to expand into Asian markets. Prospects are also being pursued in Europe currently.

### **Summary**

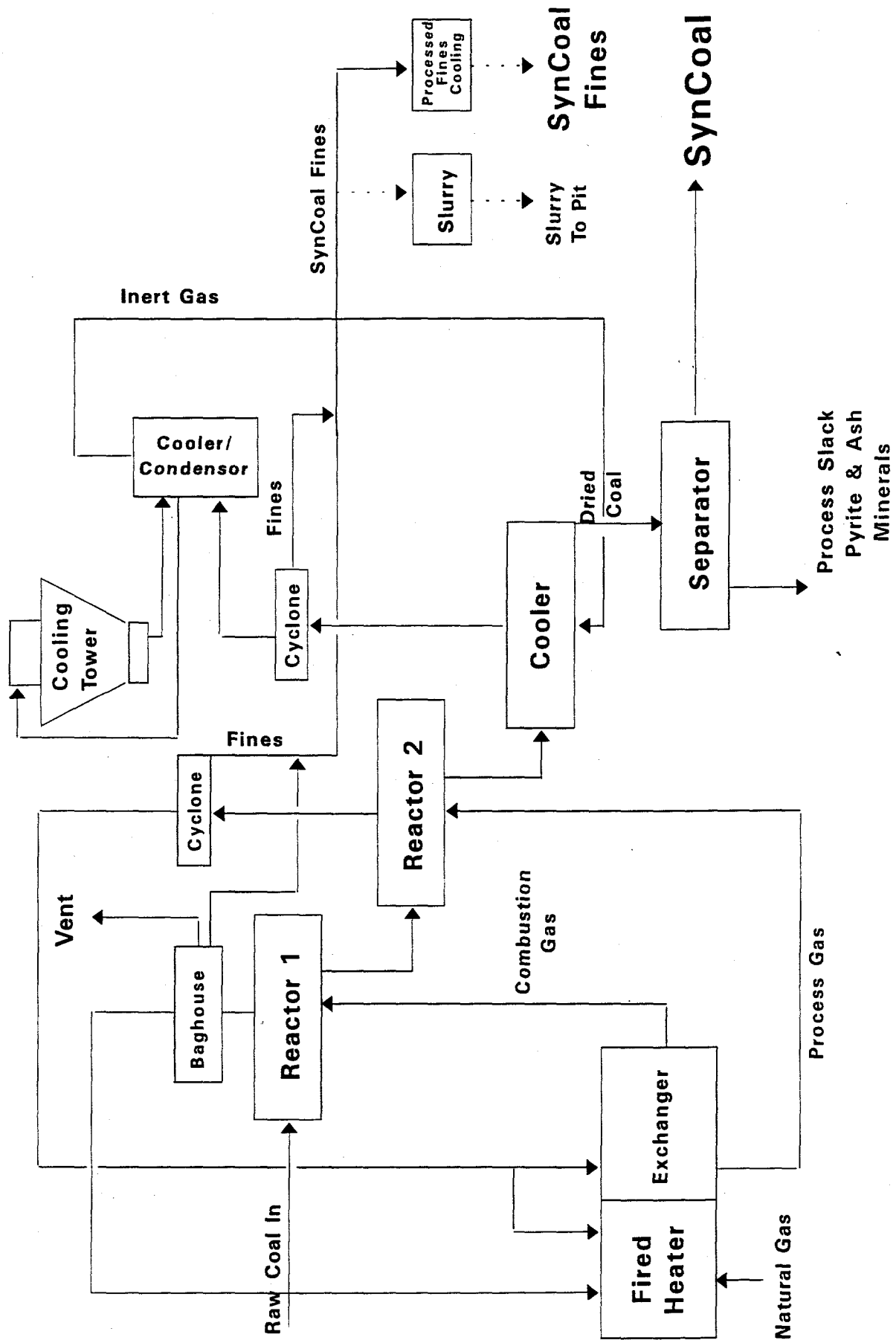
Rosebud SynCoal is continuing to advance the SynCoal® technology in a prudent and organized manner. The work to date has made SynCoal® the most advanced Low Rank Coal upgrading technology available and has put it on the cusp of commercial viability. The successful conclusion of the Center SynCoal Project and the enhanced SynCoal® delivery system and testing in Colstrip will position SynCoal® to be a viable option to enhance low rank coal fired utility operations.

# ADVANCED COAL CONVERSION PROCESS

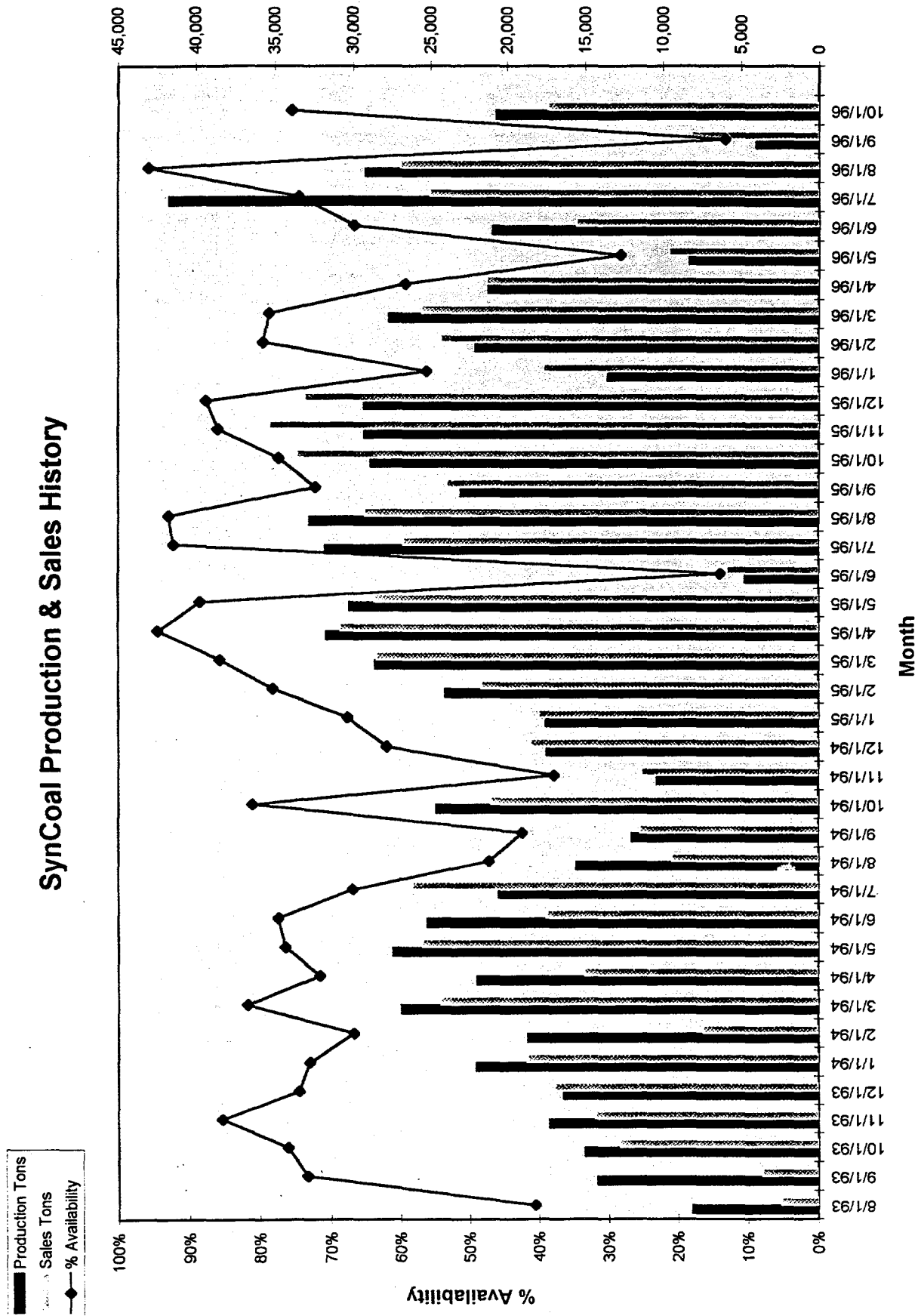
Demo Plant Location



# Process Flow Diagram



# SynCoal Production & Sales History



## ACCP MONTHLY OPERATING STATISTICS

MONTH	PRODUCTION AVAILABILITY	FORCED OUTAGE RATE	TONS PROCESSED	CAPACITY FACTOR	SHIPMENTS
Mar-92	4%	96%	700	2%	181
Apr-92	7%	89%	411	1%	212
May-92	12%	76%	2,757	7%	0
Jun-92	13%	81%	2,496	7%	214
Jul-92	7%	56%	1,436	4%	0
Aug-92	17%	60%	1,860	5%	61
Sep-92	44%	33%	8,725	24%	1,672
Oct-92	13%	63%	2,292	6%	523
Nov-92	58%	28%	6,946	19%	2,386
Dec-92	11%	80%	1,063	3%	317
Jan-93	53%	26%	8,626	23%	3,658
Feb-93	44%	18%	6,544	19%	915
Mar-93	44%	34%	6,565	17%	629
Apr-93	49%	30%	8,514	23%	745
May-93	47%	39%	9,256	24%	768
Jun-93	15%	26%	2,752	7%	199
Jul-93	0%		0	0%	655
Aug-93	41%	43%	13,427	35%	2,361
Sep-93	73%	18%	23,276	63%	3,528
Oct-93	76%	11%	24,606	64%	12,753
Nov-93	85%	14%	27,927	76%	14,349
Dec-93	74%	9%	26,009	68%	16,951
Jan-94	73%	17%	34,979	92%	19,093
Feb-94	67%	28%	29,247	85%	7,909
Mar-94	82%	14%	41,891	110%	24,627
Apr-94	72%	27%	33,686	91%	15,622
May-94	76%	8%	39,404	103%	26,415
Jun-94	77%	23%	36,657	99%	18,873
Jul-94	67%	33%	34,026	89%	26,527
Aug-94	47%	19%	24,645	64%	9,146
Sep-94	42%	35%	20,327	55%	11,408
Oct-94	81%	16%	34,908	91%	19,161
Nov-94	38%	62%	16,418	44%	11,169
Dec-94	62%	27%	25,258	66%	18,478
Jan-95	68%	32%	31,726	83%	17,695
Feb-95	78%	22%	38,325	111%	21,710
Mar-95	86%	4%	42,674	112%	28,548
Apr-95	94%	1%	47,818	129%	30,827



MONTH	PRODUCTION AVAILABILITY	FORCED OUTAGE RATE	TONS PROCESSED	CAPACITY FACTOR	SHIPMENTS
May-95	88%	5%	43,752	114%	28,674
Jun-95	14%	26%	7,142	19%	5,859
Jul-95	92%	8%	48,512	127%	26,795
Aug-95	93%	4%	48,889	128%	29,261
Sep-95	72%	28%	37,129	100%	23,954
Oct-95	77%	10%	43,316	113%	33,614
Nov-95	86%	14%	42,807	116%	35,380
Dec-95	88%	13%	47,531	124%	33,101
Jan-96	56%	44%	24,710	65%	17,662
Feb-96	79%	21%	36,280	101%	24,340
Mar-96	79%	21%	39,071	104%	25,566
Apr-96	59%	19%	30,038	81%	21,321
May-96	28%	11%	13,282	35%	9,571
Jun-96	67%	21%	31,775	85%	15,553
Jul-96	74%	26%	35,056	92%	24,998
Aug-96	96%	4%	43,832	117%	29,200
Sept-96	13%	33%	6,117	16%	8,112
Oct-96	75%	25%	33,730	90%	17,375
TOTAL			1,331,146		780,621

# **Technical Session II**

## **Advanced Industrial Systems**

# **AN UPDATE ON BLAST FURNACE GRANULAR COAL INJECTION**

**D. G. Hill**

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## **ABSTRACT**

A blast furnace coal injection system has been constructed and is being used on the furnaces at the Burns Harbor Division of Bethlehem Steel. The injection system was designed to deliver both granular (coarse) and pulverized (fine) coal. Construction was completed on schedule in early 1995. Coal injection rates on the two Burns Harbor furnaces were increased throughout 1995 and was over 200 lbs/ton on C furnace in September. The injection rate on C furnace reached 270 lbs/ton by mid-1996. A comparison of high volatile and low volatile coals as injectants shows that low volatile coal replaces more coke and results in a better blast furnace operation. The replacement ratio with low volatile coal is 0.96 lbs coke per pound of coal. A major conclusion of the work to date is that granular coal injection performs very well in large blast furnaces. Future testing will include a processed sub-bituminous coal, a high ash coal and a direct comparison of granular versus pulverized coal injection.

## **I. INTRODUCTION**

A blast furnace coal injection system has been installed at the Burns Harbor Division of Bethlehem Steel Corporation. This is the first blast furnace coal injection system in the US that has been designed to deliver granular (coarse) coal - all previously installed blast furnace coal injection systems in the US have been designed to deliver pulverized (fine) coal. Financial assistance for the coal injection system was provided by the Clean Coal Technology Program.

The use of granular coal in blast furnaces was jointly developed by British Steel and Simon-Macawber (now CPC-Macawber) and used at the Scunthorpe Works in England. The blast furnaces at Scunthorpe have about one-half the production capability of the Burns Harbor blast furnaces. Therefore, one of the main objectives of the Clean Coal Technology (CCT) test program at Burns Harbor is to determine the effect of granular coal injection on large high

productivity blast furnaces. Another objective of the CCT test program at Burns Harbor is to determine the effect of different types of US coals on blast furnace performance.

The Burns Harbor Plant produces flat rolled steel products for the automotive, machinery and construction markets. The Plant is located on the southern shore of Lake Michigan about 30 miles east of Chicago. Burns Harbor is an integrated operation that includes two coke oven batteries, an iron ore sintering plant, two blast furnaces, a three vessel BOF shop and two twin-strand slab casting machines. These primary facilities can produce over five million tons of raw steel per year. The steel finishing facilities at Burns Harbor include a hot strip mill, two plate mills, a cold tandem mill complex and a hot dip coating line.

When originally designed and laid-out, the Burns Harbor Plant could produce all the coke required for the two blast furnaces operating at 10,000 tons/day. However, improved practices and raw materials have resulted in a blast furnace operation that now can produce over 14,000 tons/day. Since the coke oven batteries are not able to produce the coke required for a 14,000 ton/day blast furnace output, other sources of coke and energy have been used to fill the gap. Over the years, coke has been shipped to Burns Harbor from other Bethlehem plants and from outside coke suppliers. In addition, auxiliary fuels have been injected into the furnaces to reduce the coke requirements. The auxiliary fuels have included coal tar, fuel oil and natural gas. The most successful auxiliary fuel through the 1980s and early 1990s has been natural gas. It is easy to inject and, at moderate injection levels, has a highly beneficial effect on blast furnace operations and performance. However, there are two significant problems with the use of natural gas in blast furnaces. One problem is the cost and the other is the amount that can be injected and, therefore, the amount of coke that can be replaced. Our process and economic studies showed that more coke could be replaced and iron costs could be reduced by injecting coal instead of natural gas in the Burns Harbor furnaces.

This led Bethlehem to submit a proposal to the DOE to conduct a comprehensive assessment of coal injection at Burns Harbor. Following an extensive review by the DOE, Bethlehem's Blast Furnace Granular Coal Injection System Demonstration Project was one of thirteen demonstration projects accepted for funding in the Clean Coal Technology Program third round of competition. The primary thrust of the project is to demonstrate commercial performance characteristics of granular coal as a supplemental fuel for steel industry blast furnaces. The technology will be demonstrated on large high productivity blast furnaces using a wide range of coal types available in the US. The planned tests will assess the impact of coal particle size distribution as well as chemistry on the amount of coal that can be injected effectively. Upon successful completion of the work, the results will provide the information and confidence needed by others to assess the technical and economic advantages of applying the technology to their own facilities.

A major consideration in evaluating coal injection in the US is the aging capacity of existing cokemaking facilities and the high capital cost to rebuild these facilities to meet emission guidelines under the Clean Air Act Amendments. The increasingly stringent environmental regulations and the continuing decline in domestic cokemaking capability will cause significant reductions in the availability of commercial coke over the coming years. Due to this decline in

availability and increase in operating and maintenance costs for domestic cokemaking facilities, commercial coke prices are projected to increase by more than general inflation. Higher levels of blast furnace injectants, such as coal, enable domestic integrated steel producers to minimize their dependence on coke.

### **Blast Furnace Process**

The ironmaking blast furnace is at the heart of integrated steelmaking operations. As shown in Figure 1, the raw materials are charged to the top of the furnace through a lock hopper arrangement to prevent the escape of pressurized hot reducing gases. Air needed for the combustion of coke to generate the heat and reducing gases for the process is passed through stoves and heated to 1500-2300°F. The heated air (hot blast) is conveyed to a refractory-lined bustle pipe located around the perimeter of the furnace. The hot blast then enters the furnace through a series of ports (tuyeres) around and near the base of the furnace. The molten iron and slag are discharged through openings (tapholes) located below the tuyeres. The molten iron flows to refractory-lined ladles for transport to the basic oxygen furnaces.

A schematic showing the various zones inside the blast furnace is shown in Figure 2. As can be seen, the raw materials, which are charged to the furnace in batches, create discrete layers of ore and coke. As the hot blast reacts with and consumes coke at the tuyere zone, the burden descends in the furnace resulting in a molten pool of iron flowing around unburned coke just above the furnace bottom (bosh area). Reduction of the descending ore occurs by reaction with the rising hot reducing gas that is formed when coke is burned at the tuyeres.

The cohesive zone directly above the tuyeres is so called because it is in this area that the partially reduced ore is being melted and passes through layers of coke. The coke layers provide the permeability needed for the hot gases to pass through this zone to the upper portion of the furnace. Unlike coal, coke has the high temperature properties needed to retain its integrity in this region and is the reason that blast furnaces cannot be operated without coke in the burden.

The hot gas leaving the top of the furnace is cooled and cleaned. Since it has a significant heating value (80-100 Btu/scf), it is used to fire the hot blast stoves. The excess is used to generate steam and power for other uses within the plant.

## **II. COAL INJECTION TECHNOLOGY**

Bethlehem decided to utilize the CPC Macawber Blast Furnace Granular Coal Injection (BFGCI) System, because unlike more widely used systems that utilize only pulverized coal, it is capable of injecting both granular and pulverized coal. Bethlehem believes that the CPC Macawber system offers a variety of technical and economic advantages which make this system potentially very attractive for application in the US basic steel industry. A schematic showing the application of the technology to the blast furnace is shown in Figure 3. Some of the advantages of this technology include:

- The injection system has been used with granular coal as well as with pulverized coal. No other system has been utilized over this range of coal sizes. Granular coal is 10-30% minus 200 mesh whereas pulverized coal is 70-80% minus 200 mesh.
- The costs for granular coal preparation systems are less than those for the same capacity pulverized coal systems.
- Granular coal is easier to handle in pneumatic conveying systems. Granular coals are not as likely to stick to conveying pipes if moisture control is not adequately maintained.
- Coke replacement ratios obtained by British Steel have not been bettered in any worldwide installation.
- System availability has exceeded 99 percent during several years of operation at British Steel.
- The unique variable speed, positive displacement CPC Macawber injectors provide superior flow control and measurement compared to other coal injection systems.

The joint development by British Steel and CPC Macawber of a process for the injection of granular coal into blast furnaces began in 1982 on the Queen Mary blast furnace at the Scunthorpe Works.(1,2) The objective of the development work was to inject granular coal into the furnace and test the performance of the CPC Macawber equipment with a wide range of coal sizes and specifications. Based on Queen Mary's performance, coal injection systems were installed on Scunthorpe's Queen Victoria, Queen Anne and Queen Bess blast furnaces and on Blast Furnaces 1 and 2 of the Ravenscraig Works. Queen Victoria's system was brought on line in November, 1984 and Queen Anne's in January, 1985. The Ravenscraig systems were started up in 1988. The success of the GCI systems at Scunthorpe and Ravenscraig led Bethlehem to conclude that the system could be applied successfully to large blast furnaces using domestic coals.

#### IV. INSTALLATION DESCRIPTION

A simplified flow diagram of the coal handling system at Burns Harbor is shown in Figure 4. The Raw Coal Handling Equipment and the Coal Preparation Facility includes the equipment utilized for the transportation and preparation of the coal from an existing railroad car dumper until it is prepared and stored prior to passage into the Coal Injection Facility; the Coal Injection Facility delivers the prepared coal to the blast furnace tuyeres.

Raw Coal Handling. Coal for this project is transported by rail from coal mines to Burns Harbor similar to the way in which the plant now receives coal shipments for the coke ovens. The coal is unloaded using a railroad car dumper, which is part of the blast furnace material handling system. A modification to the material handling system was made to enable the coal to reach either the coke ovens or the coal pile for use at the Coal Preparation Facility.

**Raw Coal Reclaim.** The raw coal reclaim tunnel beneath the coal storage pile contains four reclaim hoppers in the top of the tunnel. The reclaim hoppers, which are directly beneath the coal pile, feed a conveyor in the tunnel. The reclaim conveyor transports the coal at a rate of 400 tons per hour above ground to the south of the storage pile. A magnetic separator is located at the tail end of the conveyor to remove tramp ferrous metals. The conveyor discharges the coal onto a vibrating screen to separate coal over 2 inches from the main stream of minus 2-inch coal. The oversized coal passes through a precrusher which discharges minus 2-inch coal. The coal from the precrusher joins the coal that passes through the screen and is conveyed from ground level by a plant feed conveyor to the top of the building that houses the Coal Preparation Facility.

**Coal Preparation.** The plant feed conveyor terminates at the top of the process building that houses the Coal Preparation Facility. Coal is transferred to a distribution conveyor, which enables the coal to be discharged into either of two steel raw coal storage silos. The raw coal silos are cylindrical with conical bottoms and are completely enclosed with a vent filter on top. Each silo holds 240 tons of coal, which is a four-hour capacity at maximum injection levels. Air cannons are located in the conical section to loosen the coal to assure that mass flow is maintained through the silo.

Coal from each raw coal silo flows into a feeder which controls the flow of coal to the preparation mill. In the preparation mill, the coal is ground to the desired particle size. Products of combustion from a natural gas fired burner are mixed with recycled air from the downstream side of the process and are swept through the mill grinding chamber. The air lifts the ground coal from the mill vertically through a classifier where oversized particles are circulated back to the mill for further grinding. The proper sized particles are carried away from the mill in a 52-inch pipe. During this transport phase, the coal is dried to 1-1.5% moisture. The drying gas is controlled to maintain oxygen levels below combustible levels. There are two grinding mill systems; each system produces 30 tons per hour of pulverized coal or 60 tons per hour of granular coal.

The prepared coal is then screened to remove any remaining oversize material. Below the screens, screw feeders transport the product coal into one of four 180-ton product storage silos and then into a weigh hopper in two-ton batches. The two-ton batches are dumped from the weigh hopper into the distribution bins which are part of the Coal Injection Facility.

**Coal Injection.** The Coal Injection Facility includes four distribution bins located under the weigh hoppers described above. Each distribution bin contains 14 conical-shaped pant legs. Each pant leg feeds an injector which allows small amounts of coal to pass continually to an injection line. Inside the injection line, the coal is mixed with high-pressure air and is carried through approximately 600 feet of 1-1/2-inch pipe to an injection lance mounted on each of the 28 blowpipes at each furnace. At the injection lance tip, the coal is mixed with the hot blast and carried into the furnace raceway. The 14 injectors at the bottom of the distribution bin feed alternate furnace tuyeres. Each furnace requires two parallel series of equipment, each containing one product coal silo, one weigh hopper, one distribution bin and 14 injector systems.

## V. PROJECT MANAGEMENT

The demonstration project is divided into three phases:

Phase I	Design
Phase II	Construction and Start-up
Phase III	Operation and Testing

Phase I was completed in December 1993 and construction was completed in January 1995. Coal was first injected in four tuyeres of D furnace on December 18, 1994. The start-up period continued to November 1995 at which time the operating and testing program started. The testing of coals (Phase III) is expected to continue to July 1998.

The estimated project cost summary is shown in Table I. The total cost is expected to be about \$191 million. Additional information on project management was presented at the previous CCT Conferences. (3,4)

### Facility Start-Up

The coal injection facilities were fully started in January 1995 and by early June the coal injection rate on both furnaces had stabilized at 140 lbs/ton.(5) There were facility start-up problems in January and February, but by mid-year the coal preparation and delivery systems were operating as designed. The injection rate on C furnace was increased through the summer months and was over 200 lbs/ton for September, October and November. The injection rate on D furnace was kept in the range of 145-150 lbs/ton during the second half of the year.

In December 1995, severe coal weather caused coal handling and preparation problems that were not experienced during start-up in early 1995. The most severe problem was due to moisture condensing on the inside walls of the prepared coal silos. The moisture caked the coal and eventually blocked the injectors below the silos. As a result, coal injection on C furnace was stopped in mid-December and the coal silos were emptied and cleaned. In order to prevent condensation in the future, the top and sides of the C furnace coal silos were insulated. The D furnace silos were insulated in January 1996. The insulation has prevented any reoccurrence of blocked injectors due to caked coal.

## VI. TEST PROGRAM

The objective of the overall test program is to determine the effect of coal grind and coal type on blast furnace performance. The start-up operation was conducted with a high volatile coal from eastern Kentucky with 36% volatile matter, 8% ash and 0.63% sulfur. The coal preparation system was operated to provide granular coal throughout the start-up period. The coal injection rates and coke rates for C and D furnaces during 1995 and 1996 are shown in Figures 5 and 6, respectively.



## **Initial Results with Granular Coal**

The first comparison of interest was the blast furnace results with coal injection versus natural gas injection. A typical monthly operating period with natural gas is shown in Table II along with the first full month (April 1995) of coal injection on D furnace. The coke rate during the initial period with coal injection at 150 lbs/ton was 55 lbs greater than with natural gas at 140 lbs/ton. This was not unexpected. It has been established in the past that 1.3 to 1.4 lbs of coke are replaced by one pound of natural gas. The initial expectation for injected coal was that 0.8-0.9 lbs of coke would be replaced by one pound of coal. Also notable in Table II is the 44 lbs/ton slag volume increase that accompanies the injected coal practice. This additional slag volume is a direct result of the coal ash. Slag sulfur also increased from 0.87% to 1.09% due to the sulfur in the coal. In order to maintain hot metal chemistry control, the slag chemistry has been altered slightly to provide more sulfur removal capacity. Another item of interest is the large decrease in the hydrogen content of the top gas when coal is injected.

The next process benchmark that was important to operating personnel was the amount of injected coal necessary to return the furnace coke rate to the levels previously experienced with natural gas. This is shown by the September 1995 operating data from C furnace in Table II. After gaining experience with coal injection and establishing a steady operation at the coal preparation facility, an injection rate of 210 lbs/ton resulted in a comparable coke rate to the natural gas experience. The September operation is notable with regard to several process parameters. The wind rate has been reduced along with an increase in the oxygen enrichment level. Increasing the oxygen content of the hot blast resulted in a higher flame temperature which, in turn, enhances coal combustion in the tuyere zone. The flame temperature increased by 270 F with coal injection versus the previous practice with natural gas. Slag volume and chemistry have changed very little except for the higher sulfur content that is directly proportional to the increased injected coal rate. A decrease in the furnace permeability during this period is also apparent.

Permeability is a parameter used to show the amount of hot blast that is blown at a given pressure drop through the furnace. In general, a higher permeability means the flow of reducing gases through the furnace is smoother. The increase in coal injection from 150 to 210 lbs/ton caused a significant reduction in the furnace permeability. Figure 7 shows the effect of coal injection on permeability in both furnaces through July 1996. The reduction of furnace permeability is a major concern for higher levels of coal injection.

Table III shows the coals used during 1995 at Burns Harbor. The most important difference between the eastern Kentucky high volatile coal and the low volatile coals is the total carbon content. The effect of higher coal carbon content is shown with the blast furnace results from November 1994 and April 1995 in Table IV. The coke rate is about 50 lbs/ton lower with the low volatile coals compared to the high volatile coal.

Another advantage of low volatile coal was a substantial reduction in electrical energy at the coal grinding facility due to the softness of the coal. The Hargrove Grindability Index of the low volatile coals is in the range of 90 to 101 compared to 46 for the high volatile coal.

Table IV also shows the recent operation of July 1996 using low volatile coal. The coal rate has

increased to about 270 lbs/ton, the furnace coke rate has been reduced to 660 lbs/ton and the permeability has stabilized at 1.19. The lower blast pressure seen for the July 1996 period is also an indication of better furnace permeability. This was accomplished with increased use of blast moisture to produce more hydrogen in the bosh gas. This is shown by the increase in hydrogen content of the top gas. The increased hydrogen content results in a lower density bosh gas and, therefore, reduced gas flow resistance through the furnace stack.

### **Coke/Coal Replacement Ratio**

The quantity of furnace coke that is replaced by an injected fuel is an important aspect of the overall value of the injectant on the blast furnace operation. A detailed analysis of the furnace coke/coal replacement ratio for the C and D furnaces at Burns Harbor has been completed.

The replacement ratio for a blast furnace injected fuel is defined as the amount of coke that is replaced by one pound of the injectant. However, there are many furnace operating factors, in addition to the injectant, that affect the coke rate. In order to calculate the coke replaced by coal only, all other blast furnace operating variables that result in coke rate changes must be adjusted to some base condition. After adjusting the coke rate for changes caused by variables other than the coal, the remaining coke difference is attributed to the injected coal.

This evaluation was conducted with monthly average operating data compared to an appropriate base period for each furnace. Twenty-five months of data on both furnaces through the second quarter of 1996 were used in this evaluation.

The adjusted coke rates and the injected coal are plotted in Figure 8 along with the best fit regression line. The slope of the best fit line shows that coke/coal replacement is 0.96. This is an excellent replacement ratio and is significantly better than the 0.8-0.9 replacements reported by other coal injection operations.

The major conclusion of the test work to date is that granular coal performs very well in large blast furnaces. All other blast furnace coal injection systems use pulverized coal and some believed that pulverized coal was a requirement for large furnaces. The injection rates at Burns Harbor are not yet at the 400 lbs/ton level achieved by some, but there is nothing in the Burns Harbor experience to date that precludes higher injection rates with granular coal. The Burns Harbor furnaces will probably be limited to injection rates lower than 400 lbs/ton because of the lack of burden distribution equipment like moveable armor or a bell-less top, but this is a furnace limitation and not a coal size limitation.

### **Future Testing**

The testing of different coals will continue through 1997. The first test will be with a processed sub-bituminous coal from the Encoal Corporation in Gillette, Wyoming. The Encoal operation has also been supported by the Clean Coal Technology program. About 13,000 tons of Process Derived Fuel (PDF) from Encoal will be used in the Burns Harbor furnaces for about one week.

A trial will be conducted to determine the effect of granular versus pulverized coal. The same low volatile coal that has been injected through most of 1996 with a granular size will be pulverized to 70-80% minus 200 mesh for a one month trial. This will be the first time that a direct comparison of granular versus pulverized coal will be conducted on the same blast furnace.

Additional testing to be conducted in 1997 includes a high ash content coal and a high volatile coal. The high ash content coal will be similar to the base low volatile coal in all respects except the ash. This trial will provide a unique opportunity to determine the effect of coal ash in the blast furnace process.

The test with a high volatile coal will be a direct comparison to the base low volatile coal at a high injection rate. This test along with the high ash test will provide a sound basis for economic evaluations of alternative coal sources for all U.S. blast furnace operations with coal injection.

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2. D. S. Gathergood and G. Cooper, "Blast Furnace Injection - Why Granular Coal"? Steel Technology International, 1988.
3. D. Kwasnoski and L. L. Walter, "Blast Furnace Granular Coal Injection", Second Annual Clean Coal Technology Conference, Atlanta, GA, September 1993.
4. D. Kwasnoski and L. L. Walter, "Blast Furnace Granular Coal Injection", Third Annual Coal Technology Conference, Chicago, IL, September 1994.
5. L. L. Walter, R. W. Bouman and D. G. Hill, "Blast Furnace Granular Coal Injection", Fourth Annual Coal Technology Conference, Denver, CO, September 1995.

**TABLE I. ESTIMATED GRANULAR COAL  
INJECTION PROJECT COST SUMMARY**

	<u>\$ Million</u>
Phase I Design	5.19
Phase II Construction and Start-Up	133.85
Phase III Operation	<u>51.61</u>
Total Cost	190.65
 <u>Cost Sharing</u>	
DOE	31.26 (16.4%)
Bethlehem Steel	<u>159.39</u> (83.6%)
	190.65

TABLE II

**BURNS HARBOR BLAST FURNACE  
RESULTS - NATURAL GAS COAL INJECTION**

	<u>D Furnace November 1994</u>	<u>D Furnace April 1995</u>	<u>C Furnace September 1995</u>
Fuel Rate, lbs/ton			
Natural Gas	140	-	-
Coal	-	150	210
Coke	743	798	745
Blast Conditions:			
Reported Wind, MSCFM	171	174	164
Oxygen Enrichment, %	4.0	2.4	5.2
Moisture, Grs/SCF	6.0	16.0	8.5
Blast Pressure, psig	38.0	38.6	38.9
Flame Temperature, F	3685	3793	4062
Top Temperature, F	240	252	213
Hot Metal Analysis, %			
Silicon	.52	.56	.62
Sulfur	.040	.041	.035
Slag Analysis, %			
SiO <sub>2</sub>	37.74	36.31	36.57
Al <sub>2</sub> O <sub>3</sub>	9.64	9.70	9.50
CaO	36.50	38.21	37.71
MgO	12.20	12.08	12.31
Sulfur	0.87	1.09	1.19
Slag Volume, lbs/ton	393	437	437
Furnace Permeability	1.52	1.50	1.30
Top Gas Analysis:			
H <sub>2</sub> , %	7.33	3.05	3.13
BTU/SCF	92.8	82.6	88.1

TABLE III

## COALS USED AT BURNS HARBOR IN 1995

Coal	Eastern Ky. High Volatile	Virginia Low Volatile	Virginia Low Volatile	W. Virginia Low Volatile	W. Virginia Low Volatile
Vol. Matter, %	36.0	18.0	19.6	16.5	18.4
Ash, %	7.50	5.30	5.16	5.75	5.50
Sulfur, %	0.63	0.80	0.75	0.58	0.77
Moisture*, %	3.0	1.5	1.5	1.5	1.4
Gross Heating Value, BTU/lb	13900	14900	15029	14550	14775
Hargrove Grindability Index	46	100	101	94	90
Ultimate Analysis, %					
C	78.0	87.0	87.0	86.0	85.3
O	7.00	1.40	1.52	2.20	3.07
H	5.4	4.4	4.2	4.2	4.0

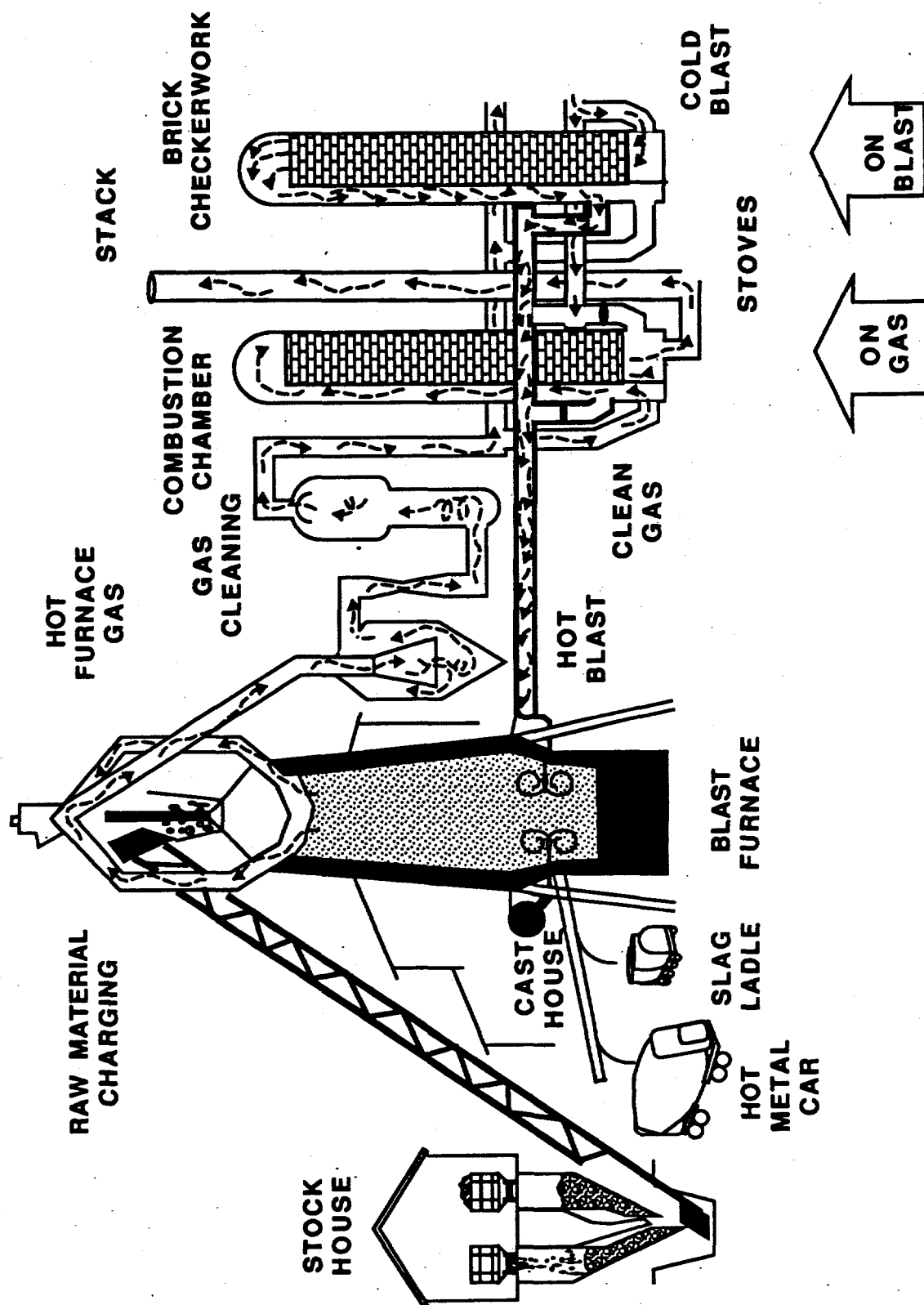
\* After drying and grinding

TABLE IV

**BURNS HARBOR C FURNACE RESULTS  
WITH COAL INJECTION**

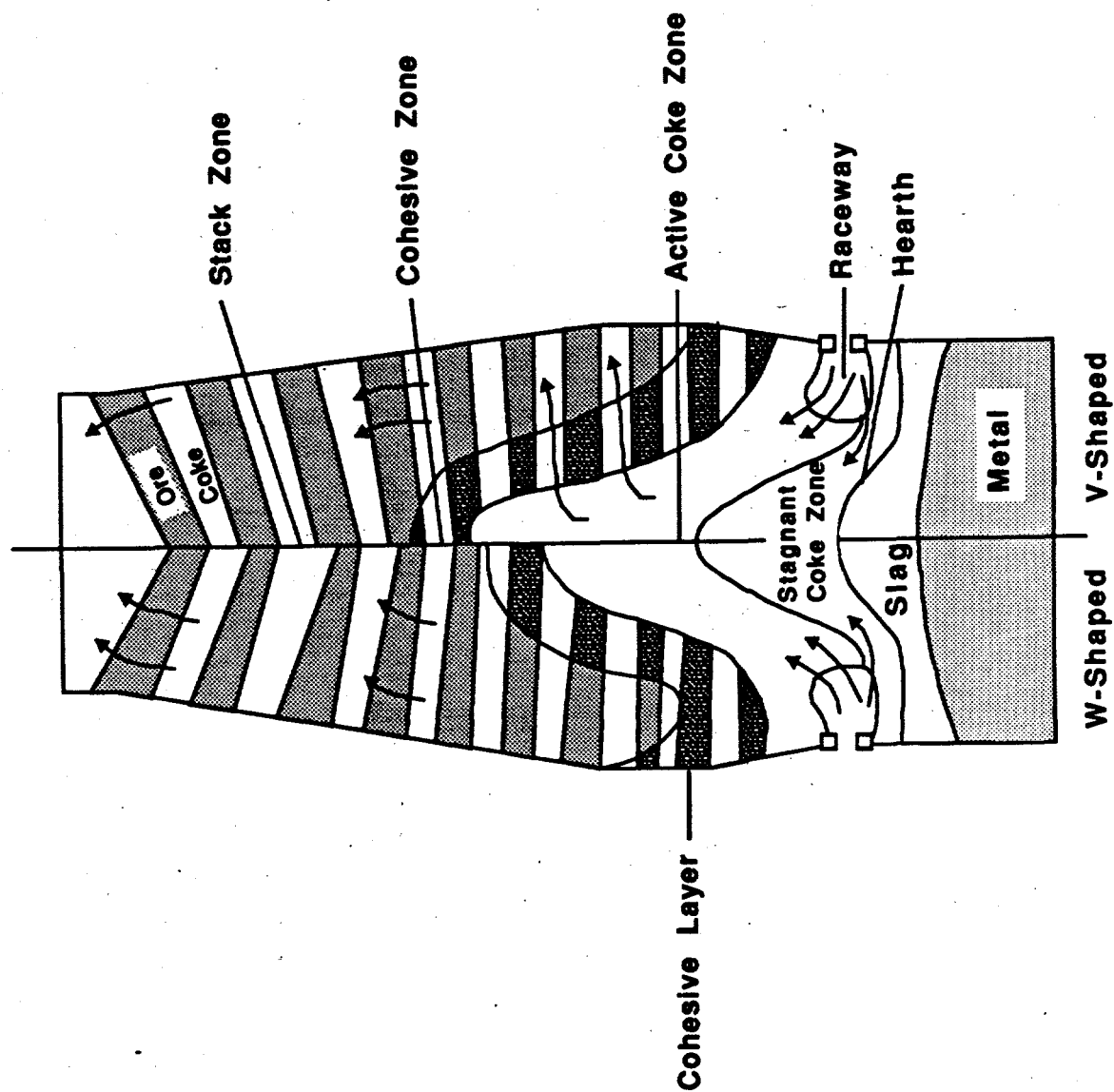
	<u>September 1995</u>	<u>November 1995</u>	<u>July 1996</u>
Coal Type	High Volatile	Low Volatile	Low Volatile
Fuel Rate, lbs/ton			
Coal	210	210	269
Coke	745	694	660
Blast Conditions:			
Reported Wind, SCFM	164	163	154
Oxygen Enrichment, %	5.2	4.6	5.6
Moisture, Grs/SCF	8.5	7.6	16.3
Blast Pressure, psig	38.9	39.4	38.6
Flame Temperature, F	4062	3996	3949
Top Temperature, F	213	210	244
Hot Metal Analysis, %			
Silicon	.62	.45	.49
Sulfur	.035	.041	.039
Slag Analysis, %			
SiO <sub>2</sub>	36.57	37.26	37.04
Al <sub>2</sub> O <sub>3</sub>	9.50	8.73	8.91
CaO	37.71	38.17	38.56
MgO	12.31	12.28	11.94
Sulfur	1.19	1.25	1.31
Slag Volume, lbs/ton	437	428	434
Furnace Permeability	1.30	1.26	1.19
Top Gas Analysis:			
H <sub>2</sub> %	3.13	3.15	4.31
BTU/SCF	88.1	84.1	89.7

**FIGURE 1**  
**THE BLAST FURNACE COMPLEX**

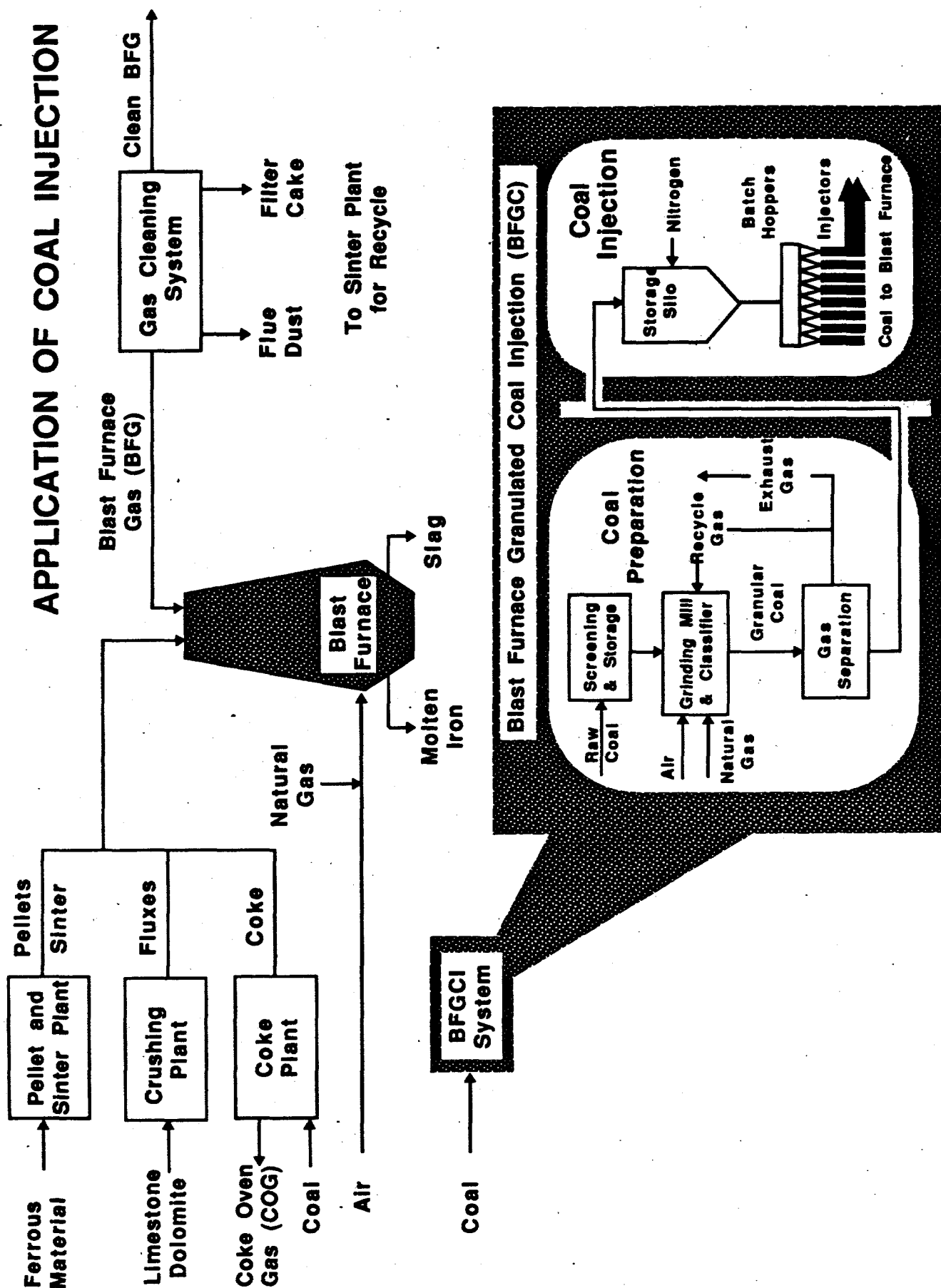




**FIGURE 2**  
**ZONES IN THE BLAST FURNACE**



**FIGURE 3**



**FIGURE 4. COAL PREPARATION AND INJECTION FACILITIES  
BURNS HARBOR PLANT**

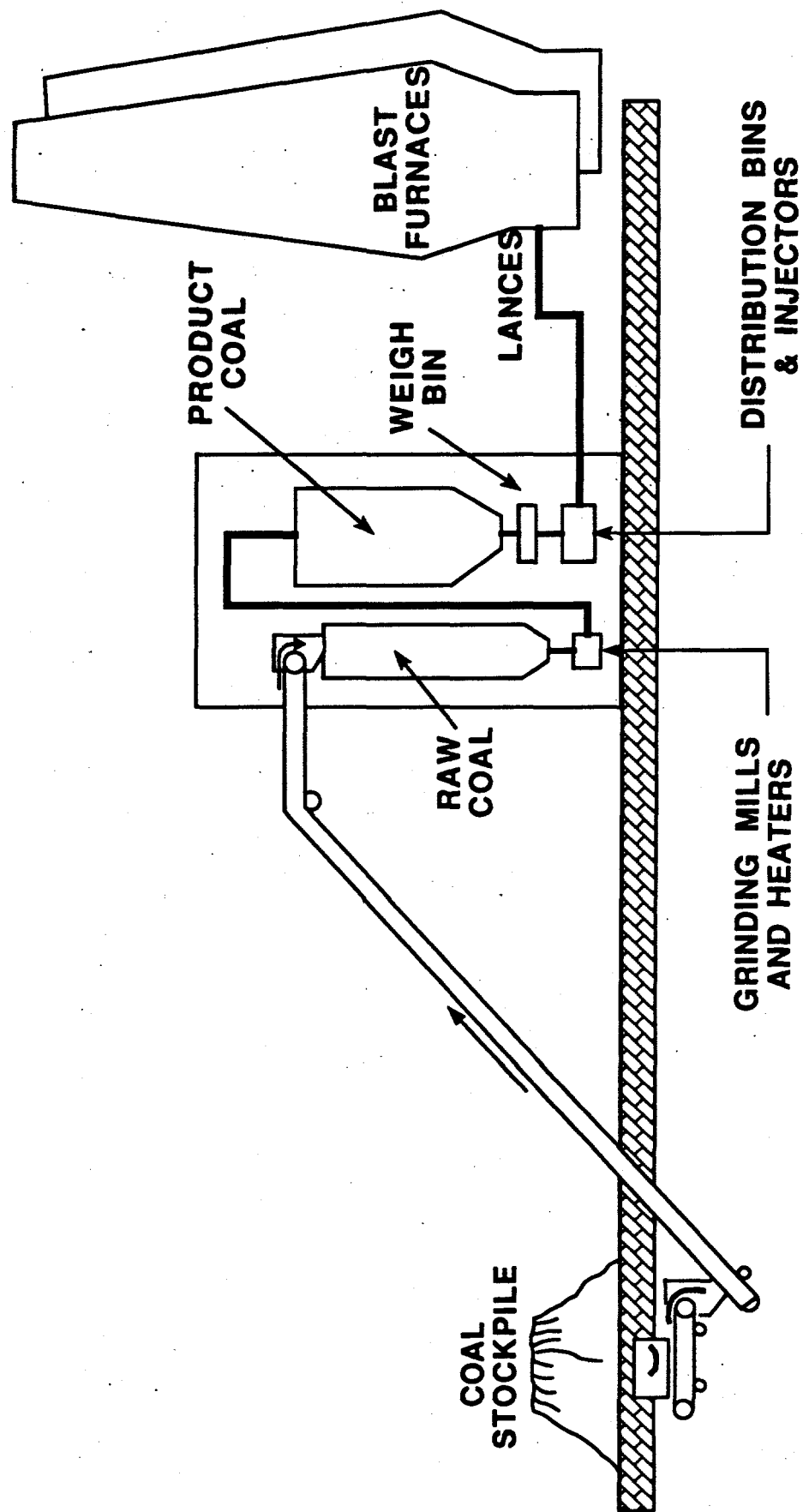


FIGURE 5

# BURNS HARBOR C FURNACE COAL and COKE RATES

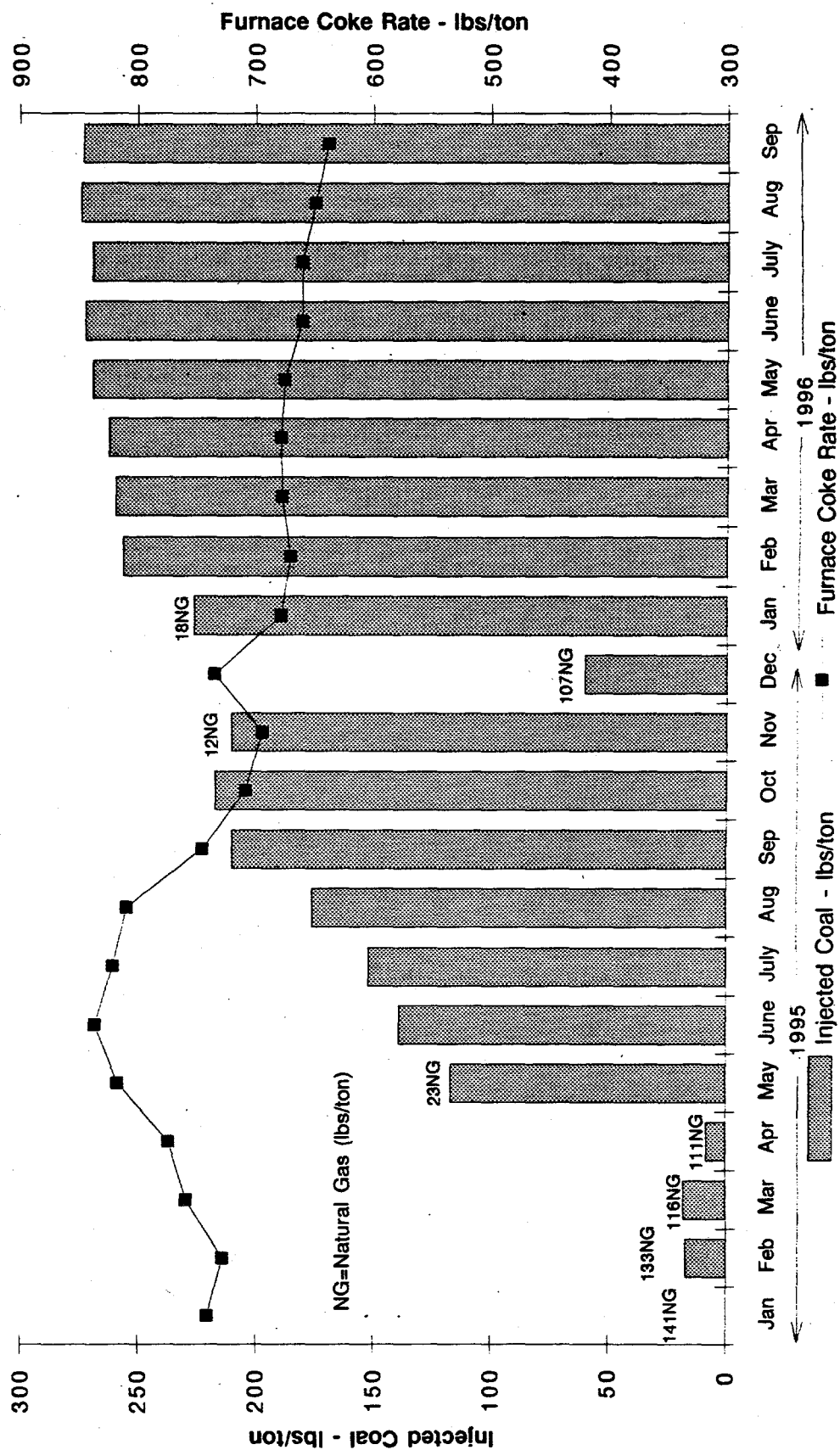


FIGURE 6

# BURNS HARBOR D FURNACE COAL and COKE RATES

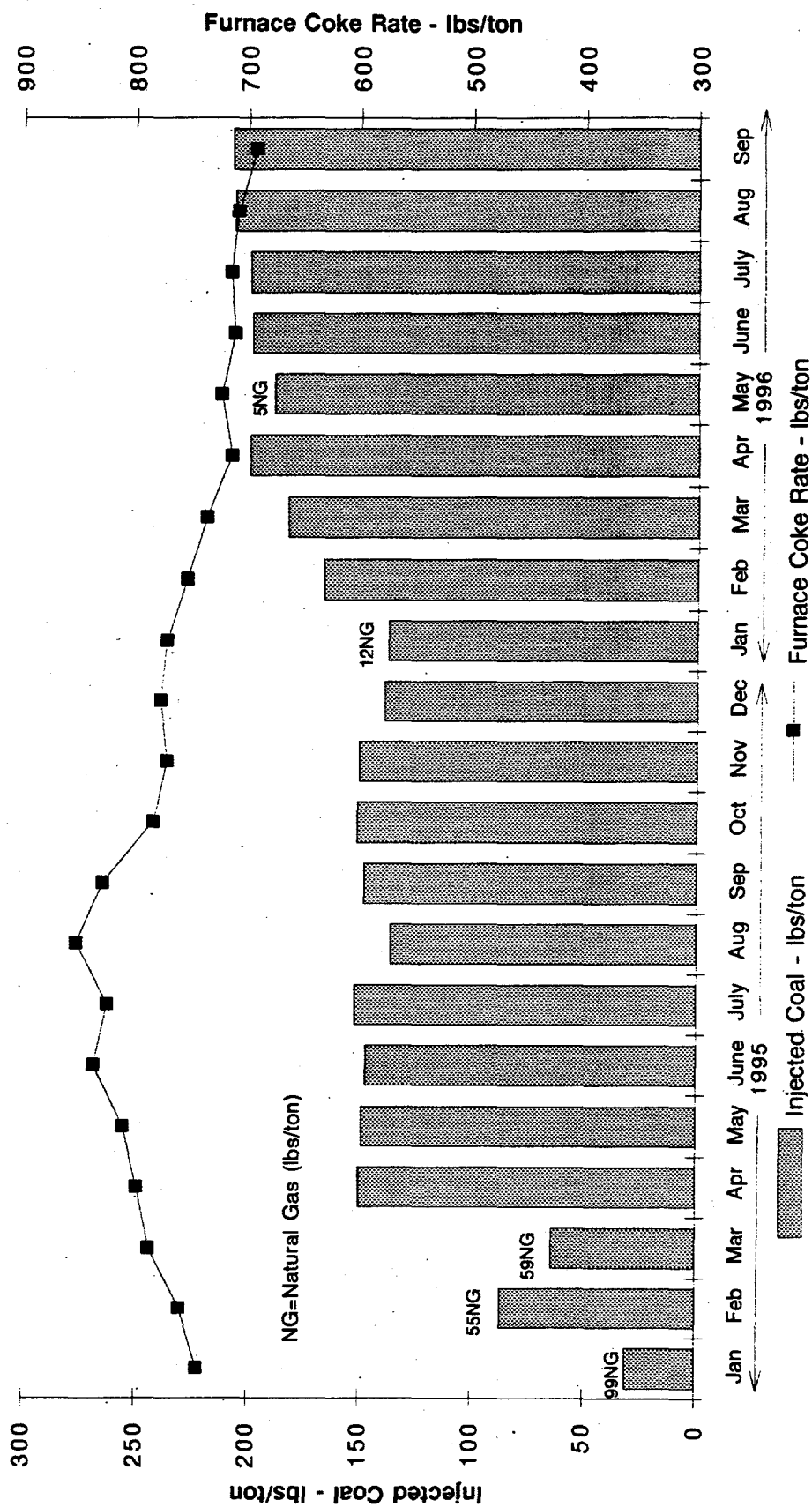


FIGURE 7

**BURNS HARBOR C & D FURNACES - INJECTED COAL RATE vs PERMEABILITY**

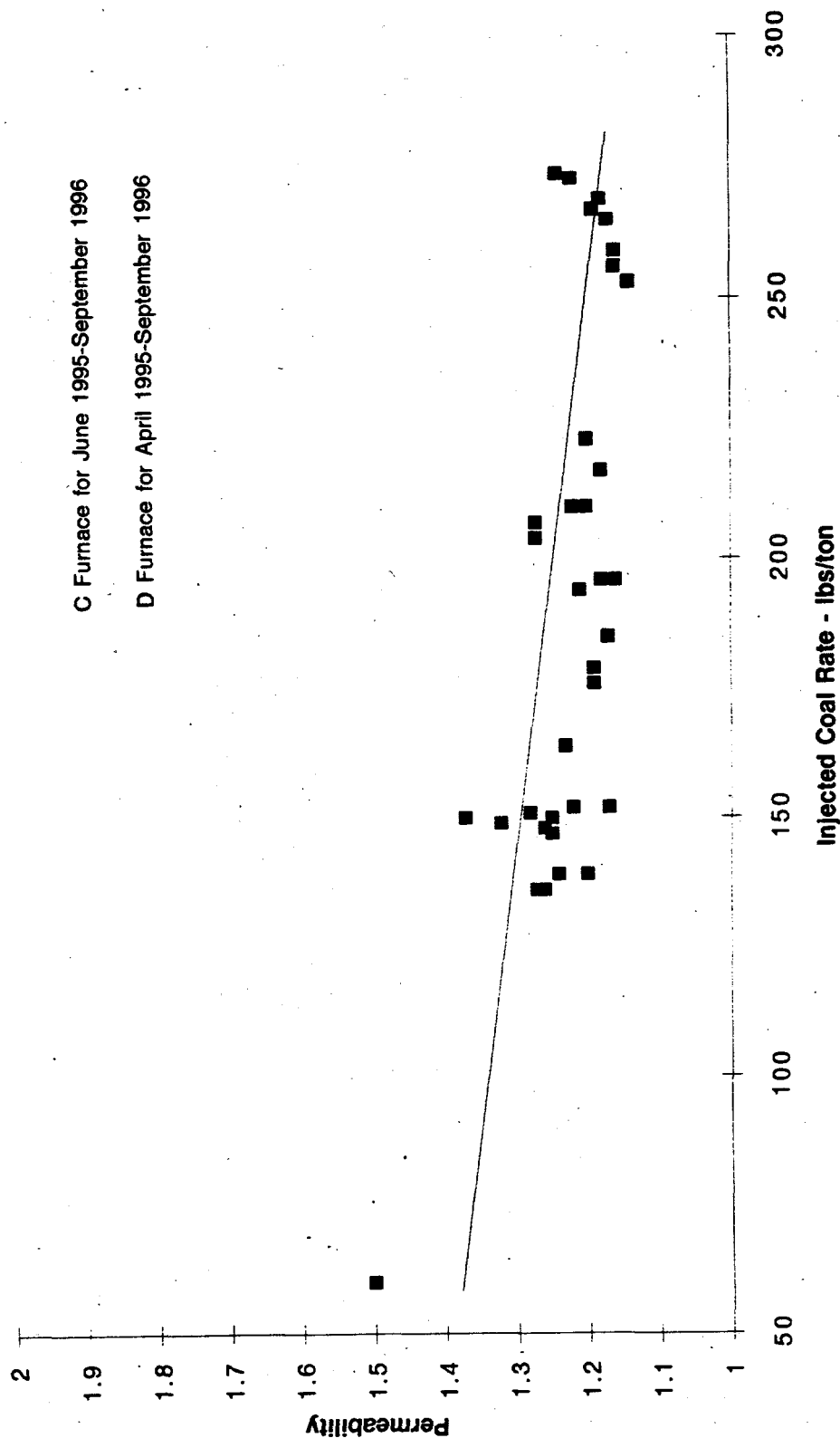
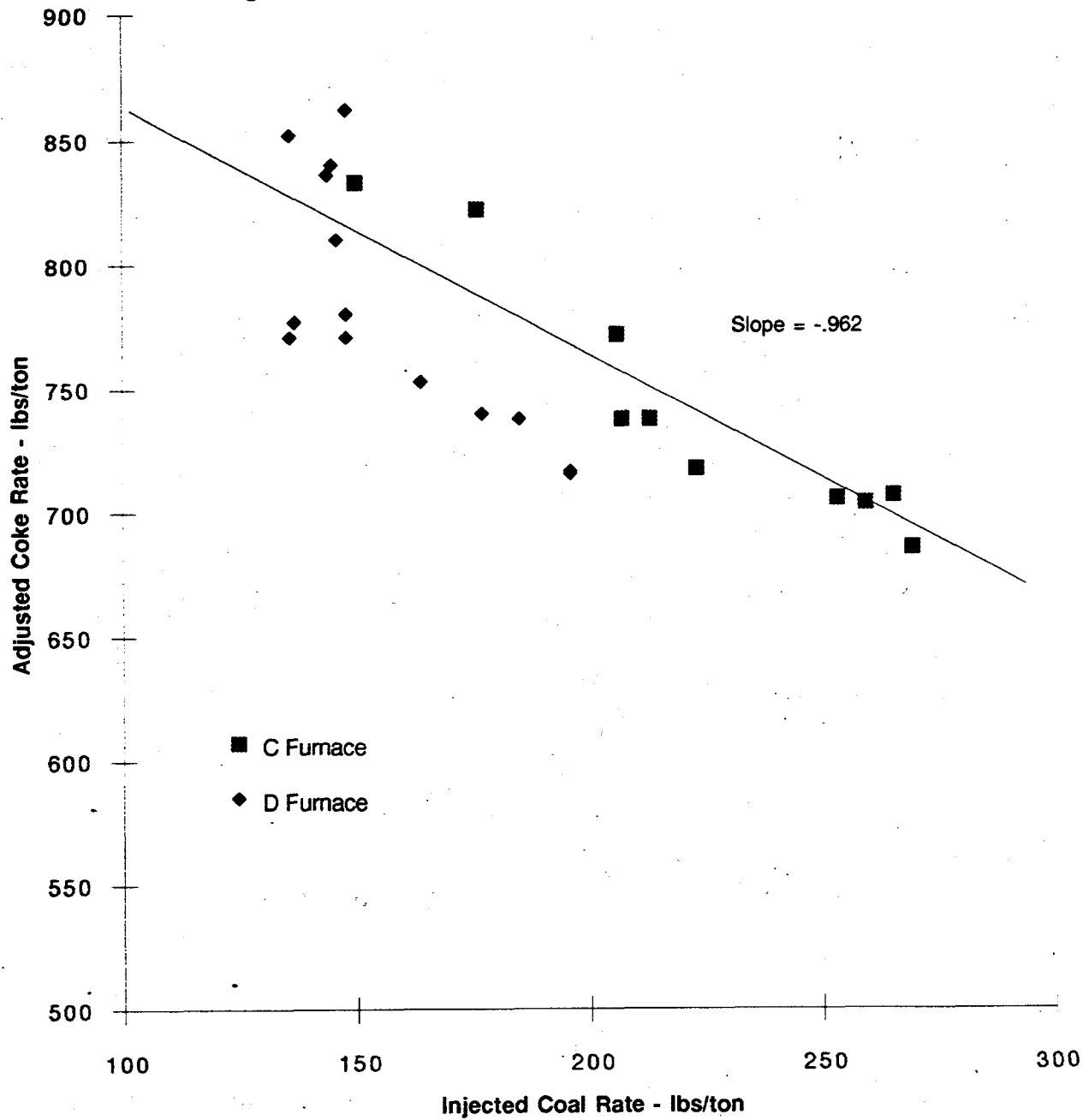
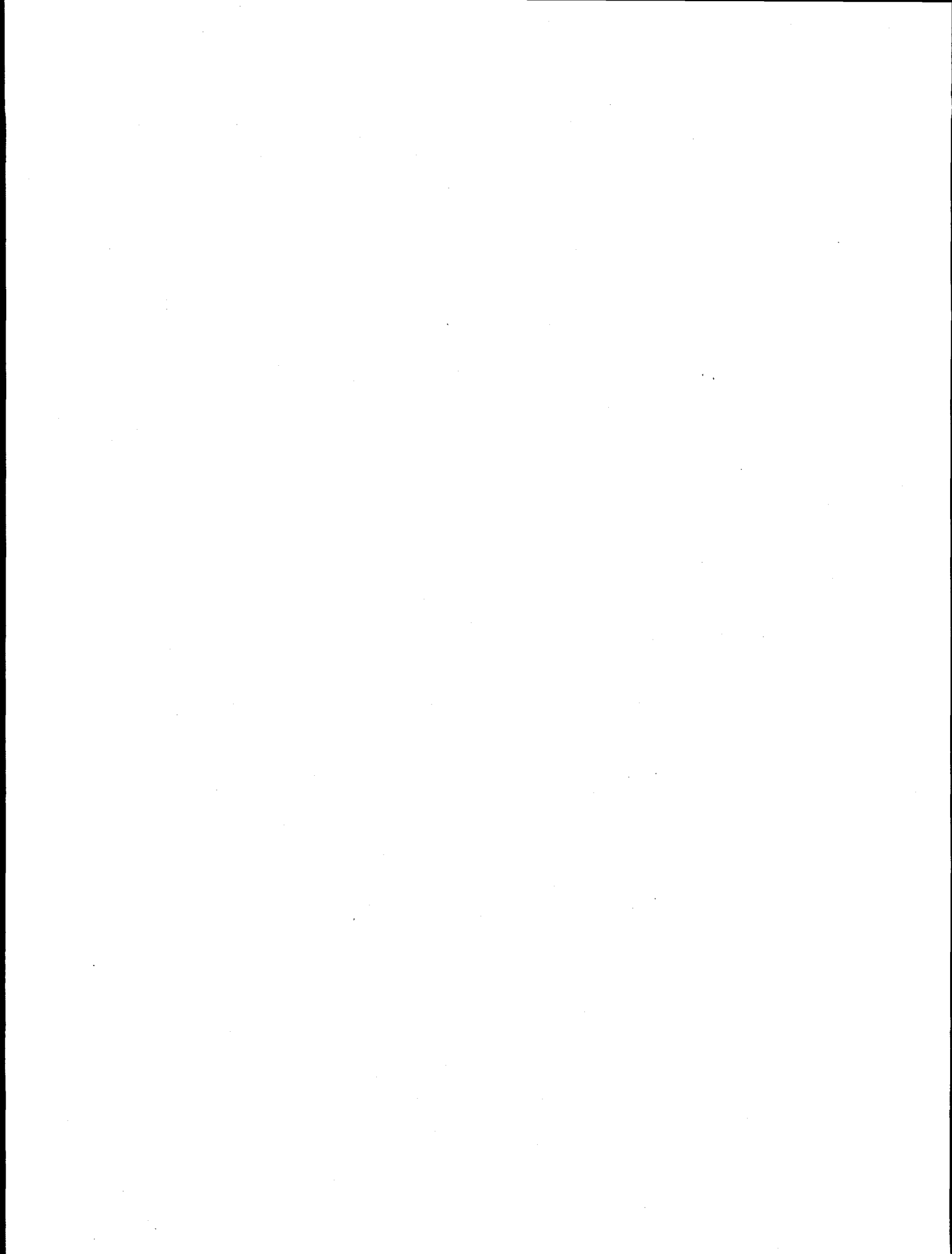


FIGURE 8

BURNS HARBOR C & D BLAST FURNACES

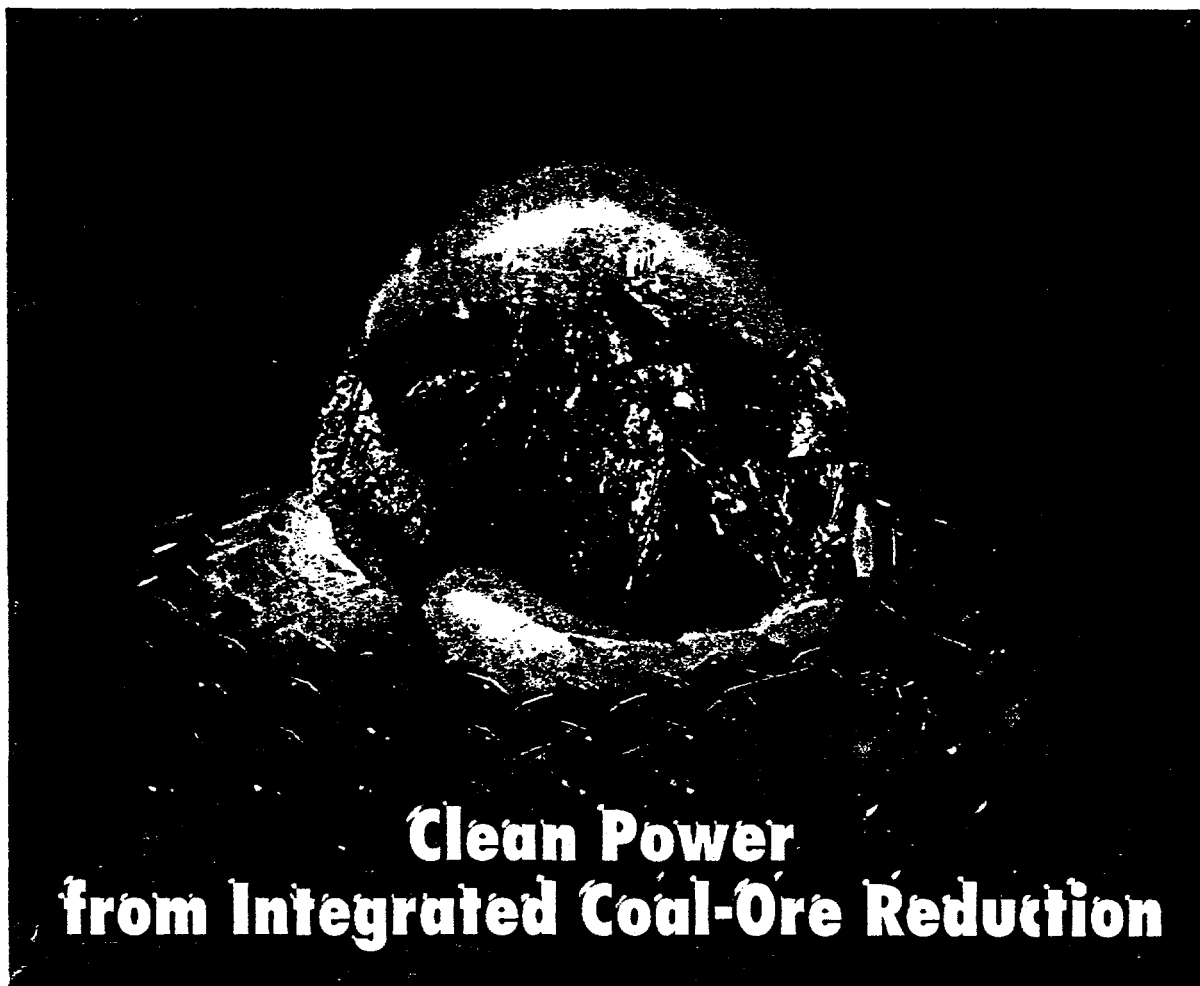
Regression Analysis - Injected Coal vs Adjusted Coke Rate







# **CPICOR™**



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## DISCLAIMER

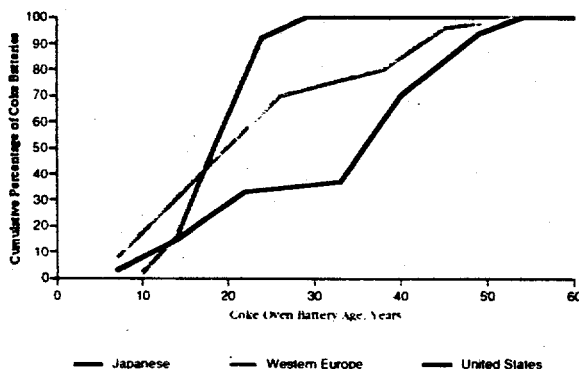
Geneva Steel, Centerior Energy Corp., and Air Products and Chemicals, Inc. currently hold no rights in or to the COREX® process or in or to the combined cycle power generation system and expressly disclaim any warranty or guarantee with respect to the information contained herein.

## BACKGROUND

A growing coke shortage is impacting the U.S. ability to produce iron and steel. Driven by environmental concerns of the sixties, the government imposed increasingly stringent requirements upon the U.S. coking industry to substantially lower the level of airborne pollutants. The U.S. steel industry, subjected to the economics of the '70s and '80s and unable to justify the building of new coke units or the environmental modifications required to save its antiquated coking batteries, purchased foreign coke (Figure 1). The impact of this policy in the mid '90s has been a rapid depletion of the world's surplus in coke production. This depletion will be further impacted as the Clean Air Act Amendments of 1990 take effect.

Age of U.S. Coke Plants

FIG. 1



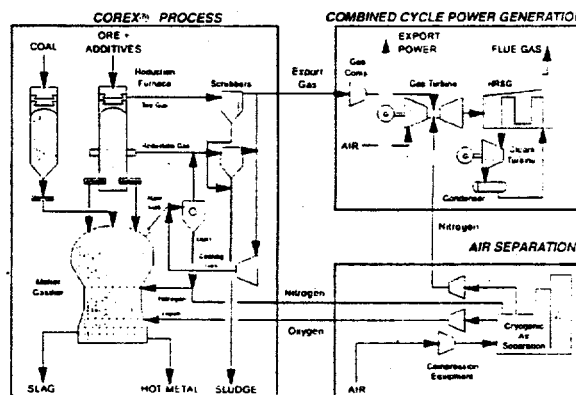
The U.S. steel industry, in order to maintain its basic iron production, is thus moving to lower coke requirements and to the cokeless or direct production of iron. The U.S. Department of Energy (DOE), in its Clean Coal Technology programs, has encouraged the move to new coal-based technology. The steel industry, in its search for alternative direct iron processes, has been limited to a single process, COREX®. The COREX® process, though offering commercial and environmental acceptance, produces a copious volume of offgas which must be effectively utilized to ensure an economical process. This volume, which normally exceeds the internal needs of a single steel company, offers a highly acceptable fuel for power generation. The utility companies seeking to offset future natural gas cost increases are interested in this clean fuel.

## INTRODUCTION

The COREX® smelting process, when integrated with a combined cycle power generation facility (CCPG) and a cryogenic air separation unit (ASU), is an outstanding example of a new generation of environmentally compatible and highly energy efficient "Clean Coal" technologies. This combination of highly integrated electric power and hot metal co-production, has been designated CPICOR™. "Clean Power from Integrated Coal/Ore Reduction." A consortium of leading companies who recognized the dilemmas of the U.S. steel and utilities industries have jointly proposed to the U.S. Department of Energy a collaborative effort to commercially demonstrate the CPICOR process using an advanced U.S. combined cycle power generation unit (Figure 2). The consortium further proposed to demonstrate optimum efficiency by combining the power generation and air separation units. The proposal was accepted for negotiation under Clean Coal V utilizing a 3,200 tons per day COREX® unit.

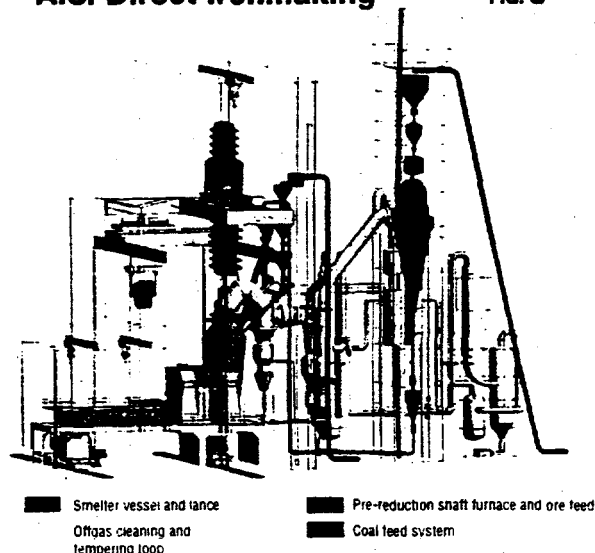
The consortium's selection of the COREX® process was based upon several factors. The U.S. urgently requires demonstration of direct iron production on a full commercial scale. The COREX®, as demonstrated by the operating unit at ISCOR and the unit under construction at Pohang, is the only process ready for upgrading to a production capacity suitable for the U.S. The Environmental Protection Agency requires an environmentally acceptable process. The COREX® process has fully demonstrated its compliance. The domestic steel industry is seeking economic operating incentives over the present coke plant/blast furnace route.

CPICOR Conceptual Flow Diagram FIG. 2



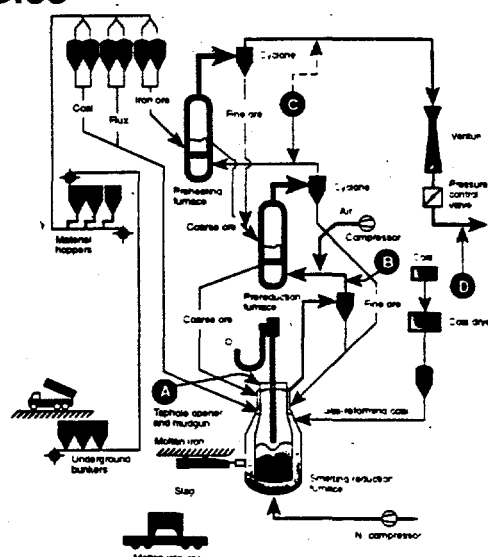
## AISI Direct Ironmaking

FIG. 5



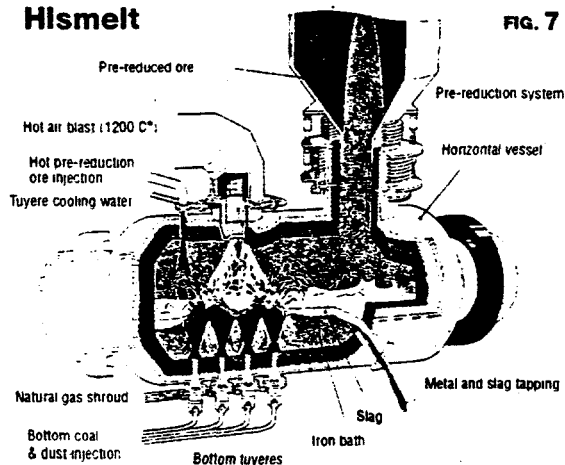
## DIOS

FIG. 6



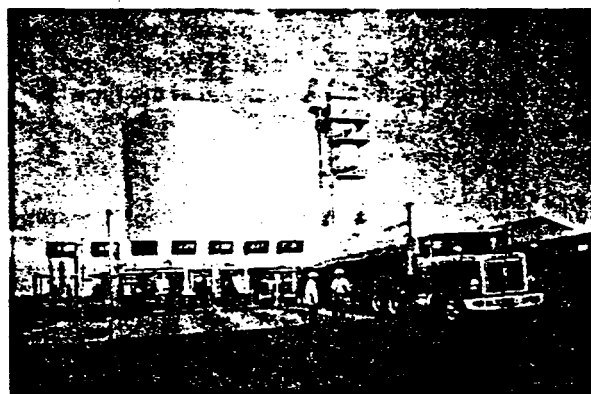
## Hismelt

FIG. 7



## ASU Facility

FIG. 8



industry, hundreds of commercial oxygen plants have been built, and presently more than 70,000 tons per day of oxygen capacity exists in the U.S. The ASU is proven, reliable, and highly efficient and will be integrated with the CCPG and COREX® within the CPICOR process. CPICOR will expand the U.S. coal base by including a wider range of coals for the simultaneous production of iron and power and will provide an integrated environmental solution for the economical revival of our steel, coal and power industries.

## PROJECT OBJECTIVES

The project objectives are to demonstrate a scale up of the COREX® and its commercial integration with the advanced combined cycle power generation system. To date, the COREX® process has demonstrated the ability to produce 330,000 tons of hot metal per year on lump ore, with the generated gas used for inplant heating purposes. To be commercially viable in the U.S., the value of the generated gas must be optimized, such as by partial integration with power generation, and the COREX®



reducing gas. The gas exits the melter/gasifier and passes through the dust separation cyclone before it is cooled to 850°C and transferred into the reduction shaft furnace. The reduction furnace is fed 5,170 TPD of iron ore and pellets and 622 TPD of raw fluxes. The charge is reduced or calcined by the ascending reducing gas. During the ascent, the sulfur contained in the gas reacts with the reduced iron and the calcined lime and dolomite. The reduced iron and the calcined fluxes are fed by water-cooled screws into the melter/gasifier. In the melter/gasifier, the reduced iron is melted by heat generated from the partial oxidation of the coal. The sulfur released during the smelting process is chemically captured in a calcium-rich, basic slag. The hot metal and slag are tapped periodically from the furnace hearth. The molten metal is sent directly to the steel mill for processing and the tapped slag (1,114 TPD) is recovered and used in the same manner as blast furnace slag.

The spent reducing gas (or top gas) leaves the reduction shaft essentially desulfurized and is quenched and cleaned through a series of wet scrubbers equipped with cyclonic separators. The cleaned export gas (1,770 MMBTU/hr) is delivered to the CCPG facility where it is compressed, mixed with air and nitrogen, and burned in a gas turbine/generator system. Process steam is generated in a heat recovery steam generator (HRSG) by extraction of heat from hot turbine exhaust gases and the combustion of surplus export gas. The steam produced in the HRSG drives an electric generator. This results in a total of 250 MW of generated power. Alternatively, a portion of the COREX® gas can be combusted within Geneva's plant for such processes as soaking pits, reheating furnaces, etc., with the major portion being used for combined cycle power generation.

In addition to demonstrating the use of COREX® gas in a CCPG unit, another key innovative feature of the CPICOR design is the potential integration of the gas turbine with the ASU. The ASU is designed to produce nitrogen and 3,000 TPD of high purity oxygen for the COREX® process. A portion of the nitrogen produced by the ASU may be delivered to the gas turbine, mixed with the compressed hot gas stream, and used to boost power output.

## **INHERENT ADVANTAGES OF CPICOR**

CPICOR technology, by virtue of its integral co-production of hot metal and power, offers a number of distinct technical and economic advantages over the competing commercial technology. The conventional method of producing hot metal from ore and coal involves two separate processes:

- 1) **Cokemaking** — Coal is heated to drive off volatile matter and produce "coke" to be used as both fuel and reducing agent in a smelting operation.
- 2) **Blast furnace smelting** — Ore, coke, limestone, and hot air are charged to reduce the ore and produce molten iron.

Approximately 30% of the coke oven gas produced during cokemaking is used to provide heat for the cokemaking operation. The excess gas is typically sent to a utility steam boiler where it is mixed with the surplus off-gas from the blast furnace to generate power. At comparable hot metal production rates, this technology generates only about one-fifth the power produced by CPICOR technology.

### ***Highly Efficient Use of Coal***

The energy efficiency of the CPICOR technology is over 30% greater than the competing commercial technology when considering only the effective production of hot metal and electric power. The higher efficiency of the CPICOR technology is due to the more effective use of the sensible heat and volatile matter than the coke-making/blasting furnace process, i.e. 55 to 40%. In addition, the CCPG achieves energy efficiencies of nearly 50% compared to a maximum of 34% with conventional coal-based power systems equipped with flue gas desulfurization.

### ***Dramatic Reduction in Emissions***

CPICOR technology is less complex and environmentally superior to conventional processes. All criteria air pollutants, particularly the acid rain precursors, SO<sub>x</sub> and NO<sub>x</sub>, are reduced by more than 85%. This reduction is due largely to the desulfurizing capability of the COREX® process, efficient control systems within the CCPG facility, and the use of oxygen in place of air in the COREX® process. The gaseous emissions from the CPICOR plant, resulting from the combustion of air and export

processes without sacrificing the flexibility for commercial operation and the reliability of power or hot metal production.

## FEASIBILITY OF CCPG INTEGRATION

Although this is the first CCPG application to be fueled with COREX® export gas, the proposed design is based on proven technology. Similarly sized and larger CCPG facilities have been designed and are currently in reliable operation today with 94% to 97% availability. The steam pressure levels selected for the CPICOR design are typical of those which have been used in power generation facilities for years. The proposed gas turbine system is a proven, reliable design with a considerable number of the candidate models currently in operation. There are many heat recovery steam generator (HRSG) units of similar design and size in operating CCPG installations. Many steam turbine/electric generator sets of the type and capacity proposed for CPICOR currently exist in electric power generation facilities and have been in operation for years. All other major equipment items for the CCPG facility are likewise based on existing technology and similarly sized units (Figure 13).

The fueling of a CCPG system gas turbine with low-BTU gas produced by the COREX® process is unique. However, fueling gas turbines with medium

and low-BTU fuel is a technology which exists commercially and is being studied, developed, and optimized by the gas turbine manufacturers. Consuming COREX® export gas in a turbine presents some technical challenges not encountered with fired boiler combustion cycles. Particulates greater than 5 microns and alkali metals can lead to turbine blade erosion. In combination with H<sub>2</sub>S and SO<sub>2</sub>, these materials can lead to hot metal corrosion of the combustor and inlet transition duct as well as blading of the turbine section. These potential problems are addressed by adequate scrubbing and filtration of the export gas in the CPICOR design. The use of proven and reliable wet scrubber technology will provide over 95% dust removal. Performance data from the ISCOR operation shows the COREX® export gas has contaminant levels within the gas turbine manufacturers' maximum specifications.

Considerable advancements have also been made in gas turbine hot section metal coatings. Cooling technologies have been developed to reduce the erosion and corrosion effects of firing offgases from processes such as COREX®. Westinghouse, Mitsubishi Heavy Industries (MHI), Siemens, ABB, General Electric, and European Gas Turbines (Ruston) all report capabilities to accept the COREX® export gas with only minor modifications to the gas turbine designs.

## Combined Cycle Statistics

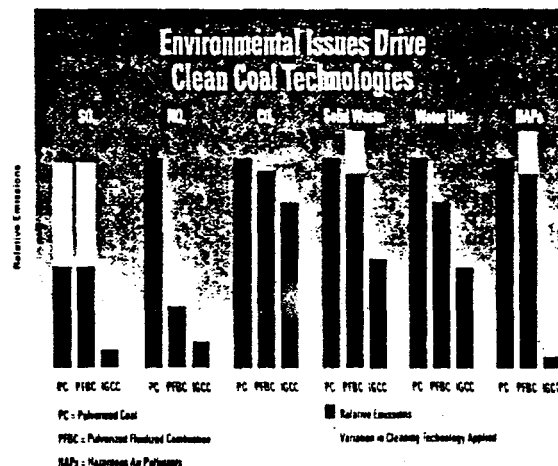
### Installed Combined Cycle Units

Installed Capacity (U.S.)	Over 66,000 megawatts
Operation Hours (U.S.)	Over 77 million hours
Power Range	Up to 350 MW per unit
Thermal Efficiencies	Up to 54+%
Availability	90 to 97%
Heat Rates	9000 to 6200 BTU/KWH

### Coal Gasification Units

Plaquemines	two 104 mw units installed 1974
Cool Water	one 120 mw unit installed 1984
Emissions	1/10 of coal fired units

FIG. 13



## COMMERCIAL OUTLOOK

CPICOR is intended to replace commercial coke oven/blast furnace technology in the production of hot metal for use in steelmaking. The best candidates for utilizing CPICOR technology are existing integrated steel plants with blast furnaces and coke ovens nearing the end of their useful lives and located where the local electric utility requires additional capacity. While commercialization of the COREX® process is driven primarily by the need for an environmentally sound source of hot metal for the steel industry, the production of electric power from the COREX® export gas is key to the economic competitiveness of the technology. Thus, commercialization will be facilitated by the ability of this project to obtain an attractive price for the power created by the plant.

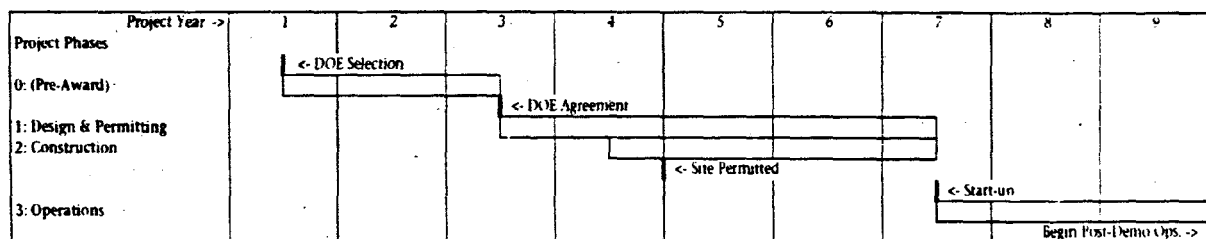
Conventional coke oven/blast furnace technology is too expensive to be utilized as replacement units or to expand domestic ironmaking capacity. Recent studies <sup>2, 3, 4</sup> conclude that no new coke batteries will be built in the United States. Of the existing 79 coke oven batteries, 40 are thirty years of age or older and are due for either replacement or major rebuilds.

As a consequence of the Clean Air Act Amendments of 1990, the emissions from existing coke ovens must be reduced substantially over the next several years. It has been estimated that the total capital investment for rebuilding or replacing current capacity could be in the range of \$4 to \$6 billion. The capital cost of coke ovens is about \$166 per ton of equivalent hot metal capacity. Coupled to the cost of a blast furnace rebuild at \$155 per ton equivalent hot metal capacity, the investment in a new COREX® facility at approximately \$255 per ton compares favorably on a capital basis.

If the iron and steel industry is to continue to produce liquid iron in the form of hot metal, a new technology must be developed and installed. Future competition to COREX® is likely to come from the new direct ironmaking processes being developed in both Japan (the DIOS process, Figure 6) and in the U.S. (the AISI process, Figure 5). Both of these processes produce iron and a lower calorific value export gas directly from iron ore and coal. However, the development of the COREX® technology is 8 to 12 years ahead of these other processes. Consequently, COREX®/CPICOR should become the technology of choice as domestic ironmaking capacity reaches the end of its useful life.

## Project Time Line

FIG. 14





# **Role Of The Liquids From Coal Process In The World Energy Picture**

**James P. Frederick  
Brent A. Knottnerus  
ENCOAL Corporation  
Gillette, Wyoming**

## **ABSTRACT**

ENCOAL Corporation, a wholly owned indirect subsidiary of Zeigler Coal Holding Company, has essentially completed the demonstration phase of a 1,000 Tons per day (TPD) Liquids From Coal (LFC™) plant near Gillette, Wyoming. The plant has been in operation for 4½ years and has delivered 15 unit trains of Process Derived Fuel (PDF™), the low-sulfur, high-Btu solid product to five major utilities. Recent test burns have indicated that PDF™ can offer the following benefits to utility customers:

- Lower sulfur emissions
- Lower NO<sub>x</sub> emissions
- Lower utilized fuel costs to power plants
- Long term stable fuel supply

More than three million gallons of Coal Derived Liquid (CDL™) have also been delivered to seven industrial fuel users and one steel mill blast furnace. Additionally, laboratory characteristics of CDL™ and process development efforts have indicated that CDL™ can be readily upgraded into higher value chemical feedstocks and transportation fuels.

Commercialization of the LFC™ is also progressing. Permit work for a large scale commercial ENCOAL® plant in Wyoming is now underway and domestic and international commercialization activity is in progress by TEK-KOL, a general partnership between SGI International and a Zeigler subsidiary.

The Project<sup>[1]</sup>, which was cost shared by the U.S. Department of Energy under Round Three of the Clean Coal Technology program, achieved its remaining long-term objectives in the past year. These included delivery and testing of pure PDF™ in a major Eastern U.S. bituminous coal boiler, operation of the plant for long periods at greater than 90% availability and processing of an alternate source coal. Plans are to continue operation of the ENCOAL® plant for several purposes:

- testing the viability of alternate commercial scale equipment
- delivery of additional test burn quantities of products
- training operators for the commercial plant
- providing additional design data for the commercial plant

A no-cost extension to the Cooperative Agreement has been approved for six months to complete the required project close-out reports. This paper covers the historical background of the Project, describes the LFC™ process and describes the worldwide outlook for commercialization.

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<sup>1</sup> Contract No. DE-FC21-90MC27339, ENCOAL Corporation, P. O. Box 3038, Gillette, WY 82717; Telefax (307) 682-7938

## **Acknowledgements**

ENCOAL Corporation wishes to acknowledge the participation of D.O.E.'s project manager, Mr. Douglas M. Jewell, whose guidance and technical advice contributed to the success of the ENCOAL® project during the design, construction and operation activities over the past six years.

## **BACKGROUND INFORMATION**

### **Objectives**

Beneficiation of low sulfur Powder River Basin (PRB) subbituminous coal is being demonstrated by the ENCOAL® Mild Coal Gasification Project using the LFC™ process. The LFC™ Technology employs a mild gasification process, that is mild pyrolysis at relatively low temperatures, to produce both liquid and solid fuels with environmentally superior properties. The demonstration plant has been in the testing and operations mode for more than 4½ years and has completed all of its original long-term goals.

ENCOAL's overall objective for the Project is to further the development of full sized commercial plants using the LFC™ Technology. In support of this overall objective, the following goals were established:

- Provide sufficient products for full-scale test burns
- Develop data for the design of future commercial plants
- Demonstrate plant and process performance

- Provide capital and operating cost data
- Support future LFC™ Technology licensing efforts

Significant progress has been made on the first four goals, and the commercialization and technology licensing efforts are in progress. This paper highlights several areas of immediate interest to potential customers and licensees. These include the status of the ENCOAL® Project, plant operating experience, plant reliability, product properties, technology development and remaining challenges. Most importantly, the status of the commercialization of the LFC™ Technology is reviewed.

### **General Description**

ENCOAL® Corporation is a wholly owned subsidiary of Bluegrass Coal Development Company, (formerly named SMC Mining Company), which in turn is a subsidiary of Zeigler Coal Holding Company. ENCOAL® has entered into a Cooperative Agreement with the United States Department of Energy (DOE) as a participant in Round III of the Clean Coal Technology Program. Under this agreement, the DOE has shared 50% of the cost of the ENCOAL® Mild Coal Gasification Project.

The Cooperative Agreement was extended in October 1994 for an additional \$18,100,000 bringing the Project total to \$90,600,000 through September 17, 1996. A no-cost extension in September 1996 moved the Cooperative Agreement end date to March 17, 1997 to allow for completion of final reporting requirements. A license for the use of LFC™ Technology has been issued to ENCOAL® from the technology owner, TEK-KOL, a general partnership between SGI International of La Jolla, California and a subsidiary of Zeigler Coal Holding Company.

The ENCOAL® Project encompasses the design, construction and operation of a 1,000 TPD commercial demonstration plant and all required support facilities. The Project is located near Gillette, Wyoming at Triton Coal Company's Buckskin Mine. Existing roads, railroad, storage silos and coal handling facilities at the mine significantly reduced the need for new facilities for the Project.

A substantial amount of pilot plant testing of the LFC™ process and laboratory testing of PDF™ and CDL™ was done.<sup>[1]</sup> The pilot plant tests showed that the process was viable, predictable and controllable and could produce PDF™ and CDL™ to desired specifications. Key dates and activities in bringing the project from the pilot plant stage to its current status are:

- Through early 1987: Development of the LFC™ process by SGI.
- Mid 1987: SMC Mining Company (SMC) joined with SGI on further development.
- Mid 1988: Feasibility studies, preliminary design, economics and some detailed design work by SMC.
- June 1988: Submittal of an application to the State of Wyoming for a permit to

- construct the plant - Approved July 1989.
- August 1989: ENCOAL® Project submitted to the DOE as part of Round III of the Clean Coal Technology Program - Selected in December 1989.
- September 1990: Cooperative Agreement signed. Contract awarded to The M. W. Kellogg Company for engineering, procurement and construction.
- October 1990: Ground breaking at the Buckskin Mine site.
- April 1992: Mechanical completion - commissioning begun.
- June 1992: First 24 hour run in which PDF™ and CDL™ were produced.
- November 1992: SMC Mining Company and its subsidiaries, including ENCOAL®, acquired by Zeigler.
- April 1993: ENCOAL® achieves two week continuous run.
- June 1993: Plant shut down for major modifications.
- December 1993: Plant recommissioned with added deactivation loop.
- July 1994: Completed 68 day continuous run - plant operational.
- September 1994: First unit train containing PDF™ shipped and burned successfully.
- October 1994: Two year extension and additional funding approved by DOE.
- April 1996: Shipped first unit train containing 100% PDF™.
- May 1996: Successfully burned PDF™ in a fully instrumented major U.S. utility boiler.

Although designed for 1000 TPD feed, the plant is currently processing 500 TPD of subbituminous PRB coal due to capacity limitations in the deactivation loop. The plant produces 250 TPD of PDF™, which has the high heat content of Eastern coals but with low sulfur content, and 250 barrels/day of CDL™, which is a low sulfur industrial fuel oil. While CDL™ is different from petroleum derived oils in its aromatic hydrocarbon, nitrogen and oxygen content, it has a low viscosity at operating temperatures and is comparable in flash point and heat content.

Not a pilot plant or a "throw-away", ENCOAL's processing plant is designed to commercial standards for a life of at least 10 years. It uses commercially available equipment as much as possible, state-of-the-art computer control systems, BACT for all environmental controls to minimize releases and a simplified flowsheet to make only two products matched to existing markets. The intent is to demonstrate the core process and not make the project overly complicated or expensive.

The ENCOAL® Project has demonstrated for the first time the integrated operation of several unique process steps:

- Coal drying on a rotary grate using convective heating
- Coal devolatilization on a rotary grate using convective heating
- Hot particulate removal with cyclones
- Integral solids cooling and deactivation
- Combustors operating on low Btu gas from internal streams

- Solids stabilization for storage and shipment
- Computer control and optimization of a mild coal gasification process
- Dust suppressant on PDF™ solids

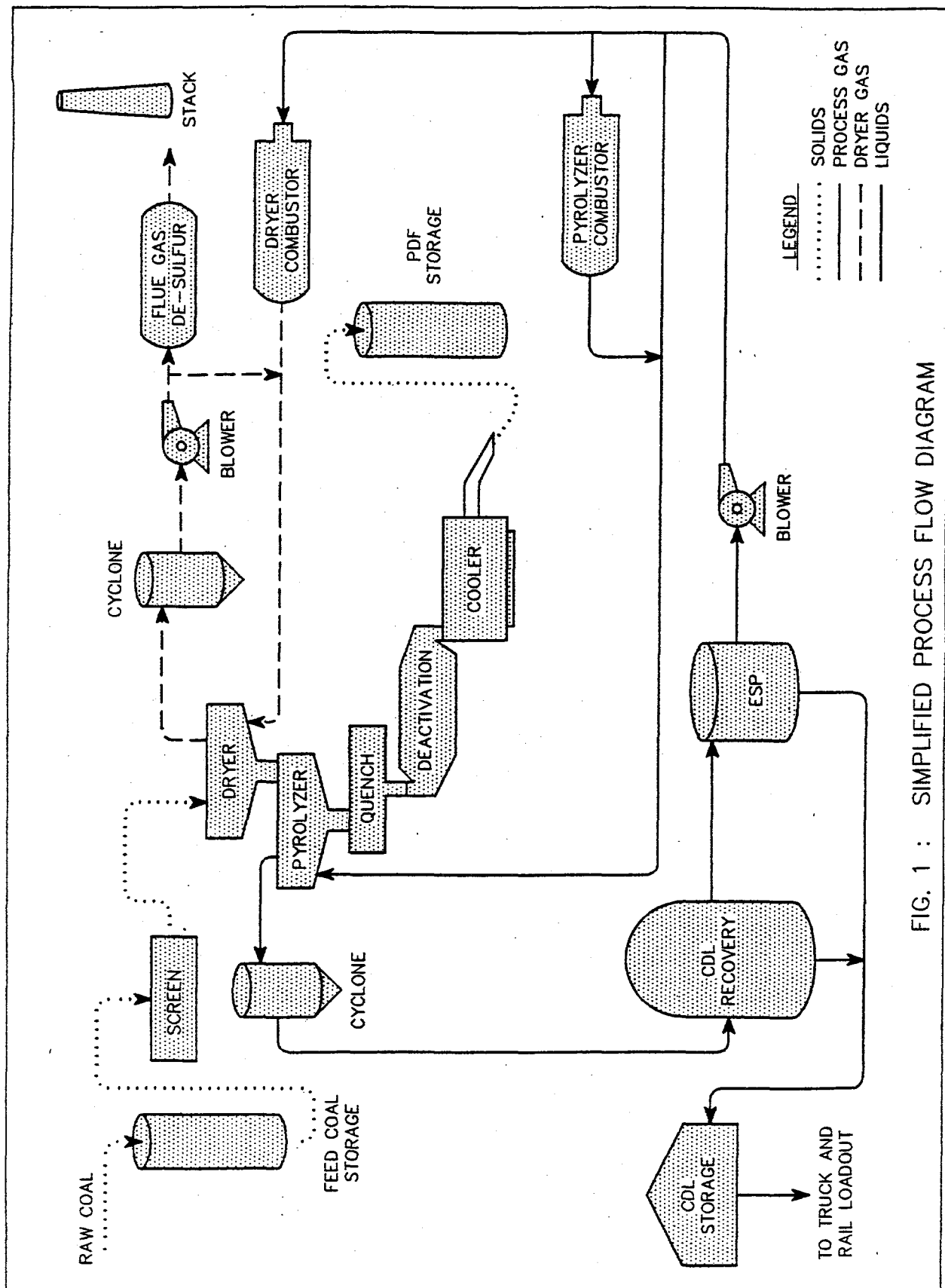
Utility test burns have shown that the fuel products can be used economically in commercial boilers and furnaces to reduce sulfur emissions significantly at utility and industrial facilities currently burning high sulfur bituminous coal or fuel oils. Ultimately, installation of commercial scale LFC™ plants should help reduce U.S. dependence on imports of foreign oil.

### **Process Description**

Figure 1 is a simplified flow diagram of ENCOAL's application of the LFC™ Technology. The process involves heating coal under carefully controlled conditions. Nominal 3" x 0" run-of-mine (ROM) coal is conveyed from the existing Buckskin Mine to a storage silo. The coal from this silo is screened to remove oversize and undersize materials. The 2" x 1/8" sized coal is fed into a rotary grate dryer where it is heated by a hot gas stream. The residence time and temperature of the inlet gas have been selected to reduce the moisture content of the coal without initiating chemical changes. The solid bulk temperature is controlled so that no significant amounts of methane, carbon monoxide or carbon dioxide are released from the coal.

The solids from the dryer are then fed to the pyrolyzer where the temperature is further raised to about 1,000°F on another rotary grate by a hot recycle gas stream. The rate of heating of the solids and their residence time are carefully controlled, because these parameters affect the properties of both solid and liquid products. During processing in the pyrolyzer, all remaining water is removed, and a chemical reaction occurs that results in the release of volatile gaseous material. Solids exiting the pyrolyzer are quickly quenched to stop the pyrolysis reaction, then transferred to a small surge bin that feeds the vibrating fluidized bed (VFB) deactivation unit.

In the VFB unit, the partially cooled, pyrolyzed solids contact a gas stream containing a controlled amount of oxygen. Termed "oxidative deactivation," a reaction occurs at active surface sites in the particles reducing the tendency for spontaneous ignition. The heat generated by this reaction is absorbed by a fluidizing gas stream which is circulated through a cyclone to remove entrained solids and a heat exchanger before being returned by a blower to the VFB. Oxygen content in the loop is maintained by introducing the proper amount of air through a control valve. Excess gas in the loop is purged to the dryer combustor for incineration.



Following the VFB, the solids are cooled to near atmospheric temperature in an indirect rotary cooler. A controlled amount of water is added in the rotary cooler to rehydrate the PDF<sup>TM</sup> to near its ASTM equilibrium moisture content. This is also an important step in the stabilization of the PDF<sup>TM</sup>. The cooled PDF<sup>TM</sup> is then transferred to a storage bin. Because the solids have little or no free surface moisture and, therefore, are likely to be dusty, a patented dust suppressant is added as PDF<sup>TM</sup> leaves the product surge bin. Patents are pending on both the oxidative deactivation and rehydration steps.

At the present time, the PDF<sup>TM</sup> is not completely stabilized with respect to oxygen and water upon leaving the plant. The PDF<sup>TM</sup> must be "finished" by a short exposure to atmospheric conditions in a layered stockpile prior to being reclaimed and shipped. In addition to atmospheric stabilized PDF<sup>TM</sup>, a stable product can be made by blending run-of-plant PDF<sup>TM</sup> with either ROM coal or the atmosphere stabilized PDF<sup>TM</sup>, but there is a Btu penalty. ENCOAL<sup>®</sup> has recently completed pilot-scale equipment tests that successfully perform this finishing step using process equipment. The design uses commercially available equipment to be installed just downstream of rotary cooler mentioned above, and will effectively stabilize PDF<sup>TM</sup> on a continuous basis. Installation of this equipment is currently scheduled in 1997.

The hot gas produced in the pyrolyzer is sent through a cyclone for removal of the particulates and then cooled in a quench column to stop any additional pyrolysis reactions and to condense the desired liquids. Only the CDL<sup>TM</sup> is condensed in this step; the condensation of water is avoided. Electrostatic precipitators recover any remaining liquid droplets and mists from the gas leaving the condensation unit.

Almost half of the residual gas from the liquid recovery unit is recycled directly to the pyrolyzer, while some is first burned in the pyrolyzer combustor before being blended with the recycled gas to provide heat for the mild gasification reaction. The remaining gas is burned in the dryer combustor, which converts sulfur compounds to sulfur oxides. Nitrogen oxide emissions are controlled via appropriate design of the combustor. The hot flue gas from the dryer combustor is blended with the recycled gas from the dryer to provide the heat and gas flow necessary for drying.

The unrecycled portion of the off-gas from the dryer is treated in a wet gas scrubber and a horizontal scrubber, both using a water-based sodium carbonate solution. The wet gas scrubber recovers the fine particulates that escape the dryer cyclone, and the horizontal scrubber removes most of the sulfur oxides from the flue gas. The treated gas is vented to a stack. The spent solution is discharged into a pond for evaporation. The plant has several utility systems supporting its operation. These include nitrogen, steam, natural gas, compressed air, bulk sodium carbonate and a glycol/water heating and cooling system. Figure 2 is a plot plan for the ENCOAL<sup>®</sup> Plant facilities including the Buckskin Mine rail loop that is used for shipping products.

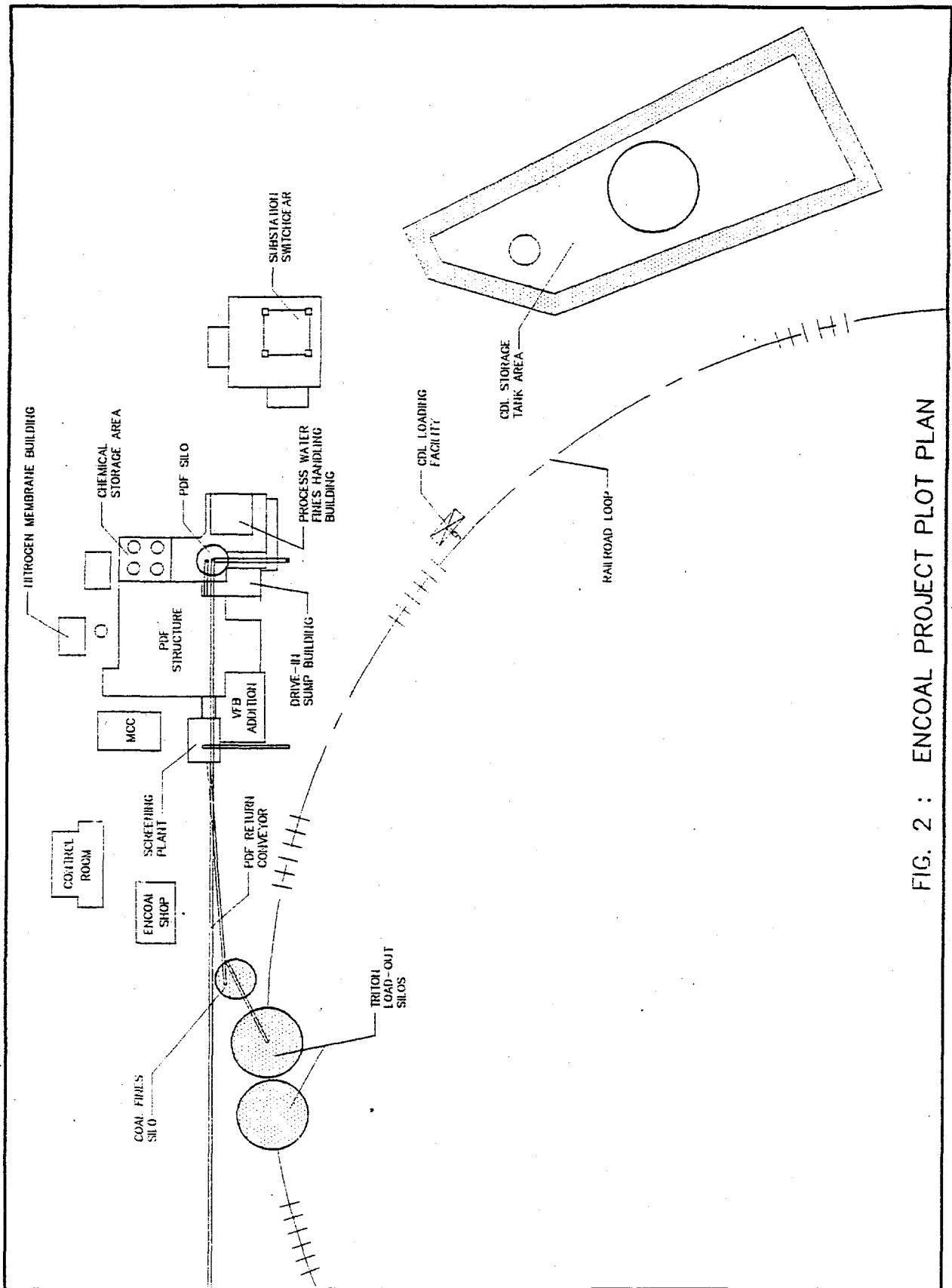


FIG. 2 : ENCOAL PROJECT PLOT PLAN



## PLANT OPERATING EXPERIENCE

### Production History

ENCOAL's LFC™ plant and facilities have now operated in an integrated mode producing PDF™ and CDL™ for more than 12,000 hours. The major pieces of equipment, including the large blowers, combustors, dryers, pyrolyzer and cooler have operated far more hours overall considering hot standby and ramping operations. This equipment has been demonstrated to operate reliably. Steady state operation exceeding 90% availability has been achieved for extended periods for the entire plant, albeit at 50% of plant capacity, and the plant is currently operational. Although some testing is still ongoing, all of the plant production of PDF™ and CDL™ is for test burns. Table 1 summarizes the plant operations over the last 4½ years.

	1992	1993	1994	1995	1996*
<b>Raw Coal Feed (Tons)</b>	5,200	12,400	67,500	65,800	59,500
<b>PDF™ Produced (Tons)</b>	2,200	4,900	31,700	28,600	30,500
<b>PDF™ Sold (Tons)</b>	-0-	-0-	23,700	19,100	32,700
<b>CDL™ Produced (Bbl)</b>	2,600	6,600	28,000	31,700	27,500
<b>Hours on Line</b>	314	980	4,300	3,400	3,200
<b>Average Length of Runs</b>	2.2	8.2	25.9	38.0	N/A**
<p>* Through November 15, 1996 ** Not Applicable; Plant in operation.</p>					

Table 1. ENCOAL® Plant Performance

Product recoveries from the feed coal have varied somewhat from the original projections. In the case of PDF™, recovery has been slightly lower. This is because more fines are generated in the process than expected and they are not all currently recovered. CDL™ recovery is higher than expected by 10-15%, apparently due to a more efficient liquid recovery system than the one used in the pilot plant.

### Product Test Burns

Commercialization of both the solid (PDF™) and liquid (CDL™) products from the ENCOAL® Plant took a major step forward in 1994. PDF™ was shipped in trainload quantities for the first time

to utility customers. The results of these shipments demonstrated that utility and industrial users can plan for test burns of PDF<sup>TM</sup> with confidence. Use of CDL<sup>TM</sup> in the industrial low sulfur residual fuel oil market was also demonstrated.

In September 1994, ENCOAL<sup>®</sup> commenced shipment of PDF<sup>TM</sup> to utility customers via the Burlington Northern railroad. Shipments made to the first customer, the Western Farmers Electric Cooperative in Hugo, Oklahoma, started at a 15% blend level and ranged up to 30%. The upper level of these blends was determined by the heat content limit in the customer's boiler. Shipments to a second customer, Muscatine Power and Water in Muscatine, Iowa, started at 40% PDF<sup>TM</sup> and ranged up to 91%. The rail cars in this shipment were capped with a small amount of ROM Buckskin coal. Capping is one way to control loss of fine material during shipment. Because the ROM coal becomes blended with the PDF<sup>TM</sup> upon unloading, it ends up as a 91% blend.

With these first shipments, ENCOAL's goals were to demonstrate its ability to coordinate with the Buckskin Mine in loading and shipping consistent blends, to ship PDF<sup>TM</sup> with dust generation comparable to or less than ROM Buckskin coal, and to ship PDF<sup>TM</sup> blends that are stable with respect to self heating. Furthermore, ENCOAL<sup>®</sup> intended to demonstrate that PDF<sup>TM</sup> could be transported and delivered to customers using regular commercial equipment. With respect to utilization, the goal for these shipments was for customers to burn trial amounts (½ unit train minimum) of PDF<sup>TM</sup> blends with minimal adjustment of equipment. These goals have all been met as reported in a more detailed test burn report<sup>[2]</sup>.

In 1995, ENCOAL<sup>®</sup> shipped two additional trains to Muscatine and initiated shipments to a third customer, Omaha Public Power District (OPPD) in Omaha, Nebraska. Three unit trains were shipped to OPPD containing approximately 25% PDF<sup>TM</sup>. This customer has been burning PRB coal in a boiler designed for bituminous coal for some time, and the increased heat content of the PDF<sup>TM</sup> blends helped increase plant output.

In 1996, ENCOAL<sup>®</sup> began shipping unit trains containing 100% PDF<sup>TM</sup> for the first time. As of the end of October 1996, two 100% PDF<sup>TM</sup> unit trains have been delivered to two separate utilities for test burns. The first was burned in Indiana-Kentucky Electric Cooperative's (IKEC) Clifty Creek Station, which is jointly owned by American Electric Power (AEP). The PDF<sup>TM</sup> was blended with Ohio high sulfur coal at the utility and burned in the Babcock & Wilcox open-path, slag-tap boiler with full instrumentation. Blends tested ranged between 70 and 90% PDF<sup>TM</sup>, and burn results indicated that even with one pulverizer out of service, the unit capacity was increased significantly relative to the base blend. More importantly, there was at least a 20% NO<sub>x</sub> reduction due to a more stable flame. Completion of this test burn achieved a major DOE Cooperative Agreement Milestone of testing PDF<sup>TM</sup> at a major U.S. utility. This goal is discussed further in an independent third party test burn report.<sup>[6]</sup> The remaining 100% PDF<sup>TM</sup> unit train was sent to Union Electric near St. Louis, MO. PDF<sup>TM</sup> shipments through October 1996 are documented in Table 2.

Coincident with PDF<sup>TM</sup> shipments was a broadening of the customer base for the liquid CDL<sup>TM</sup> product. To date, ENCOAL<sup>®</sup> has shipped CDL<sup>TM</sup> to eight different customers. With the exception of one steel mill injectant test, the CDL<sup>TM</sup> has been blended and used as fuel oil. CDL<sup>TM</sup> has proven to be acceptable in the fuel oil market through these test burns.<sup>[2]</sup> However, since the price of fuel oil is currently very low, upgrading of CDL<sup>TM</sup> into more profitable products has been studied. Initial

testing of CDL™ has shown that extraction of higher value products is both technically and economically feasible. Detailed characterization of the CDL™ and evaluation of several upgrading processes have already been completed. Other processes continue to be studied, but in general, upgrading of CDL™ will yield specialty chemical feedstocks and transportation fuels. Further work on upgrading is planned in 1997. Table 3 summarizes the CDL™ tank car shipments thus far.

DATE LOADED	CUSTOMER	BLEND (%PDF™)	TONS SHIPPED			HEAT CONTENT (Btu/lb)
			PDF™	COAL	BLEND	
09/17/94	W. Farmers	14.4	922	5,448	6,370	8,760
09/24/94	W. Farmers	21.2	1,080	4,020	5,100	8,910
10/01/94	W. Farmers	25.1	1,508	4,493	6,001	8,940
10/10/94	W. Farmers	31.9	1,603	3,241	5,024	9,310
10/24/94	W. Farmers	24.0	2,665	8,426	11,091	9,060
11/23/94	Muscatine	39.0	1,957	3,122	5,079	9,630
11/29/94	Muscatine	66.6	3,423	1,713	5,136	9,670
12/13/94	Muscatine	90.7	10,576	1,082	11,658	10,000
04/23/95	Muscatine	33.0	3,979	8,094	12,073	9,127
05/05/95	Omaha PPD	24.4	2,711	8,412	11,123	8,940
05/11/95	Omaha PPD	24.0	2,669	8,464	11,133	8,939
05/13/95	Omaha PPD	26.0	2,952	8,398	11,350	8,854
08/16/95	Muscatine	94.0	6,750	434	7,184	9,873
04/25/96	IKEC (AEP)	100.0	9,739	0	9,739	10,682
07/22/96	Union Electric	100.0	11,260	0	11,260	10,450

Table 2. Summary of Trains Shipped Containing PDF™ (Through 10/31/96)

CUSTOMER	# OF CARS	DESTINATION	USE
Dakota Gas	87	Beulah, ND	Industrial Boiler
Texpar	3	Milwaukee, WI	Small Boilers
3 M Company	14	Hutchinson, MN	Industrial Boiler
Kiesel	2	St. Louis, MO	Blend W/ #6 Oil
US Steel	2	Chicago, IL	Steel Mill Blast Furnace
Michigan Marine	18	Detroit, MI	Blend W/ #6 Oil
M&S Petroleum	40	Lake Charles, LA	Fuel Oil Blend
Baka Energy INC.	6	Houston, TX	Fuel Oil Blend

**Table 3. Summary Of CDL™ Tank Car Shipments (Through 11/15/96)**

## CHALLENGES

A detailed review of equipment and plant modifications through July 1995 has been presented<sup>[1,3,5]</sup>. Table 4 summarizes the major challenges that have been overcome and the solutions implemented.

AREA OF PLANT	DEFINITION OF PROBLEM	SOLUTION
Electrostatic Precipitators	Insulator Failures	Modified Insulators, Improved Temperature Control
Material Handling	Plugging and Spillage	Modified S-belts & Chutes
PDF <sup>TM</sup> Quenching and Steam Condenser	Oil and Coal Dust, Too Small	Added Scrubber, Added 2 Larger Exchangers
Dryer and Pyrolyzer	Sand Seal Failures	Replaced With Water Seals
Combustors	Unstable Operation	Revised Control System
Pumps and Blowers	Sizing Problems, Mostly Too Small	Replaced With Larger Equipment
Changing Process Variables	Initial Plant Design Parameters Were Off	Adjusted Operating Set Points
PDF <sup>TM</sup> Dust Collection	Dusty Conditions On Product Side of Plant - No Scrubbers	Added Two Wet Scrubbers
PDF <sup>TM</sup> Deactivation	Could Not Produce Stable PDF <sup>TM</sup> In Original Equipment	Added VFB Deactivation Loop Equipment
Process Water System	Accumulation Of Oily Fines In Process Equipment	Installed Clarifier, Floc & Vacuum Filter
Cyclone Fines Handling	Loss Of Excessive Amounts Of PDF <sup>TM</sup> In Cyclone Fines, Labor Intensive Clean-up	Recovered VFB Deactivation Fines Into PDF <sup>TM</sup> Product, Reduced Handling System
VFB Drag Conveyors	Excessive Wear and Maintenance Intensive	Redesigned High Wear Points, Modified Discharges To Reduce Plugging
Plant Operability And Maintenance	Difficult Access, Labor Intensive Clean-up, Inflexible To Operate	Piping Revisions, Access Platforms And Doors, Relocate Valves

Table 4. Summary Of Plant Modifications

Still to be solved are several challenges involving plant capacity, PDF<sup>TM</sup> deactivation, and removal of coal fines from the CDL<sup>TM</sup>. In addition, CDL<sup>TM</sup> upgrading even on the small scale of the ENCOAL<sup>®</sup> plant, appears to be economically attractive as well as something that needs to be tested before application in a large commercial plant. Data collection and designs are complete for the plant capacity improvements and PDF<sup>TM</sup> finishing projects, and work on the other projects scheduled for next year is in progress.

### **PDF<sup>TM</sup> Deactivation**

Total product deactivation remains a key challenge. At the present time, the PDF<sup>TM</sup> is not completely stabilized in the plant but has to be "finished" by a short exposure to atmospheric conditions external to the plant. ENCOAL<sup>®</sup> has recently completed pilot-scale equipment tests that successfully performed this finishing step using process equipment. The design uses commercially available equipment to be installed just downstream of the rotary cooler, and will effectively stabilize PDF<sup>TM</sup> on a continuous basis. Installation of this equipment is currently scheduled in 1997.

### **Plant Capacity**

One known bottleneck remains that prevents attainment of full design capacity of 1,000 TPD. The VFB loop is the limiting factor, since it was designed for 50% of plant capacity. A second unit was planned once the effectiveness of the PDF<sup>TM</sup> deactivation process was demonstrated. After the PDF<sup>TM</sup> finishing equipment mentioned above is installed, the addition of the second VFB may be required to reach full plant capacity.

### **CDL<sup>TM</sup> Upgrading**

The ENCOAL<sup>®</sup> plant was intentionally designed to capture a single, wide-boiling-range liquid product, CDL<sup>TM</sup>, as opposed to making multiple liquid fractions. This was done to simplify the operation, lower the capital cost and reduce the risk associated with the added complication of liquid separations. It was determined that this would be evaluated after the basic LFC<sup>TM</sup> Technology had been demonstrated. Attention has now been turned to CDL<sup>TM</sup> upgrading since the plant has moved into a production mode.

Some preliminary feasibility and design work has indicated that upgrading of the CDL<sup>TM</sup> both in the ENCOAL<sup>®</sup> plant and on a commercial scale makes economic sense; indeed it may be required to produce products that can be sold in quantity in existing markets. The M. W. Kellogg Company developed a design and cost estimate for modifying the existing plant for upgrading CDL<sup>TM</sup> in 1995. The design used information from laboratory studies and a complete CDL<sup>TM</sup> chemical characterization to develop the a workable process.

The basic concept is to produce three commercially viable streams; (1) a transportation grade fuel feedstock that would include most of the aliphatic compounds present in CDL<sup>TM</sup>, (2) a tar acid fraction that would include the cresylic acids, phenols and light aromatics and (3) a heavy residual

bottom that would be suitable as anode binder pitch. This concept is currently being considered for implementation in the ENCOAL<sup>®</sup> plant to demonstrate its potential for commercial- sized LFC<sup>™</sup> plants as well as to enhance the economics of continued operation of the existing plant.

### **CDL<sup>™</sup> Solids Removal**

The pyrolyzer loop cyclone was specifically designed to remove the coal fines from the gas stream prior to recovery of the CDL<sup>™</sup> in the quench tower and ESP's. However, the cyclone does not effectively remove all of the fines, and the CDL<sup>™</sup> consequently has 2 to 4% entrained solids. All CDL<sup>™</sup> upgrading schemes identified to date have indicated that the fines in the CDL<sup>™</sup> are undesirable. The fines must therefore be removed or reduced in quantity in order to meet customer requirements for any sale other than fuel oil. Testing of various methods of solid/liquid separation techniques is ongoing, and installation of a system at the ENCOAL<sup>®</sup> plant is scheduled in 1997.

### **PDF<sup>™</sup> Properties**

After 4½ years of operation and production of 97,900 tons of PDF<sup>™</sup>, the properties of PDF<sup>™</sup> that can be produced in the plant are fairly well defined. The variables that are controllable to some extent in the process are the heat content, volatiles, and moisture. The components dictated by the composition of the feed coal are ash, sulfur, size consist, and hardness. The LFC<sup>™</sup> process has little impact on the ash composition or ash fusion temperature. Test data have been presented in previous reports<sup>[3]</sup> that show the variability of the PDF<sup>™</sup> with process conditions. Table 5 represents the averages of the PDF<sup>™</sup> that are currently being made at the ENCOAL<sup>®</sup> plant.

PROXIMATE ANALYSIS	PLANT RUN	LAYDOWN BLEND	TARGET
Heat Content (Btu/lb)	11,112	10,682	11,400 - 11,600
Moisture (%)	9.81	10.1	8 - 9
Ash (%)	7.56	7.9	6 - 9
Volatile Matter	25.93	26.7	21 - 24
Fixed Carbon (%)	56.70	54.8	57 - 60
Sulfur (%)	0.41	0.52	0.51 Maximum
<b>OTHER</b>			
Hardgrove Grindability	47	43	45 - 50
*Sulfur/MMBtu	0.37	0.40	0.45 Maximum
*SO <sub>2</sub> /MMBtu	0.74	0.81	0.90 Maximum
Ash Mineral Analysis	Same as coal	Same as coal	Same as coal
Ash Fusion Temperature	2220°F	2220°F	2220°F

Table 5. Average Representative Properties of PDF™

### CDL™ Properties

Like PDF™, the properties of CDL™ are influenced by the pyrolyzer operation. However, the properties of CDL™ are also influenced by operation of equipment in the pyrolysis gas loop, including the pyrolyzer cyclone, the quench tower and the electrostatic precipitators. These directly affect the amount of water and sediment in the CDL™. Again, a significant amount of data has been presented in previous reports<sup>[3]</sup>, so only the following summary table is presented here. A significant amount of work has been done on the detailed chemical characterization of CDL™ for the upgrading project discussed above. This work is ongoing and will be the subject of future reports.



	<b>CDL™</b>	<b>Low Sulfur Fuel Oil</b>
API Gravity (°)	1.3 - 3.2	5
Sulfur (%)	0.3 - 0.5	0.8
Nitrogen (%)	0.6	0.3
Oxygen (%)	6.2	0.6
Viscosity @ 122°F (cs)	280	420
Pour Point (°F)	66 - 90	50
Flash Point (°F)	165	150
MBtu/gal	140	150
Water (wt %)	0.5	<1
Solids (wt %)	2 - 4	<1
Ash (wt %)	0.2 - 0.4	<1

**Table 6. Average CDL™ Quality**

## COMMERCIALIZATION

ENCOAL® Corporation has a sublicense for the LFC™ Technology from the TEK-KOL Partnership. The Partnership, owned by SGI International and a subsidiary of Zeigler Coal Holding Company, is responsible for the commercialization and licensing of the LFC™ Technology and thus is carrying out ENCOAL's obligation under the Cooperative Agreement. Under the TEK-KOL Partnership Agreement, SGI International is designated as the Licensing Contractor responsible for licensing and promoting the LFC™ Technology. Zeigler is the administrative partner responsible for preparation of lease agreements and contracts.

Commercialization of the LFC™ Technology consists of marketing the products, PDF™ and CDL™, to interested consumers at prices that will support the construction of commercial plants. Concurrently, the LFC™ Technology must be licensed to the prospective plant owners. These may or may not be the same as the consumers of the products. The technology and product marketing activities are closely interwoven and are carried out by both TEK-KOL partners. For the most part, ENCOAL® carries out all Zeigler partnership activities.

In order to determine the viability of potential LFC™ plants, TEK-KOL has already completed several detailed commercial plant feasibility studies (called Phase II studies as described previously [3]). These studies include plant design, layout, capital estimates, market assessment for co-products, operating cost assessments, and overall financial evaluation. Operation of the ENCOAL® plant provided the basis for estimating operating cost and commencing product market development, and unlike most upgrading projects, full-scale shipment and test burns made possible by the near-commercial size of the ENCOAL® plant has provided actual market information for the basis of

these studies. Operating experience of the ENCOAL<sup>®</sup> facility was also used for the design basis and capital estimates. In February 1996, TEK-KOL and Mitsubishi Heavy Industries (MHI) signed an agreement to jointly produce Design and Engineering Cost Estimates for commercial LFC<sup>™</sup> plants. This arrangement combines the scientific, engineering, and operating experience of the TEK-KOL staff with the engineering and design experience of MHI to produce a comprehensive study. To date, three detailed LFC<sup>™</sup> Phase II studies have been completed by the TEK-KOL/MHI team. These studies are discussed below.

### **Domestic Markets**

The most promising markets for the application of the LFC<sup>™</sup> Technology in the U.S. are the subbituminous coal deposits in the Powder River Basin. Close behind are the subbituminous reserves in Alaska's Beluga field, lignites in North Dakota, followed by Texas lignites near San Antonio. Testing on all of these coals has been conducted in the TEK-KOL Development Center (Center) Sample Production Unit (SPU) with favorable results.

Application of the LFC<sup>™</sup> Technology to swelling or agglomerating coals is not feasible at this time, so most of the central and eastern U.S. coals are not candidates. Removal of sulfur by the LFC<sup>™</sup> process has proven to be significant, especially when the sulfur form is highly organic, but these bituminous coals would still be too high in sulfur after processing to meet the amended clean air act requirements. Central and eastern U.S. coals are also more costly to mine than western subbituminous coal, leaving less margin for upgrading. For these reasons, central and eastern U.S. coals do not appear to be promising candidates for LFC<sup>™</sup> processing.

**Powder River Basin.** A large portion of the extensive U.S. coal reserves lie in the Powder River Basin in Montana and Wyoming. Subbituminous and low in sulfur, this coal is ideal for processing via the LFC<sup>™</sup> Technology. That is a major reason the ENCOAL<sup>®</sup> plant was located near Gillette. The southern end of the PRB in Wyoming is of special interest because the sulfur and ash are especially low. Here the PDF<sup>™</sup> product may have an increased value for metallurgical applications or as a super compliance blending material.

Overall, the PRB has the lowest mining costs in the U.S. and, being a long distance from the major utility markets, has the highest transportation costs. This combination yields a large differential value between the raw material cost and the delivered cost. The high incremental value, a well developed transportation infrastructure, qualified, available labor force and a large number of operating mines mean that the opportunities for installation of commercial LFC<sup>™</sup> plants are very good for the PRB.

A Phase II technical and economic feasibility study was completed on one potential PRB site in 1996. This study was for a commercial-size LFC<sup>™</sup> plant to be located at Triton Coal Company's North Rochelle Mine site. The site includes three 5,500 ton feed coal/day LFC<sup>™</sup> modules, a 240 MW cogeneration plant, and CDL<sup>™</sup> upgrading facilities integrated with the mine-site infrastructure. Results of the study indicated that the project has a financible rate of return (>15%) without any government subsidies, price supports, or tax credits. In other words, the LFC<sup>™</sup> products compete in current markets at current prices. However, the aid of government tax incentives would help off-

set the financial risk associated with a project of this magnitude. This study was recently refined in order to confirm the project economics, and to assemble design information for submittal of permit applications required by the State of Wyoming to allow construction to begin. An air permit application was submitted in November 1996 followed by Land Quality and Industrial Siting Permits around the end of the year.

**Alaska.** There are two promising areas in Alaska for the installation of commercial LFC™ plants, namely the Beluga fields and the Healy deposits. Both areas have extensive reserves, are largely subbituminous in nature and have low ash and sulfur. The Beluga coal is very near the Cook Inlet with the possibility of a deep water port for exports. However there is essentially no infrastructure to produce these reserves and this would be a costly venture. Current owners of the three main lease areas have not been able to attract buyers of the coal in the current market. Mine development would have to be included in any LFC™ plant venture.

At Healy, there is an existing producing mine and coal is shipped by rail to the coast for export. The Healy coal has been tested at the Center with good results. However the cost of mining is fairly high, transportation costs are high and there is no local market. The PDF™ and CDL™ from a project in this area may have difficulty competing with other locations.

**North Dakota Lignite.** Significant reserves of lignites are present in the Williston Basin of North Dakota and tests on some of them indicate good potential for LFC™ processing. Lab tests have indicated that good quality PDF™ and acceptable yields of CDL™ are produced using LFC™ Technology. Most recently, these coals have been further tested at the Center for mechanical strength during processing, also with positive results.

Overall, the economics of commercial LFC™ plants for the North Dakota lignites appear attractive. The coal seams are relatively thick and the sulfur and ash content are low, although not as low as the PRB. However, North Dakota is closer to some important markets. This coal is being considered for an alternate coal test in the ENCOAL® plant.

**Texas Lignite.** Numerous tests on Texas lignites have been conducted at the Center. With some lignites, the PDF™ quality and CDL™ recoveries have proven to be acceptable. However, other Texas lignites, although extensively available, are not considered to be viable candidates because of poor coal quality. Coal quality combined with proximity of the existing lignite mines near power plants designed to burn ROM material, makes the application of an LFC™ plant unlikely in the near future. Interest in exporting upgraded Texas lignites into other markets, or applying an LFC™ facility to replace an existing coal drying process would be two most likely scenarios for a Texas based facility.

### **International Markets**

TEK-KOL is also actively pursuing international opportunities for applying the LFC™ Technology. Primary areas of immediate interest are in China, Indonesia, and Russia. These areas have been identified by TEK-KOL as the most likely to develop in the near future, and accomplishments in these areas are discussed in more detail below. Other potential international applications for the

LFC™ Technology (*such as the Pacific Rim, Southeast Asia, India and Pakistan, Eastern Europe, and Australia*) that have previously been discussed<sup>[5]</sup>, have been identified by TEK-KOL as longer range development projects. For this reason, progress in these areas is not discussed in this paper.

**China.** China is the largest producer as well as the largest consumer of coal in the world. Over a third of the coal production occurs in the three northern provinces of Shanxi, Shaanxi and Inner Mongolia. However, due to significant transportation infrastructure problems, it is not always possible to move the coal within China to meet local needs. As a result of the extremely high economic growth in the southern and eastern coastal regions of China accompanied by a parallel demand for new electrical power, there are predictions that China may require imports of coal in the range of 10-50 million tons per year by 2010. Furthermore, the predictable result of burning such prodigious quantities of coal, much of it high in sulfur, is an environmental problem of such magnitude that it is a major concern not only of the Chinese government but also for the governments of neighboring countries and, indeed, the world.

For these reasons, China is viewed as one of the prime candidates for application of the LFC™ Technology. The LFC™ Technology offers China the opportunity:

- to more efficiently and effectively employ its vast resources of coal
- to conserve scarce and valuable railroad assets as a result of the moisture reduction aspect of the LFC™ Technology
- to vastly expand its exports into the world steam coal and metalurgical markets and, thereby, generate much needed foreign revenue
- to augment valuable and increasingly scarce petroleum assets through the production of CDL™
- to reduce the extremely severe pollution problems associated with burning high sulfur coal

The LFC™ Technology has been actively promoted in China for several years with the Ministry of Coal Industry (MOCI) and officials of regional coal mine administrations by explaining the value of employing the LFC™ Technology and developing potential commercial plant projects. Although China has huge quantities of bituminous and anthracite coal, it also has great reserves of subbituminous and lignite coals that are ideal candidates for upgrading using the LFC™ Technology. MOCI expressed keen interest in the advantage to China offered by the LFC™ Technology and representatives of SGI International have visited various mining areas in China that could be potential sites for LFC™ projects.

**Indonesia.** Approximately 93% of Indonesia's reported 36+ billion metric tons of reserves are in the form of subbituminous and lignite coal. Significantly, though, this accounts for over 97% of the identified recoverable reserves in all of the Asian countries. These reserves are split approximately 70% on the island of Sumatra and 30% on the island of Kalimantan. In fact, the Indonesian reserves have not been definitively studied yet and there exists some question as to the full extent of the identified and hypothetical reserves. On a positive note, the vast majority of the mines are open-cut operations enjoying thick seams and are mostly located near the coast or close to a navigable river, facilitating ready access to international as well as domestic markets.

Indonesia's rapid economic growth during the past decade has fueled an increase in the demand for electrical power that has grown at 11-15% per year. Furthermore, although Indonesia has been a major exporter of oil, as a result of the surging domestic growth and the limited oil reserves, it is predicted to become a net importer of petroleum by the year 2000. While a significant portion of the coal production will be destined to feed the growing domestic electrical power and industrial needs, Indonesia also requires the foreign exchange credits which will result from increasing the export market. Consequently, it is under strong pressure to better exploit its vast reserves of subbituminous and lignite coal.

Toward this end, work has been ongoing in Indonesia for over five years to promote the advantages of the LFC™ Process in answering many of Indonesia's needs. The coal industry is dominated by P.T. Tambang Batubara Bukit Asam (PTBA), the state coal mining corporation which operates under the Ministry of Mines and Energy. The structure of the industry includes the state-owned mines operated by PTBA, national companies contracted by PTBA under coal concession contract agreements, private domestic companies operating under mining concessions issued by PTBA and a few local area coal cooperatives.

Employment of the LFC™ process to upgrade low-rank coal would permit Indonesia, which is closer to Japan, South Korea, Taiwan and Hong Kong, to become very competitive in the steam coal markets. A Phase I study on some thirteen different samples indicated that several of the coals of the Tanjung Enim region of South Sumatra were good-to-excellent candidates for upgrading using the LFC™ process. Indonesia, which is short on investment capital, submitted a request to the U.S. Trade and Development Agency (TDA) for a grant for a Phase II study. This grant was approved by the TDA, and a Phase II study was completed in September 1996. This project included one to three LFC™ modules with a range of 40 to 100 MW of cogeneration, along with CDL™ upgrading facilities, transportation infrastructure, and living quarters. The study did not include the development and operation of the adjacent mine. Economics of the PTBA study were encouraging, and efforts to sign a contract with PTBA to conduct a more detailed investigation are underway.

Additionally, one Phase II study on a site adjacent to a P.T. Berau Lati Mine in East Kalimantan was completed. The study included a single LFC™ module, 40MW cogeneration plant, and a CDL™ upgrading facility that was located adjacent to the existing mine river shipping station. This one module LFC™ plant case resulted in moderate economics due to its limited throughput and relatively high operating cost. The Lati Mine coal was determined to be exceptional candidate for upgrading using the LFC™ Technology. However, local infrastructure issues, including the price of feed coal, must be resolved before the situation becomes favorable for a profitable development of a commercial LFC™ project.

Opportunities continue to be pursued in Indonesia from Aceh at the northern tip of Sumatra to lignite mines in Sulawesi. The value of the LFC™ Technology to Indonesia parallels very closely the advantages mentioned for China. Where China enjoys huge production capabilities in all forms of coal, it is especially important to Indonesia to upgrade the vast reserves of subbituminous and lignite coals in order to participate effectively in the world steam coal market. Much of Indonesian coal is already naturally low in sulfur, so the resulting PDF™ is particularly attractive to markets in Japan. Work is continuing with MHI and other Japanese firms interested in cooperating in the development of projects in Indonesia and the rest of Asia.

**Russia.** Russia accounts for about 60% of the coal production of the former Soviet Union with almost all the rest coming from Ukraine and Kazakhstan. The increasing importance of coal to the fuel and energy balance of Russia must be viewed with the understanding of the major drop in crude oil production and decreased growth rate of gas production. Representatives of the Russian coal group ROSUGOL and the Kemerovo Coal Certification Center in south central Siberia have been evaluating a project using the LFC™ Technology in the Kemerovo region. Following a visit to SGI's offices in La Jolla, California and the ENCOAL® Plant in Gillette, Wyoming, Russian representatives signed a letter of intent to proceed with Phase I and Phase II studies for an LFC™ project. The Russian delegation was particularly excited about the value added by the production of CDL™ which is so important in view of reduced oil production. The Phase I study was completed in late 1995, and indicated that the coals tested were suitable for LFC™ upgrading. Work on a Phase II study is expected to begin in 1997 pending Russian agreement to proceed. If successful, this Russian endeavor could be the first of many projects in this country with huge potential reserves.

### **Long Term Impact Of LFC™ Commercialization**

The LFC™ Technology is uniquely positioned in the world coal conversion and upgrading market to impact two widely used fossil energy forms, namely solids and liquids. Many technologies have successfully demonstrated the conversion of coal to synthetic gases which are in turn used as a clean energy source. Others have demonstrated the manufacture of hydrocarbon liquids from these synthetic gases to serve as chemical or transportation fuel feedstocks. Still other technologies have demonstrated the technical feasibility of direct conversion of coal to hydrocarbon liquids. Although not truly coal conversion, coal upgrading by removal of undesirable constituents like water, sulfur and ash has also been extensively demonstrated on a commercial scale by numerous technologies. The LFC™ Technology alone produces both an upgraded solid product and hydrocarbon liquids.

Economic conditions for typical commercial coal conversion and upgrading projects are generally absent without some form of political intervention, such as price supports, grants, subsidies or artificial market constraints. While tax credits would be helpful on the first LFC™ plant to offset risks, commercial LFC™ plants can compete in today's markets at today's prices with attractive rates of return. Therefore, countries with significant indigenous coal reserves (like the U.S.) or countries with significant investment or material supply interests (like Japan), should be able to use the LFC™ Technology to further economic growth.

Of course there are practical limits to the application of the LFC™ Technology. Some of the criteria for successful commercial projects can be generally stated as:

- Significant coal reserves - greater than 150 million ton block for a 3 module LFC™ plant
- Non caking, non agglomerating coal - like most low rank coals
- Low mining costs
- Low ash and inorganic sulfur content
- Located near navigable water or other reasonably priced accessible form of transportation

- Favorable political climate
- Markets for products for products at acceptable prices

There are many coal deposits in the world today which meet all of these criteria.

Consumers of solid and liquid energy products, which more and more is a world-wide market, should see significant advantages in the products from commercial LFC™ plants. The benefits for the consumer can be summarized as:

- Reduced dependence on petroleum based liquid products and the widely variable prices in that market
- Reduced environmental impact from the burning of PDF™ and CDL™ in the form of lower SO<sub>x</sub> and NO<sub>x</sub> as demonstrated by test burns. LFC™ plants are also very environmentally benign
- Lower fuel costs for power plants and industrial boilers on a fully utilized basis
- Long term, stable fuel supply
- Unique characteristics for metallurgical and ferroally markets
- For consumers with coal reserves, increased use of domestic resources

Given the widespread availability of qualified candidate coals and the numerous benefits that accrue to consumers of the LFC™ products, commercialization of the LFC™ Technology should be able to make a major long term positive impact on the world energy picture. TEK-KOL and the commercial LFC™ plant development team are actively pursuing these opportunities.

## **FUTURE WORK**

The next step in the Project is to continue to deliver high quality, pure PDF™ to utility customers and potential steel industry and ferroally users for test burns. These deliveries will aid in the development of future PDF™ markets and help secure product contracts for commercial LFC™ plants. Work on installing PDF™ finishing equipment, plant capacity upgrades, and CDL™ solids removal systems are expected in 1997. Installation and operation of these systems will provide the operation data and experience important for the final design and construction of a commercial LFC™ facility.

The goal is to maintain better than 90% availability on the plant this year and complete any remaining major plant modifications by the end of 1997. Efforts to commercialize the LFC™ Technology will continue both at home and abroad. The evaluation of CDL™ upgrading will also continue and a decision made about proceeding with an ENCOAL® plant modification.

## **CONCLUSIONS**

The ENCOAL® Project has completed most of its goals. Essentially all the major Cooperative Agreement Milestones have been met, and final reporting requirements will be completed in early 1997. The debugging phase is complete and steady state operation has been achieved. The LFC™

Technology is essentially demonstrated and marketable PDF™ and CDL™ are being produced.

Significant quantities of both products have been shipped and successfully used by customers, thus proving them to be acceptable fuel sources in today's markets. Efforts to commercialize the LFC™ Technology, both domestically and internationally, are in progress.

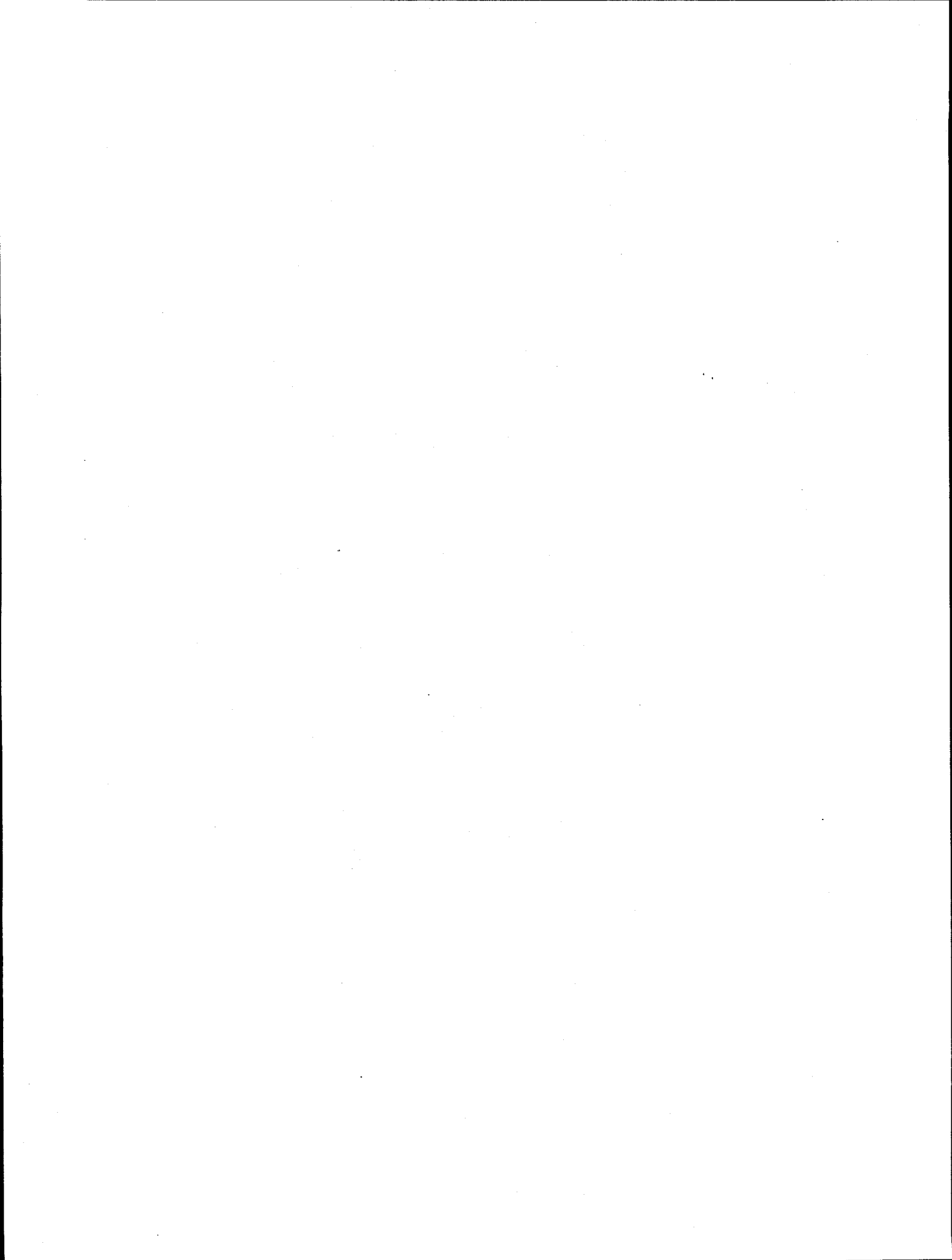
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## GLOSSARY

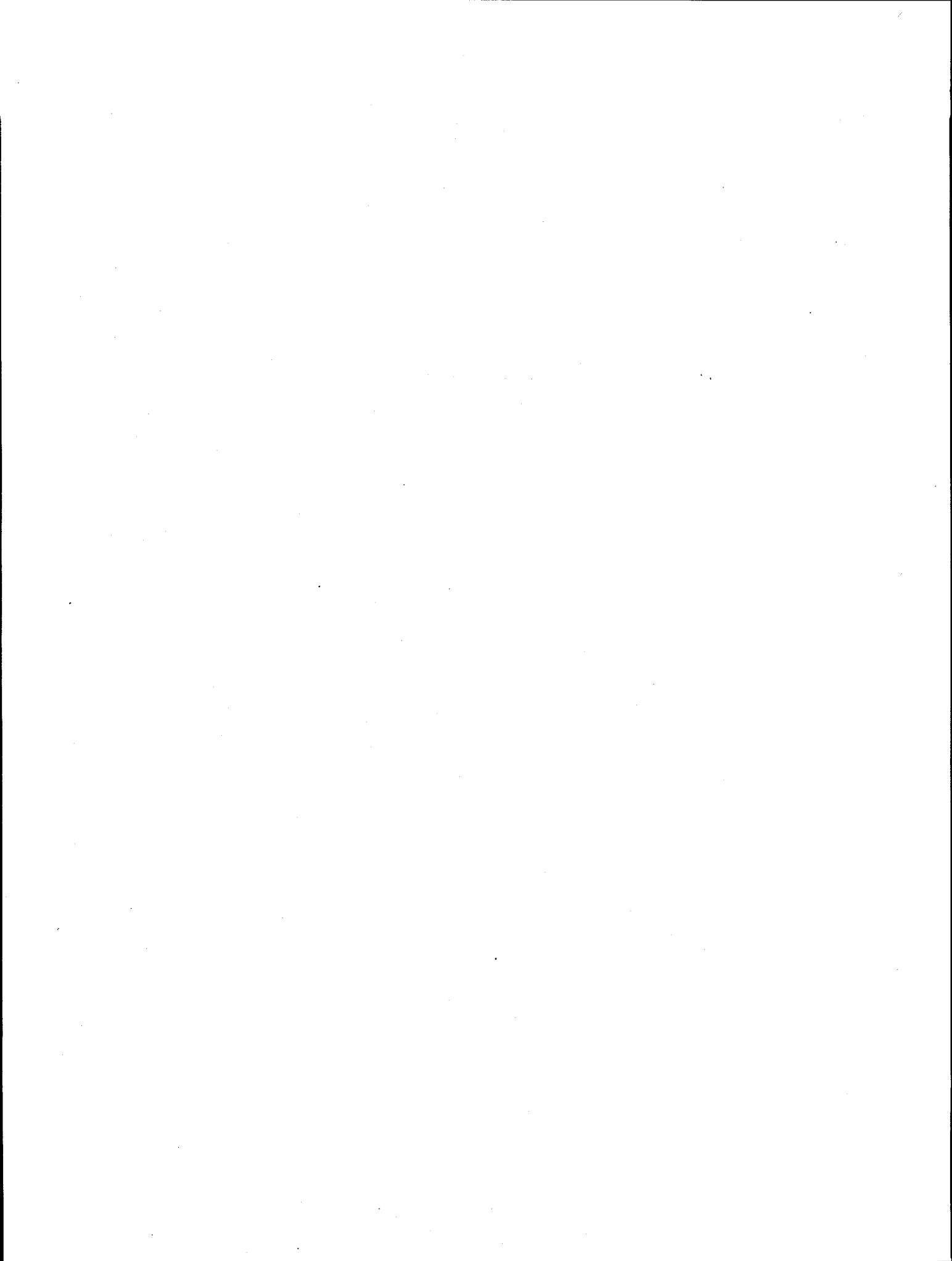
AEP	American Electric Power
AS	American Society of Testing Methods
°API	American Petroleum Institute measure of oil density
BACT	Best Available Control Technology
Btu	British Thermal Unit
Center	TEK-KOL Development Center in Perrysburg, Ohio
CDL™	Coal Derived Liquid
CO	Carbon Monoxide
CH <sub>4</sub>	Methane
DOE	U. S. Department of Energy
ENCOAL®	ENCOAL® Corporation, a wholly owned subsidiary of Bluegrass Coal Development Co., which is a wholly owned subsidiary of Zeigler Coal Holding Co.
EPA	U.S. Environmental Protection Agency
ESP	Electrostatic Precipitators
IKEC	Indiana-Kentucky Electric Cooperative
lb.	Pound
LFCT™	Liquid From Coal
MHI	Mitsubishi Heavy Industries, Hiroshima, Japan
MMBtu	Million British Thermal Units
MOCI	Ministry of Coal Industry
MT	Metric Tonnes
N/A	Not Available
NO <sub>x</sub>	Nitrogen Oxides
OPPD	Omaha Public Power District, Omaha, Nebraska
OSHA	Occupational Safety & Health Administration
PDF™	Process Derived Fuel
PRB	Powder River Basin
ROM	Run-of-mine
S-Belt	Vertical conveyor with flexible sidewalls and rubber buckets
SGI	SGI International, La Jolla, CA
SMC	SMC Mining Company, Evansville, IN ( <i>name changed to Bluegrass Coal Development Co.</i> )
SO <sub>2</sub>	Sulfur Dioxide
SPU	Sample Production Unit, TEK-KOL Development Center
Std. Dev.	Standard Deviation
TEK-KOL	A general partnership between SGI International and a subsidiary of Zeigler Coal Holding Company
TGA	Thermogravimetric analysis, procedure for analyzing coal and PDF™
TPD	Tons Per Day
vs.	Versus
WP&L	Wisconsin Power and Light
wt.	Weight
#	Pound



**Coal-Diesel Demonstration Project  
at University of Alaska, Fairbanks**

**Robert Wilson  
Arthur D. Little, Inc.**

**Paper Not Available**



# **Technical Session III**

## **Advanced Cleanup Systems**

**THE CLEAN COAL TECHNOLOGY PROGRAM  
10 MWe DEMONSTRATION OF GAS SUSPENSION ABSORPTION  
FOR FLUE GAS DESULFURIZATION**

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**ABSTRACT**

AirPol Inc., with the cooperation of the Tennessee Valley Authority (TVA) under a Cooperative Agreement with the United States Department of Energy, installed and tested a 10 MWe Gas Suspension Absorption (GSA) Demonstration system at TVA's Shawnee Fossil Plant near Paducah, Kentucky. This low-cost retrofit project demonstrated that the GSA system can remove more than 90% of the sulfur dioxide from high-sulfur coal-fired flue gas, while achieving a relatively high utilization of reagent lime.

This paper presents a detailed technical description of the Clean Coal Technology demonstration project. Test results and data analysis from the preliminary testing, factorial tests, air toxics tests, 28-day continuous demonstration run of GSA/electrostatic precipitator (ESP), and 14-day continuous demonstration run of GSA/pulse jet baghouse (PJBH) are also discussed within this paper.

## I. INTRODUCTION

AirPol, with the assistance of the Tennessee Valley Authority (TVA), demonstrated the Gas Suspension Absorption (GSA) technology in the Clean Coal Technology project entitled "10 MW Demonstration of Gas Suspension Absorption." AirPol performed this demonstration under a Cooperative Agreement awarded by the United States (U.S.) Department of Energy (DOE) in October 1990. This project was selected in Round III of the Clean Coal Technology Program.

This project was the first North American demonstration of the GSA system for flue gas desulfurization (FGD) for a coal-fired utility boiler. This low-cost retrofit project achieved the expected target, which was to remove more than 90% of the sulfur dioxide ( $\text{SO}_2$ ) from the flue gas while achieving a high utilization of reagent lime. TVA furnished its Center for Emissions Research (CER) as the host site and provided operation, maintenance, and technical support during the project. The CER is located at the TVA's Shawnee Fossil Plant near Paducah, Kentucky.

The experience gained by AirPol in designing, fabricating, and constructing the GSA equipment through the execution of this project will be used for future commercialization of the GSA technology. The results of the operation and testing phase will be used to further improve the GSA system design and operation.

The specific technical objectives of the GSA demonstration project were the following:

- Demonstrate  $\text{SO}_2$  removal in excess of 90% using high-sulfur U.S. coal.
- Optimize design and operating parameters to maximize the  $\text{SO}_2$  removal efficiency and lime utilization.
- Compare the  $\text{SO}_2$  removal efficiency of the GSA technology with existing spray dryer/electrostatic precipitator (SD/ESP) technology.

DOE issued an amendment to the Cooperative Agreement to include the additional scope of work for air toxics testing and also the operation and testing of a 1 MWe fabric filter pilot plant in cooperation with TVA and the Electric Power Research Institute (EPRI). The two-fold purpose of this additional work was the following:

- Determine the air toxics removal performance of the GSA technology.
- Compare the  $\text{SO}_2$ , particulate, and air toxics removal performance between GSA/ESP and GSA/fabric filter systems.

The fabric filter used in this project is a pulse-jet baghouse (PJBH) which can treat flue gas removed either upstream or downstream of the ESP. The testing of the PJBH was conducted for both configurations.

The total budget for the project with the added scope of work was \$7,720,000; however, the project cost was under the budget. The favorable variance resulted mainly from actual material and construction costs being much lower than the original estimate. The performance period of the project, including the air toxics measurements, PJBH testing, and report preparation was from November 1990 to June 1995.

AirPol began the design work on this project in November 1990, shortly after award of the Cooperative Agreement by DOE in October 1990. At the outset of the project, access to the site at the CER was delayed for one year by TVA to allow the completion of another project. That caused a one-year delay in this Clean Coal Technology project. The design phase of the GSA project was completed in December 1991. The fabrication and construction of the GSA unit was completed ahead of schedule in early September 1992. The planned operation and testing of the demonstration unit were conducted from late October 1992 to the end of February 1994.

## **II. HISTORY OF THE GSA TECHNOLOGY**

The GSA process is a novel concept for FGD that was developed by AirPol's parent company, F.L. Smidth miljo a/s in Copenhagen, Denmark. The process was initially developed as a cyclone preheater system for cement kiln raw meal (limestone and clay). This innovative system provided both capital and energy savings by reducing the required length of the rotary kiln and lowering fuel consumption. The GSA system also showed superior heat and mass transfer characteristics and was subsequently used for the calcination of limestone, alumina, and dolomite. The GSA system for FGD applications was developed later by injecting lime slurry and the recycled solids into the bottom of the reactor to function as an acid gas absorber.

In 1985, a GSA pilot plant was built in Denmark to establish design parameters for SO<sub>2</sub> and hydrogen chloride (HCl) absorption for waste incineration applications. The first commercial GSA unit was installed at the KARA Waste-to-Energy Plant at Roskilde, Denmark, in 1988. Currently, there are seventeen GSA installations in Europe; 15 are municipal solid waste incinerator applications, and two are industrial applications (cement and iron ore reduction).

With the increased emphasis on SO<sub>2</sub> emissions reduction by electric utility and industrial plants as required by the Clean Air Act Amendments of 1990, there is a need for a simple and economic FGD process, such as GSA, by the small to mid-size plants where a wet FGD system may not be feasible. The GSA FGD process, with commercial and technical advantages confirmed in this demonstration project, will be a viable alternative to meet the needs of utility and industrial boilers in the U.S.

## **III. GSA FGD PROCESS DESCRIPTION**

The GSA FGD system, as shown in the Figure 1 Process Flow Diagram, includes:

- A circulating fluidized bed reactor.



- A separating cyclone incorporating a system for recycling the separated material to the reactor.
- A lime slurry preparation system which proportions the slurry to the reactor via a dual-fluid nozzle.
- A dust collector which removes fly ash and reaction products from the flue gas stream.

The flue gas from the boiler air preheater is fed into the bottom of the circulating fluidized bed reactor, where it is mixed with the suspended solids that have been wetted by the fresh lime slurry. The suspended solids consist of reaction products, residual lime, and fly ash. During the drying process in the reactor, the moisture in the fresh lime slurry, which coats the outer surface of the

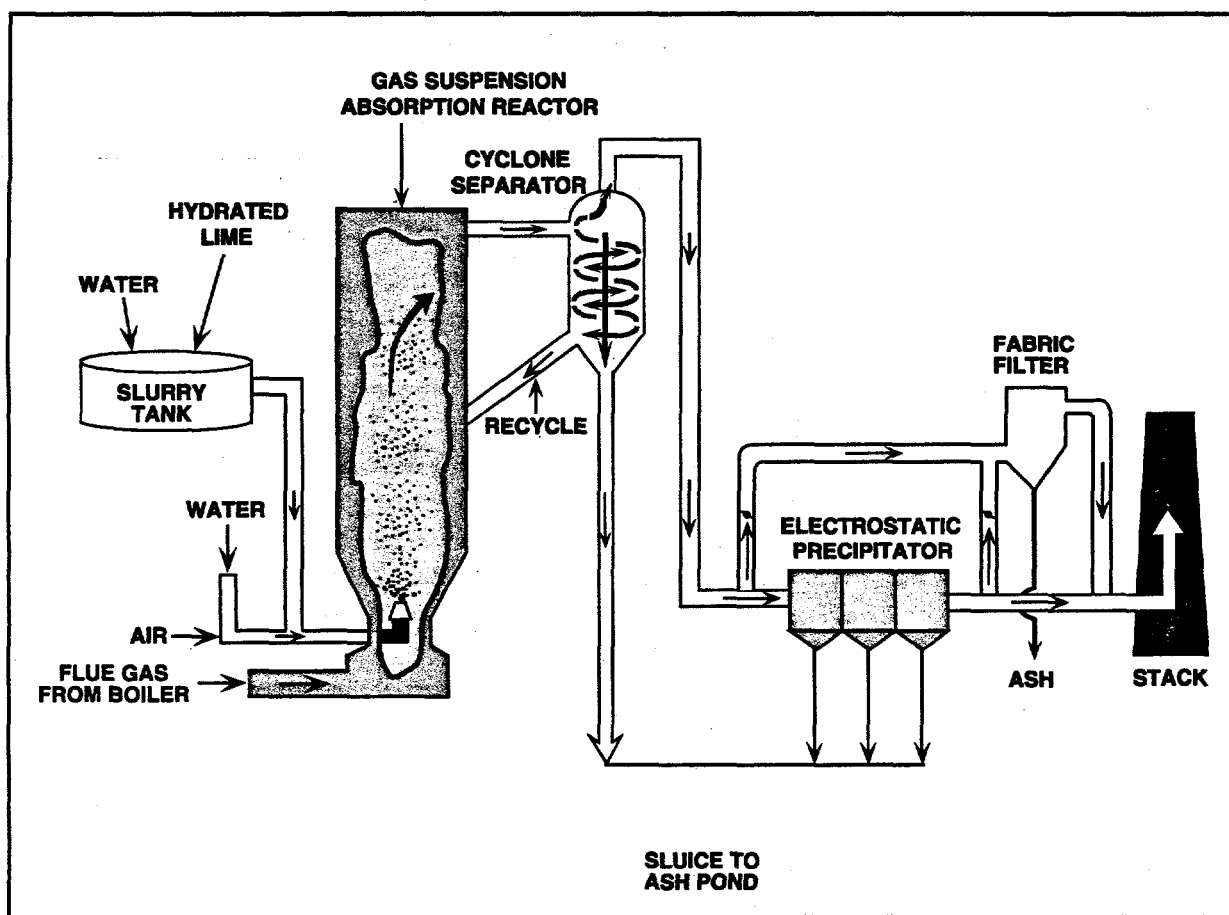


Figure 1. Gas Suspension Absorption Process Flow Diagram

suspended solids, evaporates. Simultaneously, the lime particles in the slurry undergo a chemical reaction with the acid components of the flue gas,  $\text{SO}_2$  and  $\text{HCl}$ , capturing and neutralizing them.

The partially cleaned flue gas flows from the top of the reactor to the separating cyclone and then to an ESP (or a fabric filter), which removes the dust and ash particles. The flue gas, which has now been cleaned, is released into the atmosphere through the stack.

The cyclone separates most of the solids from the flue gas stream. Approximately 95% to 99% of these collected solids are fed back to the reactor via a screw conveyor, while the remaining solids leave the system as a byproduct material. Some of these solids recirculated to the reactor are still reactive. This means that the recirculated lime is still available to react and neutralize the acid components in the flue gas.

The pebble lime is slaked in a conventional, off-the-shelf system. The resulting fresh slaked lime slurry is pumped to an interim storage tank and then to the dual-fluid nozzle. The slurry is diluted with trim water prior to being injected into the reactor.

### **Automatic Process Adjustment**

An effective monitoring and control system automatically ensures that the required level of  $\text{SO}_2$  removal is attained while keeping lime consumption to a minimum. This GSA control system, which is shown in Figure 2, incorporates three separate control loops:

1. Based on the flue gas flow rate entering the GSA system, the first loop continuously controls the flow rate of the recycled solids back to the reactor. The large surface area for reaction provided by these fluidized solids and the even distribution of the lime slurry in the reactor, provide for the efficient mixing of the lime with the flue gas. At the same time, the large volume of dry material prevents the slurry from adhering to the sides of the reactor.
2. The second control loop ensures that the flue gas is sufficiently cooled to optimize the absorption and reaction of the acid gases. This control of flue gas temperature is achieved by the injection of additional water along with the lime slurry. The amount of water added into the system is governed by the temperature of the flue gas exiting the reactor. This temperature is normally set a few degrees above flue gas saturation temperature to insure that the reactor solids will be dry so as to reduce any risk of acid condensation.

3. The third control loop determines the lime slurry addition rate. This is accomplished by continuously monitoring the  $\text{SO}_2$  content in the outlet flue gas and comparing it with the required emission level. This control loop enables direct proportioning of lime slurry feed according to the monitored results and maintains a low level of lime consumption.

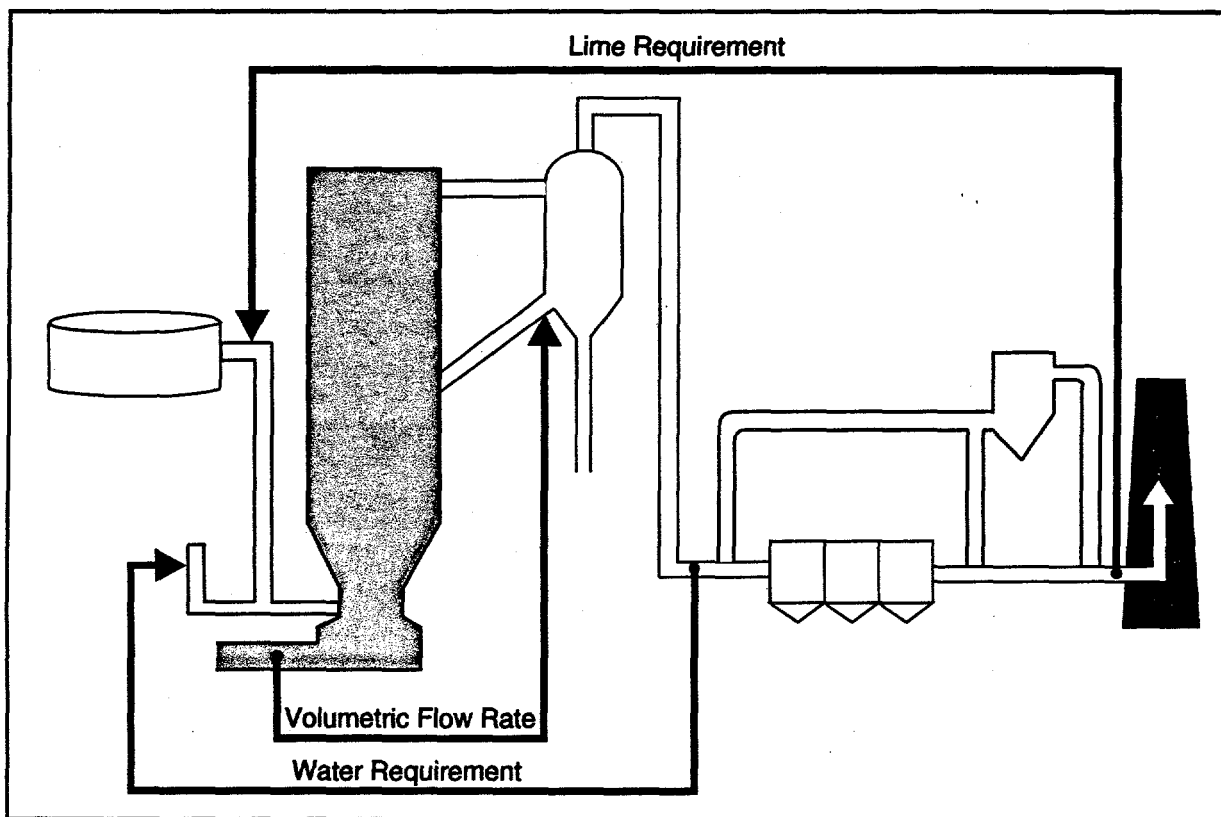


Figure 2. Gas Suspension Absorption Control System

#### IV. COMPARISON OF GSA PROCESS WITH COMPETING TECHNOLOGY

Simplicity is the key feature of the GSA system. The advantages of the GSA system over competing technologies are summarized as follows:

##### Slurry Atomization

The major difference between GSA and competing technologies lies in the way the reagent is introduced and used for  $\text{SO}_2$  absorption. A conventional semi-dry scrubber:

- Requires a costly and sensitive high-speed rotary atomizer or a high-pressure atomizing nozzle for fine atomization,
- Absorbs  $\text{SO}_2$  in an "umbrella" of finely atomized slurry with a droplet size of about 50 microns,
- May require multiple nozzle heads or rotary atomizers to ensure fine atomization and full coverage of the reactor cross section, and
- Uses recycle material in the feed slurry necessitating expensive abrasion-resistant materials in the atomizer(s).

The GSA process, on the other hand,

- Uses a low-pressure, dual-fluid nozzle,
- Absorbs  $\text{SO}_2$  on the wetted surface of suspended solids with superior mass and heat transfer characteristics,
- Uses only one spray nozzle for the purpose of introducing slurry and water to the reactor, and
- Uses dry injection of recycle material directly into the reactor, thereby avoiding erosion problems in the nozzle or technical limitations on the amount of solids that can be recycled.

### **Simple and Direct Method of Lime/Solid Recirculation**

The recirculation of used lime is the trend for semi-dry scrubbing systems. The recirculation of solids in the GSA system is accomplished using a feeder box under the cyclone, which introduces the material directly into the reactor. The recirculation feature commonly used in most other semi-dry processes has an elaborate ash handling system to convey and store the ash. The method of introducing the recirculated material is usually by mixing it with the fresh lime slurry. The presence of ash in the lime slurry may cause sediment problems in the slurry lines and excessive nozzle wear.

### **High Acid Gas Absorption**

The GSA reactor is capable of supporting an extremely high concentration of solids (recirculated material) inside the reactor, which acts like a fluidized bed. This concentration will normally be as high as 200-800 grains/scf. These suspended solids provide a large surface area for contact between the lime slurry (on the surface of the solids) and the acidic components in the flue gas. This high contact area allows the GSA process to achieve levels of performance that are closer to those of a wet scrubber, rather than a dry scrubber. Since drying of the solids is also greatly enhanced by the characteristic large surface area of the fluidized bed, the temperature inside the reactor can be

reduced below that of the typical semi-dry scrubber. This lower operating temperature facilitates acid gas removal in the GSA system.

### **Low Lime Consumption / Minimum Waste Byproduct Residue**

The design of the GSA reactor allows for more efficient utilization of the lime slurry because of the high internal recirculation rate and precise process control. The higher lime utilization (up to 80%) lowers the lime consumption, thereby minimizing one of the major operating costs. In addition, the lower lime consumption reduces the amount of byproduct generated by the system.

### **Low Maintenance Operation**

Unlike typical semi-dry scrubbers, the GSA system has no moving parts inside the reactor, thus ensuring relatively continuous, maintenance-free operation. The orifice diameter of the GSA injection nozzle is much larger than that used in a conventional semi-dry process, and there is little chance for it to plug. Nozzle wear is also minimized. Should the need for replacing the nozzle arise, it can be replaced in a few minutes. The cyclone also has no moving parts. Both the reactor and the cyclone are fabricated from unlined carbon steel.

The GSA process also has few pieces of equipment. Most of the equipment is in the lime slurry preparation area, which typically is an off-the-shelf item, and the technology is well known.

### **No Internal Buildup**

By virtue of the fluidized bed inside the reactor, the inside surface of the reactor is continuously "brushed" by the suspended solids and is kept free of any buildup. Internal wall buildup can be a problem with the conventional semi-dry scrubber. There is also no wet/dry interface on any part of the equipment and this avoids any serious corrosion problem.

### **Modest Space Requirements**

Due to the high concentration of suspended solids in the reactor, the reaction occurs in a relatively short period of time. A high flue gas velocity of 20 to 22 feet per second as compared to 4 to 6 feet per second for a semi-dry scrubber, as well as the shorter residence time of 2.5 seconds as compared to 10 to 12 seconds for a semi-dry scrubber, allow for a smaller diameter reactor which leads to a considerable reduction in space requirements.

### **Short Construction Period**

The compact design of the GSA unit requires less manpower and time to be erected as compared to the typical semi-dry scrubbers. Despite the relatively complicated tie-ins and extremely constrained work space, the retrofit GSA demonstration unit at the TVA's CER was erected in three and a half months.

### **Heavy Metals Removal**

Recent test results from waste incineration plants in Denmark indicate that the GSA process is not only effective in removing acidic components from the flue gas but is also capable of removing heavy metals, such as mercury, cadmium, and lead. This heavy metal removal capability of the GSA process at the CER was confirmed by the air toxics tests.

## **V. PROJECT STATUS AND KEY MILESTONES**

The project schedule and tasks involved in the design, construction, and operation and testing phases are as follows:

Phase I - Engineering and Design		Start - End
1.1	Project and Contract Management	11/01/90-12/31/91
1.2	Process Design	11/01/90-12/31/91
1.3	Environmental Analysis	11/01/90-12/31/91
1.4	Engineering Design	11/01/90-12/31/91
Phase II - Procurement and Construction		
2.1	Project and Contract Management	01/01/92-09/30/92
2.2	Procurement and Furnish Material	01/01/92-04/30/92
2.3	Construction and Commissioning	05/01/92-09/30/92
Phase III - Operating and Testing		
3.1	Project Management	10/01/92-12/31/94
3.2	Start-up and Training	10/01/92-10/14/92
3.3	Testing and Reporting	10/15/92-06/30/95

The parametric optimization tests were completed on schedule in August 1993. Following the air toxics testing, which was finished in October 1993, there was a 28 day, around-the-clock demonstration run from the later October to late November 1993 and a 14-day, around-the-clock PJBH demonstration run from late February to mid-March 1994. All testing has been completed and the project reports have been prepared.

## VI. TEST PLAN

A test plan was prepared to depict in detail the procedures, locations, and analytical methods to be used in the tests. All of the following objectives were achieved by testing the GSA system:

- Optimization of the operating variables.
- Determination of stoichiometric ratios for various SO<sub>2</sub> removal efficiencies.
- Evaluation of erosion and corrosion at various locations in the system.
- Demonstration of 90% or greater SO<sub>2</sub> removal efficiency when the boiler is fired with high-sulfur coal.
- Determination of the air toxics removal performance.
- Evaluation of the PJBH performance in conjunction with the GSA process.

### Optimization Tests

The optimization of the SO<sub>2</sub> removal efficiency in the GSA system was accomplished through the completion of a statistically-designed factorial test plan. For each test series, the GSA system was set to operate at a certain combination of operating parameters. The results of these test series were analyzed statistically to determine the impact of the parameters, thus arriving at the optimum operating point for the GSA process at the various operating conditions expected in future applications. Operating parameters studied in this phase of the project were the following:

- Inlet flue gas flow rate
- Inlet flue gas temperature
- Inlet dust loading
- Solids recirculation rate
- Stoichiometric ratio
- Approach-to-saturation temperature
- Coal chloride level

## **Data Collection**

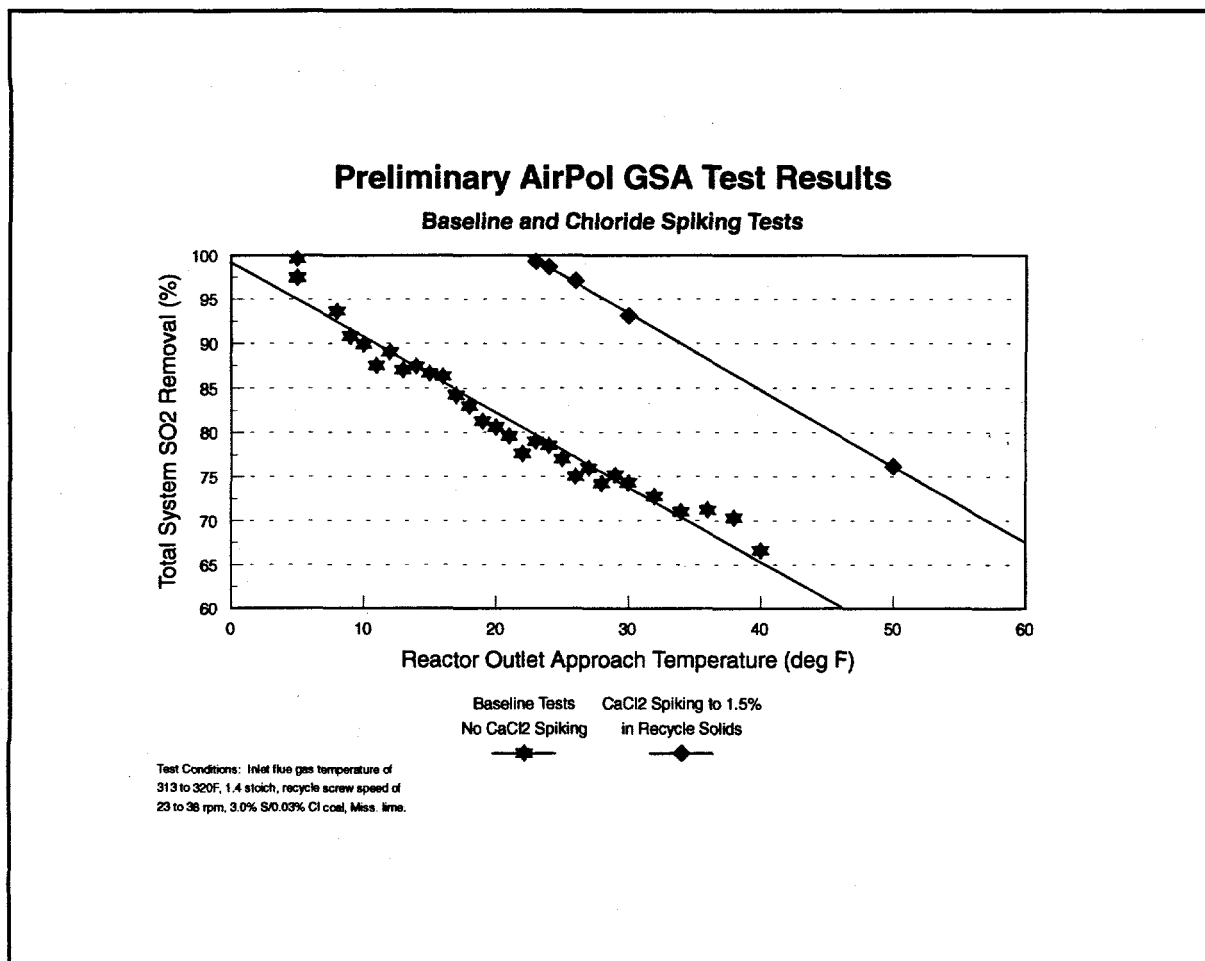
The following data were sampled and recorded during the tests by either the computerized data sampling and recording system (via field mounted instruments) or by manual field determinations:

- Inlet flue gas flow into the system
- SO<sub>2</sub> loading at the system inlet, SO<sub>2</sub> loading at the ESP inlet and outlet
- Flue gas temperature at the system inlet, the reactor outlet, and the ESP outlet
- Particulate loading at the ESP inlet and outlet
- Fresh lime slurry flow rate and composition (for lime stoichiometry calculation)
- Water flow rate
- Wet-bulb temperature at the reactor inlet (for approach-to-saturation temperature calculation)
- Coal analysis (proximate and ultimate)
- Lime analysis
- Byproduct rate and composition
- Water analysis
- Power consumption

## **VII. PRELIMINARY TESTING**

Immediately after the dedication of the AirPol GSA demonstration plant in late October 1992, a series of preliminary tests was begun. The purpose of these tests was to investigate the operating limits of the GSA system as installed at the CER. The results from several of the preliminary tests completed at the CER in November and December were very interesting, and these results were used as the basis for the design of the factorial test program. During one of the preliminary tests, the approach-to-saturation temperature in the reactor was gradually decreased and the overall system (reactor/cyclone and ESP) SO<sub>2</sub> removal efficiency was monitored over this four-day test. The overall system SO<sub>2</sub> removal efficiency increased from about 65% to more than 99% at the closest approach-to-saturation temperature (5°F). The other conditions, which remained constant, were 320°F inlet flue gas temperature, 1.40 moles Ca(OH)<sub>2</sub>/mole inlet SO<sub>2</sub> for the lime stoichiometry, and essentially no chloride in the system. The SO<sub>2</sub> removal results from this test are shown in Figure 3.





**Figure 3. Preliminary AirPol GSA Test Results**

The data from this test show that the SO<sub>2</sub> removal efficiency increased dramatically as the flue gas temperature in the reactor more closely approached the saturation temperature of the flue gas, with the incremental increases in SO<sub>2</sub> removal becoming more and more significant as the approach-to-saturation temperature declined. The ability of the GSA system to operate at this close approach-to-saturation temperature without any indication of plugging problems was surprising. Later analysis showed that the moisture level in the solids remained below 1%.

A second extended test was run during December 1992. This test was run at the same conditions as the previous test, except that in this test, calcium chloride was added to the system to simulate the combustion of a high-chloride (about 0.3%) coal. Previous work by TVA at the CER had demonstrated that spiking these semi-dry, lime-based FGD processes with a calcium chloride solution adequately simulated a high chloride coal application. Again, the approach-to-saturation temperature was gradually decreased over a four-day period with all other conditions held constant and the overall system SO<sub>2</sub> removal efficiency was monitored. The results from this second test, which are included in Figure 3 above, show that the presence of chlorides enhances SO<sub>2</sub> removal.

The overall system SO<sub>2</sub> removal efficiency for the chloride-spiked tests increased from about 70% at the high approach-to-saturation condition to essentially 100% at the closer approach-to-saturation temperature (23°F). No attempt was made to operate the system at the close approach-to-saturation temperatures used in the first test because the SO<sub>2</sub> removal efficiency was approaching 100%. In addition, there were initially some concerns about the secondary effect of calcium chloride addition. Calcium chloride is an ionic salt that tends to depress the vapor pressure of water in the system and thus, slows the evaporation of water from the slurry. Calcium chloride is also a hygroscopic material, which means it has the ability to absorb moisture from the humid flue gas. The increased moisture in the "dry" solids allows more reaction with SO<sub>2</sub>, but also increases the potential for plugging in the system. The easiest method for mitigating this potential for plugging is to increase the approach-to-saturation temperature in the reactor. However, the moisture levels in the solids during this test remained below 1%, even at the closest approach-to-saturation temperature.

Another interesting finding from the preliminary testing is that the GSA process is capable of supporting a very high level of recirculation material in the reactor. This high solid concentration inside the reactor is the reason for the superior drying characteristics of the GSA system. Based on the results from these initial tests, the recycle rate back to the reactor was doubled prior to starting the factorial testing.

## VIII. FACTORIAL TESTING

The purpose of the statistically-designed factorial test program was to determine the effect of process variables on the SO<sub>2</sub> removal efficiency in the reactor/cyclone and the ESP.

Based on the successful preliminary testing, the major process design variables were determined, levels for each of these variables were defined, and an overall test plan was prepared. The major variables were approach-to-saturation temperature, lime stoichiometry, fly ash loading, coal chloride level, flue gas flow rate, and recycle screw speed. Two levels were determined for nearly all of the variables, as shown in Table 1 below. The one exception was the approach-to-saturation temperature where three levels were defined, but the third level was run only for those tests at the lower coal chloride level.

Major Variables and Levels for Factorial Testing Table		
Variable		Level
Approach-to-saturation temperature	°F	8 <sup>a</sup> , 18, and 28
Ca/S	moles Ca(OH) <sub>2</sub> /mole inlet SO <sub>2</sub>	1.00 and 1.30
Fly ash loading	gr/acf	0.5 and 2.0
Coal chloride level	%	0.02 and 0.12
Flue gas flow rate	kscfm	14 and 20
Recycle screw speed	rpm	30 and 45
<sup>a</sup> 8°F level run only at the low-chloride level		

**Table 1.** Major Variables and Levels for Factorial Testing

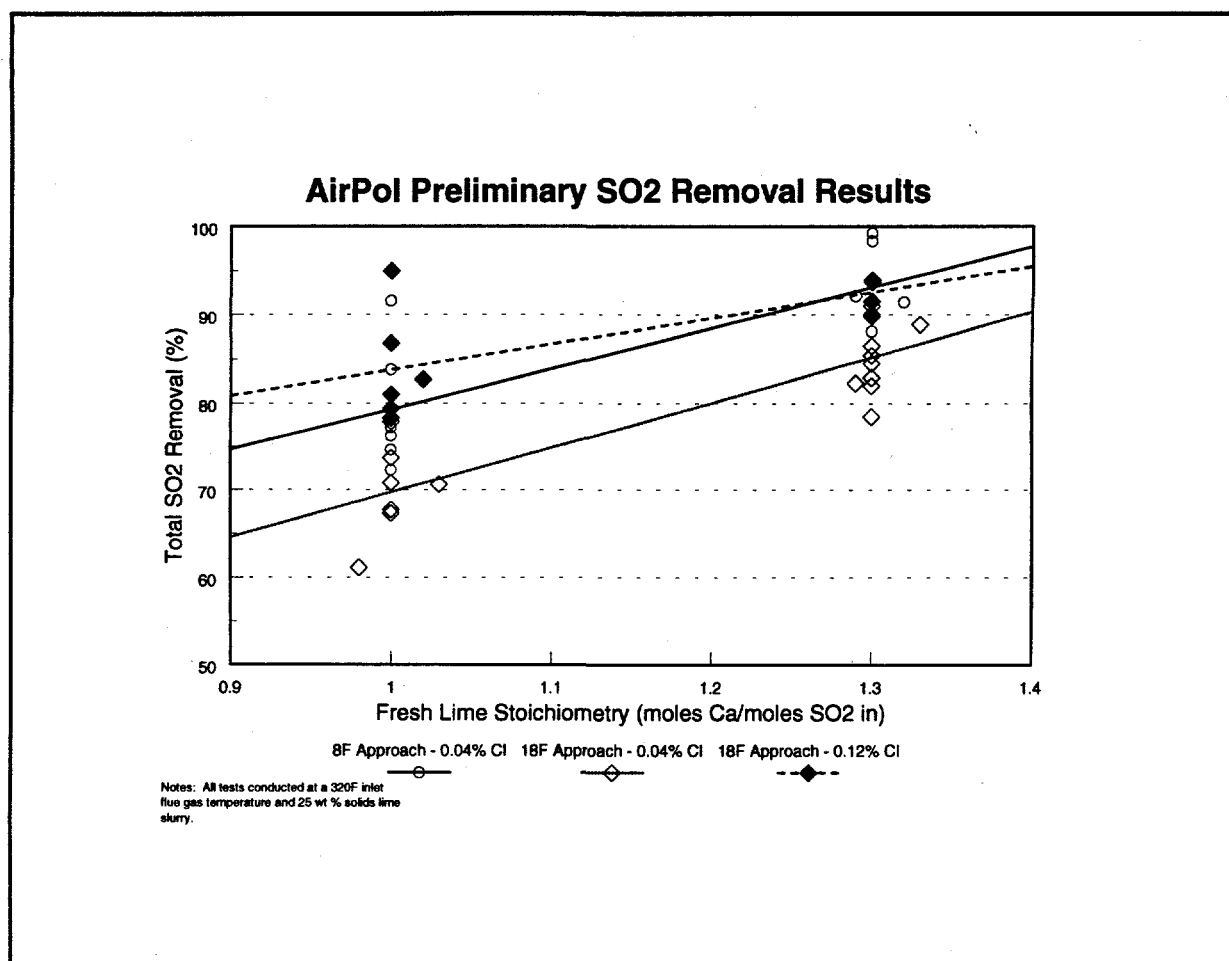
Although the preliminary chloride spiking tests had not been run at an approach-to-saturation temperature below 23°F, the decision was made to complete these chloride-spiking factorial tests at an 18°F approach-to-saturation temperature. There was some risk in this decision because the water evaporation rate decreases at the higher chloride levels. However, based on previous test work at the CER, the expectation was that at the lower chloride levels in this test plan, equivalent to a coal chloride level at 0.12%, the GSA system could operate at the 18°F approach-to-saturation temperature condition.

## IX. RESULTS OF FACTORIAL TESTING

### SO<sub>2</sub> Removal Efficiency

The overall system SO<sub>2</sub> removal efficiency results from these factorial tests have been analyzed, and several general relationships have become apparent. First, as was expected based on previous testing at the CER, significant positive effects on the SO<sub>2</sub> removal efficiency in the system came from increasing the lime stoichiometry and other factors such as increasing the coal chloride level or decreasing the approach-to-saturation temperature. Increasing the recycle rate resulted in higher SO<sub>2</sub> removal, but the benefit appeared to reach an optimum level, above which further increases in the recycle rate did not seem to have a significant effect on SO<sub>2</sub> removal. Increasing the flue gas flow rate had a negative effect on the SO<sub>2</sub> removal in the system.

The overall system SO<sub>2</sub> removal efficiency during these tests ranged from slightly more than 60% to nearly 95%, depending on the specific test conditions. The higher SO<sub>2</sub> removal efficiency levels were achieved at the closer approach-to-saturation temperatures (8 and 18°F), the higher lime stoichiometry level (1.30 moles Ca(OH)<sub>2</sub>/mole inlet SO<sub>2</sub>), and the higher coal chloride level (0.12%). The lower SO<sub>2</sub> removal efficiency levels were achieved at the higher approach-to-saturation temperature (28°F), the lower lime stoichiometry level (1.00 mole Ca(OH)<sub>2</sub>/mole inlet SO<sub>2</sub>), and the lower coal chloride level (0.02-0.04%). The data from these factorial tests completed at these conditions are shown in Figure 4. The slight scatter in the data in this figure is due to the variations



**Figure 4.** Overall System SO<sub>2</sub> Removal Results from the GSA Factorial Testing

in the other major process variables in these tests (i.e. flue gas flow rate, recycle screw speed, etc.). Most of the SO<sub>2</sub> removal in the GSA system occurs in the reactor/cyclone, with only about 2 to 5 percentage points of the overall system removal occurring in the ESP. There was substantially less SO<sub>2</sub> removal in the ESP than in previous testing at the CER, but the overall system SO<sub>2</sub> removal efficiencies appear to be comparable with the GSA process for most test conditions.

As one would expect, the lime stoichiometry level, which was tested at 1.00 and 1.30 moles  $\text{Ca}(\text{OH})_2$ /mole inlet  $\text{SO}_2$ , seems to have the most significant effect on the  $\text{SO}_2$  removal efficiency in the GSA system.

The approach-to-saturation temperature, which was evaluated at three levels of 8, 18, and 28°F for the low coal chloride conditions and the two levels of 18 and 28°F for the higher coal chloride condition, appears to be the second most important variable in the GSA system in terms of the overall system  $\text{SO}_2$  removal efficiency.

The third most important variable seems to be the chloride level in the system. Two coal chloride levels were tested, the baseline coal chloride level of 0.02 to 0.04% and the equivalent of a 0.12% coal chloride level. The higher chloride level was achieved by spiking the feed slurry with a calcium chloride solution.

One of the most surprising results of this factorial testing was the ability of the GSA system to operate at an 8°F approach-to saturation temperature at the low-chloride condition without any indication of plugging. This is even more impressive given the very low flue gas residence time in the reactor/cyclone. The second interesting result of this testing was the ability of the GSA system to operate at the 18°F approach-to-saturation temperature at the higher chloride level. In the preliminary testing at a much higher coal chloride level (0.3%), the lowest approach-to-saturation temperature tested was 23°F. No operating problems were encountered in the tests completed at the 0.12% coal chloride level and 18°F approach-to-saturation temperature conditions. In fact, the average moisture level in the solids remained below 1.0% in all of these factorial tests, even at the higher coal chloride level.

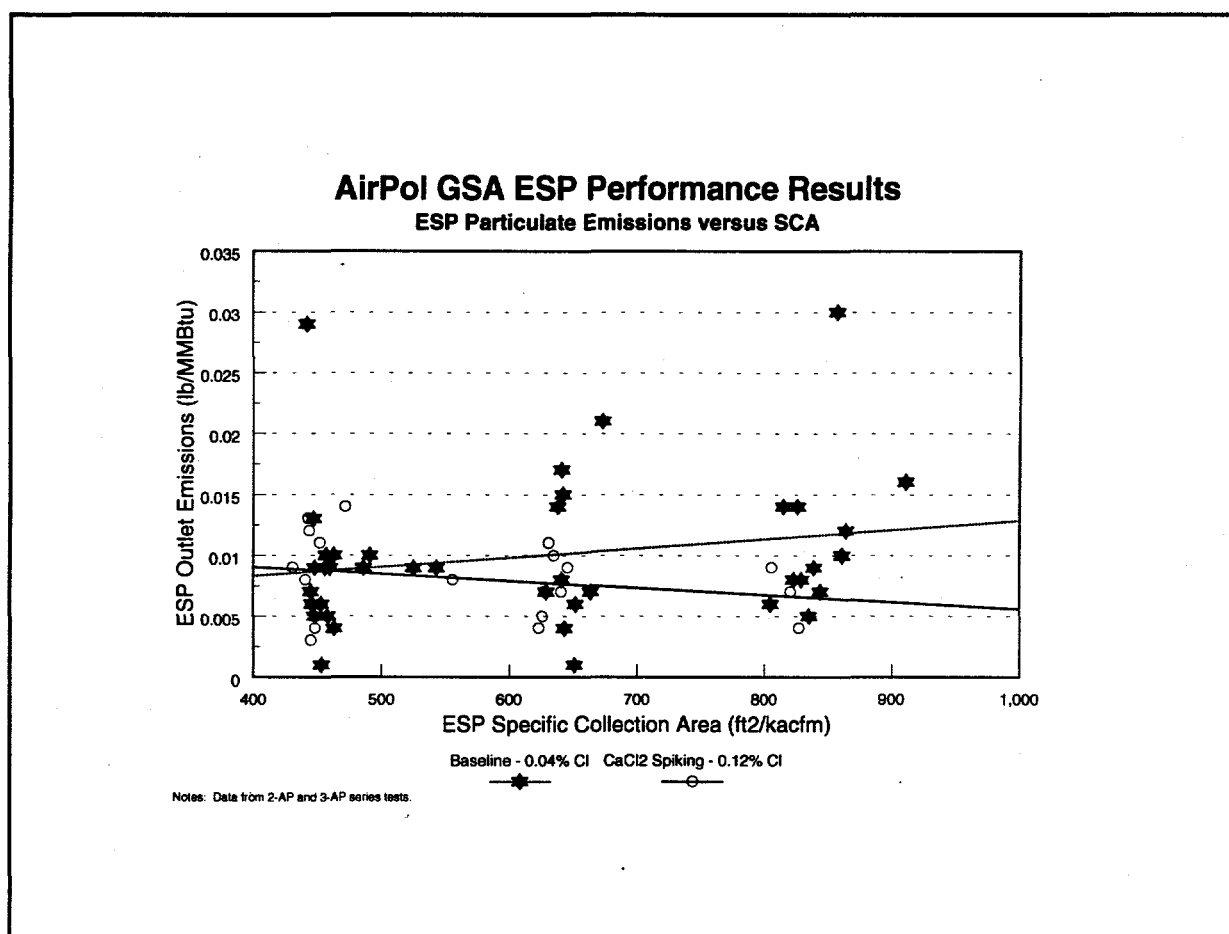
### **ESP Performance**

The ESP installed at the CER is a relatively modern, 4-field unit with 10-inch plate spacing, similar in design to several full-scale ESPs installed on the TVA Power System. This unit has 23-foot-high plates with 8 parallel gas passages. The specific collection area (SCA) of the unit is about 440  $\text{ft}^2/\text{kacfm}$  under the cooled, humidified flue gas conditions downstream of the reactor/cyclone. (For the untreated flue gas at 300°F, i.e., in a fly-ash-only application, the SCA of this ESP is about 360  $\text{ft}^2/\text{kacfm}$ .)

The particulate removal performance of this ESP was determined for each of the factorial tests, even though this was not the primary focus of the testing. The most important result of this particulate testing was that the emission rate from the ESP was substantially below the New Source Performance Standards (NSPS) for particulates (0.03 lb/MBtu) at all of the test conditions evaluated as shown in Figure 5. The typical emission rate was 0.010 lb/MBtu. The particulate removal efficiency in the ESP for nearly all of the tests was above 99.9% and the outlet grain loadings were below 0.005 gr/acf.

However, during the testing there were disturbing indications of low power levels in the first field of the ESP, particularly in those tests involving chloride spiking. In some of these chloride-spiking tests completed at the high flue gas flow rate (20,000 scfm), the power level in the first field was only about 5% of the normal level, effectively meaning that the first field had "collapsed." Even with these low power levels in the first field of the ESP, the particulate removal efficiencies were still 99.9+ percent and the emission rate was in the range of 0.010 lb/MBtu. The cause of these low power levels in the first field of the ESP is being investigated. These low power levels could be the result of a number of factors, including plate-wire alignment problems as observed in a recent internal inspection.

One surprising result of this ESP testing was that there was no significant improvement in the ESP performance with increasing SCA. For some of these tests, the SCA in the ESP approached 800 ft<sup>2</sup>/kacfm and the flue gas velocity in the ESP dropped below 2.0 ft/sec and yet the emission rate remained in the same range as in the other tests, i.e., 0.010 lb/MBtu.



**Figure 5. ESP Performance Results from the GSA Factorial Testing**

### **Pulse Jet Baghouse Performance**

Although not part of the original GSA project, TVA and EPRI had co-funded the installation of a 1-MWe PJBH pilot plant at the CER to be operated in conjunction with the existing GSA demonstration. Later, AirPol and DOE joined in the operation and testing of this PJBH pilot plant program. The PJBH pilot plant, which was started up in late January, can pull a slipstream of flue gas from either the ESP inlet or outlet, as shown in Figure 1. In the first series of factorial tests, the PJBH pilot plant pulled flue gas from the ESP inlet and, thus, treated flue gas with the full particulate loading (3 to 5 gr/acf) from the GSA reactor/cyclone. The inlet flue gas flow rate was about 5,000 acfm, which corresponds to an air-to-cloth ratio (A/C) of 4.0 acfm/ft<sup>2</sup> in the PJBH. During the second series of factorial tests, the PJBH pilot plant pulled flue gas from the ESP outlet. The same inlet flue gas flow rate was treated (5,000 acfm), but two-thirds of the bags were removed prior to this testing and thus, the A/C for these tests was 12 acfm/ft<sup>2</sup>.

The cleaning of the bags in the PJBH was pressure-drop-initiated during this testing with the cleaning cycle beginning whenever the tubesheet pressure drop reached 6 inches of water. The cleaning continued until the tubesheet pressure drop had declined to about 4-1/2 inches of water. The bags were cleaned by a low-pressure, high-volume, ambient air stream delivered by a rotating manifold.

### **SO<sub>2</sub> Removal Efficiency for Reactor/Cyclone/PJBH System**

The SO<sub>2</sub> removal efficiency in the reactor/cyclone/PJBH system was typically about 3-5 percentage points higher than that achieved in the reactor/cyclone/ESP system at the same test conditions. This higher SO<sub>2</sub> removal efficiency in the PJBH system was not unexpected given the intimate contact between the SO<sub>2</sub>-laden flue gas and the solids collected on the outside of the bags as the flue gas passed through the filter cake and the bags before being discharged to the stack. However, it should be noted that most of the SO<sub>2</sub> removal occurred in the reactor/cyclone; the PJBH SO<sub>2</sub> removal efficiency, based on the inlet SO<sub>2</sub> to the reactor, contributed less than 8 percentage points to the overall system SO<sub>2</sub> removal efficiency during this testing.

### **Particulate Removal**

The particulate removal efficiency in the PJBH was 99.9+ percent for all of the tests completed with the full dust loading from the GSA reactor/cyclone. The emission rate for all of these tests was well below the New Source Performance Standards for particulates and was typically in the range of 0.010 lb/MBtu.

## X. AIR TOXICS TESTING

The air toxics tests, which followed the factorial tests, were conducted during September and October, 1993. The objectives of these tests were to:

- Determine emissions and net removal efficiency of hydrogen chloride (HCl), hydrogen fluoride (HF), total particulate matter and trace metals. The trace metals included antimony (Sb), arsenic (As), barium (Ba), beryllium (Be), cadmium (Cd), chromium (Cr), cobalt (Co), lead (Pb), manganese (Mn), mercury (Hg), nickel (Ni), selenium (Se) and vanadium (V).
- Evaluate the impact of the particulate control device configuration (ESP alone, PJBH alone, or ESP plus PJBH in series) on final emissions of hazardous air pollutants.
- Compare the emissions of HCl, HF and trace metals with and without the injection of lime slurry.

The tests were conducted in two configurations, i.e. with the PJBH in series and parallel with the ESP. Two test conditions were evaluated for each configuration: baseline, with no lime introduction into the system; and demonstration, with lime slurry injection. Three simultaneous sampling runs were performed at each of the four permutations. The streams sampled are shown in Table 2.

Type of Sample	Location
Gaseous	GSA inlet, ESP inlet, ESP outlet, PJBH inlet and PJBH outlet
Aqueous	Lime slurry and trim water
Solid	Coal, GSA cyclone, ESP field 1, ESP field 2,3,4, PJBH hopper and re-injected fly-ash

**Table 2.** Sampling Locations For The Air Toxics Tests.

All of these tests were completed while the boiler was burning the high-sulfur (2.7%), low-chloride Andalex coal and were run at the high flue gas flow rate (20,000 scfm) and the high fly ash loading (2.0 gr/acf) test conditions. The baseline tests were performed at 270°F GSA reactor inlet temperature to protect the acrylic bags in the PJBH. The demonstration tests operated at 320°F GSA reactor inlet temperature, with a 12°F approach to saturation temperature at the GSA outlet.

## XI. RESULTS OF AIR TOXICS TESTING

Tables 3 and 4 present the removal efficiencies and uncertainties of the baseline and demonstration case with varying ESP and baghouse configurations. Removal efficiencies for beryllium and nickel were not determined due to analytical laboratory error. The removal efficiency for most trace metals



is generally over 90 percent. Caution is required when reviewing the removal efficiency of antimony, since most of the antimony measurements were below detection limits. Mercury concentration was also low. Only trace levels of mercury, i.e. close to the method detection limits, could be detected in the baseline and parallel tests. The removal efficiency for mercury appears to fall in the 50%-95% range.

The GSA/ESP arrangement indicated average removal efficiencies of greater than 99 percent for arsenic, barium, chromium, lead and vanadium. Removal efficiencies are significantly less than 99 percent for antimony, manganese, mercury and selenium. Lower removals for mercury and selenium are expected because of the volatility of these metals.

The GSA/PJBH configuration showed 99+ percent removal efficiencies for arsenic, barium, chromium, lead, manganese, selenium and vanadium. Cadmium removal was much lower with this arrangement than any of the other arrangements in both baseline and demonstration tests. Mercury removal efficiency for this arrangement was lower than with the GSA/ESP arrangement.

The removal of HCl and HF was dependent on the utilization of lime slurry and was relatively independent of particulate control device configuration. The removal efficiencies are greater than 98% and 96% for HCl and HF, respectively.

Parameter	GSA + ESP Series		GSA + ESP Parallel		GSA + FF Parallel		GSA + ESP + FF Series	
	Reported RE (%)	Total Uncertainty (%)	Reported RE (%)	Total Uncertainty (%)	Reported RE (%)	Total Uncertainty (%)	Reported RE (%)	Total Uncertainty (%)
Antimony	89.71	18.38	96.91	13.49	97.68	14.17	89.67	17.10
Arsenic	98.74	8.17	98.48	8.14	99.83	8.10	99.98	8.11
Barium	98.37	7.81	99.58	7.92	99.54	7.92	99.69	7.77
Cadmium	97.42	10.99	86.98	11.31	71.40	13.11	94.03	10.85
Chromium	99.09	8.63	98.14	9.24	99.46	9.23	99.65	8.47
Cobalt	98.38	9.55	98.24	9.52	98.68	9.51	98.66	9.51
Lead	98.79	9.47	97.36	9.16	99.51	9.16	99.69	9.35
Manganese	99.20	9.13	98.28	9.36	99.57	9.24	99.77	9.13
Mercury	79.15	38.24	66.38	11.71	31.97	527.49	94.45	14.26
Selenium	73.05	28.46	81.56	35.36	99.93	9.49	99.11	10.41
Vanadium	98.73	13.98	98.71	13.00	99.07	12.90	99.17	13.74
Particulate	99.59	9.70	99.52	4.16	99.86	4.16	99.90	9.70
HCl	---	---	---	---	7.71	478.82	-12.38	370.75
HF	---	---	---	---	22.08	488.02	-73.24	248.95

Table 3. Baseline Tests Removal Efficiencies and Uncertainties

Parameter	GSA + ESP Series		GSA + ESP Parallel		GSA + FF Parallel		GSA + ESP + FF Series	
	Reported RE (%)	Total Uncertainty (%)	Reported RE (%)	Total Uncertainty (%)	Reported RE (%)	Total Uncertainty (%)	Reported RE (%)	Total Uncertainty (%)
Antimony	84.72	37.99	98.78	14.24	98.65	14.20	95.01	18.59
Arsenic	99.96	8.37	96.36	47.79	99.98	8.24	99.99	8.37
Barium	99.63	8.80	92.72	90.19	99.49	9.44	99.74	8.81
Cadmium	98.68	10.77	93.27	64.71	78.63	20.31	97.37	11.73
Chromium	99.48	9.58	95.11	58.92	99.50	8.85	99.66	9.60
Cobalt	98.66	9.48	94.27	64.39	98.91	9.47	99.13	9.62
Lead	99.88	9.08	92.08	107.09	99.61	9.51	99.88	9.07
Manganese	92.44	33.45	95.58	53.68	99.13	10.18	99.87	9.67
Mercury	88.27	24.72	-38.89	1918.94	49.23	136.17	90.16	27.34
Selenium	76.87	88.86	99.81	10.34	99.80	10.32	99.96	10.18
Vanadium	99.18	13.87	93.37	75.62	99.00	12.50	99.46	13.90
Particulate	99.86	3.63	96.63	43.05	99.94	4.00	99.96	3.62
HCl	---	---	---	---	99.96	11.85	98.71	13.03
HF	---	---	---	---	96.82	14.67	98.99	12.81

Table 4. Demonstration Tests Removal Efficiencies and Uncertainties

## **XII. DEMONSTRATION RUN**

### **28-day GSA/ESP Demonstration Run**

The 28 day demonstration run, with GSA operating in conjunction with ESP only, started on October 25, 1993 and ended on November 24, 1993. This demonstration run began with the boiler burning the high-sulfur (2.7%), low-chloride Andalex coal and test conditions of: 320°F inlet flue gas temperature; 18°F approach-to-saturation temperature; 1.5 gr/acf fly ash injection; 0.12 percent coal chloride level; 20,000 scfm flue gas flow rate; and 30 rpm recycle screw speed. The SO<sub>2</sub> control mode was engaged for this run with an overall system SO<sub>2</sub> removal efficiency set-point of 91 percent. Due to some problems encountered in obtaining the test coal, a switch was made to burning a higher-sulfur (3.5%) coal for a period of time. The Ca/S ratio averaged 1.40 - 1.45 moles of Ca(OH)<sub>2</sub>/mole inlet SO<sub>2</sub> during this demonstration run.

The demonstration run showed that all three of the major objectives were successfully achieved.

- The overall system SO<sub>2</sub> removal efficiency averaged 90-91 percent, i.e., very close to the set-point. The switch to the higher-sulfur coal demonstrated the flexibility of the GSA system
- The particulate removal efficiency was good at an average of 99.9+ percent, with an emission rate below 0.015 lbs/MBtu.
- The GSA system demonstrated the reliability of this technology by remaining on-line for the entire 28-day period that the boiler was operating.

### **14-day PJBH Demonstration Run**

The purpose of the 14-day demonstration run was to demonstrate that the GSA system (reactor/cyclone/PJBH), as installed at the CER, could operate reliably and continuously, while simultaneously achieving 90+ percent SO<sub>2</sub> removal and maintaining the PJBH outlet emissions below the NSPS for particulates.

The specific design test conditions for this run were the same as those used for the previous 28-day GSA demonstration, except that the fly ash addition rate was reduced slightly from 1.5 to 1.0 gr/acf. This demonstration run was successfully completed in March 1994, and the following observations were made.

- The overall system (reactor/cyclone/PJBH) SO<sub>2</sub> removal efficiency averaged more than 96 percent during the entire 14-day demonstration run.

- The average Ca/S level during this run ranged from about 1.34 to 1.43 moles  $\text{Ca(OH)}_2$ /mole inlet  $\text{SO}_2$ .
- The PJBH particulate removal efficiency averaged 99.99+ percent. The emission rate was 0.001 to 0.003 lbs/MBtu.

### XIII. ECONOMIC EVALUATION

Under the scope of this project, Raytheon Engineers & Constructors prepared an economic evaluation of the GSA FGD process using the same design and economic premises that were used to evaluate about 30-35 other FGD processes for the Electric Power Research Institute. The relative process economics for the GSA system were evaluated for a moderately difficult retrofit to a 300-MW boiler burning a 2.6 percent sulfur coal. The design  $\text{SO}_2$  removal efficiency was 90 percent.

The resulting capital cost estimate (in 1990 dollars) is shown in Table 5 together with the estimate for the conventional wet limestone, forced-oxidation (WLFO) scrubbing system. The total capital requirement of \$149/kW for the GSA process is substantially lower than the \$216/kW for the WLFO system. The significant reduction in capital is primarily due to lower costs in the  $\text{SO}_2$  absorption area.

Total Capital Investment Comparison (1990 \$, 300-MW, 2.6% S coal)		
	\$/kW	
<u>Area</u>	<u>GSA</u>	<u>WLFO</u>
Reagent Feed	25	37
$\text{SO}_2$ Removal	38	71
Flue Gas Handling	18	24
Solids Handling	5	7
General Support	1	2
Additional Equipment	4	4
Total Process Capital	91	145
Total Capital Requirement	149	216

Table 5. Total Capital Investment Comparison

The levelized annual revenue requirements for the two processes (in 1990 dollars) are shown in Table 6. The levelized annual requirement for the GSA process is somewhat lower than that for the WLFO system. The principal operating cost for the GSA process is the cost of the pebble lime.

<b>LEVELIZED COSTS</b>		
<b>(300-MW, 2.6% S coal, 15-year levelizing)</b>		
	<b>Mills/kWh</b>	
<u>Fixed Costs</u>	<u>GSA</u>	<u>WLFO</u>
Operating Labor	0.52	0.66
Maintenance	1.49	1.74
Administrative and Support Labor	<u>0.34</u>	<u>0.41</u>
	2.35	2.81
<u>Variable Costs</u>		
Raw Material	1.82	0.65
Solids Disposal	0.86	0.57
Water	0.01	-
Steam	-	0.55
Electricity	<u>0.47</u>	<u>1.16</u>
	3.16	2.93
<u>Fixed Charge (Capital)</u>	<u>5.40</u>	<u>7.30</u>
Total	10.91	13.04

**Table 6. Levelized Costs**

#### **XIV. COMMERCIALIZATION**

One of the objectives of this demonstration project was for AirPol to establish its capability in designing, fabricating, and constructing the GSA system so that the demonstrated technology can be effectively commercialized for the benefit of the U.S. electric utility and industrial markets. The progress of this demonstration project matches very well with the development of the utility FGD market. The GSA technology is now being commercialized in order to meet the Phase II Clean Air Act Amendments (CAAA) compliance requirements.

During the course of designing the demonstration unit, an effort was made by AirPol to standardize the process design, equipment sizing, and detailed design so that the installation of a commercial unit can be accomplished within a relatively short time frame. Furthermore, equipment design was simplified, resulting in reduced material and construction costs. With the confidence that the GSA system is capable of achieving the required levels of performance, AirPol has developed a standard design of scale-up units.

The successful effort from the project has resulted in a commercial application in Ohio. AirPol has a GSA system for a 50 MWe municipal boiler burning Ohio coal as its first commercial utility installation in the United States. The state of Ohio, in conjunction with the Ohio Coal Development Office, awarded the city of Hamilton a grant to install a GSA system in the city's municipal power plant. In order to meet the requirements of the CAAA, it has been necessary to burn relatively expensive, low-sulfur coal in this plant. The installation of the GSA will allow the city to meet environmental regulations while using high-sulfur Ohio coal for power generation.

The pollution control equipment in existence at Hamilton was a hot-side electrostatic precipitator (ESP). This precipitator was undersized from inception, and never worked well. Several alternatives for this ESP were considered in connection with the installation of the GSA:

- (1) Install the GSA upstream of the ESP and extend the unit to attain sufficient capacity.
- (2) Use the ESP as primary dust collector upstream of the GSA with a new final dust collector
- (3) Demolish the ESP and replace it with the GSA and new final dust collector
- (4) Leave the ESP in place, de-energize it, and connect the GSA with final collector downstream

The fourth alternative was finally selected and the GSA was connected to the existing exhaust stack downstream of the ID fan. A long duct from the stack crosses a roadway and drops down and enters the GSA reactor. After passing through the reactor and cyclone, the flue gas enters a fabric filter of the pulse-jet type and continues to a new ID fan that returns the cleaned flue gas to the exhaust stack just above the point where it left to enter the GSA.

A lime preparation system adjacent to the GSA with lime silo, slurry tank with agitator, and a slurry pump produces a lime slurry of 20% concentration that is pumped to the reactor. The by-product collected in the fabric filter is gathered in screw conveyors and transported pneumatically to a by-product silo, from where it is removed by truck to landfill.

The technical data for the Hamilton installation are as follows:

Boiler Capacity	50 MWe
Type of Boiler	Pulverized Coal
Type of Coal	Ohio 3% + sulfur
Gas Volume	224,728 ACFM
Gas Temperature	320 °F
Moisture Content	7.8 % by Volume
Oxygen Content	4.6 % by Volume
Particulate Content	2 gr/SCFD
SO <sub>2</sub> Content	2,612 PPMd
SO <sub>2</sub> Removal	90 % Design

Another GSA installation for a fossil fuel boiler is being installed in Kaohsiung, Taiwan. The installation is in a sugar refinery where two oil fired boilers each have a dedicated GSA. The larger boiler has a steam generating capacity of 100 TPH, while the smaller one generates 35 TPH.

Both GSA units are equipped with fabric filters, dual ID fans, and a gas recirculating system. The reason for the dual fans and the recirculation is that both boilers have great load swings, and in order to attain the required SO<sub>2</sub> removal efficiency, the GSA must run with at least 50% of design gas volume. When the boiler capacity is reduced below the design capacity, a portion of the flue gas is recirculated via a separate fan from the outlet of the fabric filter to the inlet of the GSA reactor to maintain minimum design gas flow.

Both systems operate with hydrated lime, and calcium chloride is added from a storage tank in order to enhance the acid gas absorption. Due to the fact that the oil firing generates minimal amounts of particulates, by-product from the by-product silo is returned to the reactor to create particulate nuclei for lime slurry.

The cleaned flue gases from the two systems enter an existing masonry exhaust stack. Before the gases reach the stack they pass a steam heated coil that increases the gas temperature to reduce the visible steam plume from the stack.

The technical data for the two plants are as follows:

Boiler Steam Rating	100 TPH	35 TPH
Type of Fuel	Oil #6	Oil #6
Gas Volume	97,554 ACFM	27,748 ACFM
Gas Temperature	298 °F	280 °F
Moisture Content	13.36 % by Volume	3.5% by Volume
Oxygen Content	1.96 % by Volume	1.74 %by Volume
Particulate Content	2.4 gr/SCFD	2.3 gr/SCFD
SO <sub>2</sub> Content	510 PPMd	517 PPMd
SO <sub>2</sub> Removal	80 % Design	80 % Design



In addition to the Hamilton and Taiwan installations, approximately 20 GSA plants for refuse and hazardous waste incineration are in operation, most of them in Europe. Some of these installations have very sophisticated control equipment for NO<sub>x</sub>, furans, and dioxins with extremely low outlet concentrations.

#### **XV.      DISCLAIMER**

Reference in this report to any specific commercial product, process, or service is to facilitate understanding and does not necessarily imply its endorsement or favoring by either DOE or TVA.

# **SMALL, MODULAR, LOW-COST COAL-FIRED POWER PLANTS FOR THE INTERNATIONAL MARKET**

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## **ABSTRACT**

This paper presents recent operating results of Coal Tech's second generation, air cooled, slagging coal combustor, and its application to power plants in the 1 to 20 MW range. This 20 MMBtu/hour combustor was installed in a new demonstration plant in Philadelphia, PA in 1995. It contains the combustion components of a 1 MWe coal fired power plant, a 17,500 lb/hour steam boiler, coal storage and feed components, and stack gas cleanup components. The plant's design incorporates improvements resulting from 2000 hours of testing between 1987 and 1993 on a first generation, commercial scale, air cooled combustor of equal thermal rating. Since operations began in early 1996, a total of 51 days of testing have been successfully completed. Major results include durability of the combustor's refractory wall, excellent combustion with high ash concentration in the fuel, removal of 95% to 100% of the slag in the combustor, very little ash deposition in the boiler, major reduction of in-plant parasitic power, and simplified power system control through the use of modular designs of sub-systems and computer control. Rapid fuel switching between oil, gas, and coal and turndown of up to a factor of three was accomplished. All these features have been incorporated in advanced coal fired plant designs in the 1 to 20 MWe range. Incremental capital costs are only \$100 to \$200/kW higher than comparable rated gas or oil fired steam generating systems. Most of its components and subsystems can be factory assembled for very rapid field installation. The low capital, low operating costs, fuel flexibility, and compatibility with very high ash fuels, make this power system very attractive in regions of the world having domestic supplies of these fuels.

## **I. INTRODUCTION**

This paper updates the results of work performed on Coal Tech's commercial scale 20 MMBtu/hour air cooled, slagging coal combustor since the last report at the 1995 Clean Coal Conference [1]. During the past year, a second generation, 20 MMBtu/hr combustor has been placed in operation in a coal combustion system. It incorporates all the features of Coal Tech's new low power cost, solid fuel plant. The central feature of this plant is an air cooling combustor whose wall heat transfer loss

is recuperated to the combustion air, making this heat available to the thermodynamic cycle. A portion of the  $\text{SO}_2$  and  $\text{NO}_x$  emissions are controlled inside the combustor, which is designed for new and retrofit boiler applications. Coal Tech's development of the air cooled combustor began in the late 1970's in a 1 MMBtu/hr air cooled, cyclone combustor [2], continued in the mid 1980's in a 7 MMBtu/hr water cooled, cyclone combustor [3], and was followed by 2000 hours of operation of a first generation, 20 MMBtu/hr, air cooled combustor between 1987 and 1994 [4-7]. The latter facility was located in an industrial heating plant in Williamsport, PA. Fuels tested include coal, coal water slurry, refuse derived fuel, oil, and gas. Test operations to 1991 were sponsored in part by the United States Department of Energy Clean Coal Technology Program [4].

Subsequent testing under another DOE sponsored project began in 1992 [5-7]. The first phase focused on improving combustor durability and combustor operation under automatic computer control. Several hundred hours of operation over a 7 month period in 1993 were implemented without requiring any internal refurbishment of the combustor walls.

The second phase of this project began in 1994 and is currently in progress. The results of prior testing were incorporated in the design of a new coal fired power plant using a second generation combustor rated at 20 MMBtu/hr and capable of generating up to 1 MW of electric power. The combustion parts of the plant were fabricated and installed in 1995 at an industrial site in Philadelphia, PA. The subsystems of the plant were designed to take advantage of the unique features of the air cooled combustor. This includes an oil design flat bottom boiler that was modified for real time removal of any ash or slag carried over from the air cooled cyclone combustor. It also includes a coal processing system that produces coarsely pulverized coal (50% passing 100 mesh compared to previous operation at 70% passing 200 mesh). This greatly reduces the capital and operating cost of the coal handling system. All the auxiliary subsystems, such as combustor cooling and combustion air supply, fuel supply, and cooling circuits were modularized to reduce capital cost and operating and maintenance costs. As part of this latter effort, the power requirements for the 20 MMBtu/hr combustor were reduced by two-thirds compared to the prior unit. Some features of this new plant were described at the Clean Coal Conference in Denver, CO in 1995 [1].

Test operations began in early 1996, and to date 51 days of testing have been completed. Results have substantially exceeded design performance. For example, the amount of bottom fly ash deposits in the boiler has been so low that its real time removal has not yet been necessary.

This paper summarizes the recent test results and discusses the use of this new design for low cost power plants in the 1 to 20 MWe range. This power system is especially attractive in regions with local deposits of high ash coals.

### **Coal Tech's Advanced Air Cooled, Cyclone Coal Combustor**

The cyclone combustor is a high temperature ( $> 3000^\circ\text{F}$ ) device in which a high velocity swirling gas is used to burn crushed or pulverized coal. Figure 1 shows a schematic of Coal Tech's patented, air cooled combustor. Gas and oil burners rapidly preheat the combustor and boiler during startup. Pulverized coal and powdered sorbent for  $\text{SO}_2$  control are injected into the combustor in an annular

region enclosing the gas/oil burners. Air cooling is accomplished by flowing combustion air through tubes on the outside of a ceramic liner in the combustor. This cooling air provides over 90% of the combustion air in the combustor, and it is introduced tangentially in a swirling manner into an annulus enclosing the fuel injection cylinder in the combustor, (see figure 1). The ash and reacted sorbent melt on the liner and the resultant slag is drained through a tap at the downstream end of the combustor.

Nitrogen oxide emissions are reduced by operating the combustor in a fuel rich mode, with final combustion taking place in the boiler. Operations in the first generation 20 MMBtu/hr combustor in Williamsport yielded, under optimum conditions, about two-thirds  $\text{NO}_x$  reductions to 0.26 lb of gas/MMBtu, or 200 ppm (at 3 %  $\text{O}_2$ ) at about 70% of stoichiometric air/fuel ratio in the combustor and high combustion efficiencies. The stoichiometric ratio for the combustor/boiler was between 1.25 and 1.5. Sulfur emissions are controlled primarily by sorbent injection into the combustor. Measurement of  $\text{SO}_2$  levels at the stack gas outlet from this previous boiler yielded average  $\text{SO}_2$  reductions of 50% to 70%, and as high as 85%, with calcium hydrate injected into combustor at Ca/S mol ratios of 3 to 4. Particulate emissions were controlled in part by slag retention in the combustor. It was augmented with a wet particle scrubber which reduced the particle emissions to as low as 0.26 lb of solids /MMBtu.

## **II. THE SECOND GENERATION, 20 MMBTU/HR COMBUSTOR/BOILER PLANT IN PHILADELPHIA, PA**

The design of this plant was based on the results of tests in the 20 MMBtu/hr air cooled combustor in Williamsport, PA, and on various site specific combustor applications studies for power plants in the 1 to 20 MW range that were performed in the past several years [6,7].

It was originally planned to install the 20 MMBtu/hr combustor/boiler at the new site with an atmospheric back pressure turbine to generate about 500 kW of power from the 17,500 lb/hr, 250 psig boiler. Sale of this power would partially defray the cost of more extensive durability tests on the combustor/boiler system. However, due to excellent progress this year in the combustor test effort, it was decided to eliminate the power generation step and proceed to commercial introduction of the technology.

To meet the particle emission standard for Philadelphia, a baghouse was required in place of the wet particle scrubber that was used in Williamsport. The latter's best performance resulted in a particle emission of 0.26 lb/MMBtu, which was below the Williamsport standard of 0.4 lb/MMBtu. The Philadelphia standard is 0.06 lb/MMBtu. The manufacturer of the baghouse has stated that particle emissions of less than 0.03 lb/MMBtu can be readily achieved under the operating conditions existing in the present facility.

Figure 2 is a side view of the new 20 MMBtu/hr combustor/boiler installation in Philadelphia. PA. Its total size is such that it can be shipped by tractor trailer to any site. Figure 3 shows a plan and side view of the Philadelphia facility. It includes provision for a 25 ton raw coal delivery and storage

area, a low cost coal mill, a 4 ton pulverized coal storage bin, sorbent storage bins, pneumatic coal and sorbent delivery, a boiler, the combustor and its auxiliary subsystems, specifically, water cooling, oil, gas, combustion air, cooling air, compressed air, slag removal, and the stack system, including the baghouse, and induced draft fan. The entire system is controlled by programmable logic controllers (PLC) and computer process control. Performance parameters are measured and recorded on a computer. Combustion gases,  $O_2$ ,  $CO$ ,  $NO_x$ , and  $SO_2$ , are measured in the boiler radiant furnace section, boiler stack outlet, and baghouse outlet. Novel Features and Operating Experience of the Second Generation Combustor/Boiler Facility.

The facility was designed to include the major features that will be incorporated in Coal Tech future commercial power plants in the 1 to 20 MWe range. Therefore, the primary design objective was to minimize capital, operating and maintenance costs.

Capital cost is minimized by factory assembly of major subsystems of the plant. Oil/gas designed boilers are compatible with the air cooled, coal combustor. These boilers are factory assembled for thermal ratings of up to 200 MMBtu/hr. Air cooled combustors can be fabricated up to 150 MMBtu/hr. The combustor's auxiliary subsystems are assembled in modules and attached to the combustor support structure. Therefore, the combustor and boiler can be shipped from the factory in two modules.

Another important capital cost saving results from the fuel flexibility and rapid shift among the various fuels. This sharply reduces the need for on site fuel storage.

Air cooling operation was much improved in the present combustor to the point where gas and oil fuel consumption for heatup and cooldown of the combustor was reduced by about a factor of two from the quantities used in the Williamsport combustor. Another major result of the improved air cooling was a factor of two reduction in the cooling fan power requirement. In addition, the quantity of compressed air flow required to operate the facility was sharply reduced. Finally, the use of a baghouse in place of the wet particle scrubber sharply reduced the induced stack fan power. As a result, the total power used in the Philadelphia plant was reduced to one-third of the level required in the first generation Williamsport facility.

The improved combustor operation reduced the combustion gas temperature at the boiler outlet an average of 100°F to 150°F for identical coal firing and sootblowing conditions in the previous combustor. Additional cooling of the stack gases was added to allow the use of substantially lower cost bags for the baghouse and to further reduce the stack fan power.

The combustor is a higher maintenance component than the boiler. It is, therefore, essential to minimize downtime when it requires refurbishment. Consequently, the current combustor design allows its removal from all its auxiliary sub-systems and from the boiler in less than 1 day.

A high maintenance item has been the combustor's slag tap assembly, primarily during the initial tests. Subsequent modifications were made which have sharply reduced maintenance to this item. The relay controlled system used in the previous combustor system was replaced with programmable logic controllers (PLC). The PLC assure that the combustor's fuel supply and the boiler's steam

supply operate with all safety interlocks functioning. The previous computer process control software was upgraded to account for the changes in the design of the present combustor. As the test effort proceeded, it was found that the combustor could be controlled with a much simpler procedure than was used for the previous combustor, and the software was changed accordingly.

With these improvements, the personnel needed to operate the facility was reduced from an average of six used in the Williamsport facility to two or three, depending on the specific test objectives.. Based on this experience, it is anticipated that a fully commercial plant can be operated with substantially fewer personnel than are used in a conventional coal fired plant.

### **20 MMBtu/hr Combustor Operation in the Philadelphia Facility**

As soon as the present combustor was placed into operation, its exhibited performance was far superior to the earlier unit. Areas of improvement include combustion efficiency, slag retention, wall materials durability, and length of heatup and cooldown.

Slag retention, which is a key measure of slagging combustor performance, improved substantially. In the earlier 20 MMBtu/hr combustor, only one-half to two-thirds of the injected coal ash and sorbent minerals was converted to slag in the combustor. The balance of the ash and sorbent was blown out of the combustor as dry fly ash. Furthermore, over one-half of the slag formed in the combustor flowed out of the exit nozzle to the boiler floor, thus limiting the run time of the combustor. Although provision has been made to remove slag carryover from the combustor to the boiler by installing a combustor/boiler transition section, in the operations to date, the amount of slag carried over from the combustor to the boiler ranged from 0% to 5% of the total slag. Slag retention was also substantially better than before, averaging two-thirds of the injected mineral matter, which includes coal ash and sorbents.

Combustor refractory liner durability is another major performance parameter. Chemical reactions between the liquid slag and the combustor refractory wall can rapidly deplete the latter. However, by control of the combustor wall temperature, a layer of frozen slag can form on the combustor's refractory wall which maintains the integrity of the wall. Much progress had been made in perfecting this wall replenishment technique in the earlier 20 MMBtu/hr combustor. Replenishment of the refractory liner by injection of fly ash with the coal and sorbent proved to be very effective in the earlier combustor. In the present combustor, the combustor wall replenishment procedure has been further improved. Consequently, it has not been necessary to reline the combustor wall with refractory in the operations to date.

The cooling and combustion air distribution and control scheme was substantially modified for the present combustor in order to simplify the combustion and combustor wall cooling process. The new scheme has proven to be much simpler to control, and the need for the previous complicated computer control has been eliminated.

To minimize nitrogen oxide emissions it is necessary to operate the combustor under fuel rich conditions. Final combustion occurs in the furnace section of the boiler where the CO and H<sub>2</sub> rich combustor gas exhaust is mixed with additional air to complete combustion. Optimum NO<sub>x</sub> reduction occurs at about 70% stoichiometric air/fuel ratio in the combustor. [3,4]. However, operation of the earlier 20 MMBtu/hr combustor at this condition resulted substantially reduced combustion efficiency [3,4].

The three methods of measuring combustion efficiency in the slagging combustor are based on carbon in the slag, CO in the stack gases, and carbon in the stack fly ash. Under fuel rich conditions, significant amounts of carbon in the slag indicates poor combustion inside the combustor. In the present combustor, combustion efficiency, based on carbon in the slag, has been over 99% in almost all the tests including at fuel rich operation as low as 75% stoichiometric air/fuel ratio. Since carbon monoxide is an air pollutant, it is essential that it be minimized in the combustion process. The CO concentration in the stack was generally in the 200 ppm range which corresponds to better than 99% combustion efficiency.

Both these measurements of combustion efficiency do not account for unburned carbon that is carried over to the stack baghouse. Due to the difficulty in obtaining real time sampling of the baghouse fly ash, the carbon content in the fly ash was determined from random grab samples taken from all the ash collected on the day of testing. The carbon content of the ash ranged from 20% to 50% (dry basis). Since on average about one-third of mineral matter injected reported to the baghouse, one can compute the conversion of the solid carbon in the coal to CO<sub>2</sub> and CO in the combustor from the amount of unburned carbon in the baghouse fly ash. This yielded a carbon conversion greater than 90% for most of the tests. In several tests small quantities of fly ash in the stack were collected in a filter. Analysis of the carbon content in one of these tests yielded a carbon conversion of 94%.

The stoichiometric ratio in the combustor (SR1) ranged from fuel rich to fuel lean ( $0.75 < SR1 < 1.1$ ). Final combustion air was added at the combustor outlet into the boiler which yielded a stoichiometric ratio in the boiler furnace (SR2) in the range from 1.3 to 1.8.

Several bituminous coals were tested having higher heating values (HHV) in the range of 12,000 to 13,700 Btu/lb, ash contents in the 11% to 15% range, and sulfur contents in the range from 1.18% to 3.7%. The bulk of the tests were performed with 3+% sulfur coal.

The initial test effort this year has been focused on overall combustor performance, with lesser emphasis on SO<sub>2</sub> and NO<sub>x</sub> control. The most recent tests have focused on SO<sub>2</sub> control and excellent results have been achieved, especially in low sulfur coal. The analysis of these data is incomplete, and the results will be presented at the Conference. Both limestone and hydrated lime were injected into the combustor for slag conditioning and sulfur removal. Previous results in the 20 MMBtu/hr combustor in Williamsport showed that limestone was much less effective than calcium hydrate for sulfur capture. With calcium hydrate injection into the previous combustor, excellent sulfur capture results were achieved, where a maximum reduction in the 85% range was measured [5,7].

The degree of sulfur capture in the present combustor was found to be very sensitive to combustion conditions, the method and quantity of sorbent injection, and the mineral matter injection rate. In recent tests at high slag mass flow rates firing 3+% sulfur coal, SO<sub>2</sub> reductions measured in the end wall of the boiler furnace and in the boiler gas outlet at the stack were in the range of 60% to 75% at Ca/S mol ratios of under 3. Similar reductions were measured with injection of calcium hydrate into the boiler furnace near the combustor gas inlet to the boiler. In this case, the Ca/S mol ratios were in the range of 3.5 to 4.9. However, in the latter case, a substantial amount of the hydrate fell to the floor of the boiler furnace. Therefore, the Ca/S mol ratio is not an accurate measure of calcium utilization in this case.

In very recent tests with 1.5% sulfur coal, the SO<sub>2</sub> reductions were substantially higher. Reductions in the range of 75% to as high as 95% were measured. When expressed in lb/MMBtu, SO<sub>2</sub> emissions as low as 0.22 lb/MMBtu were measured. This is well below the 0.5 lb/MMBtu SO<sub>2</sub> emission standard for Philadelphia, and it near the 0.2 lb/MMBtu that is one of the current test objectives. The reductions were higher at the end wall of the boiler furnace than in the stack at the outlet of the boiler. No conclusive explanation for this behavior has been found. It is suspected the higher SO<sub>2</sub> at the boiler outlet may due to blowby of combustion gases through gaps in between the boiler tubes on the convective tube side of the boiler. This reduces the reaction time of sorbent with combustion gas in the boiler furnace. This matter should be clarified when all the data are analyzed.

One interesting result has been finding relatively high sulfur concentrations (10% to 20%) in the slag in several of the tests. With high slag mass flow rates, as obtained with high ash coals or by injecting additional ash, it may be possible to encapsulate all the coal sulfur in the slag. Tests in which additional metal oxide powder was injected into the combustor at up to 40% injected mineral mass flow rates have been very recently completed. The analysis of the resultant slags is not yet complete.

NO<sub>x</sub> emissions are controlled by operating the combustor fuel rich. Maximum reductions to as low as 0.26 lb/MMBtu were measured in the previous combustor at stoichiometric ratio (SR1) in the range of 0.7. In the present combustor, the tests were performed at less fuel rich conditions. At slightly fuel rich conditions, the NO<sub>x</sub> ( reported as NO<sub>2</sub>) has been in the range of 0.36 to 0.7 lb/MMBtu. The 0.36 lb/MMBtu value was measured at SR1 equal to 0.9.

The above discussion presents a general overview of the performance of the present second generation 20 MMBtu/hr combustor. Its performance has been found to be superior to the previous combustor, especially in the areas of combustion efficiency, slag retention and removal from inside the combustor, durability of the internal combustor wall, and simplicity of control and operation. Considerable data on SO<sub>2</sub> on NO<sub>x</sub> control have been measured, but the analysis is not complete. Based on the results obtained to date, the best results achieved in the current combustor (namely SO<sub>2</sub> reductions to 0.22 lb/MMBtu and NO<sub>x</sub> reduction to 0.36 lb/MMBtu) match those measured in the prior unit (SO<sub>2</sub> reductions to 0.34 lb/MMBtu and NO<sub>x</sub> reduction to 0.26 lb/MMBtu).



## II. APPLICATION OF THE AIR COOLED SLAGGING COMBUSTOR TO 1 TO 20 MWe POWER PLANTS

The present second generation combustor facility was designed as a prototype for a low cost, modular, coal fired commercial power plant. As such, considerable attention was given to incorporate novel designs for all the components of the present facility in order to minimize cost. Key elements in the plant that result in cost saving are:

- The air cooled, coal combustor is compatible with compact boilers designed for oil firing, which have about one-half the volume of conventional coal fired boilers.
- Modular design of the combustor and its auxiliary subsystems.
- Fuel flexibility and rapid switching between fuels, which minimizes the need for on-site fuel storage, especially bulky coal storage.
- Optimization of stack gas particulate control system to minimize baghouse cost, and stack fan power.
- Combustion of coarsely pulverized coal to allow use of low cost coal mills.
- Automated microprocessor control of the plant.

All these factors were incorporated in the design of the present facility in Philadelphia. The best way of demonstrating the cost impact of a power plant based on Coal Tech's combustion system technology is to compute the incremental cost of this plant versus a comparable size gas or oil fired plant. The additional components/subsystems required by the Coal Tech plant are coal storage, processing, and feed systems, the air cooled combustor, the stack gas cleanup system, additional fans and blowers, and additional controls. All these items use Coal Tech designs that optimize performance and cost. The gas and oil fuel components and storage are used only for startup, cooldown and emergencies. Therefore, their thermal rating and fuel consumption is only a few percent of the total energy used by the coal fired plant. In computing the added cost of this coal plant versus a conventional oil or gas plant, the incremental cost of the oil/gas burners and oil/gas storage and delivery components in the latter plant must be subtracted from the cost of the coal system. All power generation components, including the steam loop, generators, electric power distribution, such as turbine-generator, electric power distribution, are common to both plants.

A series of cost estimates were developed for several thermal ratings ranging from 20 MMBtu/hr to 125 MMBtu/hr. One combustor at 125 MMBtu/hr can produce about 10 MWe. Two combustors at this rating are attached to one boiler yield 20 MWe. For a steam generating plant only, the incremental cost of this coal plant over a conventional oil/gas plant is in the range of \$10 to \$25/lb/hr of steam, with the cost decreasing as the thermal rating increases. For a power plant, this incremental cost is in the \$100 to \$200/kW range.

The following example shows the economic benefit of this system. For a 125 MMBtu/hr coal fired steam plant with this combustor an incremental cost of \$14/lb/hr of steam is obtained. Applying this to 7000 hour/year operation, a coal-oil or gas cost differential of \$0.85/MMBtu, a 10%-90% equity-debt ratio, and a 4 year amortization, one obtains an internal rate of return (IRR) of 43%. For the smallest plants, a higher coal-oil/gas differential is required to yield a similarly high IRR. The low cost of this coal fired power technology allows great flexibility in achieving excellent rates of return.

This technology is especially attractive for the international market in regions with domestic coal reserves but no domestic oil or gas reserves. In this case additional factors enter in the power plant analysis, such as import restrictions on clean fuels. The compatibility of this combustor with high ash fuels which are readily found in many international markets, further adds to the attractiveness of this technology. In the first generation combustor, operations with additional ash injection to 50% mineral matter was successfully implemented [5,7]. In the current combustor the total mineral matter injection rate has to date been as much as 40% of the total solid fuel injected.

#### **IV. CONCLUSIONS**

The results of the effort to date on the second generation 20 MMBtu/hr air cooled, slagging coal combustor facility have confirmed the performance and economic benefits of this technology. Very rapid progress in the test effort since the facility became operational at the beginning of 1996 have accelerated its commercial development schedule.

The compact and modular design of the plant in the 1 to 20 Mwe range allows factory fabrication and assembly of its subsystems and shipment of the modules to the site for rapid assembly. These features make it attractive for steam and power generation at industrial sites in the US and overseas.

#### **V. ACKNOWLEDGMENTS**

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Figure 1  
Coal Tech's Air Cooled Combustor

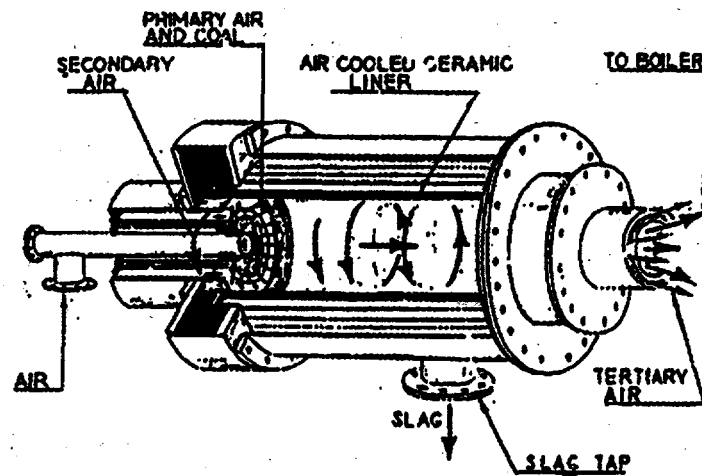
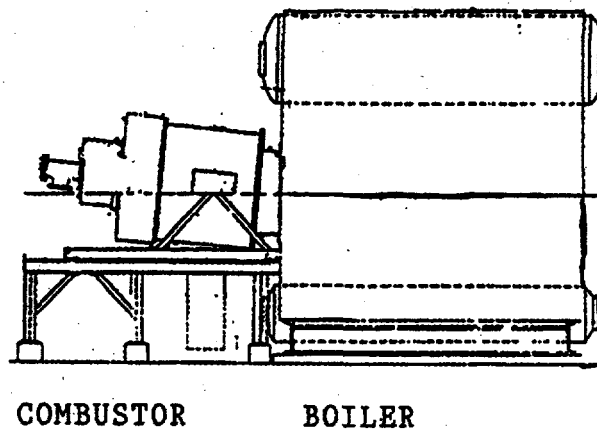


Figure 2:  
20 MMBtu/hr Combustor-Boiler Installation at  
Philadelphia Plant



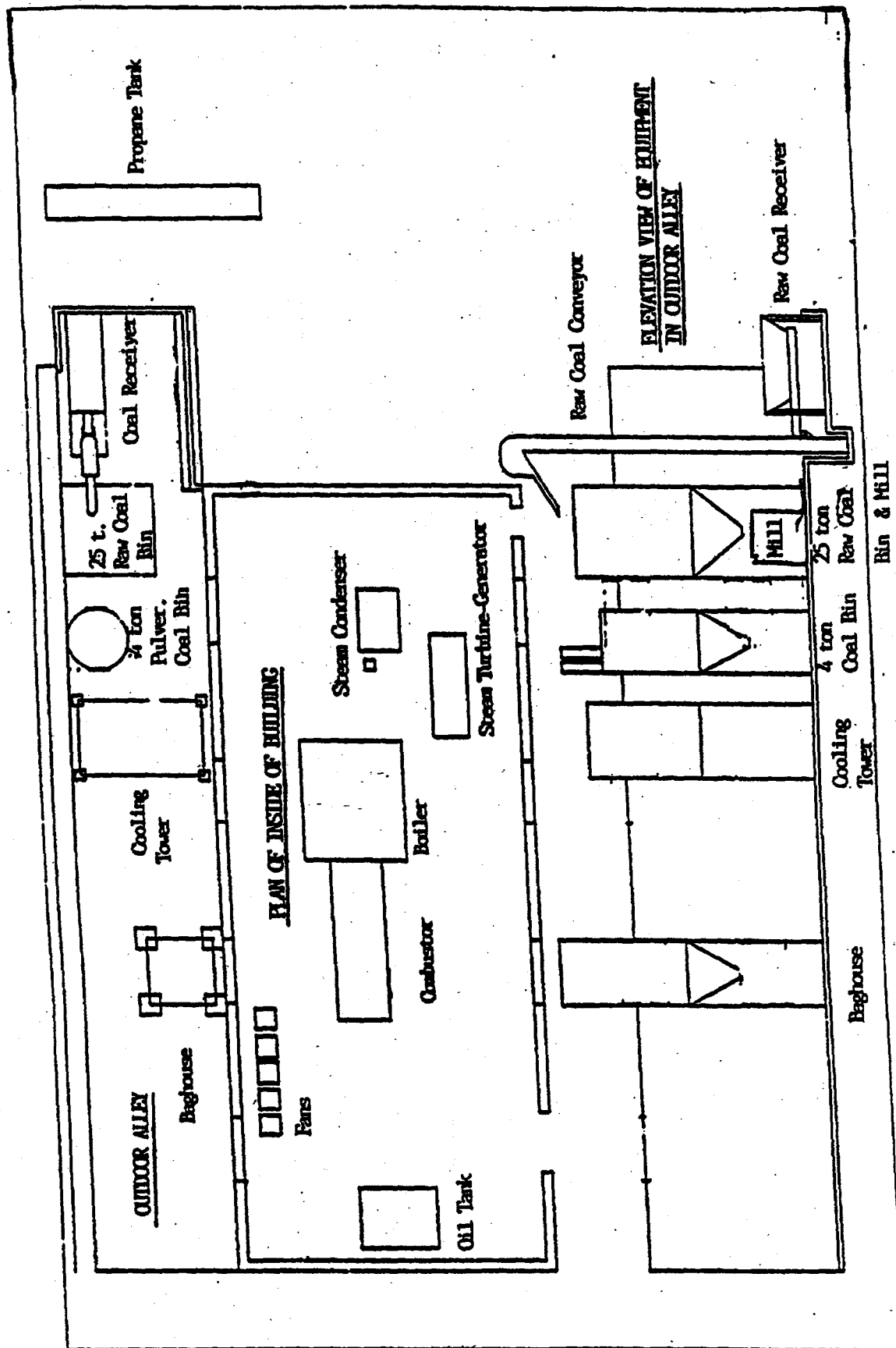


Figure 3: Plot Plan of the 20 MMBtu/hr Combustor-Boiler Test Site in Philadelphia.

# **CHIYODA THOROUGHbred CT-121 CLEAN COAL PROJECT AT GEORGIA POWER'S PLANT YATES**

## ***PHASE II RESULTS***

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### **ABSTRACT**

The Chiyoda Thoroughbred CT-121 flue gas desulfurization (FGD) process at Georgia Power's Plant Yates completed a two year demonstration of its capabilities in late 1994 under both high- and low-particulate loading conditions. This \$43 million demonstration was co-funded by Southern Company, the Electric Power Research Institute and the DOE under the auspices of the U.S. Department of Energy's Round II Innovative Clean Coal Technology (ICCT) program.

The focus of the Yates Project was to demonstrate several cost-saving modifications to Chiyoda's already efficient CT-121 process. These modifications included: the extensive use of fiberglass reinforced plastics (FRP) in the construction of the scrubber vessel and other associated vessels, the elimination of flue gas reheat through the use of an FRP wet chimney, and reliable operation without a spare absorber module.

This paper will focus on the testing results from the last trimester of the second phase of testing (high-ash loading). Specifically, operation under elevated ash loading conditions, the effects of low- and high-sulfur coal, air toxics verification testing results and unexpected improvements in byproduct gypsum quality are discussed.

### **I. INTRODUCTION**

The demonstration at Georgia Power's Plant Yates involved the retrofit construction of a CT-121 wet-limestone scrubber to an existing 100 MW pulverized coal-fired boiler. The principle difference between the CT-121 process and more common spray tower-type FGD systems is the use of a single process vessel, Chiyoda's patented Jet Bubbling Reactor® (JBR), in place of the usual spray tower/reaction tank/thickener arrangement. Initial startup of the process occurred in October 1992, and the demonstration project was completed in December 1994. Process operation continues with the CT-121 scrubber as an integral part of the site's Phase I Clean Air Act compliance plan.

Several of the latest evaluations that comprised the CT-121 demonstration project are discussed in this paper. In the last trimester of testing the CT-121 process was operated under moderate-ash inlet loading conditions while process reliability and availability were continuously evaluated. Additionally, exceptional concurrent particulate removal efficiencies were measured under moderate-particulate loading conditions, which was consistent with particulate removal efficiencies observed in earlier measurements under both high- and low-particulate loading conditions.

Parametric testing was also conducted under moderate-ash loading conditions while burning both high- and low-sulfur coals. The data gathered were regressed and multi-variable regression models were developed to provide an accurate prediction of the scrubber's SO<sub>2</sub> removal efficiency under the most likely future operating conditions. As part of the moderate-particulate removal evaluation, limited air toxics measurements were also performed for the second time. The purpose of this additional testing was to evaluate air toxics removal across the CT-121 under elevated ash loading conditions as well as to validate or controvert the findings of an earlier air toxics testing effort that was sponsored by DOE in June of 1993<sup>1</sup>.

A brief discussion of findings on the properties of the gypsum stack (not contaminated with flyash) following one year of dormancy is also included in this paper. An analysis of the chloride content showed that chloride levels in the gypsum decreased over time without any specific action by the project team. This finding increases the possible uses of the unwashed gypsum produced by this process. An indicator of public acceptance was the granting of a Plant Food License to Georgia Power for the non-ash gypsum at Plant Yates, by the State of Georgia's Department of Agriculture in October of 1996.

In general, the Yates CT-121 process performed well, exhibiting excellent SO<sub>2</sub> removal efficiency, particulate removal and consistent reliability. In addition to these successes, several possible process improvements were identified during the demonstration that could improve future designs of an already superior process.

## **II. FACILITY AND OPERATING DESCRIPTION**

The Yates plant site is comprised of seven coal-fired boilers, all Phase I affected units, with a total rated capacity of 1,250 MW. Plant Yates' 100 MW Unit 1 is the source of flue gas for the CT-121 process. All of the flue gas from Unit 1 is treated by the CT-121 wet FGD process with no provision for flue gas bypass. During the low flyash phase of parametric testing in 1992 and 1993, the existing ESP for Unit 1 was used for particulate control. The design efficiency for this ESP is 98%. In March, 1994, the ESP was fully deenergized at the start of high-particulate parametric testing, and partially energized to a target efficiency of 90% between June 1994 and November 1994.

A simplified site diagram for the Yates CT-121 retrofit is presented in Figure 1.

The scrubber demonstration facility equipment can be divided into five major subsystems:

- Boiler / ESP
- CT-121 scrubber / wet chimney
- Limestone preparation circuit
- Byproduct gypsum stacking area
- Process control system.

The central feature of the process is Chiyoda's unique absorber design, called a Jet Bubbling Reactor (JBR), which combines concurrent chemical reactions of limestone dissolution,  $\text{SO}_2$  absorption/neutralization, sulfite oxidation, gypsum precipitation and gypsum crystal growth together in one vessel. A cut-away view of the JBR is illustrated in Figure 2. Since much of the undesirable crystal attrition and secondary nucleation associated with the large centrifugal pumps in conventional FGD systems is eliminated in the CT-121 design, large easily dewatered gypsum crystals are consistently produced. This design also significantly reduces the potential for gypsum scale growth, a problem that frequently occurs in natural-oxidation FGD systems.

In the Yates installation (Figure 1), the flue gas enters the scrubber's inlet gas cooling section down-stream of the boiler's induced draft (I.D.) fan. This fan also serve as the scrubber's booster fan. Here the flue gas is cooled and saturated with a mixture of pond water and JBR slurry. From the gas cooling section, the flue gas enters an enclosed plenum chamber in the JBR formed by the upper deck plate and lower deck plate. Sparger tube openings in the floor of the inlet plenum force the inlet flue gas below the level of the slurry reservoir in the jet bubbling zone (froth zone) of the JBR as shown in Figure 3. After bubbling through the slurry, where all the concurrent reactions occur, the gas flows upward through large gas riser tubes that bypass the inlet plenum. Entrained liquor in the cleaned gas disengages in a second plenum above the upper deck plate due to a drastic velocity reduction and the cleaned gas passes to the 2-stage, chevron-style, horizontal-flow mist eliminator, then on to a wet FRP chimney.

"A closed-circuit, wet ball mill limestone preparation system is used to grind raw (3/4x0) limestone. The particle size of the ground limestone is small enough (90% passing a #200 mesh screen) to ensure that it is dissolved easily and that the amount of unreacted limestone in the JBR can be minimized or eliminated.

The JBR slurry reservoir provides about 36 hours of solid-phase residence time, depending on the  $\text{SO}_2$  pick-up rate. The slurry from the JBR is pumped intermittently to a gypsum slurry transfer tank (GSTT) for JBR slurry level control and slurry density control. In the GSTT, the slurry is diluted for pumping to a Hypalon®-lined gypsum (or gypsum/ash) stacking area for gravity dewatering and storage. Gypsum stacking is a disposal technique that involves filling a diked area with slurry for gravity sedimentation. Over time, this area fills with settled solids. The filled area is then partially excavated to increase the height of the containment dikes. The repetitive cycle of



sedimentation, excavation, and raising of perimeter dikes continues on a regular basis during the active life of the stack. Process water is naturally decanted, stored in a surge pond and then returned to the CT-121 process. There is no blowdown or discharge from the Yates CT-121 process.

During normal operation of the FGD system, the amount of  $\text{SO}_2$  removed from the flue gas is controlled by varying the JBR pressure drop ( $\Delta P$ ) or slurry pH. However, changing  $\Delta P$  is easier and quicker to respond to changing conditions since it is done by adjusting the JBR liquid level. Higher liquid levels result in increased  $\text{SO}_2$  removal because of increased contact time between the incoming flue gas and the scrubbing slurry. The pH can also be varied to affect  $\text{SO}_2$  removal with higher pH resulting in increased removal efficiency. Boiler load and flue gas  $\text{SO}_2$  concentration also affect removal efficiency, but are less controllable.

One of the most unique aspects of the CT-121 installation at Plant Yates is the wide use of fiberglass reinforced plastics (FRP) in several of the vessels to avoid the traditional corrosion damage associated with closed-loop FGD systems. Two of the vessels (the JBR and the limestone slurry storage tank) were constructed on site since their large size precluded shipment. The JBR inlet transition duct, where the flue gas is cooled prior to contacting the sparger tubes as a wet-dry interface, is also made completely of FRP. The inlet transition was discovered to be an area susceptible to erosion during high ash testing but homogeneous appliqué filler materials, Duromar® and Duromix®, now offer robust protection to exposed FRP surfaces at Plant Yates. A distinct advantage of the FRP construction was that it eliminated the need for a flue gas prescrubber, traditionally included in flue gas scrubber systems to remove chlorides that cause significant corrosion in alloys (fiberglass is mostly unaffected by inorganic acid attack and chlorides).

### **III. PROJECT OBJECTIVES**

To evaluate the effectiveness of the Yates CT-121 design advances, the following test objectives of the two year demonstration program were established:

- Demonstrate long-term reliable operation of the CT-121 FGD system;
- Evaluate particulate removal efficiency of the JBR and system operation at normal and elevated particulate loadings;
- Correlate the effects of pH and JBR gas-side pressure drop ( $\Delta P$ ) on system performance;
- Correlate the effect of limestone grind on system performance;
- Evaluate the impact of boiler load on system performance;
- Evaluate the effects of alternate fuels and reagents on system performance;
- Evaluate equipment performance and construction material reliability; and
- Monitor solids properties, gypsum stack operation and possible impacts of the gypsum stack on ground water.

Many of these objectives were investigated during this last trimester of the second phase of the

demonstration project, also known as the High-Particulate Auxiliary Test block. Two of the test periods in this test block provided data relevant to the focus of this paper:

- High-Particulate Alternate Coal Tests which evaluated scrubber performance under elevated particulate loading conditions while burning high-sulfur (3.4%) coal;
- High-Particulate Alternate Limestone Tests which evaluated an alternate limestone reagent source, while under elevated particulate loading, burning low-sulfur coal (1.2 % S).

Particulate and air toxics removal testing were also conducted during the Alternate Limestone testing. The data from the parametric portion of this test period was regressed to develop a predictive performance model for the conditions at which the testing was conducted, since these conditions are the most likely scenario for post-demonstration operation

#### **IV. RESULTS**

The CT-121 scrubber at Plant Yates continued to prove itself a very viable and cost effective technology for use in Clean Air Act, Title IV compliance. It exhibited excellent availability, maintained greater than 97% limestone utilization, and demonstrated the ability to exceed 98% SO<sub>2</sub> removal efficiency with high sulfur coal, while at maximum boiler load. The flexibility of the CT-121 process was also demonstrated through the use of a wide range of coals, varying from 1.2% to 4.3% sulfur content.

##### **Operating Statistics**

The duration of the demonstration, including the startup and shake-down phase, was 27 months, or approximately 19,000 hours. The low-particulate test phase (including shake-down) consisted of 11,750 hours, during which time the scrubber was operated for 8,600 hours. The remaining 7,250 hours of the demonstration included 5,510 hours of operation at elevated particulate loading. Complete operating statistics for the entire demonstration project are detailed in Table 1. The "high-ash" test period actually consisted of a high-ash loading period (during the Parametric Test block) in which the ESP was completely deenergized, and a moderate-ash loading period (during the Long-Term and Auxiliary Test blocks) during which the ESP was partially de-energized to simulate a more realistic scenario: a CT-121 retrofit to a boiler with a marginally performing particulate collection device. The moderate-ash loading condition resulted in better availability than did the high-ash loading condition.

	Low-Ash Test Phase	High-Ash Test Phase	Demonstration Project Duration (Cumulative)
Total Hours in Test Period	11,750	7,250	19,000
Scrubber Available Hours	11,430	6,310	18,340
Scrubber Operating Hours	8,600	5,210	13,810
Scrubber Called Upon Hours	8,800	5,490	14,290
Reliability <sup>1</sup>	0.98	0.95	0.96
Availability <sup>2</sup>	0.97	0.95	0.97
Utilization <sup>3</sup>	0.73	0.72	0.75

1. Reliability = Hours scrubber operated divided by the hours called upon to operate.

2. Availability = Hours scrubber available divided by the total hours in the period.

3. Utilization = Hours scrubber operated divided by the total hours in the period.

**Table 1. Summary of Operating Statistics**

### **Effect of Inlet SO<sub>2</sub> Concentration**

The SO<sub>2</sub> removal efficiency of the scrubber was measured under five different inlet SO<sub>2</sub> concentration ranges; three during this most recent testing. The coal burned by Unit 1 for a majority of the testing was a blend of Illinois No.5 and No.6 bituminous coal that averaged 2.4% sulfur (as burned), except for a brief, unplanned period when 3.0% sulfur coal was burned. A 4.3% sulfur bituminous coal was burned during the Low-Particulate Alternate Coal Test block, and a 3.8% sulfur coal was burned for the High-Particulate Alternate Coal Test block. The High-Particulate Alternate Limestone Test (last test of the demonstration project) coincided with Plant Yates' compliance-driven transition to a low sulfur coal (approximately 1.2% S). This provided the scrubber project an opportunity to evaluate a fifth coal source.

The effect of inlet SO<sub>2</sub> concentration on SO<sub>2</sub> removal efficiency is quite significant. Figure 4 illustrates the decrease in SO<sub>2</sub> removal as inlet SO<sub>2</sub> concentration increased for the coal sources evaluated. Performance of the scrubber was outstanding during the low-sulfur coal burn. It should be noted that the low-sulfur coal tested limited the JBR pH to a maximum of 3.8 because of Aluminum-Fluoride-inhibited limestone dissolution (Al-F blinding). The Al-F blinding stems from the low-ionic strength of the scrubbing liquor, the elevated ash loading to the JBR and the coal trace metals concentrations. A maximum operating pH of 3.75 was chosen to ensure that near-complete limestone utilization was maintained in the scrubber. The test data from 1000 ppm (inlet SO<sub>2</sub> concentration) operations indicates that SO<sub>2</sub> removal efficiency did not decline at a slightly lower pH.

The evaluation of five different inlet SO<sub>2</sub> concentrations demonstrates the flexibility of the CT-121 process as well as its exceptional SO<sub>2</sub> removal capability, even when burning fuels with a very high sulfur content. This is even more impressive considering that the maximum designed sulfur

content for the demonstration unit was only 3.0%, and that this limit was exceeded by 43% in one test period. Other test data shows that even higher SO<sub>2</sub> removal efficiencies are achievable at higher pH values.

### **Particulate Removal Efficiency**

Because of the torturous path taken by the flue gas during treatment in the JBR, an effort was made to quantify particulate removal. Consequently, the ability of the CT-121 process to remove flyash particulate was evaluated several times throughout the demonstration. Particulate loading measurements were made at the inlet and outlet of the scrubber under three different conditions of inlet mass loading, summarized in Table 2. The discussion here will focus on the particulate removal capabilities of the scrubber under only the moderate-ash loading conditions.

<b>Condition</b>	<b>ESP Energization</b>	<b>ESP Collection Rate</b>	<b>ESP Outlet-JBR Inlet JBR Inlet Mass Loading (lb/MMBtu)</b>
<i>1</i>	Full	High	<i>Low (0.02-0.10)</i>
<i>2</i>	Partial	Moderate	<i>Moderate (0.20-0.50)</i>
<i>3</i>	Off	Low	<i>High (5.00-5.50)</i>

**Table 2. ESP Configuration during Particulate Testing**

Measurements of particulate removal across the JBR (Condition 2, Table 2) were made near the minimum and maximum nominal boiler loads (50 and 100 MW), and at low and high JBR  $\Delta P$  settings (10 and 18 in.WC). The test conditions and results are shown in Table 3. As shown in Table 3, at all tested inlet particulate loadings, boiler loads, and JBR pressure drops the JBR exhibited excellent particulate removal efficiency, ranging from 97.7% to 99.3%.

Although the outlet particulate loading varied from 0.005 to 0.029 lb/MMBtu, analytical results indicate that from 20 to 80 percent of outlet particulate is sulfate (SO<sub>4</sub>). Based on the calcium analyses performed on the same material, it is believed that the measured sulfate originated from gypsum carryover and acid mist carryover, so it is scrubber-generated. This finding reduces the estimate of actual ash mass loading at the outlet of the scrubber (actual fugitive emissions) to approximately 70% of the amount captured, measured and recorded during outlet testing.

Test I.D.	Approximate ESP Efficiency (%)	JBR $\Delta P$ (in. WC)	Boiler Load (MW)	JBR Inlet Mass Loading (lb/MMBtu)	JBR Outlet Mass Loading <sup>1,2</sup> (lb/MMBtu)	JBR Removal Efficiency (%)
AL2-1	90	18	100	1.288	0.029	97.7
AL2-2	90	10	100	1.392	0.010	99.3
AL2-3	90	18	50	0.325	0.005	98.5
AL2-4	90	10	50	0.303	0.006	98.0

<sup>1</sup> Federal U.S. NSPS is 0.03 lb/MMBtu for units for which construction began after 9/18/78

<sup>2</sup> Plant Yates Unit 1's permitted emission limit for existing units is 0.24lb/MMBtu (40% opacity)

**Table 3. Particulate Removal Testing - Summary of Results**

### **Particulate Removal Efficiency by Particle Size**

The particle size distribution of the scrubber inlet and outlet particulate matter was measured at all four test conditions as shown in Table 3. The results of these analyses indicate that excellent particulate removal efficiency occurred in most of the measured size ranges (cut-points). Figure 5 illustrates the particulate removal efficiency of the scrubber by comparing inlet and outlet mass loading at different particle size cut-points (shown using a logarithmic scale). The inlet data were combined for both 50 MW tests and for both 100 MW tests to simplify the plots since inlet conditions were identical in each case.

As observed in the plots, the 100 MW case showed better particulate removal efficiency than the 50 MW case at most cut-points. One possible explanation is based on the mechanism of particulate removal in the scrubber. Because the velocity of flue gas is higher at higher loads, the particulate has more momentum and is more likely to come into contact with the wet/dry interface as each flue gas "bubble" rises through the slurry.

As was reported during earlier particulate removal tests, and again observed in Figure 5, the best removal efficiencies were observed for particle sizes greater than 10 $\mu$ m. At all test conditions, there was greater than 99% particulate removal efficiency of particles in this size bin. In some cases, efficiency exceeded 99.99%. As the particle size decreased, there was a drop in observed particulate removal efficiency, but over 90% efficiency was observed at all particle sizes between 1 $\mu$ m and 10 $\mu$ m. Between 0.5 $\mu$ m and 1 $\mu$ m, the particulate removal dropped to sometimes negligible values. In this range, it is believed that acid mist carryover offset the ash particulate removal, resulting in poor particulate removal values. Analyses of the outlet catch indicated that an average of 30% of the outlet particulate can be attributed to gypsum and acid mist carryover. Below about 0.5 $\mu$ m, the particulate removal efficiency increased to above 90%. Also observed in Figure 7 was a higher particulate removal efficiency at the higher JBR  $\Delta P$  values. This increase in removal efficiency ranged from 1 decade (90%), at the largest particle sized, to less than 1/10th of

a decade (10%) at the 0.5 $\mu$ m cut-point. The increased particulate removal at the higher JBR  $\Delta$ P in this size range results from a deeper sparger tube submergence depth and therefore, a longer gas-phase residence time allowing more opportunity for the particulate to be captured in the slurry.

## V. AIR TOXICS TESTING

The Yates CT-121 ICCT Project had two opportunities to measure its air toxics removal potential (also referred to as HAP or hazardous air pollutants). In 1993, Yates was chosen by the DOE as one of its eight coal-fired sites for an air toxics study<sup>1</sup> conducted on EPA's behalf in support of Clean Air Act Title II requirements for subsequent health risk determinations. In late 1994, the Yates ICCT Project expanded its scope of work to duplicate portions of that 1993 effort, in an attempt to validate the DOE's 1993 results. The results are both interesting and mutually supportive. However, the fossil fuel sources between the two tests were radically different and an exact comparison of results can not be easily made.

In 1993, the DOE was hoping to investigate three issues;

- Air toxics characterizations/penetrations in fossil fuel systems (fuel/boiler/ESP);
- Air toxics removal potential for postcombustion equipment (ESP/wet scrubber);
- Air toxics emissions factors in lb/10<sup>12</sup> BTU.

From the 1993 results, the DOE concluded that:

- As much as 99% of the HAPs of interest are in the particulate phase;
- Specie removal across the ESP was proportional to total particulate removal;
- Uncertainty was high because most measurements were near the minimum analytical detection limits;
- Special difficulties were encountered with selenium, mercury and chromium (Cr<sup>6+</sup>).

The 1994 air toxics sampling conducted as part of the ICCT Project by Radian at Plant Yates was performed to address the technical difficulties encountered during the 1993 tests; specifically:

- Selenium sampling and analysis;
- Mercury partitioning and speciation;
- Flyash penetration of the FGD process; and
- Source apportionment (origin of exiting and particulate matter),

as well as to be able to compare emissions and removals from a radically different coal source within the same boiler/ESP/scrubber flue gas pathway.

In comparing the results of the two efforts from a macro-perspective, several observations

emerge that may effect the use of air toxics data in further rulemaking and health risk determinations:

- The 1993 effort saw significantly more measurement error than the 1994 effort;
- The Chiyoda CT-121 JBR is highly efficient at HAP removal;
- Sampling is very sensitive to ANY error (e.g.: Contamination) at these near-minimum detection level measurements; and
- Source apportionment identifies a significant emission contribution from particulate generated within the wet scrubbing process.

The uncertainty in the 1994 testing data is generally lower than that of the 1993 testing data (i.e., sampling procedures improved). Secondly, due to the larger uncertainty evident in some species in 1993, the accuracy of any calculated emission factors would likewise be suspect. It is apparent from the data that some species can be measured with much lower uncertainty than others. Fairly low uncertainty were found for arsenic, vanadium, and lead. Conversely, antimony, chromium, manganese, and nickel all had unacceptably large measurement confidence intervals, sometimes the confidence interval was 10 times larger than the measurement itself. Calculated removal efficiencies from the 1994 tests are shown in Figure 6. It is prudent to remind ourselves that extrapolation of admittedly uncertain data does not lend itself to producing certain results for emission factor estimation or subsequent health effects determinations. Caution should be emphasized in the use of these and any similar air toxics measurement data.

## VI. GYPSUM QUALITY

The gypsum stacking area at Plant Yates had three separate cells for segregated impoundment; a "clean" gypsum stack area, a gypsum/ flyash stack area, and a recycle water pond. During Phase I (the low-ash test phase) of this demonstration project, the "clean" gypsum stack was used to dewater and store the pure-gypsum byproduct; decanted clear process water was collected in the common pond area and returned to the process. There was no blowdown, discharge or water treatment of scrubber process water. During the high-ash test phase (Phase II), the segregated gypsum/ash area was used for stacking the ash/gypsum mixture. Since these stacks are physically separated "cells", the original "clean" gypsum stack then, sat idle during the later ash/gypsum phase of testing.

The gypsum slurry deposited in the both areas was originally with a high chloride content, due to the closed loop nature of the scrubber's operation, with liquid phase chloride concentrations calculated to be as high as 35,000 ppm at equilibrium. Because of these high chloride concentrations, any slurry-deposited gypsum solids would normally require washing in order to satisfy requirements of the gypsum wallboard or cement manufacturing industries. Core samples of the "clean" stack that were taken after the stack had been idle for over a year indicated a surprising result: the chloride concentration in the gypsum had decreased from about 6000 ppm, measured 3 months after Phase I completion, to less than 50 ppm less than one year later. Table 4 presents chloride data for the gypsum stack.

There are two likely reasons for this decrease in chloride concentration in the gypsum in the "clean" gypsum stack. The first is that the rainfall that occurred over the idle year washed the gypsum and decreased the chloride concentration. The rate of chloride decrease over time, or as a function of rainfall, was not measured because this was an unplanned (and at the time, unknown) benefit of the gypsum stacking technique. The other reason lies in the fact that a majority of the chloride content in the gypsum solids is due to the chlorides in the water entrained in the gypsum solids. Core samples from the gypsum stack typically indicated that the solids content was approximately 83 wt.% on average shortly after the stack was idled. After one year, the solids content had increased to an average of 90 wt.% at a depth of 3 feet. Although this decrease in entrained water played some role in decreasing the chloride concentration in the gypsum, it is likely that rainwater washing of the stack was the predominant cause of the decrease in chloride concentration. This is further evidenced by the data presented in Table 4 that shows free moisture did not decrease at the 6 foot level, although chloride concentration did.

Of interesting note, was the 1996 issuance of a Plant Food Permit to Georgia Power that will allow the unrestricted sale of ash-free gypsum from the Yates Project to meet the unfilled demand for agricultural gypsum of 1 million+ ton/year in Georgia alone.

Dike	Inactive Period	Sample Depth (ft)	Chloride (ppm)	Moisture (%)
West	>90 days	4	930	16.0
	>90 days	8	7610	17.5
	>90 days	9.5	5720	17.7
	>90 days	14.5	5540	15.1
	>400 days	1	60	8.1
	>400 days	3	40	9.2
	>400 days	6	20	12.0
	>90 days	10	5740	14.5
	>90 days	13.5	5610	17.4
South	>90 days	16.5	6710	17.4
	>400 days	1	20	8.0
	>400 days	3	20	11.0
	>400 days	6	20	18.3

Table 4. Chloride and Moisture Levels in "Clean" Gypsum Stack decline over time

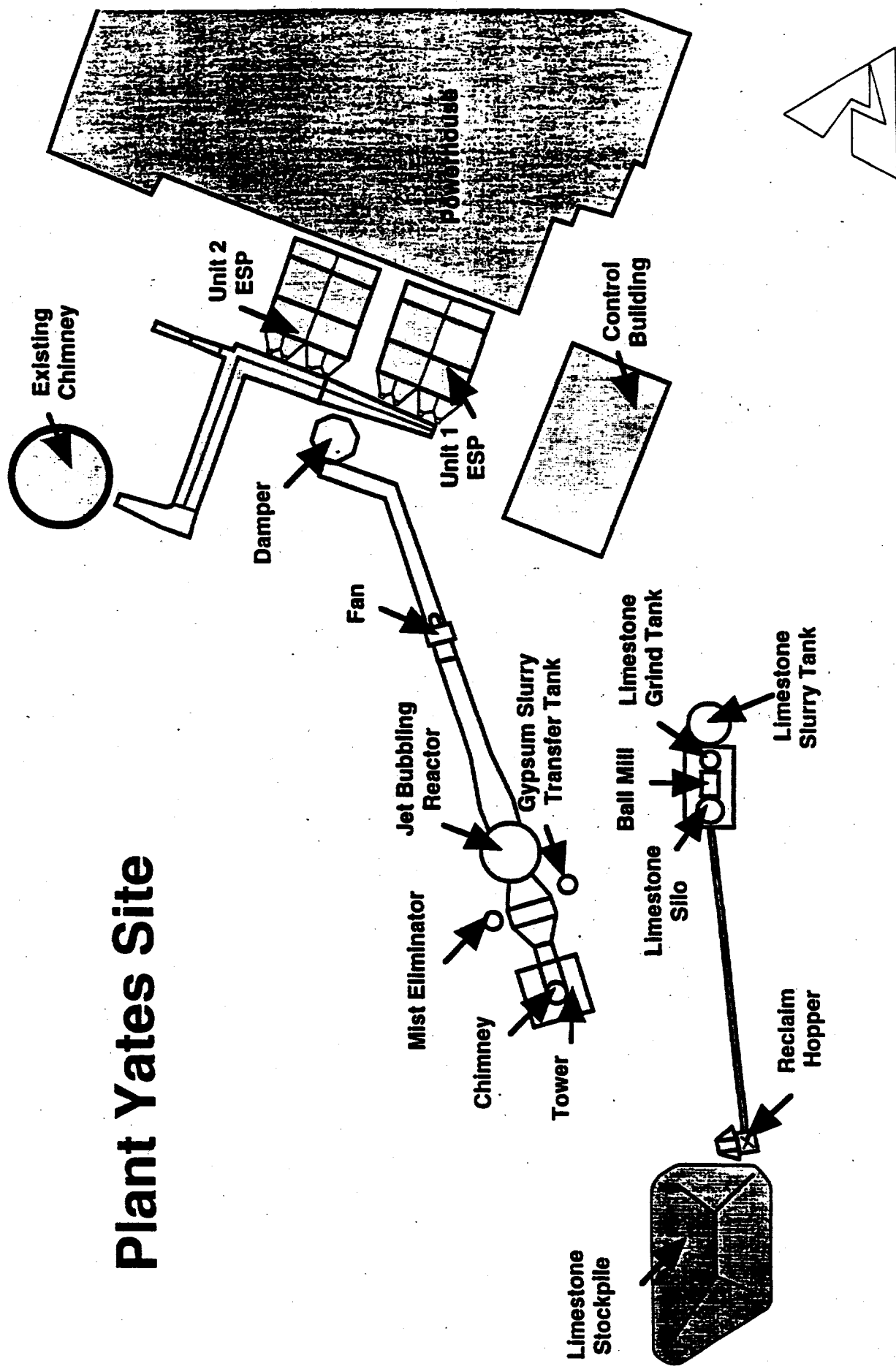


## VIII. SUMMARY

Chiyoda's CT-121 FGD process was very successfully tested at conditions far beyond design expectations. From an operating standpoint, the process was reliable, showed consistently high removals ( $\text{SO}_2$ , particulate, air toxics), was energy efficient and reagent efficient. From a chemical engineering standpoint, the mass transfer interactions are robust and resilient, only limited at conditions far beyond design parameters. This would allow a designer / operator to install a cost effective CT-121 system that would give consistently excellent service, even in periods of difficult operating conditions.

1. "A study of Toxic Emissions from a Coal-Fired Power Plant Utilizing an ESP While Demonstrating the ICCT CT-121 FGD Project," Radian Corporation, Final Report for U.S. DOE, Contract No., DE-AC22-93PC93253. June 16, 1994

# Plant Yates Site



Chiyoda CT-121 Project

The Southern Company

FIGURE 1

# Chiyoda's Jet Bubbling Reactor

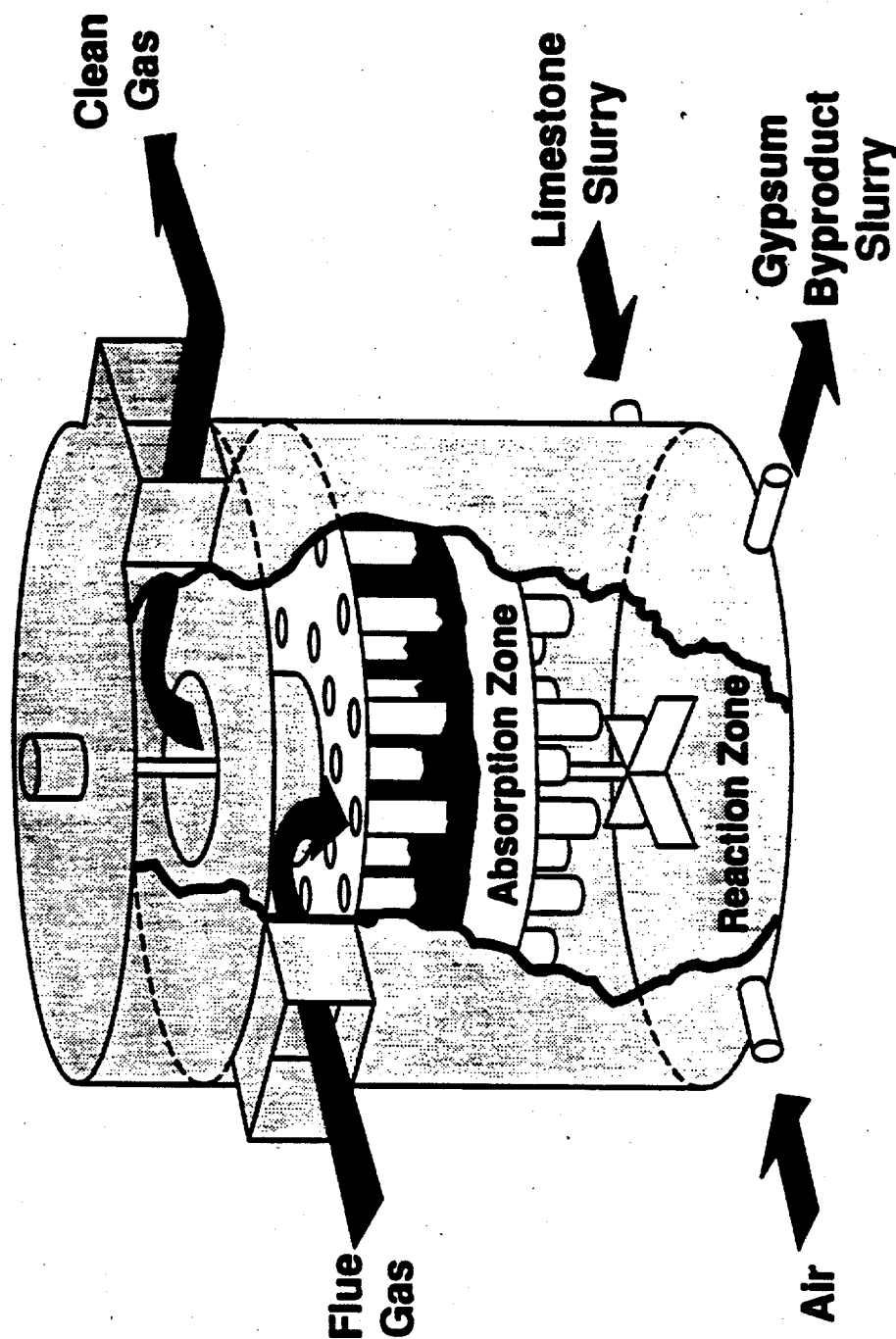
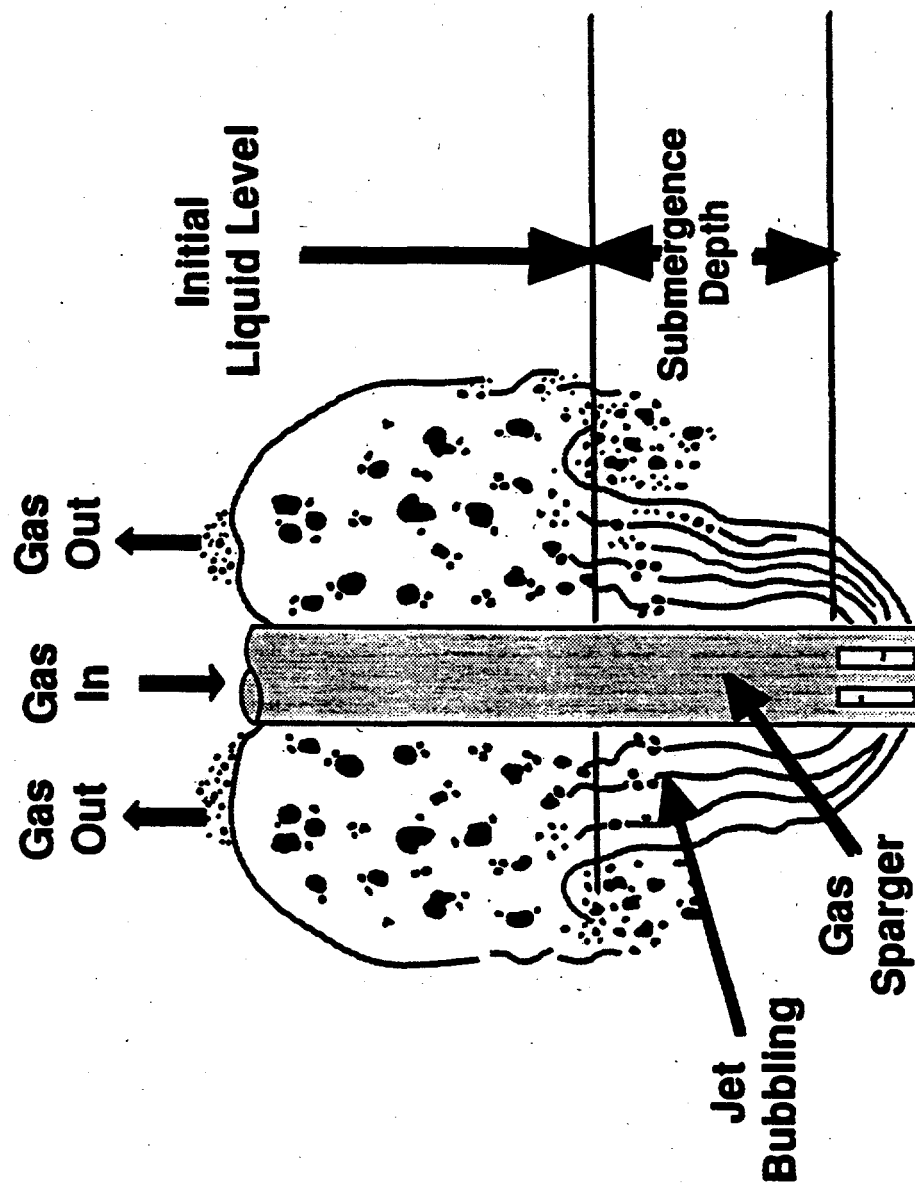


FIGURE 2

# CT-121 Gas Sparger Action



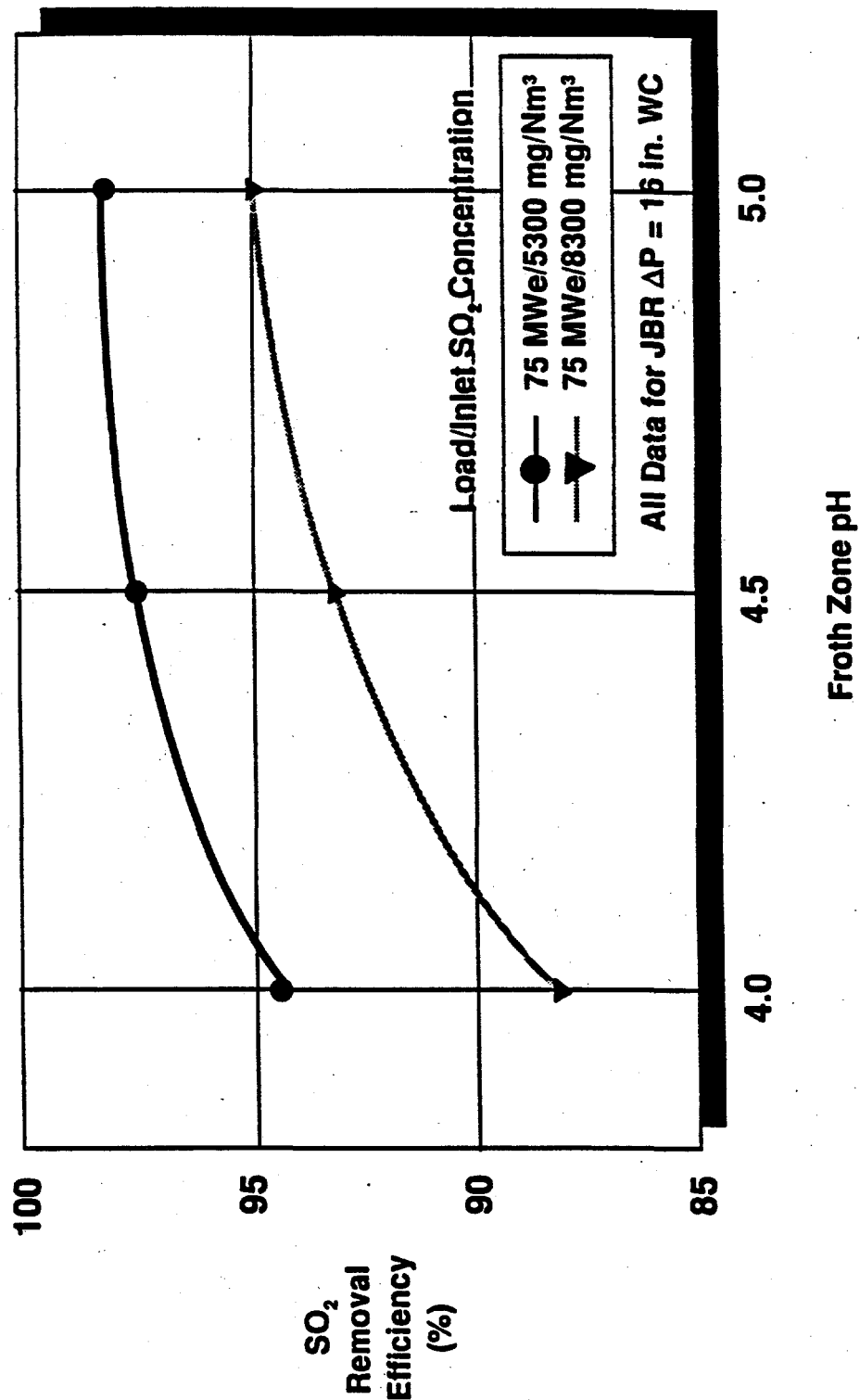
410123

Chiyoda Corporation

Chiyoda CT-121

FIGURE 3

# High-Sulfur Coal Effects on SO<sub>2</sub> Removal Efficiency

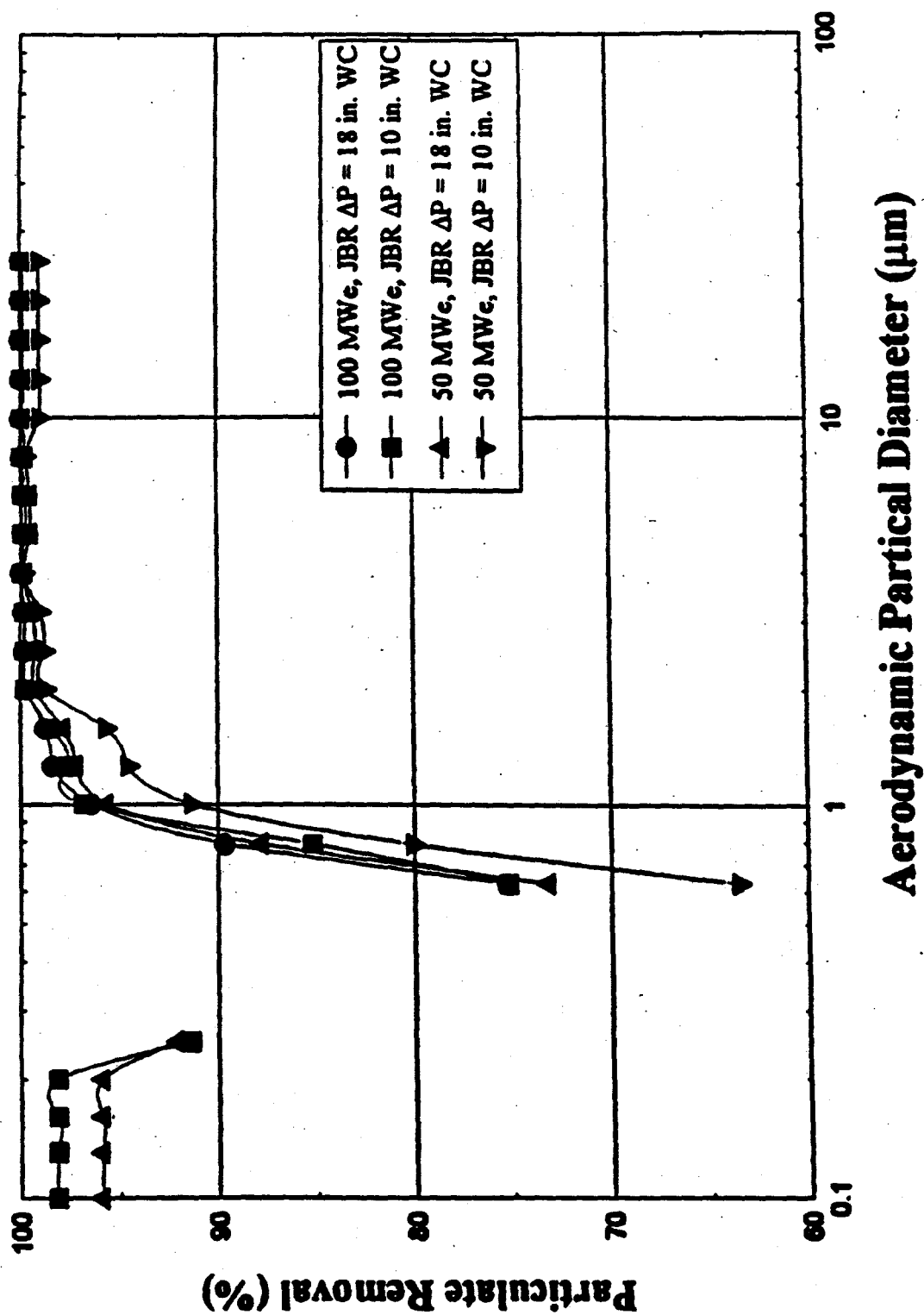


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Chiyoda Corporation

Chiyoda CT-121

FIGURE 4

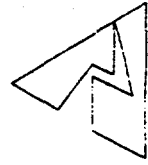
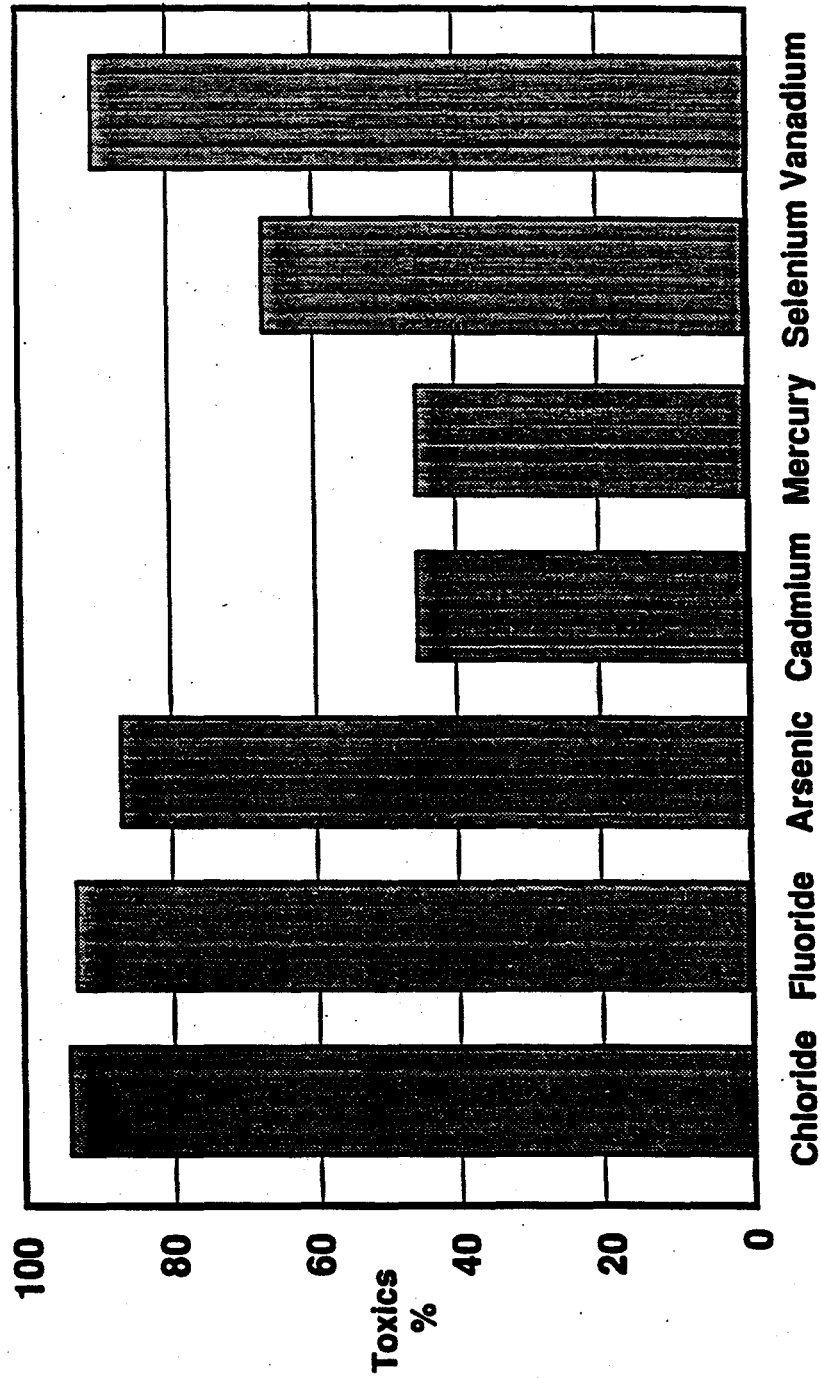


**Moderate-Ash Scrubber Particulate Removal Efficiency**

**FIGURE 5**

# Air Toxics Removal

Yates CT-121 Project (JBR Components Only)



410123

The Southern Company

Chiyoda CT-121 Project

FIGURE 6

# **THE HEALY CLEAN COAL PROJECT AN OVERVIEW**

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## **ABSTRACT**

The Healy Clean Coal Project, selected by the U.S. Department of Energy under Round III of the Clean Coal Technology Program is currently in construction. The project is owned and financed by the Alaska Industrial Development and Export Authority (AIDEA), and is cofunded by the U.S. Department of Energy. Construction is scheduled to be completed in August of 1997, with startup activity concluding in December of 1997. Demonstration, testing and reporting of the results will take place in 1998, followed by commercial operation of the facility. The emission levels of NO<sub>x</sub>, SO<sub>2</sub> and particulates from this 50 megawatt plant are expected to be significantly lower than current standards. The project status, its participants, a description of the technology to be demonstrated, and the operational and performance goals of this project are presented herein.

## **BACKGROUND**

In September 1988, Congress provided \$575 million to conduct cost-shared Clean Coal Technology (CCT) projects to demonstrate technologies that are capable of retrofitting or repowering existing facilities. To that end, a Program Opportunity Notice (PON) was issued by the Department of Energy (DOE) in May 1989, soliciting proposals to demonstrate innovative energy efficient technologies that were capable of being commercialized in the 1990's, and were capable of (1) achieving significant



reductions in the emissions of sulfur dioxide and/or the oxides of nitrogen from existing facilities to minimize environmental impacts such as transboundary and interstate pollution and/or (2) providing for future energy needs in an environmentally acceptable manner.

In response to the PON, DOE received 48 proposals in August 1989. After evaluation, 13 projects were selected in December 1989 as best furthering the goals and objectives of the PON. The projects were located in ten states and represented a variety of technologies.

One of the 13 projects selected for funding is the Healy Clean Coal Project proposed by the Alaska Industrial Development and Export Authority (AIDEA). The project will demonstrate the combined removal of SO<sub>2</sub>, NO<sub>x</sub>, and particulates from a new 50 megawatt electric coal-fired power plant using both innovative combustion and flue gas cleanup technologies. AIDEA will own the Project, perform as DOE grant recipient, administer state funds, obtain financing through sale of bonds, and manage the Project. The architect/engineer for the project is Stone & Webster Engineering Corporation. Fairbanks utility Golden Valley Electric Association (GVEA) will operate the facility and pay for power generated under terms of a power sales agreement.

#### **TECHNOLOGY TO BE DEMONSTRATED**

Coal provided by the Usibelli Coal Mine, adjacent to the project site, will be pulverized and burned at the new facility to generate high-pressure steam. The high-pressure steam will be supplied to a steam turbine generator to produce electricity. Emissions of SO<sub>2</sub> and NO<sub>x</sub> from the plant will be controlled using TRW's Entrained Combustor with limestone injection in conjunction with a boiler designed by Foster Wheeler. Further SO<sub>2</sub> and particulate removal will be accomplished using the Activated Recycle Spray Dryer Absorber System and Bag Filter developed by Joy Environmental Equipment, Inc.

The TRW Entrained Combustor is designed to operate under fuel-rich conditions, utilizing two staged combustion to minimize NO<sub>x</sub> formation. These conditions are obtained using a precombustor for heating the fuel-rich main combustor for partial combustion with combustion completion occurring in the boiler. The first and second stages of combustion produce a temperature high enough to generate a slag

(liquid ash) while reducing the fuel-bond nitrogen to molecular nitrogen (N<sub>2</sub>). The third and final stage of combustion in the boiler occurs at a combustion temperature maintained below the temperature that will cause thermal NO<sub>x</sub> formation.

The combustor is also used to reduce SO<sub>2</sub> emissions by the injection of pulverized limestone into the hot gases as they leave the combustor and enter the furnace. This technique changes the limestone into lime (flash calcination), which reacts with the sulfur compounds in the exhaust gas to form calcium sulfate. SO<sub>2</sub> is removed with combustor and boiler bottom ash. The flue gas, which contains the remaining sulfur compounds, calcium sulfate, and other solid particles leaves the boiler and passes through a spray dryer absorber and a bag filter for further SO<sub>2</sub> and particulate removal prior to exiting through the stack.

The innovative concept to be demonstrated in SO<sub>2</sub> removal is the reuse of the unreacted lime, which contains minimal fly ash, in the second-stage spray dryer SO<sub>2</sub> removal. The majority of fuel ash is removed in the combustor in the form of slag. A portion of the ash collected from the spray dry absorber vessel and the bag filter are first slurried with water, chemically and physically activated, and then atomized in the spray dryer absorber vessel for second-stage SO<sub>2</sub> removal. Third stage SO<sub>2</sub> and particulate removal occurs in the bag filter as the flue gas passes through the reactive filter cake in the bags.

The use of limestone in the combustor, combined with the recycle system, replaces the more expensive lime required by commercial spray dryer absorbers, reduces plant wastes, and increases SO<sub>2</sub> removal efficiency when burning high- and low-sulfur coals.

The integrated process is expected to achieve SO<sub>2</sub> removal greater than 90%, a reduction in NO<sub>x</sub> emissions to 0.2 pounds per million Btu. The integrated process is suited for new facilities or for repowering or retrofitting existing facilities. It provides an alternative technology to conventional pulverized coal-fired boiler flue gas desulfurization (FGD) and NO<sub>x</sub> reduction processes, while lowering overall operating costs and reducing the quantity of solid wastes.

The demonstration project is under construction adjacent to the Golden Valley

Electric Association (GVEA) existing Healy No. 1 pulverized coal-fired power plant near Healy, Alaska. Subbituminous coals from the adjacent Usibelli Coal Mine (UCM) will be the fuels. The primary fuel to be fired is a blend of run-of-mine (ROM) and waste coals. ROM coal is a subbituminous coal with a higher heating value (HHV) range of 7500-8200 Btu/lb, a low average sulfur content of 0.2 percent, and an average ash content of 8 percent. The waste coal is either a lower grade seam coal or ROM contaminated with overburden material having an HHV range, average sulfur content, and average ash content of approximately 5,000-9,000 Btu/lb, 0.15 percent, and 20 percent respectively. The project will demonstrate the ability of slagging combustors to utilize low quality coals effectively. It is anticipated that coal consumption will average 330,000 tons annually over the 40 year plant life.

### **PROJECT STATUS**

The projected project cost is about \$267 million with \$117.3 being a grant from the U.S. Department of Energy, and the remainder a combination of state grant, interest earnings, contributions from project participants, AIDEA bonds, and power sales. Construction of the HCCP began in the Spring of 1995 and is scheduled for completion in late 1997. The construction is on schedule, with startup activities planned for the fall of 1997. Demonstration testing and reporting of the results, scheduled to commence upon completion of construction, will take place in 1998. Following completion of the demonstration test program, the plant will be operated and maintained as a commercial electric generation plant.

# **Coal Reburning for Cost-Effective NO<sub>x</sub> Compliance**

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**R. G. Rock**  
**Eastman Kodak Company**

**D. T. O'Dea**  
**New York State Electric and Gas**

**Scott Hunsicker**  
**Fuller Power Corporation**

**J. U. Watts**  
**U.S. Department of Energy**  
**Pittsburgh Energy Technology Center**

## **ABSTRACT**

This paper presents the application of micronized coal reburning to a cyclone-fired boiler in order to meet RACT emissions requirements in New York State. Discussed in the paper are reburning technology, the use of a coal micronizer, and the application of the technology to an Eastman Kodak unit. The program is designed to demonstrate the economical reduction of NO<sub>x</sub> emissions without adverse impact to the boiler.

## **I. INTRODUCTION**

The Eastman Kodak Company's Kodak Park Site is one of the largest industrial parks in the nation, spanning an area in excess of 1300 acres. There are over two hundred buildings on the site that produce thousands of different photographic and chemical products. Supporting production are two power plants containing a total of fourteen boilers. Kodak has an agreement with the New York State Department of Environmental Conservation (NYSDEC) in which it states that Kodak will install coal or natural gas reburning systems on all four of its cyclone boilers. Kodak has recently completed installation of a natural gas reburning system on #43 Boiler which is located on the western side of the Kodak Park Site facility. The upgrades of the three remaining boilers (#15, #41, and #42) are planned for the 1996 thru 1998 time frame. #15 Boiler is located apart from the other three cyclone

boilers, on the eastern section of the facility, approximately three miles from #43 Boiler. The original schedule for upgrades was #43 by 1996, #41 and #42 by 1997, and #15 by 1998.

In September 1996, New York State Electric and Gas (NYSEG) presented Kodak with an alternative: NYSEG and Kodak could work together with the United States Department of Energy (DOE) to complete the upgrade of #15 Boiler if Kodak would use micronized coal as the reburn fuel instead of natural gas. This proposal was attractive to Kodak for three reasons: (1) there is no natural gas main pipeline in eastern Kodak Park Site; (2) natural gas is currently more than twice the cost of coal; and (3) DOE would co-fund the cost of installing the new system. The project will enable Kodak to meet the terms and conditions of the Kodak/DEC agreement in a more economical and timely fashion.

### **Eastman Kodak #15 Boiler**

Kodak's #15 Boiler, installed in 1956, is a cyclone-fired unit located at Kodak Park in Rochester, New York (see Figure 1). Supplied by Babcock & Wilcox Co., the unit contains two cyclone furnaces on the front wall firing crushed Eastern Bituminous coal. It typically operates at steam generation rates between 300,000 to 400,000 lb/hr; peak generation rate is 440,000 lb/hr. The cyclone furnaces operate at a very high heat release rate, creating molten slag which is captured on the cyclone walls and flows to a slag tap at the bottom of the furnace. Particulate control is maintained by an electrostatic precipitator.

In February 1996, EER performed a baseline test and measured  $\text{NO}_x$  emissions at  $1.21 \text{ lb}/10^6 \text{ Btu}$  for full load and  $0.92 \text{ lb}/10^6 \text{ Btu}$  for low load. Baseline CO emissions were 56 ppm and 34 ppm at full and low loads respectively. The results correlated closely with Kodak's belief that the baseline  $\text{NO}_x$  emissions are  $1.25 \text{ lb}/10^6 \text{ Btu}$  and baseline CO is less than 100 ppm.

### **Coal Reburning Technology for $\text{NO}_x$ Control**

Coal Reburning is a  $\text{NO}_x$  control technology whereby  $\text{NO}_x$  is reduced by reaction with hydrocarbon fuel fragments [1]. A typical application of coal reburning to a coal-fired boiler is illustrated in Figure 2. No physical changes to the main burners (cyclone furnaces in this case) are required. The burners are simply turned down and operated with the lowest excess air commensurate with acceptable lower furnace performance considering such factors as flame stability, carbon loss, and ash deposition.

The technology involves reducing the levels of coal and combustion air in the burner area and injecting reburn fuel (micronized coal) above the burners followed by the injection of overfire air (OFA) above the reburn zone. This three-zone process creates a reducing area in the boiler furnace within which  $\text{NO}_x$  created in the primary zone is reduced to elemental nitrogen and other less harmful nitrogen species. Each zone has a unique stoichiometric ratio (ratio of total air in the zone

to that theoretically required for complete combustion) as determined by the flows of coal, burner air, reburn fuel, and OFA. The descriptions of the zones are as follows:

- **Primary (burner) Zone:** Coal is fired at a rate corresponding to 75 to 90 percent of the total heat input.  $\text{NO}_x$  created in this zone is slightly lower than normal operation due to the lower heat release and the reduced excess air level.
- **Reburn Zone:** Reburn fuel (micronized coal) is injected above the main burners through wall ports. The reburn fuel consumes the available oxygen and produces hydrocarbon fragments ( $\text{CH}$ ,  $\text{CH}_2$ , etc.) which react with  $\text{NO}_x$  from the lower furnace, reducing it to elemental nitrogen,  $\text{N}_2$ . Optimum  $\text{NO}_x$  reduction performance is typically achieved when the reburn zone is operated at about 90% of stoichiometric, which is slightly fuel rich (reducing) [2].  $\text{NO}_x$  reduction can be adjusted by varying the reburn fuel injection rate, typically over the range of 10-25% of total boiler heat input. To minimize the reburn fuel required to achieve fuel rich conditions in the reburn zone, EER's design utilizes injectors rather than burners, which would have introduced additional air [3].
- **Burnout (exit) Zone:** The oxygen required to burn out the combustibles from the reburn zone is provided by injecting air through overfire air ports positioned above the reburn zone. These ports are similar to conventional overfire air ports except that they are positioned higher in the furnace so as to maximize the residence time for  $\text{NO}_x$  reduction occurring in the reburn zone. OFA is typically 20 percent of the total air flow. OFA flow rate and injection parameters are optimized to minimize CO emissions and unburned carbon-in-fly ash.

The concept of  $\text{NO}_x$  reduction via reactions with hydrocarbon fuels has been recognized for some time [4]. The work has progressed from analysis and pilot-scale tests [2] through several full-scale demonstrations including three installations on coal-fired utility boilers as part of the U.S. Department of Energy's Clean Coal Technology Program [5] and a commercial installation at New York State Electric and Gas' Greenidge Plant [6].

### **Goals of Micronized Coal Reburning Demonstration**

The objective of the coal reburning demonstration is to evaluate the applicability of the technology to full-scale cyclone-fired boilers for reduction of  $\text{NO}_x$  emissions. The project goals are:

- Reduce  $\text{NO}_x$  emissions at full load from the current established baseline of  $1.25 \text{ lb}/10^6 \text{ Btu}$  to  $0.60 \text{ lb}/10^6 \text{ Btu}$ .
- Maintain CO emissions at or below 100 ppm.
- Minimize the impact on boiler efficiency.

- Reduce  $\text{NO}_x$  without serious impact to cyclone operation, boiler performance or other emissions streams.
- Demonstrate a technically and economically feasible retrofit technology.
- Demonstrate the advantages of micronized coal reburning over conventional coal reburning.

Several derived benefits can be realized with coal reburning. From an economic standpoint, coal reburning is less expensive to install and costs less to operate than selective catalytic reduction. With micronized coal as the reburn fuel, the utilization of the fuel is enhanced which results in reduced carbon-in-ash when compared to conventional coal reburning, which also reduces particulate loading to the ESP. These benefits outweigh the additional power requirements associated with operation of the micronizers.

## II. PROCESS DESIGN

The application of reburning to a particular boiler requires careful consideration of the furnace flow field characteristics and the boiler design when developing reburning system specifications. To optimize the emissions control performance and to minimize any negative impacts of the retrofit, it is necessary to develop a design that achieves rapid and uniform mixing of the reburn fuel and overfire air streams, but minimizes the extent of modifications to the boiler heat release and heat absorption profiles.

### Controlling Process Parameters

Since the early 1980's, EER has extensively evaluated the reburning process at bench, pilot and full-scale to identify the parameters that control process performance. The results of these studies have shown that the most critical parameters are: primary  $\text{NO}_x$  level; reburn zone temperature and residence time; reburn zone stoichiometric ratio; and mixing of the reburn fuel and overfire air with the bulk furnace gases.

*Reburn Zone Stoichiometric Ratio:* The impact of this parameter on the  $\text{NO}_x$  emissions achievable with various reburn fuels is shown in Figure 3 [7]. As shown in the figure, overall  $\text{NO}_x$  reductions are highest when the ratio is approximately 0.9. To minimize the amount of reburn fuel needed to reach the optimum ratio, the primary combustion zone is operated as close to stoichiometric as possible. It should be noted, however, that with cyclone-fired boilers reducing the stoichiometric ratio in the primary zone will disrupt the slagging characteristics of the cyclone. Therefore, the fuel-to-air ratio in this area remains relatively unchanged.

*Furnace Temperatures and Residence Times:* As defined above, the reburn zone is that area of the boiler situated between the reburn fuel injectors and overfire air injectors. The amount of time required for the flue gas to pass thru this area is referred to as the residence time. The locations of injectors are selected using the following criteria:

- High temperatures in the reburn zone are preferred in order to maximize the rate of  $\text{NO}_x$  reduction. This suggests that the reburn fuel be injected just downstream of the primary zone.
- The temperature in the burnout zone must be high enough to allow oxidation of carbon monoxide and hydrocarbon fragments from the reburn zone to occur readily.
- The residence time must be of sufficient duration for the reactions to occur. EER has evaluated a number of reburning systems and concluded that a residence time of 0.2 to 0.5 seconds will achieve high efficiency  $\text{NO}_x$  reduction.

*Mixing:* Pilot-scale studies of the reburning process have shown the importance of effective mixing in both the reburn and burnout zones [8]. Effective mixing of the reburn fuel optimizes the process efficiency by making the most efficient use of the available furnace residence time. Effective mixing of the overfire air reduces carbon monoxide emissions and unburned carbon or soot.

### **Design Approach**

The final design was established on the basis of small-scale flow modeling, thermal heat transfer computer analysis, and operation of a pilot-scale micronizer using EER's Boiler Simulator Furnace. The reburn fuel and overfire air injection elevations were selected to provide the maximum amount of residence time possible in the reburn zone in order to maximize the  $\text{NO}_x$  control performance. This approach involved injection of the reburn fuel at an elevation in the furnace just above the exit of the cyclones and injection of overfire air at a distance downstream of the coal injectors that would provide for a maximum bulk residence time (Figure 4).

The reburn fuel is pneumatically transported to the boiler using recycled flue gas (FGR) as the carrier medium. The fuel is then introduced into the boiler thru injectors that are designed to rapidly mix the small quantity reburn fuel with the furnace gases. FGR is particularly suited as a carrier gas in lieu of air since it consists of a very low level of  $\text{O}_2$ . Note that any  $\text{O}_2$  introduced as carrier gas must be consumed by additional reburn fuel. The use of FGR minimizes this fuel requirement.

## **III. SYSTEM DESCRIPTION**

### **Coal Micronizer**

Preparation of the reburning fuel is performed using a MicroMill system supplied by Fuller Mineral Process Inc. The MicroMill is a patented centrifugal-pneumatic mill that works on the principle of particle-to-particle attrition. Coal is conveyed with a hot air stream into the cone area, creating a vortex of air and coal particles. As the diameter of the cone section of the mill becomes larger, the air to coal velocity decreases. The coal assumes a position in the cone based on each particle's size and weight. Particles of similar size will form bands of material with the larger particles at the



bottom of the cone. Smaller particles will move through these bands and enter the vortex created by the rotating blades in the rotational impact zone of the mill. As these smaller particles collide with the

larger particles, size reduction occurs. When a particle's size is small enough to attain the required velocity, it passes through the blades located in the scroll section of the mill and exits the mill to a static classifier.

A static classifier is used for final particle size distribution. Oversized material falls through a rotary air lock and back into the feed airstream of the mill. Stripping air provided to the classifier can be adjusted to fine tune the classifier collection efficiency allowing larger or smaller particles to pass to the boiler.

The MicroMill system can fit in approximately a thirteen foot by nine foot area and is only about twelve feet high. The mill's overall size and weight made it an ideal choice for Kodak's tight space limitations and its modular construction makes it easy to perform maintenance. The mill is designed with wear resistant materials in areas contacting the feed being processed to minimize maintenance. When maintenance is required, the cone can be unbolted, lowered on the pivot pin and rotated for access to the rotor, wear liners and replaceable blades.

The MicroMill is supported by Fuller's extensive research and development facilities which includes a full scale MF3018 MicroMill for product testing and demonstration. The Kodak feed materials were tested on this unit to determine expected capacity, fineness and power consumption. In the lab a capacity of three tons per hour at 86% passing  $44\mu$  was obtained. The limiting factor in the laboratory was motor HP. The motor for the project was increased from 150 horsepower to 200 horsepower; thus higher capacities are expected in the field. Power consumption expected for the mill is about 37.3 KW/ton of material processed. In addition, the fineness required for the application is 80% passing  $44\mu$ , which will further increase the capacity of the system. Flexibility has been designed into the system to provide a higher fineness product or greater capacity at a lower fineness.

The two-mill system for the Kodak project includes:

- Mill and motor
- Classifier
- Recycle and feed rotary airlock
- Blow through tee and feed piping
- Classifier and mill air control valves.
- Air flow meter

The mill is equipped with a water-cooled bearing jacket, vibration sensor, bearing RTD's and a proximity switch. The bearing jacket will allow the use of Kodak's uncooled flue gas as a transport medium. By utilizing the water cooled jacket the need for expensive flue gas cooling equipment was eliminated.

## **Coal Transportation and Injection**

The coal transportation system is shown in Figure 5. The slipstream for flue gas is extracted from the boiler just downstream of the precipitator and is boosted by a single fan to feed both coal micronizers. FGR is used to transport coal to the boiler and also boost its injection momentum to ensure that the reburn fuel is mixed effectively in the furnace.

Two coal micronizers with classifiers are used in the system. Each micronizer is supplied coal from a bunker thru a screw feeder. The FGR system assists in the micronizing process and in operation of the classifiers. The mills are capable of operating singly or as a pair, although, due to capacity limitations, both may be required to produce the targeted  $\text{NO}_x$  reduction.

The micronized coal exiting the two mills is merged into a single 18-inch pipe for transportation to the boiler. The line is then divided into eight 6-inch segments by a coal flow splitter supplied by EER.

The splitter is designed to apportion the coal into equal segments without incurring any pressure drop. Upstream of the splitter is a coal rope breaker (RopeMaster<sup>®</sup>) supplied by Rolls-Royce/International Combustion, which will enhance the splitter's effectiveness. Downstream of the splitter are eight FlowMastEER<sup>®</sup> dampers designed by EER that are used to perform final adjustments to the coal flow balance. The dampers can also be used to create flow biasing.

Eight micronized coal injectors are installed, six on the rear wall and one on each side wall near the rear wall. The injectors utilize the considerable momentum provided by the FGR transport gas plus additional design features to enhance coal penetration. Each injector is equipped with a variable swirl device to control the mixing characteristics of each fuel jet as it enters the furnace. Adjustments will be made during initial startup to optimize the injector effectiveness. The coal injectors were designed by EER specifically for this project.

## **Overfire Air System**

Located on the front wall are four overfire air injectors. These injectors utilize EER's Second Generation dual-concentric overfire air design. This is EER's second application of this concept [9]. The injectors are designed to provide good jet penetration as well as good lateral dispersion across the boiler depth and width. Each injector is equipped with an integral damper to maintain the desired injection velocity as load changes and a swirler which, when adjusted, provides for optimum mixing in the burnout zone.

## **Controls**

Kodak installed a new Coen burner management system and replaced the complete boiler control system with a Westinghouse WDPF distributed digital control system. The new controls operate both the existing equipment and the micronized coal reburning system, with all normal start/stop/modulate operator actions occurring in the control room. Critical operations are

interlocked to prevent inadvertent operation of equipment when such operation may present an operating hazard or other undesirable condition. The controls are designed to shut down the reburning system while maintaining operation of the boiler. Kodak's insurance carrier, Factory Mutual, has approved this control arrangement. Previous to this project, EER reburning retrofits were approved by Factory Mutual and Hartford Steam.

### **Operation**

During operation of the reburning system, the total fuel to the boiler is the sum of the fuel to the cyclones plus the fuel to the reburn injectors. Any change in the amount of reburn fuel must be balanced by an converse change in the fuel to the cyclones. During normal operation, the boiler generates steam at rates between 300,000 and 400,000 lb/hr. The lower limit of 300,000 lb/hr is based on the amount of bottom ash required to prevent slag freezing. The range of reburn fuel injection is based on the following two factors:

- The minimum reburn fuel injection rate is based in lower operational limit of the coal preparation equipment (coal feeder, micronizer, classifier, etc.).
- The maximum reburn fuel injection rate is that amount required raise the boiler from the cyclone minimum operating level (300,000 lb/hr steam) to the boiler maximum operating level (400,000 lb/hr steam). Note that the minimum cyclone operating level may be lower than 300,000 lb/hr during reburning since reburn fuel ash also contributes to the bottom ash total. The maximum amount of reburn fuel that can be injected is estimated to be 25% of the total heat input.

At boiler full load with maximum operation of reburning, load can be reduced by lowering the injection rate of the reburn fuel. The load on the cyclones would remain the same. This capability is described in Figure 6.

## **IV. SUMMARY**

The coal reburning installation at Eastman Kodak Company will permit Kodak to meet RACT emissions requirements in New York State. The project, conducted under the auspices of the U.S. Department of Energy's Clean Coal Technology Program, is designed to demonstrate the economic advantages of using coal micronizer technology versus conventional coal reburning. Testing of the system will verify the target goals of NO<sub>x</sub> emissions reduction and determine the full range of operation, including turndown capabilities. The testing will also be used to develop a database of technical information that can be applied to similar boilers.

Coal reburning is less expensive to install and costs less to operate than selective catalytic reduction (SNCR). Using coal as the reburn fuel results in economical reburn fuel selection, decreased primary mill capacity, no additional chemical/catalyst cost, and no ammonia slip normally associated with

SNCR. With micronizer technology feature, the utilization of the reburn fuel is enhanced which results in reduced carbon-in-ash when compared to conventional coal reburning, which also reduces particulate loading to the ESP.

This paper has focused on reburning technology, a description of the project and its inherent benefits including. Future papers will present the results of extensive testing.

## V. ACKNOWLEDGMENTS

The following organization is acknowledged for their contributions to the project: New York State Energy Research and Development Authority and the U.S. Department of Energy's Clean Coal Technology Program. EER would also like to acknowledge the support of Rolls-Royce/International Combustion and Parsons Power Group Inc.

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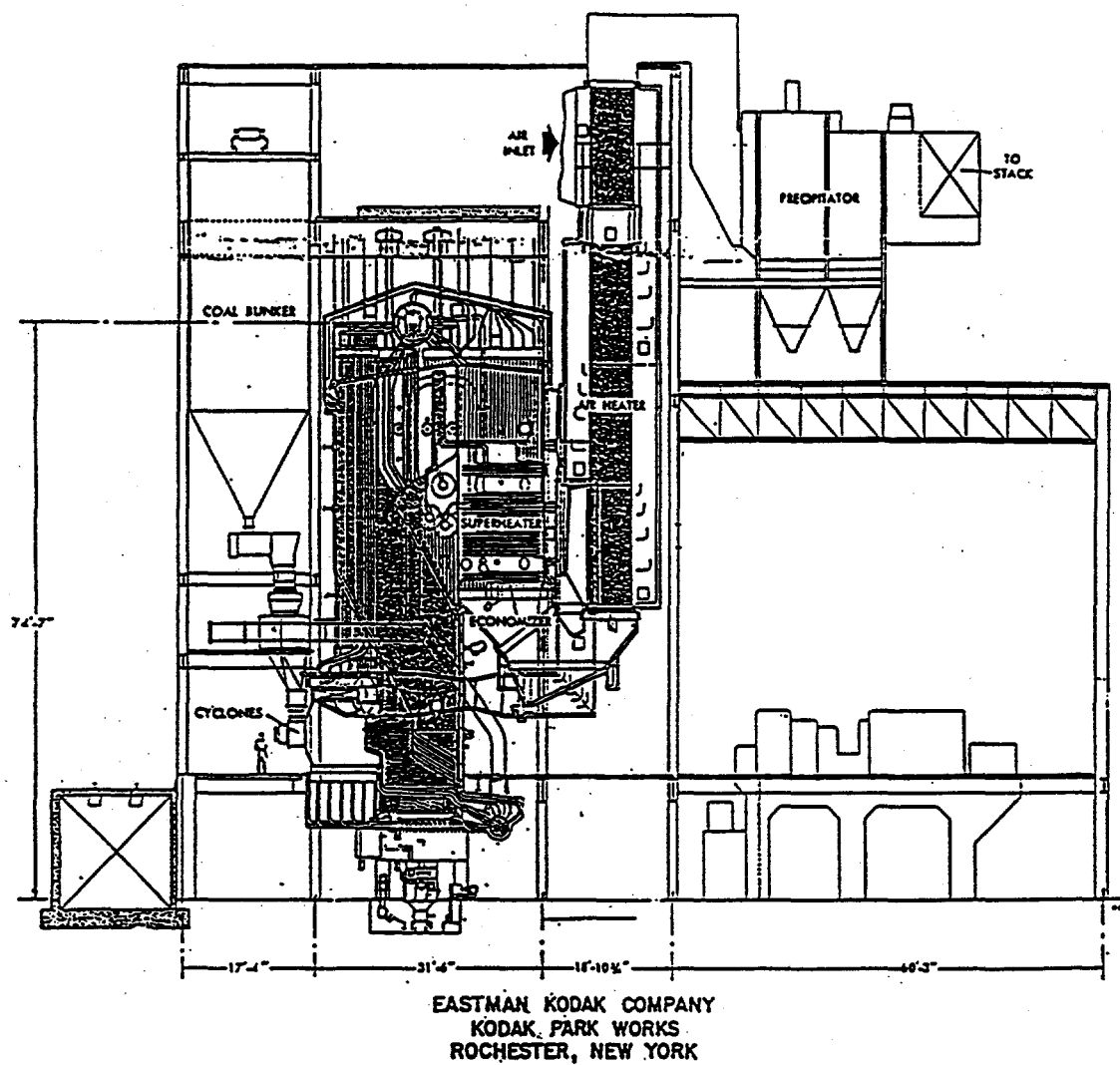


Figure 1. Kodak #15 Boiler.

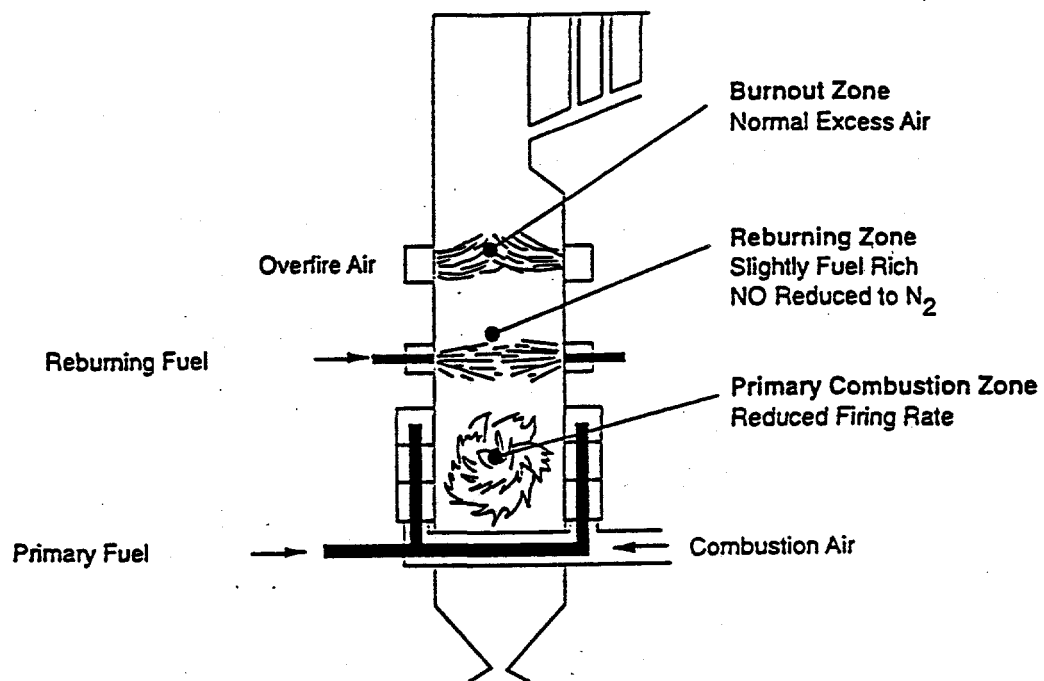


Figure 2. Application of reburning technology to a utility boiler.

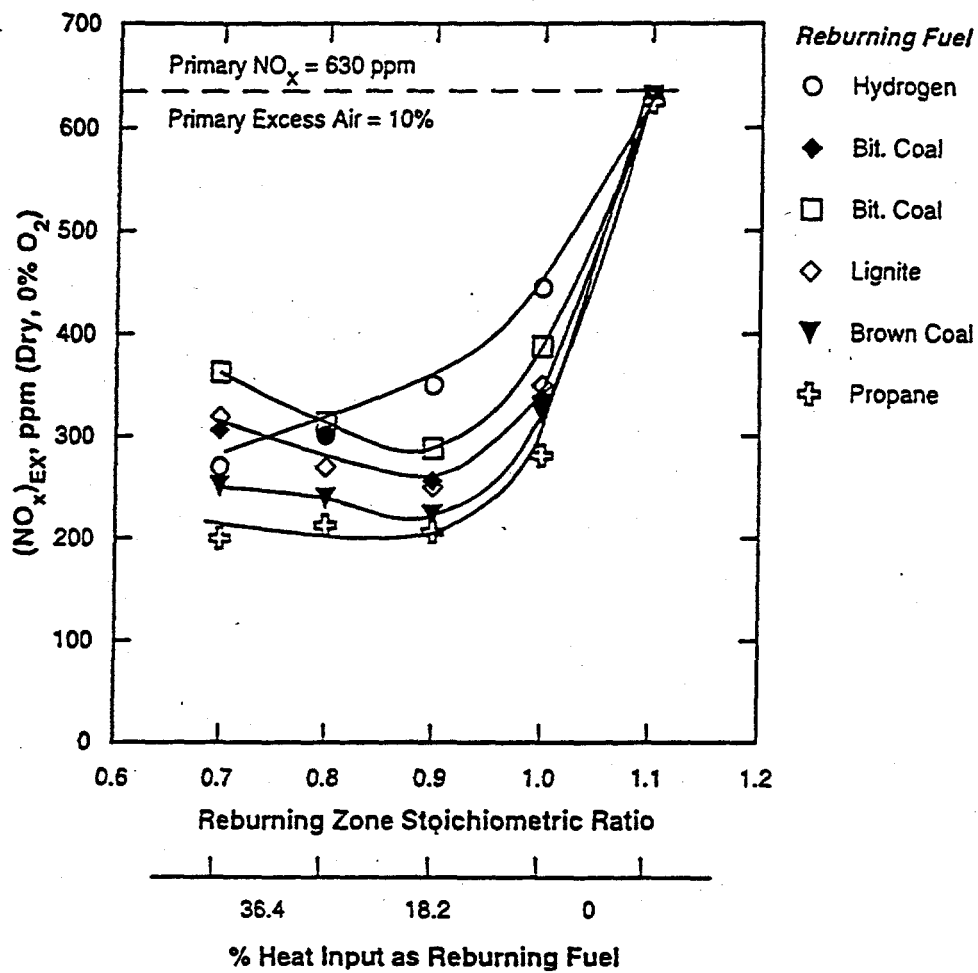


Figure 3. Impact of reburning zone stoichiometric ratio and reburning fuel type on reburning performance.



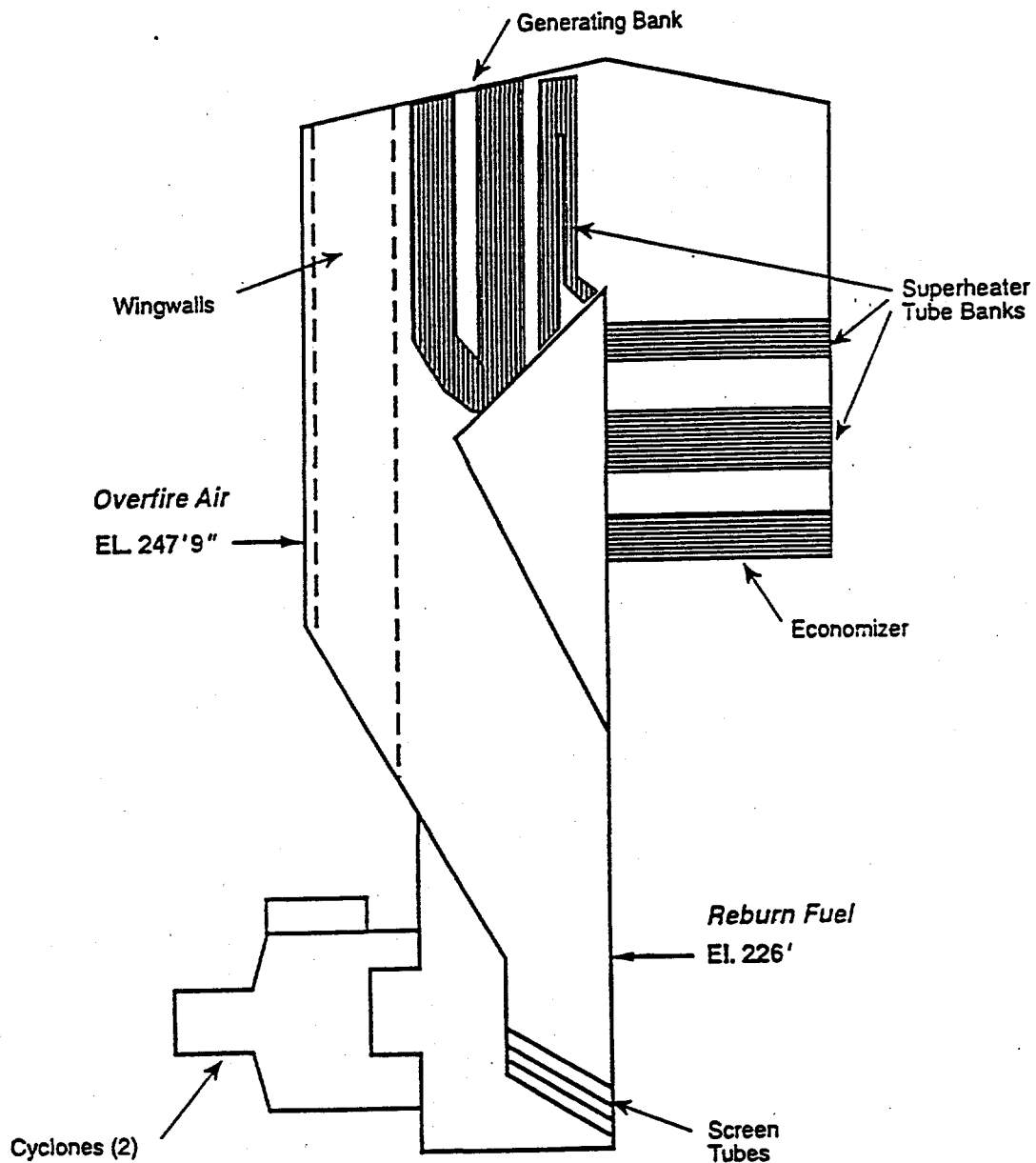


Figure 4. Reburn fuel and overfire air injection elevations for Kodak #15 Boiler.

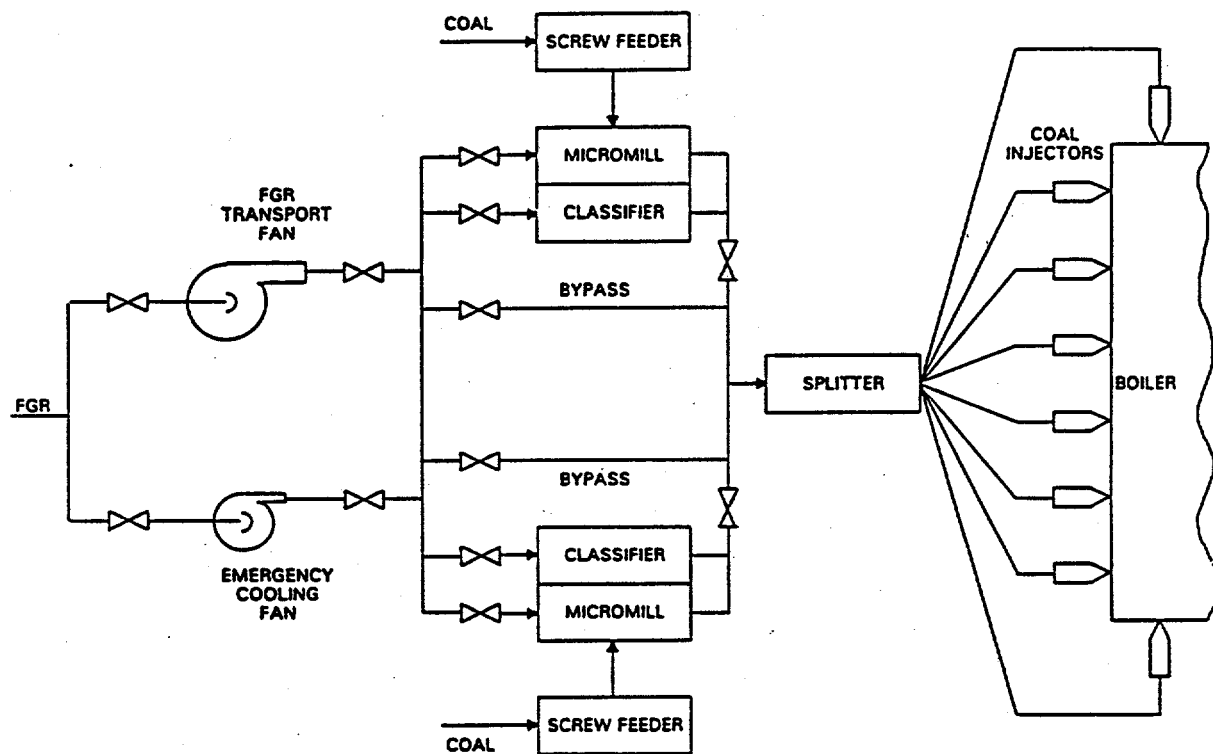


Figure 5. Micronized Coal Feed System

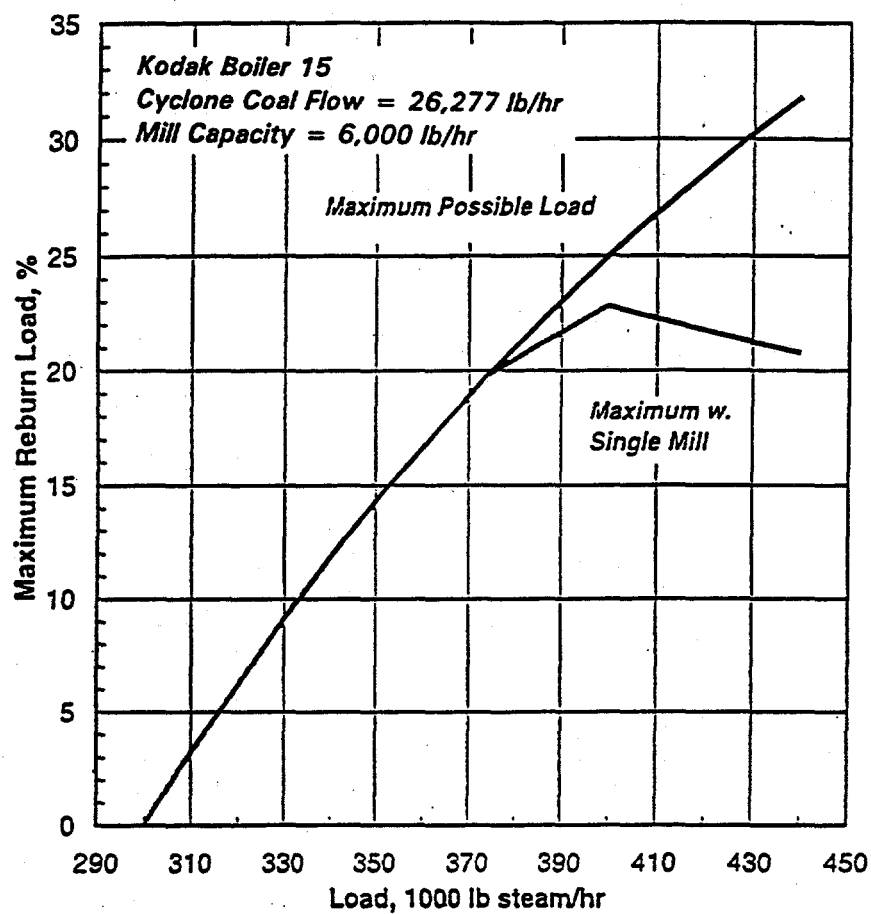


Figure 6. Maximum level of reburning system operation achievable while maintaining minimum coal flow to cyclones.

# THE NOXSO CLEAN COAL PROJECT

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## ABSTRACT

The NOXSO Clean Coal Project will consist of designing, constructing, and operating a commercial-scale flue-gas cleanup system utilizing the NOXSO Process. The process is a waste-free, dry, post-combustion flue-gas treatment technology which uses a regenerable sorbent to simultaneously adsorb sulfur dioxide ( $\text{SO}_2$ ) and nitrogen oxides ( $\text{NO}_x$ ) from flue gas from coal-fired boilers. The NOXSO plant will be constructed at Alcoa Generating Corporation's (AGC) Warrick Power Plant near Evansville, Indiana and will treat all the flue gas from the 150-MW Unit 2 boiler. The NOXSO plant is being designed to remove 98% of the  $\text{SO}_2$  and 75% of the  $\text{NO}_x$  when the boiler is fired with 3.4 weight percent sulfur, southern-Indiana coal. The NOXSO plant by-product will be elemental sulfur.

The elemental sulfur will be shipped to Olin Corporation's Charleston, Tennessee facility for additional processing. As part of the project, a liquid  $\text{SO}_2$  plant has been constructed at this facility to convert the sulfur into liquid  $\text{SO}_2$ . The project utilizes a unique burn-in-oxygen process in which the elemental sulfur is oxidized to  $\text{SO}_2$  in a stream of compressed oxygen. The  $\text{SO}_2$  vapor will then be cooled and condensed. The burn-in-oxygen process is simpler and more environmentally friendly than conventional technologies. The liquid  $\text{SO}_2$  plant produces 99.99% pure  $\text{SO}_2$  for use at Olin's facilities.

The \$82.8 million project is co-funded by the U.S. Department of Energy (DOE) under Round III of the Clean Coal Technology program. The DOE manages the project through the Pittsburgh Energy Technology Center (PETC).

## INTRODUCTION

The NOXSO Process is a waste-free, dry, post-combustion flue-gas cleanup technology which uses a regenerable sorbent to simultaneously adsorb  $\text{SO}_2$  and  $\text{NO}_x$  from flue gas from coal-fired utility and industrial boilers. In the process, the  $\text{SO}_2$  is converted to a saleable sulfur by-product (liquid  $\text{SO}_2$ ,

elemental sulfur, or sulfuric acid) and the  $\text{NO}_x$  is converted to nitrogen and oxygen. Since  $\text{SO}_2$  and  $\text{NO}_x$  removal occur at normal flue-gas temperatures (downstream of the combustion air preheater), the NOXSO Process is equally suited for retrofit as well as new installations.

Process development began in 1979 with laboratory-scale tests and progressed to pre-pilot-scale tests (3/4-MW) and a life-cycle test. Each of these test programs [1,2,3] has provided data necessary for the process design. Tests of the  $\text{NO}_x$  recycle concept, which is inherent to the NOXSO Process, have been conducted on small boilers at PETC and at the Babcock & Wilcox (B&W) Research Center in Alliance, Ohio [4].

A 5-MW Proof-of-Concept (POC) pilot-plant test at Ohio Edison's Toronto Plant in Toronto, Ohio, was completed in 1993 [5]. Based on more than 7,000 hours of operation with flue gas, it was demonstrated the process can economically remove more than 95% of the acid rain precursor gases from the flue-gas stream.

The NOXSO Clean Coal Project is the final step in commercialization of the technology. The project was selected during Round III of the DOE Clean Coal Technology Program and is managed through PETC. NOXSO Corporation is the project participant, project manager and technology supplier. The project is being hosted by AGC at their Warrick Power Plant (WPP) near Evansville, Indiana. Morrison Knudsen Corporation is providing engineering services. Projex Inc. is managing construction of the facility.

Final processing of the sulfur by-product to make liquid  $\text{SO}_2$  will be completed at Olin Corporation's Charleston, Tennessee facility. The  $\text{SO}_2$  plant which utilizes a unique burn-in-oxygen process for converting sulfur to liquid  $\text{SO}_2$  is complete. The burn-in-oxygen process is simpler and more environmentally friendly than conventional technologies. Midwest Technical, Inc. provided engineering services. Projex, Inc. managed construction of the facility.

Design and procurement activities are currently being conducted for the NOXSO plant. Preliminary construction activities were completed during the fall of 1996, with full-scale construction scheduled to begin in February 1997. Mechanical completion will occur in June 1998. After commissioning and start-up, the plant will be operated for two years as part of the Clean Coal Project.

Meanwhile, mechanical completion, testing and start-up of the liquid  $\text{SO}_2$  plant was achieved in December 1997. Feedstock sulfur will be purchased on the market until the start-up of the NOXSO plant, at which time the NOXSO plant will be the sole source of feedstock for the liquid  $\text{SO}_2$  plant. Operating and environmental data will be collected during the plant's operation.

Funding for the \$82.8 million project will be provided by the DOE, NOXSO, AGC, Warrick County, the Southern Indiana Gas and Electric Company (SIGECO), the Gas Research Institute (GRI), W.R. Grace, and the Electric Power Research Institute (EPRI). NOXSO will raise most of its project funds through the sale of revenue bonds issued and guaranteed by the state of Indiana. The guarantee is made possible by state legislation signed into law on March 28, 1995. NOXSO will repay the bonds from revenue generated by the sale of  $\text{SO}_2$  allowances and by the sale of liquid  $\text{SO}_2$  to Olin during a thirteen-year time period which includes the two-year demonstration operation period.

## **II. THE NOXSO COMMERCIAL DEMONSTRATION PLANT**

The objective of the NOXSO Clean Coal Technology Project is to design, construct, and operate a NOXSO plant at commercial scale. At the completion of this project, the performance, operability, reliability, construction cost, and operating cost data will be available to assist utilities in making decisions regarding the choice of flue-gas cleanup technology.

### **Host Site Information**

The WPP is owned by AGC and operated by the Southern Indiana Gas and Electric Company (SIGECO). The plant supplies electricity to Alcoa's adjacent Warrick Operations aluminum facility and to the utility grid. The WPP consists of three coal-fired steam electric generating units (Units 1, 2, and 3), each rated at 150 MW, and Unit 4, rated at 300 MW. Unit 4 is jointly owned by AGC and SIGECO. Approximately 80% of the electric power generated at WPP is used by Warrick Operations with the remainder being sent to the utility grid.

As shown in Figure 1, the WPP is located in Warrick County, about 15 miles east of Evansville, Indiana, on Indiana Route 66. The WPP and Warrick Operations are located on approximately 600 acres of land between Indiana Route 66 and the Ohio River.

High sulfur Squaw Creek coal with composition as shown in Table 1 will be burned in Unit 2 after the NOXSO plant is installed. Squaw Creek coal is currently blended with a low sulfur coal for use in Units 1, 2, and 3 to satisfy the Warrick County State Implementation Plan (SIP) limit of 5.11 pounds SO<sub>2</sub> per million Btu of heat input.

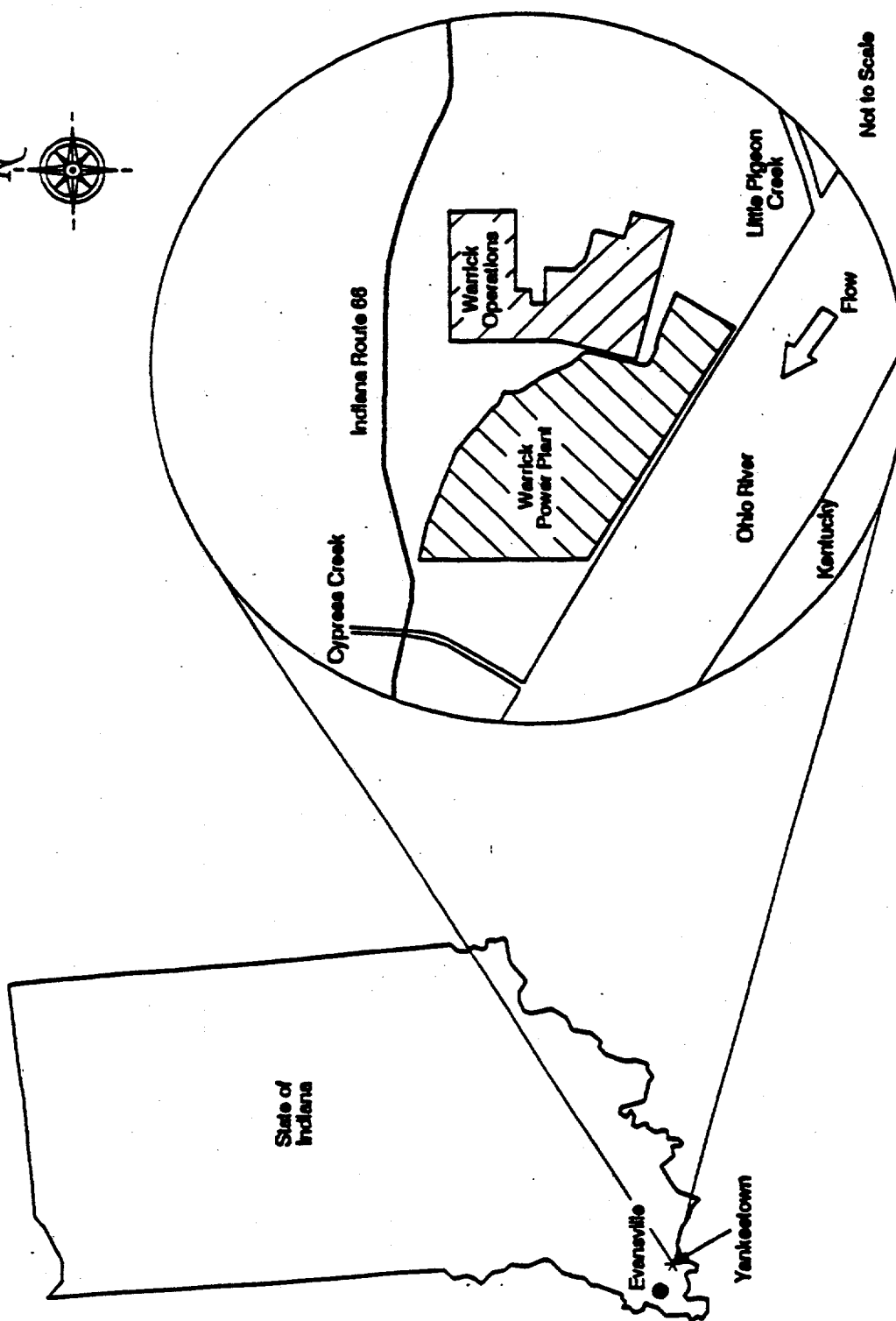


Figure 1. Warrick Power Plant and Warrick Operations Site Location.

Parameter	Weight Percent (%)
Moisture	12.92
Carbon	62.02
Hydrogen	4.58
Nitrogen	1.22
Chlorine	0.05
Sulfur	3.39
Ash	8.23
Oxygen	7.60
Higher Heating Value (HHV) (Btu/lb)	11,307

**Table 1. Squaw Creek Coal - Ultimate Analysis**

AGC has opted-in WPP Units 1, 2, and 3 to the Acid Rain Program of the Clean Air Act (CAA) Amendments of 1990. The Opt-In Program (40 CFR Part 72) allows nonaffected sources, such as AGC's WPP Units 1, 2, and 3, to enter the SO<sub>2</sub> portion of the acid rain program and receive SO<sub>2</sub> emission allowances.

Table 2 shows the design parameters for Unit 2. The wall-fired unit built by Babcock & Wilcox Company (B&W) was placed into service in 1964. The boiler is a natural circulation, Carolina-type radiant unit with 16 circular coal burners arranged in a 4-by-4 grid on a single furnace wall. Coal is reduced from 3/4 inches (in) to 60% less than 200 mesh by B&W EL-76 ball and race pulverizers.

Boiler Manufacturer	Babcock & Wilcox
Operation Date	1964
Primary Fuel	Coal
Start-up Fuel	Natural gas with co-fire
Boiler Type	Wall-fired, natural circulation, Carolina-type radiant unit
Nameplate Rate	144 MW
Steam Flow	1,000,000 lb/hr
Steam Temperature	1,005°F
Design Pressure	1,975 psig
Turbine/Generator Set	160 MW
Existing Burners	16 wall-fired burners
Particulate Control	Western Precipitator electrostatic precipitator designed for 1.83 grains/acfm outlet dust for 688,600 acfm flue gas at 710°F

**Table 2. Unit 2 Design Parameters**



## NOXSO Process Description

The NOXSO Process is a dry, post-combustion flue-gas treatment technology which will use a regenerable sorbent to simultaneously adsorb  $\text{SO}_2$  and  $\text{NO}_x$  from the flue gas from Unit 2 of AGC's WPP. In the process, the  $\text{SO}_2$  will be converted to liquid  $\text{SO}_2$  and the  $\text{NO}_x$  will be reduced to nitrogen and oxygen. The NOXSO plant is designed to remove 98% of the  $\text{SO}_2$  and 75% of the  $\text{NO}_x$ . Details of the NOXSO Process are described with the aid of Figure 2.

Flue gas from the power plant is drawn through two flue-gas booster fans which force the air through two fluid-bed adsorbers and a baghouse before passing to the power plant stack. For simplicity, only one adsorption train is shown in Figure 2. Water is sprayed directly into the adsorber fluid beds as required to lower the temperature to 250-275°F by evaporative cooling. The fluid-bed adsorber contains active NOXSO sorbent. The NOXSO sorbent is a 1.2 mm diameter stabilized  $\gamma$ -alumina bead impregnated with sodium. The baghouse removes sorbent which may be entrained in the flue gas and directs it to the fly ash sluicing system.

Spent sorbent from the adsorbers flows into a dense-phase conveying system which lifts the sorbent to the top bed of the sorbent heater vessel. The sorbent flows through the four-stage fluidized-bed sorbent heater in counterflow to the heating gas which heats the sorbent to the regeneration temperature of approximately 1150°F.

In heating the sorbent, the  $\text{NO}_x$  is driven off and carried to the power plant boiler in the  $\text{NO}_x$  recycle stream. The  $\text{NO}_x$  recycle stream is cooled from approximately 360°F to 140°F in the feedwater heater. This heat-exchanger heats a slip stream of the power plant's feedwater, thereby reducing the amount of extraction steam taken from the low pressure turbine, enabling the generation of additional electricity. The cooled  $\text{NO}_x$  recycle stream replaces a portion of the combustion air. The presence of  $\text{NO}_x$  in the combustion air suppresses the formation of  $\text{NO}_x$  in the boiler resulting in a net destruction of  $\text{NO}_x$ .

The heated sorbent is transported through an L-valve to the steam disengaging vessel. Transport steam is separated from the sorbent to reduce the volume of the regenerator off-gas stream. Sorbent gravity flows into the regenerator where it is contacted with natural gas. Through a series of chemical reactions, the sulfur on the sorbent combines with the methane and forms  $\text{SO}_2$  and  $\text{H}_2\text{S}$ . Additional regeneration occurs in the steam treater section of the regenerator when the



sorbent is contacted with steam, converting the remaining sulfur on the sorbent to  $H_2S$ . The regenerator off-gas stream is directed to a sulfur recovery plant where the  $H_2S$  and  $SO_2$  are converted to elemental sulfur. Tail gas from the sulfur recovery plant will be oxidized and recycled back through the adsorbers to remove any residual sulfur compounds.

High temperature sorbent exiting the regenerator is conveyed with an L-valve to the four-stage fluidized-bed sorbent cooler. The sorbent flows counter to the ambient air which cools the sorbent. Regenerated sorbent exits the cooler at  $320^{\circ}F$ . The sorbent is then conveyed through an L-valve to the sorbent surge tank before being returned to the adsorber, completing the sorbent cycle.

Ambient air which is forced through the sorbent cooler by the heater-cooler fans exits the sorbent cooler at approximately  $950^{\circ}F$ . This preheated air then enters the air heater where it is heated to approximately  $1340^{\circ}F$ . The high temperature air is used in the sorbent heater to heat the sorbent to the regeneration temperature of  $1150^{\circ}F$ .

### **NOXSO Plant Description**

The Demonstration Plant will be located in a generally unoccupied area of the plant yard south of Unit No. 2. This area requires minimal site preparation and provides adequate space for the NOXSO plant while offering a convenient tie-in point for the flue-gas ductwork, see Figure 3, since the existing flue-gas plenum and plant stacks are located on the south side of the power plant. This location also provides plant access from the south for sorbent and nitrogen delivery while the sulfur recovery unit is accessible by rail and road. The general arrangement is shown in Figure 4.

The NOXSO plant will take up an area approximately  $250' \times 200'$  in size, just south of Precipitator Road, which is an east-west running plant access road south of the power plant. The analyzer and control building is located to the east of the NOXSO plant while the sulfur recovery unit is situated to the west, at the southern end of the battery limits.

The locations of the major process vessels within this area are chosen to minimize the amount of ductwork required to deliver and return the flue gas, and to minimize the horizontal distances that the sorbent must travel between vessels. Thus, the adsorption trains, including booster fans, adsorbers, and baghouses, are situated furthest north within the battery limits. The adsorption trains are shown in the foreground of Figure 4. The adsorbers, like the regenerator and sorbent cooler, are self-supporting vessels through the use of vessel skirts which reduce the overall amount of structural steel required.

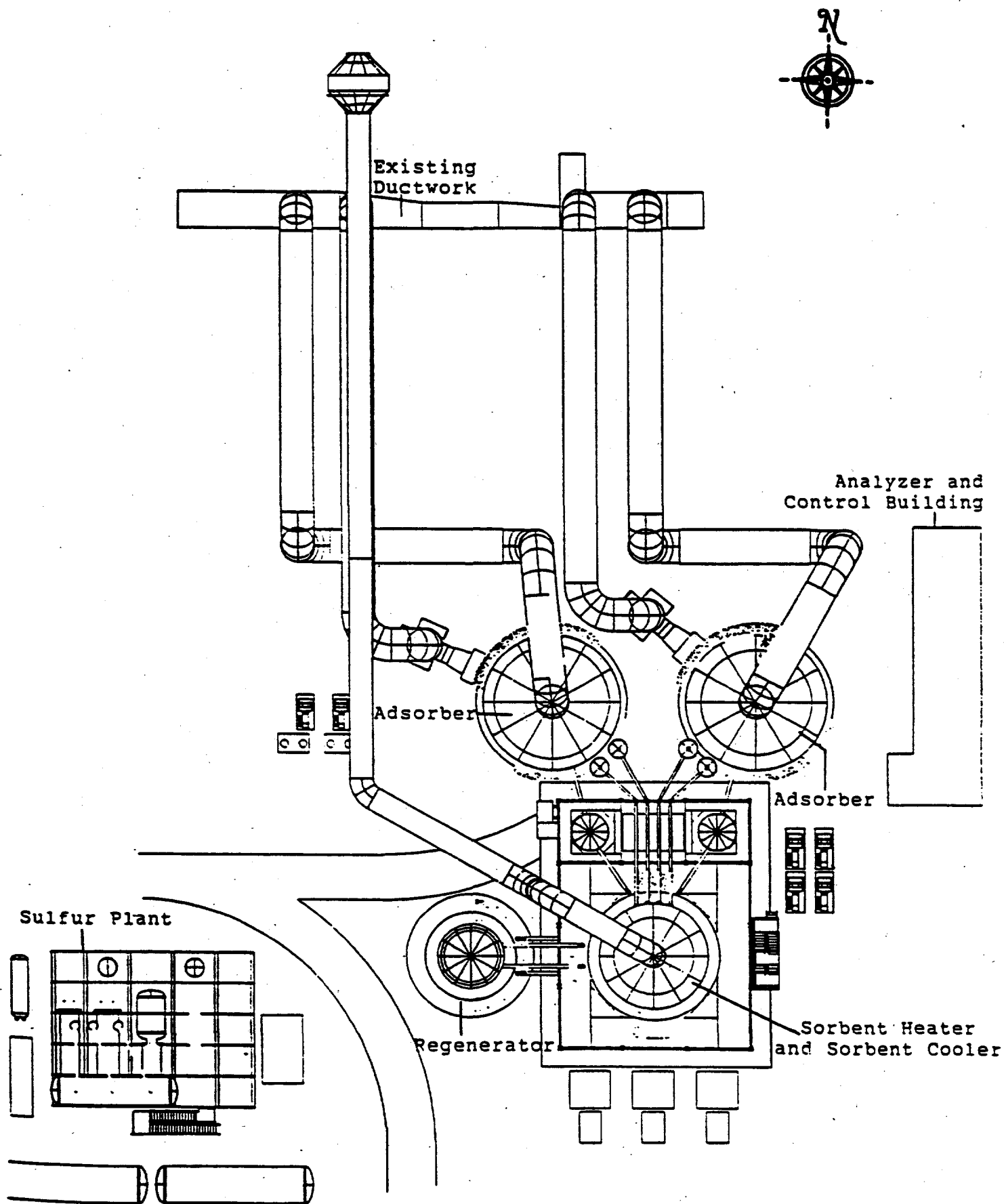


Figure 3. Plant Layout

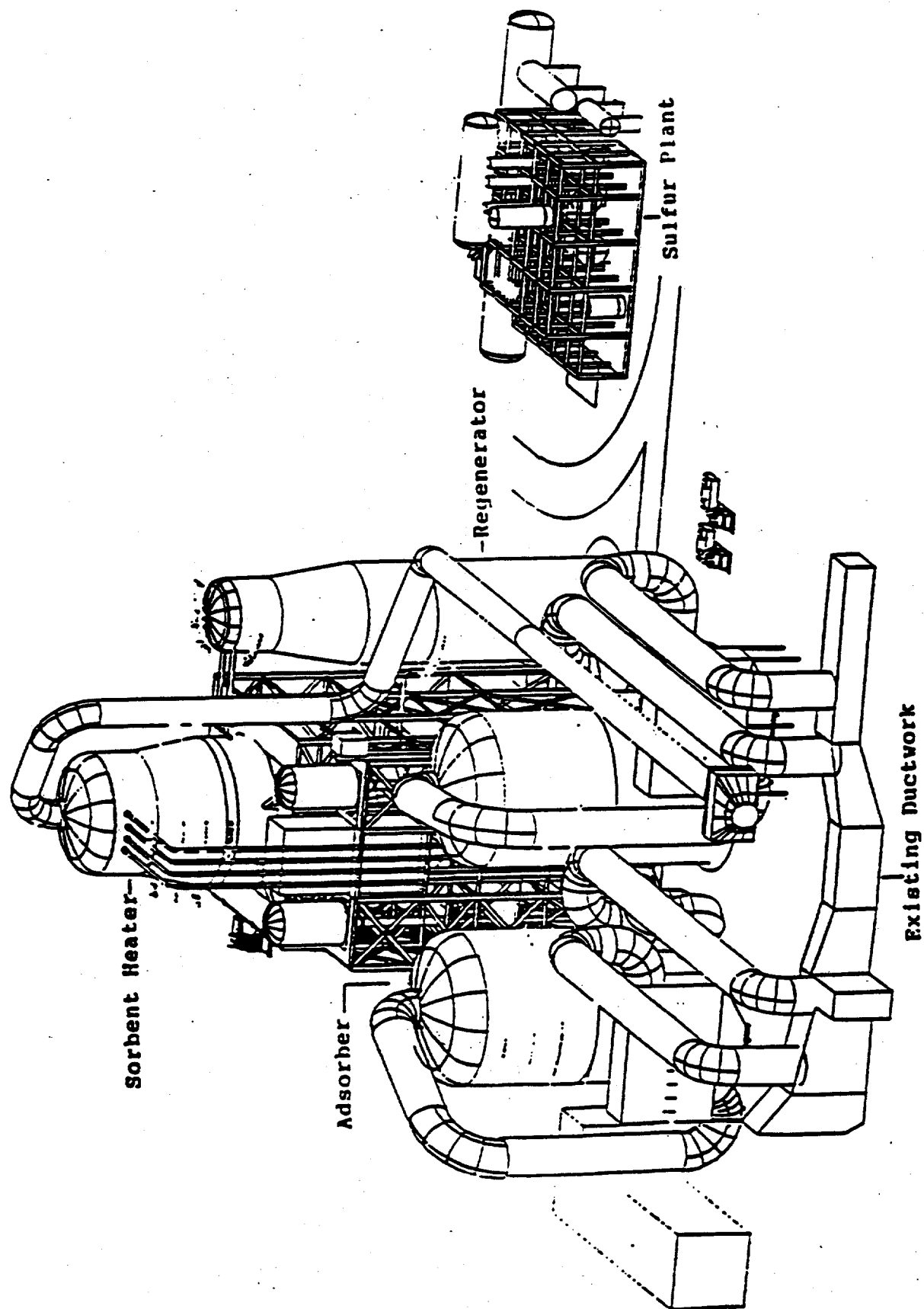


Figure 4. General Arrangement

The regeneration train, consisting of the sorbent heater, steam disengaging vessel, regenerator and sorbent cooler, is just south of the adsorption trains. The sorbent cooler and sorbent heater are in a stacked arrangement, so that the heat energy recovered by the fluidizing air in the sorbent cooler may be used in the sorbent heater. The sorbent cooler, hidden by the structural tower in Figure 4, is skirt supported on the ground, while the sorbent heater is supported 95' in the air at its base by the sorbent heater tower. This tower is centered behind and situated as close as possible to the two adsorbers to minimize the horizontal distance that the sorbent must travel between the two trains.

The regenerator and steam disengaging vessel are in a stacked arrangement to allow gravity flow of the sorbent between the two vessels. Again, to minimize the horizontal sorbent conveying distance, the regenerator is situated as close as possible to the sorbent heater tower. The regenerator is located on the west side of the tower because of space availability for the sulfur recovery unit, which is to the west of the regenerator. It is essential to position the sulfur recovery unit as close as possible to the regenerator to limit the distance of the steam-traced, regenerator off-gas line.

### III. THE LIQUID SO<sub>2</sub> FACILITY

As discussed previously, the purpose of the NOXSO Clean Coal Project is to demonstrate the NOXSO flue-gas treatment system in a fully integrated commercial scale operation. The NOXSO plant will reduce SO<sub>2</sub> and NO<sub>x</sub> emissions from Alcoa Generating Corporation's Warrick Power Plant Unit 2. The removed sulfur will be processed into elemental liquid sulfur. In addition, as part of the project, a liquid SO<sub>2</sub> plant has been constructed at Olin Corporation's Charleston, Tennessee facility to convert the sulfur into liquid SO<sub>2</sub>.

#### Host Site Information

Figure 5 is a site plan of the Olin Charleston Plant (OCP). There are five basic areas within the plant: administration, including process technology and product quality/environmental control buildings; chlor-alkali, consisting of chlorine/caustic soda production facilities, Reductone® (sodium hydrosulfite) production facilities, hydrochloric acid production facilities, boiler house, and water treatment; HTH® Dry Chlorinator (calcium hypochlorite) production facilities and associated warehousing; rubber services, and associated warehousing; and maintenance facilities.

As shown in Figure 6, OCP is located in Bradley County, in southeastern Tennessee about 12 miles northeast of Cleveland, Tennessee. Charleston, Tennessee, the closest town to the site, is 1.5 miles southeast of the plant. The OCP consists of roughly 975 acres between Lower River Road and the Hiwassee River (which flows to the northwest). Liquid SO<sub>2</sub> is a primary feedstock at the OCP where it is used to produce sodium hydrosulfite which is sold to the paper industry where it is used as a bleach for paper and clay.

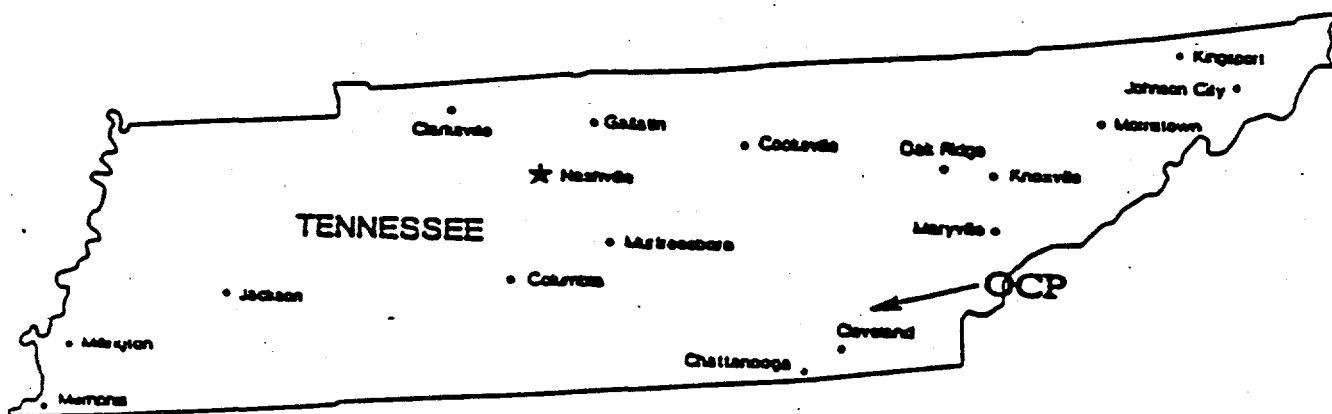


Figure 5. OCP Location





## **Liquid SO<sub>2</sub> Process Description**

The liquid SO<sub>2</sub> facility consists of two components, the liquid SO<sub>2</sub> plant and a cryogenic air separation plant. The facility is located on less than an acre of Olin property east of the existing switchgear building. Figure 7 presents the site plan for the liquid SO<sub>2</sub> facility detailing its relationship within Olin's plant site.

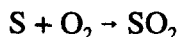
The SO<sub>2</sub> plant, the primary aspect of the liquid SO<sub>2</sub> facility, is an advanced liquid SO<sub>2</sub> production process designed for ease of operation and maintenance and to minimize process waste streams and emissions to the environment. Reliable operation of a 9,000 tpy commercial unit over the last five years has demonstrated and proven the technology. In the basic process, molten sulfur is oxidized to SO<sub>2</sub> vapor in compressed oxygen. The SO<sub>2</sub> vapor is then separated from vaporized sulfur and condensed. Key resources, including molten sulfur, oxygen (O<sub>2</sub>), and caustic, are fed to the process. The process in turn produces liquid SO<sub>2</sub>, steam, and sodium sulfite.

The cryogenic air separation plant provides 99.5% O<sub>2</sub> to the liquid SO<sub>2</sub> plant. The oxygen is produced by liquefying air and then using fractional distillation to separate it into its components. The air separation plant requires inputs of air, electricity, and cooling water and produces, in addition to the O<sub>2</sub>, a small amount of pure nitrogen (N<sub>2</sub>).

### **Liquid SO<sub>2</sub> Plant Description**

The facility will have the operating capacity to produce about 125 tpd (45,000 tpy) of liquid SO<sub>2</sub>. Figure 8 presents a basic flow diagram of the liquid SO<sub>2</sub> process. Primary unit operations are numerically labelled on this figure and referenced in the following discussion. Liquid sulfur at about 270°F is continuously pumped from two 250 ton capacity sulfur storage tanks (1) to the sulfur day tank (2). Sulfur flows by gravity from the day tank to the SO<sub>2</sub> reactor (3). The sulfur level in the reactor is controlled by equalization with the level in the sulfur day tank.

During start up the sulfur in the reactor is electrically heated to about 600°F. Oxygen is then injected into the sulfur through a submerged sparger. The sulfur at the reactor operating pressure, 80 psig and 600°F, is above the auto-ignition temperature. The following reaction occurs:



The reaction is spontaneous and exothermic. The reactor temperature rises to about 1100°F, the boiling point of sulfur at 80 psig. The production rate of SO<sub>2</sub> is controlled by the oxygen feed rate to the reactor.

The vapor stream of SO<sub>2</sub> and sulfur is cooled in the sulfur condenser (4) to about 270°F. The condenser is cooled by generating steam at 35 psig. Most of the sulfur vapor condenses and the mixture of condensed sulfur, which flows by gravity, and SO<sub>2</sub> vapor is returned to the molten sulfur day tank. The liquid sulfur drops out in the sulfur day tank and is recycled to the reactor.

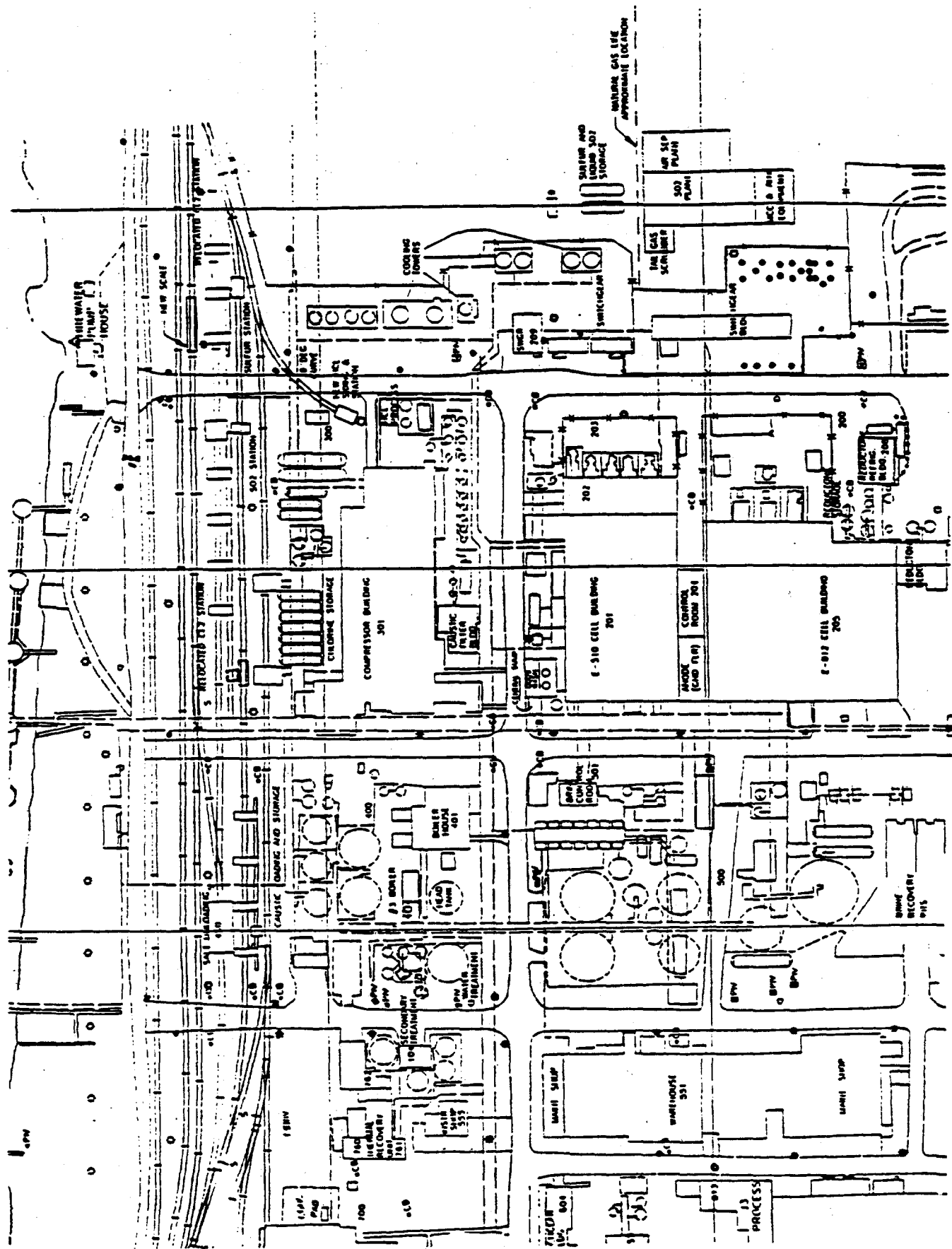


Figure 7. Liquid SO Plant Location

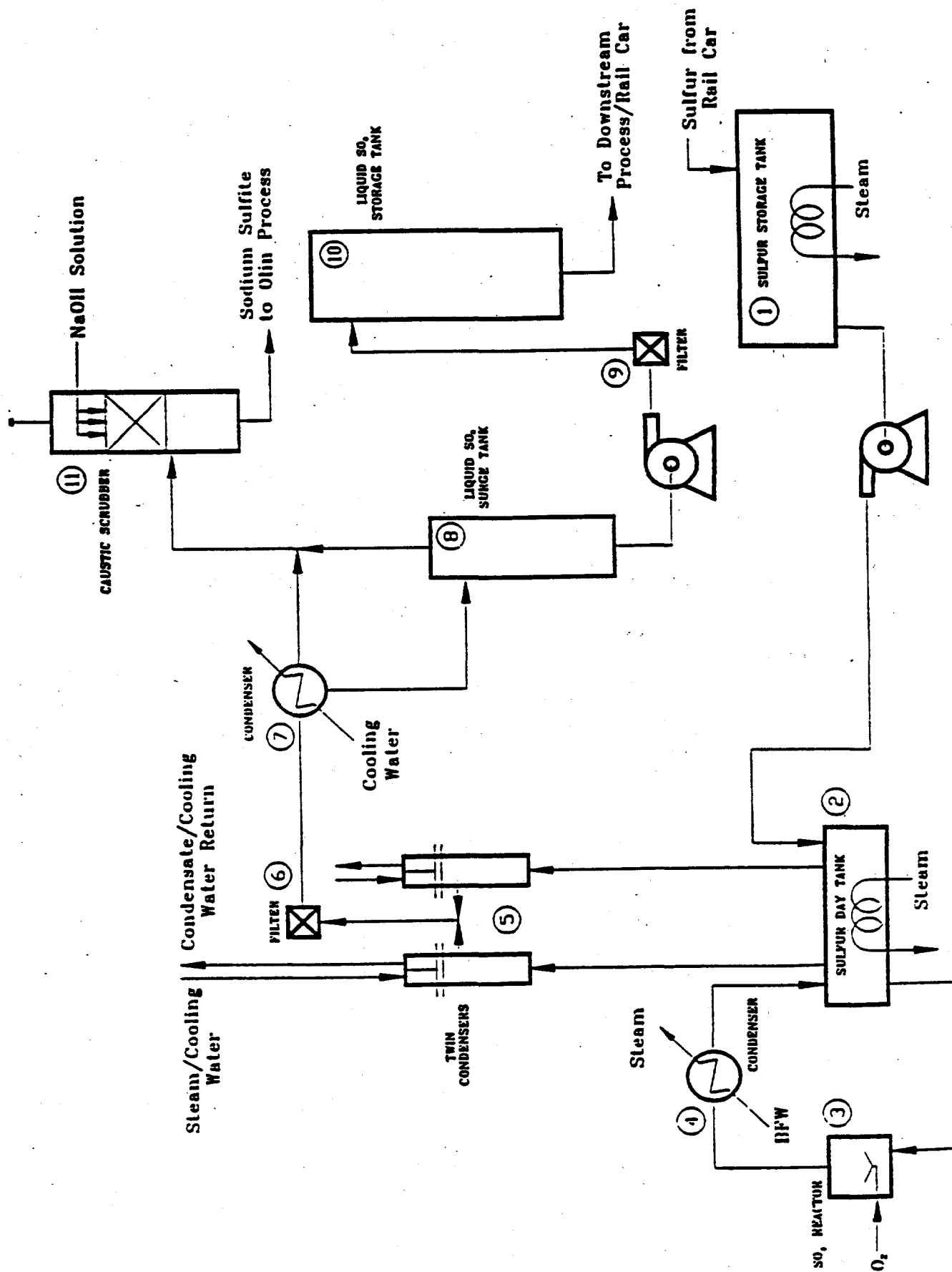


Figure 8. Liquid Sulfur Dioxide Process

The SO<sub>2</sub> vapor does not condense at 270°F and is not significantly soluble in molten sulfur. It is further cooled in the twin condensers (5) to remove additional trace amounts of sulfur. The condensers operate in a two step repeating cycle. In the first step, the condenser cools the SO<sub>2</sub> to 120°F using cooling water. Entrained liquid sulfur and remaining sulfur vapor will collect as a solid on the condenser tube walls. In the second step, the condenser gas outlet is blocked and the sulfur is melted using low pressure steam. The molten sulfur will drain by gravity back to the sulfur day tank. The condensers will alternate between these modes of operation; one condenser will remove sulfur while the second condenser is regenerated using steam.

After filtration (6), the SO<sub>2</sub> vapor is condensed in the SO<sub>2</sub> condenser (7) using cooling water. At the system pressure of 80 psig, the SO<sub>2</sub> condenses at about 104°F. The liquid SO<sub>2</sub> will flow to the liquid SO<sub>2</sub> surge tank (8). From the surge tank it will be pumped through a filter (9) to remove any entrained particulate then to a 200-ton capacity liquid SO<sub>2</sub> storage tank (10). From the storage tank, the liquid SO<sub>2</sub> will be pumped to an existing process liquid SO<sub>2</sub> feed tank or to rail cars for shipment.

A vent stream from the SO<sub>2</sub> condenser and liquid SO<sub>2</sub> surge tank contains non-condensibles, trace amounts of nitrogen and argon introduced to the sulfur reactor with the oxygen, and SO<sub>2</sub> vapor. The SO<sub>2</sub> vapor is removed from the vent stream in a caustic scrubber (11). A sodium hydroxide (NaOH) solution is used to remove the SO<sub>2</sub> vapor from the gas stream. The sodium sulfite formed from the reaction of NaOH and SO<sub>2</sub> will be used by Olin to neutralize a chlorine waste stream from an existing Olin process.

### **Air Separation Plant Description**

Figure 9 presents a basic flow diagram of the air separation plant used to supply O<sub>2</sub> to the liquid SO<sub>2</sub> process. Primary unit operations are numerically labelled on this figure and referenced in the following discussion. As mentioned previously, oxygen is produced by liquefying air and then using fractional distillation to separate the liquefied air into its components. The three fundamental steps in this process are purification, refrigeration, and rectification.

#### ***Purification***

Atmospheric air contains dirt, water vapor, and carbon dioxide (CO<sub>2</sub>) which must be removed from the compressed air stream to prevent plugging of downstream process equipment. The atmospheric air passes through an intake filter (1) to remove entrained particulate and is compressed to 125 psig in a centrifugal compressor (2). After compression the air is cooled in a direct contact after cooler (3) using cooling water. Carbon dioxide, water vapor, and gaseous hydrocarbons are then removed by adsorption on activated alumina and molecular sieve (4). Parallel units are used, like the twin condensers from the liquid SO<sub>2</sub> process, one bed will be regenerated while the other is online. The adsorbents are regenerated using heat and a nitrogen purge gas generated downstream.

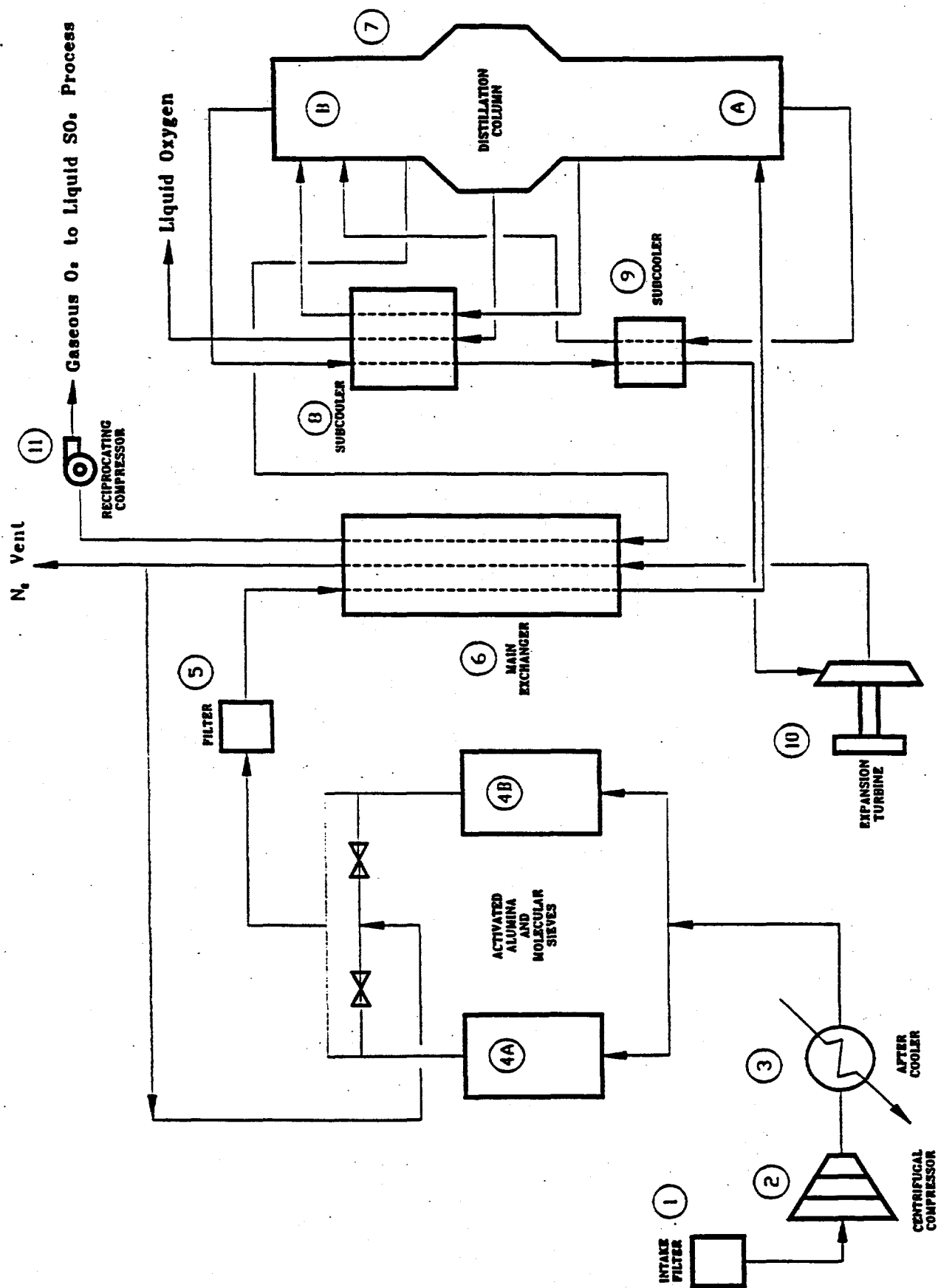


Figure 9. Air Separation Plant

## **Refrigeration/Rectification**

The purified air passes through a pipeline filter (5) and enters the main exchanger (6) where it is cooled by heat exchange with outgoing gaseous oxygen and waste gas. After the main exchanger, the purified air enters the bottom section of the lower column (7A) of the distillation column (7). The lower column operates at about 60 psig while the upper column (7B) of the distillation column operates at about 5 psig. Rectification, vapor - liquid contacting, occurs in the distillation column. As the incoming air rises up the column, it contacts a descending liquid. Since oxygen has a higher boiling point than nitrogen, as the vapor ascends it becomes richer in nitrogen while as the liquid descends it becomes richer in oxygen. Cold nitrogen rich vapor is withdrawn from several places within the distillation column and used to cool recycle streams in the subcoolers (Units 8 & 9). Heat energy is also removed from the system by expanding the nitrogen rich vapor in the expansion turbine (Unit 10), thereby doing work and lowering the temperature. Pure oxygen vapor is withdrawn from the bottom of the upper column. This vapor is warmed in the main exchanger and compressed using a reciprocating compressor (11) to the required operating pressure.

## **Process Alternatives/Advantages**

Traditional, older processes used to produce liquid SO<sub>2</sub> from sulfur involve burning the sulfur in air. The resulting gas stream can contain, at best, 16-18 vol.% SO<sub>2</sub> with the balance being mainly nitrogen, oxygen, and water from the combustion air. The SO<sub>2</sub> must then be separated from the other combustion gases. This is done by stripping the SO<sub>2</sub> from the gas stream using either water or an organic solvent like dimethylaniline. Regardless of which stripping liquor is used, these processes are more complex and have greater environmental impacts. To illustrate, the burn in air with water stripping process.

Process advantages of the burn-in-oxygen liquid SO<sub>2</sub> process include the following:

- Process gas at a lower temperature, 1100°F versus about 2500°F.
- Production of lower pressure steam, 35 psig versus 600 psig.
- No acidic wastewater stream which must be neutralized.
- Smaller volume tail gas stream which economically allows for the use of a more efficient scrubber resulting in lower SO<sub>2</sub> emissions.
- No spent acid stream which must be reclaimed or disposed of.
- No solvent emissions or disposal of solvent.

In addition, due to the lower process gas temperature and steam pressure, and simplicity of the process the liquid SO<sub>2</sub> process is inherently more reliable and safe to operate

#### IV. ACKNOWLEDGEMENT

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## **Integrated Dry NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System Performance Summary**

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### **ABSTRACT**

The Integrated Dry NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System was installed at Public Service Company of Colorado's Arapahoe 4 generating station in 1992 in cooperation with the U.S. Department of Energy (DOE) and the Electric Power Research Institute (EPRI). This full-scale 100 MWe demonstration combines low-NO<sub>x</sub> burners, overfire air, and selective non-catalytic reduction (SNCR) for NO<sub>x</sub> control and dry sorbent injection (DSI) with or without humidification for SO<sub>2</sub> control. Operation and testing of the Integrated Dry NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System began in August 1992 and will continue through 1996. Results of the NO<sub>x</sub> control technologies show that the original system goal of 70% NO<sub>x</sub> removal has been easily met and the combustion and SNCR systems can achieve NO<sub>x</sub> removals of up to 80% at full load. Duct injection of commercial calcium hydroxide has achieved a maximum SO<sub>2</sub> removal of nearly 40% while humidifying the flue gas to a 20°F approach to saturation. Sodium-based dry sorbent injection has provided SO<sub>2</sub> removal of over 70% without the occurrence of a visible NO<sub>2</sub> plume. Recent test work has improved SNCR performance at low loads and has demonstrated that combined dry sodium injection and SNCR yields both lower NO<sub>2</sub> levels and NH<sub>3</sub> slip than either technology alone.

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<sup>1</sup> Currently with BTU Services, Visalia, California



## INTRODUCTION

Public Service Company of Colorado (PSCC) was selected by DOE for a CTT-III project in December 1989 to demonstrate an Integrated Dry NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System. The demonstration project is taking place at PSCC's Arapahoe Unit 4, a 100 MWe top-fired unit which fires a low sulfur (0.4%) Colorado bituminous coal as its main fuel, but also has 100% natural gas capability. Figure 1 shows a boiler elevation drawing.

The Integrated Dry NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System combines five major control technologies to form an integrated system to control both NO<sub>x</sub> and SO<sub>2</sub> emissions. The system uses low-NO<sub>x</sub> burners, overfire air, and urea injection to reduce NO<sub>x</sub> emissions, and dry sorbent injection using either sodium- or calcium-based reagents with (or without) humidification to control SO<sub>2</sub> emissions. The goal of the project was to reduce NO<sub>x</sub> and SO<sub>2</sub> emissions by up to 70%. The combustion modifications were expected to reduce NO<sub>x</sub> by 50%, and the SNCR system was expected to increase the total NO<sub>x</sub> reduction to 70%. Dry Sorbent Injection was expected to provide 50% removal of the SO<sub>2</sub> emissions while using calcium-based reagents. Because sodium is much more reactive than calcium, it was expected to provide SO<sub>2</sub> removals of up to 70%. Figure 2 shows a simplified schematic of the Integrated Dry NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System at Arapahoe Unit 4.

The total cost of this innovative demonstration project is estimated to be \$27,411,000. Funding is being provided by DOE (50%), PSCC (43.7%), and EPRI (6.3%). DOE funding is being provided as a zero interest loan and is expected to be paid back from the proceeds obtained during commercialization of the technology over a 20-year period which begins at the conclusion of the demonstration project.

Construction began in July 1991 and was completed in August 1992. The test program began in August 1992 and all low sulfur coal testing was scheduled for completion in June 1994. Addition of the new SNCR injection location and alternate lance design tests will extend the test program through December 1996. Project completion is currently scheduled for February 1997.

Prior publications presented results of the performance of the individual technologies (1-11). This paper will provide a brief overview of the individual technologies and their performance, but will focus on results from recent test activities. These recent activities have included: 1) testing of a new SNCR injection location to improve low load performance; 2) long term performance of the integrated system; and 3) recent results of ammonia adsorption in the ash.

## TECHNOLOGY DESCRIPTION

This section will provide a brief description of the technologies used in the integrated NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System with emphasis on the SNCR and sodium dry sorbent injection system. The reader is referred to prior publications for more complete descriptions of the low NO<sub>x</sub> combustion system and calcium dry sorbent system with humidification.<sup>(1-11)</sup>

## Low NO<sub>x</sub> Combustion System

B&W's DRB-XCL® (Dual Register Burner-aXially Controlled Low-NO<sub>x</sub>) burner had been successfully used to reduce NO<sub>x</sub> emissions on wall-fired boilers but had never been used in a vertically-fired furnace. The burner utilizes dual registers to control near burner mixing and a sliding air damper to control air flow to each individual burner independent of swirl. Twelve of these burners were installed on the roof of Arapahoe Unit 4. The low NO<sub>x</sub> combustion system also incorporated three B&W dual zone NO<sub>x</sub> ports which were added to each side of the furnace approximately 20 feet below the boiler roof. These ports can inject up to 28% of the total combustion air through the furnace sidewalls.

Arapahoe Unit 4 was originally designed with the ability to fire 100% natural gas. Natural gas firing capabilities were maintained with the DRB-XCL® burners by installing a gas ring header at the tip of the burner. However, the burner is not specifically designed to be a low-NO<sub>x</sub> burner with natural gas firing.

## Selective Non-Catalytic Reduction

The purpose of the SNCR system at Arapahoe was two-fold. First, to further reduce the final NO<sub>x</sub> emissions obtained with the combustion modification so that the goal of 70% NO<sub>x</sub> removal could be achieved. Second, the SNCR system is an important part of the integrated system interacting synergistically with the dry sodium injection system. During this program, it was shown that when both systems are used simultaneously, both NO<sub>2</sub> emissions from the sodium system and NH<sub>3</sub> slip from the SNCR system are reduced.

When the SNCR system was originally designed and installed, it incorporated two levels of wall injectors with 10 injectors at each level. These two separate levels were intended to provide load following capability. The locations of these two levels were based on flue gas temperature measurements made with the original combustion system. However, the retrofit low-NO<sub>x</sub> combustion system resulted in a decrease in the furnace exit gas temperature of nominally 200°F. This decrease in temperature moved the cooler injection level out of the SNCR temperature window. With only one operational injection level, the load-following performance of the system was compromised.

Two approaches were pursued to improve the low load performance of the SNCR system. First, short-term testing showed ammonia to be more effective than urea at low loads. Although ammonia was more effective than urea, it remained desirable to store urea due to safety concerns. A system was installed that allows on-line conversion of urea into ammonia compounds. The on-line conversion system improved low load performance, but the improvement was not as large as desired at the lowest load (60 MWe).

More recently, NOELL, Inc. (the original supplier of the SNCR system) suggested an additional injection location in a higher temperature region of the furnace. Because no unit outages were planned, the only option for incorporating an additional injection level was to utilize two existing (but unused) sootblower ports in conjunction with NOELL's Advanced Retractable Injection

Lances (ARILs). This location was chosen because the ports existed, not because the temperatures were ideal for SNCR.

Figure 3 shows a diagram of the SNCR system installed at Arapahoe Unit 4. The system uses NOELL's proprietary dual-fluid injection nozzles to distribute the urea uniformly into the boiler. A centrifugal compressor is used to supply a large volume of medium-pressure air to the injection nozzles. The large quantity of air helps to atomize the urea solution as well as provide energy to rapidly mix the atomized solution with the combustion products. The SNCR system includes the option of passing the urea solution over a proprietary catalyst which converts the urea to ammonia-based compounds. This ammonia conversion system was not utilized during the current series of tests described in this paper.

Figure 4 shows the location of the new ARIL lances relative to the two original SNCR injection locations. Level 2 is the location that became unusable as a result of the flue gas temperature decrease after the low- $\text{NO}_x$  combustion system retrofit. The ARIL system consists of two retractable lances and two retractable lance drive mechanisms. Each lance is nominally 4 inches in diameter and approximately 20 feet in length. Each lance has a single row of nine injection nozzles spaced on two-foot centers. A single division wall separates the Arapahoe Unit 4 furnace into east and west halves, each with a width of approximately 20 feet. When each lance is inserted, the first and last nozzles are nominally one foot away from the division and outside walls, respectively.

Each injection nozzle is composed of a fixed air orifice (nominally one-inch in diameter), and a replaceable liquid orifice. The liquid orifices are designed for easy removal and cleaning, because they can become plugged. This ability to change nozzles also allows adjustments in the chemical injection pattern along the length of the lance in order to compensate for any significant maldistributions of flue gas velocity, temperature, or baseline  $\text{NO}_x$  concentration.

Two separate internal liquid piping circuits are used to direct the chemical to the individual injection nozzles in each lance. The four nozzles near the tip of the lance are supplied by one circuit, and the remaining five are supplied by the other. This provides the ability to bias the chemical flow between the "inside" and "outside" halves of each side of the furnace in order to compensate for various coal mill out-of-service patterns. Each lance is also supplied with a pair of internal thermocouples for detecting inside metal temperatures at the tip of the lance.

The retractable lance drive mechanisms were supplied by Diamond Power Specialty Co. (DPSC). The drives are Model IK 525's which have been modified for the liquid and air supply parts. Both remote (automatic) and/or local (manual) insertion and retraction operations are accomplished with the standard IK electric motor and gearbox drive system. A local control panel is provided on each side of the boiler, attached to each ARIL lance drive mechanism. Each panel contains a programmable logic controller for the lance install/retract sequencing and safety interlocks. Each lance can be rotated either manually at the panel, or automatically by the control system during load-following operation. One of the key features of the ARIL lance system is its ability to rotate the lances. As will be discussed, this feature provides a high degree of flexibility in optimizing SNCR performance by varying the flue gas temperature at the injection location by simply rotating the lance.

In addition to NOELL's ARIL lances, an alternate lance design, supplied by Diamond Power Specialty Company, was also evaluated. This alternate lance design represented a simplification to the original ARIL design. The liquid solution is injected through a single pressure atomizer located in the air supply pipe ahead of the lance. This eliminates the internal liquid piping, and spraying at the lance inlet provides evaporative cooling to help cool the lance. In addition, the design prevents air and liquid from being injected in the local region around the boiler when the lances are retracted.

### **Dry Sorbent Injection**

PSCC designed and installed a dry sorbent injection system that can inject either calcium- or sodium-based reagents into the flue gas upstream of the fabric filter. Figure 5 shows a simplified flow diagram of the equipment. The reagent is fed through a volumetric feeder into a pneumatic conveying system. The air and material then pass through a pulverizer where the material can be pulverized to approximately 90% passing 400 U.S. Standard mesh. The material is then conveyed to the duct and evenly injected into the flue gas. After the original results suggested that the duct flue gas temperature was too low for effective  $\text{SO}_2$  removal with sodium bicarbonate, the dry injection system was modified to allow injection of sodium-based compounds at the entrance to the air heater where the flue gas temperature is approximately  $600^\circ\text{F}$ . The pulverizer can be bypassed allowing calcium hydroxide to be fed from the silos and injected either ahead of the fabric filter or into the boiler economizer region where the flue gas temperature is approximately  $1000^\circ\text{F}$ .

To improve  $\text{SO}_2$  removal with calcium hydroxide, a humidification system capable of achieving a  $20^\circ\text{F}$  approach to saturation temperature has been installed. The system was designed by B&W and includes 84 I-Jet humidification nozzles which can inject up to 80 gpm of water into the flue gas ductwork. The humidifier is located approximately 100 feet ahead of the fabric filter and there is no bypass duct.

### **Balance of Plant**

Besides the major environmental equipment, the project also included required upgrades to the existing plant. A new distributed control system was installed to control the boiler and other pollution control equipment added as part of the integrated system. The fly ash collection system was also converted from a wet to a dry collection system to allow dry collection of the ash and injection waste products. A Continuous Emissions Monitor (CEM) system was installed at Arapahoe Unit 4 to collect data for the extensive test program. This monitor allows measurements of  $\text{N}_2\text{O}$ ,  $\text{NH}_3$ ,  $\text{NO}_2$ , and  $\text{H}_2\text{O}$  in addition to the more common pollutants.

### **RESULTS**

Fossil Energy Research Corporation (FERCo) of Laguna Hills, California, has been performing all testing of the Integrated Dry  $\text{NO}_x/\text{SO}_2$  Emissions Control System. The test program is nearing

completion and the individual testing of the low-NO<sub>x</sub> burners, overfire air, urea injection, calcium duct injection, calcium economizer injection, and sodium injection has been completed. Testing of the SNCR lances and the complete integrated system while firing low-sulfur coal is in progress. In addition to efficiency and emissions measurements, four tests were conducted to determine baseline and removal capabilities of the system for many common air toxic emissions. Prior papers (1-11) also discussed the performance of the individual technologies. This paper will overview these prior results and focus on the performance of the new SNCR injection lances and the performance of the integrated system. In addition, the paper will also present some recent information on NH<sub>3</sub> absorption on fly ash and the impacts on fly ash handling.

## **Review of NO<sub>x</sub> and SO<sub>2</sub> Reduction Performance**

This section will provide a brief overview of the NO<sub>x</sub> and SO<sub>2</sub> reductions from the individual technologies used in the integrated system. The reader is referred to prior publications for more detailed discussions (1-11).

### **Low-NO<sub>x</sub> Combustion System Performance**

The following section describes the performance of the low-NO<sub>x</sub> combustion system (low-NO<sub>x</sub> burners and OFA ports).

#### ***Low-NO<sub>x</sub> Burners***

Figure 6 compares the Arapahoe Unit 4's NO<sub>x</sub> emissions before and after the retrofitting of the low-NO<sub>x</sub> combustion system. Note, NO<sub>x</sub> (NO + NO<sub>2</sub>) and NO are used interchangeably when discussing the performance of the combustion system, since NO<sub>2</sub> levels are very low for this unit. The original combustion system produced nearly uniform NO<sub>x</sub> emissions of 800 ppmc (corrected to 3% O<sub>2</sub>, dry) or 1.1 lb/MMBtu across the boiler's load range. The low-NO<sub>x</sub> combustion system reduced NO<sub>x</sub> emissions by over 63 percent, to less than 300 ppmc, across the boiler's load range. Note that all testing was conducted under steady-state conditions and with the careful supervision of test technicians to achieve the maximum possible reduction in NO<sub>x</sub> emissions. Under load following conditions, NO<sub>x</sub> emissions were about 10 to 25% higher. Additional modifications to the control system and additional operator training may decrease the difference.

#### ***Overfire Air***

At full load, opening the OFA control dampers to full open (maximum OFA) diverts about 24% of the total combustion air to the OFA ports and at low load (60 MWe) it diverts about 28% of the combustion air. At maximum OFA, the low-NO<sub>x</sub> combustion system reduces NO<sub>x</sub> emissions by 62 to 69% across the load range. Since the OFA ports are located in a very hot part of the boiler, a significant amount of cooling air is required and the minimum amount of OFA is limited to about 15% of the total combustion air at full load and about 8% at low load. At minimum OFA, the retrofitted combustion system reduces NO<sub>x</sub> emissions by 60 to 63%. Arapahoe Unit 4

cannot be tested at 0% OFA, but the small difference in  $\text{NO}_x$  reduction between maximum and minimum OFA indicates that the low- $\text{NO}_x$  burners are responsible for most of the  $\text{NO}_x$  reduction.

### **SNCR System Performance**

As mentioned previously, in addition to reducing  $\text{NO}_x$  emissions significantly, the low- $\text{NO}_x$  combustion system also reduced the temperature of flue gas at the furnace exit by about  $200^\circ\text{F}$ . Since SNCR systems are very sensitive to changes in flue-gas temperatures, this reduction made the flue-gas temperature too cold for one row of injection nozzles, so all testing was performed using the row of injection nozzles originally designed for loads below 80 MWe. Figure 7 shows the SNCR performance achievable over the load range for a 10 ppm  $\text{NH}_3$  slip limit with this single row of injectors. At full load,  $\text{NO}_x$  reductions of 45% are achieved. However, the performance decreased dramatically as the load decreased; at 60 MW,  $\text{NO}_x$  removals were limited to 11% for a 10 ppm  $\text{NH}_3$  slip.

### **Calcium-Based Economizer Injection Performance**

$\text{SO}_2$  removal has been less than expected with calcium hydroxide injection at the economizer. Initial testing at a Ca/S molar ratio of 2.0 without humidification resulted in  $\text{SO}_2$  removals in the range of 5 to 8%. Note that the stoichiometric ratio for the Ca/ $\text{SO}_2$  reaction is 1.0, since one mole of Ca reacts with one mole of  $\text{SO}_2$  to form calcium sulfate,  $\text{CaSO}_4$ . It was found that the sorbent distribution was very poor, and only approximately one-third of the flue gas was being treated. New nozzles that increased reagent distribution only increased the  $\text{SO}_2$  removal to 15% at a Ca/S molar ratio of 2.0. Although distribution of the calcium reagent is far from perfect, it appears that high levels of  $\text{SO}_2$  removal are not possible at Arapahoe Unit 4 using the current  $\text{Ca}(\text{OH})_2$  material, even in areas with high Ca/S molar ratios.

### **Calcium-Based Duct Injection Performance**

Higher  $\text{SO}_2$  removal was achieved with duct injection of calcium hydroxide and humidification with  $\text{SO}_2$  removals approaching 40% at Ca/S ratios of 2.0 and approach to saturation temperatures of 20 to  $30^\circ\text{F}$ . These levels of  $\text{SO}_2$  removal are consistent with the prior DOE study at Ohio Edison's Edgewater Station.<sup>[13]</sup> Immediately after this test, problems developed with the dry fly ash transport system, and it is suspected that the low approach temperature contributed to this problem. Then, after a short period of 24 hour/day testing during load following operation, fabric filter pressure drop significantly increased due to the buildup of a hard ash cake on the fabric filter bags which could not be cleaned during normal reverse air cleaning. The heavy ash cakes were caused by the humidification system, but it was not possible to determine if the problem was caused by operation at a  $30^\circ\text{F}$  approach temperature or a short-excursion to a lower approach temperature caused by a rapid decrease in boiler load.

## **Sodium-Based Injection**

Sodium-based reagents are much more reactive than calcium-based sorbents and can achieve significantly higher SO<sub>2</sub> removals during dry injection.<sup>[14,15]</sup> Figure 8 shows the SO<sub>2</sub> removal for dry sorbent injection for sodium bicarbonate and sodium sesquicarbonate. In Figure 8, SO<sub>2</sub> removals are plotted as a function of Normalized Stoichiometric Ratio (NSR). This corresponds to the amount of sodium compound injected relative to the amount of sodium required to form sodium sulfate, Na<sub>2</sub>SO<sub>4</sub> (i.e., two moles of Na per mole SO<sub>2</sub>). Sodium bicarbonate provided the highest SO<sub>2</sub> removal and was also the most efficient reagent in terms of sodium utilization. Flue gas temperature at the fabric filter inlet duct at Arapahoe Unit 4 varies from 250 to 280°F. The dry sorbent injection (DSI) system was originally designed for duct injection before the fabric filter only. However, initial testing with sodium bicarbonate showed that SO<sub>2</sub> removal was erratic, which was attributed to the low flue gas temperatures. The DSI system was modified to inject sodium sorbents at the air heater inlet where the flue gas temperature is approximately 600°F. It should be noted that sodium sesquicarbonate does not exhibit this slow reaction rate when injected ahead of the fabric filter.

A major disadvantage of sodium-based injection is that it converts some existing NO in the flue gas to NO<sub>2</sub>. In addition, during the conversion process a small amount of the total NO<sub>x</sub>, 5 to 15%, is removed. However, the net NO<sub>2</sub> exiting the stack is increased. While NO is a colorless gas, small quantities of the brown/orange NO<sub>2</sub> can cause a visible plume to develop. The chemistry of the conversion is not well understood but it is generally accepted that NO<sub>2</sub> increases as SO<sub>2</sub> removal increases. Figure 8 shows that NO<sub>2</sub> emissions are generally higher with sodium bicarbonate, although a significant amount of data scatter exists. The threshold NO<sub>2</sub> level that forms a visible plume is site specific; at Arapahoe Unit 4, a visible plume appears when NO<sub>2</sub> concentrations reach 30 to 35 ppm. Also, the NO<sub>2</sub> levels were found to depend on conditions in the fabric filter with NO<sub>2</sub> levels increasing dramatically after each cleaning cycle.<sup>(11)</sup>

## **SNCR Lance Performance Results**

The recent test work has focused on the performance of the SNCR lances, both the NOELL ARIL lances and a comparison of the performance of the alternate DPSC lance to the ARIL lance.

### **ARIL Lances**

Prior to incorporating the ARIL lances into the SNCR control system, a series of parametric tests was conducted to define the optimum injection angle at each load. As shown in Figure 4, each lance can rotate to inject urea into a different region of the furnace in order to follow the SNCR temperature window as the boiler load changes. The minimum injection angle is 22° (0° corresponds to injection vertically downward), at which point the chemical is injected parallel to the tube wall located below the lances. Smaller injection angles are not used to avoid direct liquid impingement on these tubes. An injection angle of 90° corresponds to injection

straight across the furnace toward the front wall, and an angle greater than 90° results in injection of the solution in a direction up toward the roof-mounted burners.

While the primary focus of the parametric tests was to define the injection angle versus load, the tests also investigated the effects of:

- coal mill out-of-service patterns
- coal mill biasing
- biasing the urea flow along the length of the lances
- independent adjustment of the injection angles for each lance

The results of these tests are described below.

### ***Effect of Lance Angle***

One of the primary attributes of the ARIL lance system is the inherent flexibility of accessing the optimum flue gas temperature location by simply rotating the lance. Figure 9 shows the effect of varying the lance injection angle at loads of 43 and 50 MWe. All of the tests shown in these figures were performed at a N/NO<sub>x</sub> ratio of 1.0, with two mills in service. At 43 MWe, varying the injection angle had little effect on NO<sub>x</sub> removal, and the maximum removal occurred at an angle of 35 degrees (Figure 9a). However, Figure 9a shows that the lance angle had a large effect on NH<sub>3</sub> slip; decreasing from 46 ppm at an angle of 22° to under 5 ppm at an angle of 135°. This overall behavior at 43 MWe suggests that, on average, injection is occurring just on the high side of the SNCR temperature window. In fact, the optimum temperature, in terms of NO<sub>x</sub> removal, appears to correspond to an angle of 35°. However, since it is desirable to maintain the NH<sub>3</sub> slip less than 10 ppm, an injection angle of 90° is a more appropriate operating angle at this load.

At a slightly higher load of 50 MWe (Figure 9b), the effect of lance injection angle was markedly different. At this load, where the average flue gas temperatures were higher, injection angle had little effect on NH<sub>3</sub> slip. However, at the higher temperature, lance angle had a large effect on NO<sub>x</sub> removal. The relative insensitivity of the NH<sub>3</sub> slip and large sensitivity of the NO<sub>x</sub> removal to lance angle suggests that at 50 MWe, chemical injection is occurring far on the high side of the SNCR temperature window for injection angles ranging from 22° to 135°.

The results at 43 and 50 MWe shown in Figure 9 illustrate how varying lance angle can be used to optimize the SNCR performance over the load range. As the load increases, the preferred injection angle will decrease. Again, the minimum angle is 22°, where the chemical is injected parallel to the tube sheet located below the lances.

### ***Performance over the Load Range***

The SNCR performance using the ARIL lances over the load range from 43 to 80 MWe is shown in Figure 10. Note that for this particular lance location, the flue gas temperatures are too high



for the lances to be effective at loads greater than 80 MWe. As the load increases, the preferred lance angle decreases in order to inject the urea into a lower temperature region.

As discussed above, at 43 MWe with an angle of  $90^\circ$ , injection occurred on average just on the high temperature side of the window. At  $N/NO_x = 1$ ,  $NO_x$  removals were 35% with less than 10 ppm  $NH_3$  slip. At 50 MWe, a  $45^\circ$  injection angle was on average at a better location in the SNCR window, with  $NO_x$  removals of 40% and  $NH_3$  slip less of 5 ppm at  $N/NO_x = 1$ . As the load increased to 60 MWe, a decrease in lance angle to  $34^\circ$  resulted in SNCR performance similar to a load of 43 MWe. At higher loads of 70 and 80 MWe, injection was clearly occurring on the high side of the temperature window. Note that the  $NH_3$  slip at 80 MWe was higher than the slip at 70 MWe even though the chemical was injected into a region of higher overall temperature (i.e., compare the  $NO_x$  removals at 70 and 80 MWe in Figure 10). This effect was a result of temperature stratification in the furnace, and the way in which the stratification varies with different coal mill patterns. This effect is discussed in more detail below. However, comparing Figures 9 and 10 to the low load performance of the wall injectors in Figure 7 clearly shows that the lances have markedly improved the low-load performance of the SNCR system.

### ***Effect of Boiler Operation on SNCR Performance***

As mentioned above, local changes in temperature due to variations in boiler operating parameters (excess  $O_2$ , mill pattern, mill biases, etc.) can have a major impact on SNCR performance. This is particularly true at Arapahoe Unit 4 where the 12 burners are located on the roof of the furnace. Each of the four coal mills feeds three burners, two burners on one side of the furnace and a single burner on the other side of the furnace. Since the furnace has a division wall, there is an imbalance in heat release across the furnace, and a corresponding variation in flue gas temperature, when only three mills are in service. These temperature variations impact the performance of both the wall injectors and the ARIL lances. In this paper, the effect will be illustrated by looking at the performance of the ARIL lances with varying mill out-of-service patterns. During normal operation, Arapahoe Unit 4 operates with four mills in service over the load range from 80 to 110 MWe (although the unit can operate up to 100 MWe with only 3 mills). From 60 to 80 MWe, the unit typically operates with three mills in service. Below 60 MWe, the unit is usually operated with only two mills in service.

Figure 11 shows the effect of various mill out-of-service (OOS) patterns on east/west imbalances across the furnace. The bottom of Figure 11 shows a plan view of the in-service burners (numbered) and out-of-service burners (filled circles) for a given mill pattern. Note that the left side of these figures corresponds to the west wall of the furnace (adjacent to burners 1, 2 and 3), and the right side corresponds to the east wall (adjacent to burners 10, 11 and 12). With either A mill or C mill out-of-service, more heat release occurs on the east side of the furnace, while the west side has more heat release with either B mill or D mill out-of-service.

The change in lance metal temperatures provides a general indication of changes in flue gas temperatures on the east and west sides of the furnace. As seen in Figure 11, the changes in lance metal temperatures reflect the variations in heat release in the furnace with differing mill out-of-service patterns. Correspondingly, the  $NO_x$  removals and  $NH_3$  slip levels also reflect these

variations in temperature. For instance,  $\text{NH}_3$  slip decreased on the west side when D mill was out-of-service, since more coal was fired (and the flue gas temperatures were higher) on the west side. The lance metal temperatures also indicated that, in general, the east side of the furnace was hotter than the west side. Figure 12 shows the overall impact of various mill out-of-service patterns on SNCR performance at 60 MWe. As can be seen,  $\text{NO}_x$  removals varied from 30% to 52% (@  $\text{N}/\text{NO}_x = 1.5$ ) depending on which particular mill was out-of-service. Comparably, the  $\text{NH}_3$  slip varied from under 5 ppm to over 30 ppm with different mill-in-service patterns. This behavior made overall optimization of the SNCR system quite challenging.

In addition to the temperature variations that occur with the various mill out-of-service patterns, day-to-day variations can occur as a result of changes in the performance of the individual mills, or changes in any other variables which affect the flue gas temperature distribution. Three operational changes were investigated to deal with these types of temperature variations.

- varying the urea flow along the length of each lance
- independently varying each lance angle
- biasing the in-service coal mills

Varying the urea flow between the two liquid zones in each lance provided minor improvements in the performance of the SNCR system. Independently varying the lance angles as a function of the mill-in-service pattern also provided minor improvements. Unfortunately the implementation of either of these strategies would significantly complicate the automatic control system. On the other hand, biasing the in-service coal mills, which is relatively easy to implement, resulted in major improvements in the performance of the SNCR system. Arapahoe Unit 4 is equipped with four  $\text{O}_2$  monitors at the economizer exit. Biasing the coal mills to provide a balanced  $\text{O}_2$  distribution at this location is a fairly simple exercise for the boiler control operator. Figure 13 shows the improvements in SNCR performance that can be achieved by biasing the coal mills. These tests were performed at a load of 60 MWe with both lances at an injection angle of  $22^\circ$  and A mill out-of-service. The "biased" condition in Figure 13 corresponds to a negative 10% bias on B mill and D mill, and a positive 10% bias on C mill. This has the net effect of moving coal from the east side to the west side of the furnace to compensate for A mill being out-of-service (see bottom of Figure 11). Biasing the mills increased  $\text{NO}_x$  removals from nominally 27% to 42% at an  $\text{NH}_3$  slip limit of 10 ppm.

### ***Overall System Performance***

The parametric tests were conducted to determine at which loads the ARIL lances should be used, as well as the optimum injection angle for each of these loads. Based on the parametric tests, the control system has been set up to operate with the Level 1 wall injectors at loads above 80 MWe. Below 80 MWe, the ARIL lances are used. Figure 14 compares the  $\text{NO}_x$  removal over the load range for injection at the two locations with an  $\text{NH}_3$  slip limit of 10 ppm. It is evident that the installation of the ARIL lances has improved low-load performance of the SNCR system. Currently,  $\text{NO}_x$  removals of more than 30% are achievable over the load range with less than 10 ppm  $\text{NH}_3$  slip. The minimum  $\text{NO}_x$  removal of 30% occurs at 80 MWe, which corresponds to the point where the temperature becomes too high for the ARIL lances and too low for the Level 1

injectors. With continuing operation of the system, it is anticipated that further optimization will take place as the operators gain more experience balancing the furnace.

### **Alternate Lance Design**

While the  $\text{NO}_x$  removal performance of the ARIL lances has been good, their location in the furnace has resulted in some operational problems. At this particular location in the furnace, the lances are exposed to a large differential heating between the top and bottom surfaces. The top surface receives a high radiant load from the burners, while the bottom of the lance radiantly communicates with the relatively cold tube wall immediately below. This uneven heating pattern causes a great amount of thermal expansion along the upper surface, and the lances bend downward toward the tubes. Within 30 minutes of insertion, the tip of each lance would drop by approximately 12 to 18 inches. Within less than six weeks of operation, the lances became permanently bent, making insertion and retraction difficult. This was partially addressed by adding additional cooling slots at the end of the lance.

An alternate lance design supplied by Diamond Power Specialty Company (DPSC) was evaluated during this test period. As mentioned previously, this design sprays the urea solution through a single atomizer at the entrance to the lance. This provides evaporative cooling to supplement the air cooling. The evaporative cooling was expected to help minimize the lance bending discussed above. This alternate lance design was evaluated by installing a single lance on the west side of the boiler in place of one of NOELL's ARIL lances. The two different lance designs were compared during a nominal three week test program.

Overall, the DPSC lance performed mechanically well. The lance exhibited less bending than the ARIL lance, and eliminated air injection on the outside of the boiler.

Figure 15 compares the performance of the ARIL and DPSC lances on the west side of the furnace. In terms of  $\text{NO}_x$  reduction and  $\text{NH}_3$  slip performance, the DPSC lance was not quite as good as the ARIL lance. With the B mill OOS, the ARIL lance yielded 42 percent  $\text{NO}_x$  removal with less than 5 ppm slip on the west side (60 MWe,  $\text{N}/\text{NO}_x = 1$ ). Under comparable conditions, the DPSC lance yielded 36 percent  $\text{NO}_x$  removal and less than 5 ppm slip. This slight difference in performance is primarily attributable to the urea distribution along the lance. The ARIL lance uses a separate liquid circuit with individual liquid orifices at each air nozzle. This results in a fairly uniform liquid distribution along the length of the lance. The DPSC lance, on the other hand, sprays the urea solution into the cooling air stream at the inlet to the lance. Impingement on the walls and incomplete evaporation results in the liquid tending to be carried toward the far end of the lance, with part of the urea exiting as a stream of liquid rather than a finely atomized spray. In fact, this explains why the optimum angle for the DPSC lance is  $34^\circ$  compared to  $22^\circ$  for the ARIL lance at 60 MWe. The higher temperature associated with the  $34^\circ$  angle is needed to evaporate the liquid stream. In addition, the feed tube geometry of the DPSC lance created an additional pressure drop, restricting the amount of cooling air flow. This resulted in less penetration of the air jets, although this was partially compensated for by the unatomized portion of the urea solution, which carried the urea farther into the furnace before decomposing and releasing the reactive nitrogen components.

Overall, the results of the short test program of the DPSC lance were sufficiently positive that a second DPSC lance has been ordered. An additional three weeks of testing is planned.

### Integrated System Performance

An important part of the test program was demonstrating the integrated performance of the various  $\text{NO}_x$  reduction and  $\text{SO}_2$  removal technologies. In particular, a key element of the program was documenting the synergistic benefits of simultaneous operation of the SNCR and sodium-based dry sorbent injection system. When operated together, it was expected that the SNCR system would reduce  $\text{NO}_2$  emissions from the sodium DSI system, while the sodium DSI system would in turn reduce  $\text{NH}_3$  slip from the SNCR system.

Ideally, it would have been desirable to parametrically evaluate the merits of the integrated system over a range of operating conditions. This was not entirely possible for a number of reasons. With sodium-based dry sorbent injection,  $\text{NO}_2$  levels are not only dependent on the amount of sodium injected, but also depend on the particulate in the fabric filter and the cleaning intervals.<sup>(11)</sup> Likewise, the time required for  $\text{NH}_3$  levels to stabilize at the exit of the fabric filter, both before and after sodium injection, was greater than the 10-hour a day period during which the load from Arapahoe 4 could be blocked. Thus, characterizing the integrated performance relied on a limited number of parametric tests followed by a series of "long term" tests under normal load following conditions. During these "long-term" tests, the  $\text{NO}_x$  reduction and  $\text{SO}_2$  removal systems were operated in automatic while the unit was operated according to system dispatch requirements. Data were collected at regular intervals using a data logger. No effort was made to set up specific test conditions, as these tests were designed to simulate operation of these systems once they are turned over to the plant at the completion of this program.

The results of a parametric test with sodium sesquicarbonate injection and the SNCR system are shown in Figure 16. During these tests, the DSI system was started first, followed by the SNCR system. For this test, the DSI system was set at an NSR of 2.0 (i.e., 4 moles of sodium per mole of  $\text{SO}_2$ ) and the SNCR system at  $\text{N}/\text{NO}_x = 0.6$ . Following the start of the DSI system, the  $\text{SO}_2$  removal stabilizes at nominally 70% removal and the  $\text{NO}_x$  removal at 12%. This level of  $\text{NO}_x$  removal is consistent with previous tests of the DSI system. The  $\text{NO}_2$  levels increased to only about three ppm at the point that the SNCR system was started. With the SNCR system started, the  $\text{NO}_x$  removal increased to 35 to 40% and the  $\text{NO}_2$  levels remained constant at three ppm. Even following a cleaning cycle, the  $\text{NO}_2$  levels did not increase with the SNCR system in operation. Just before 1800 hours, the SNCR system was turned off and an immediate increase in  $\text{NO}_2$  was noted.

Figure 17 shows the results of a parametric test with sodium bicarbonate injection ahead of the air preheater. With sodium bicarbonate injection alone at an NSR of 1.1,  $\text{NO}_2$  levels on the order of 50 ppm are expected (see Figure 8). For the test results shown in Figure 16, the SNCR system was started at  $\text{N}/\text{NO} = 1.1$  nominally two hours before the DSI system. As can be seen, the  $\text{NO}_2$  levels remained near zero for the entire test. Further, it can be seen that following the start of the DSI system, the  $\text{NH}_3$  slip levels continued to decrease.

The results shown in Figures 16 and 17 clearly show that there is a synergistic benefit of operating the SNCR and sodium-based DSI systems simultaneously.

Because of the difficulties encountered running these short term integrated tests, the balance of the integrated tests were run under normal load following conditions. During these tests the integrated system was operated 24 hours per day. Figure 18 shows the data collected during one 24-hour period (February 25, 1996). During these tests, the integrated system was utilizing sodium sesquicarbonate injection ahead of the fabric filter, and the SNCR system was load following with both the wall injectors and ARIL lances.

On this day, the boiler load was nearly constant for the first 17 hours of the day. The  $N/NO_x$  ratio and  $NH_3$  emissions were also relatively steady during this time. At 1600 hours, the DSI system was started with a 75 percent  $SO_2$  removal setpoint with the hope that the load would remain steady and it would be possible to assess the beneficial effects of running the integrated system. Although, the load increased significantly about two hours after the DSI system was started, it eventually settled back down to a level similar to the level before the increase. Figure 18 shows that the average  $NH_3$  emissions with and without sodium injection were similar, which was expected since the  $NH_3$  trim control was functioning during both of these tests. However, the results also show that there was a substantial increase in the  $N/NO_x$  ratio. Since the SNCR control system was set to maintain the  $NH_3$  emissions within the range of 7 to 8 ppm, it should have increased the urea injection rate if the DSI system reduced  $NH_3$  emissions. A temporary increase was expected as a result of the load swing, but the  $N/NO_x$  ratio should have returned to the pre-swing level within two to three hours (as was seen after the "morning demand peak" between 0800 and 0900 hours). When the DSI system was started at 1600 hours, there was an immediate 10 percent increase in the  $NO_x$  removal, which is consistent with the increases seen during sodium-based DSI-only tests. After this initial  $NO_x$  removal increase, there was another slower increase (amounting to nominally 10 to 15 percent removal) which occurred as the  $N/NO_x$  ratio increased. Although the scaling of the data makes it difficult to see, Figure 18 indicates that the  $N/NO_x$  ratio basically doubled after the DSI system was started. The increase in  $N_2O$  emissions (from nominally 8 to 16 ppm), confirms that the  $N/NO_x$  ratio was increased by roughly a factor of two. These results clearly indicate that there was a substantial reduction in the stack  $NH_3$  slip, when the SNCR and DSI systems were run concurrently.

Figure 19 shows data collected during the 24-hour period on March 4, 1996. The DSI system was operated for the entire period and the SNCR system was started at 1420 hours. The boiler load was fairly steady at this time, and was low enough for the control system to insert the ARIL lances. Although the DSI feedrate was not very consistent, Figure 18 shows that there was nominally a 50 percent reduction in the  $NO_2$  emissions when urea injection began. The load remained steady for nearly four hours; then it increased for the usual "evening demand peak" at 1800 hours. When the lances retracted, the  $N/NO_x$  ratio dropped as demanded by the control system, and the  $NO_2$  emissions were also seen to decrease. By 1900 hours, the  $NO_2$  emissions had been reduced to near-zero levels. This effect is due to the difference in the  $NH_3$  emissions between injection at the Level 1 and ARIL locations. Although effort was made to set up the SNCR control system such that the  $NH_3$  slip was limited to 10 ppm throughout the load range, the Level 1 location is "cooler" overall than the ARIL location; thus injection at Level 1 is more sensitive to variations in the flue gas temperature profile. Therefore, in general, urea injection at

the Level 1 location results in higher  $\text{NH}_3$  slip levels at the fabric filter inlet. Since the  $\text{NH}_3$  emissions are generally higher with urea injection at the Level 1 location, it would be expected that the reduction in stack  $\text{NO}_2$  emissions would also be higher (relative to injection at the ARIL location). The hypothesis is further supported by the decrease in  $\text{NO}_2$  emissions seen when the urea injection switched from the lances to Level 1 at 1800 hours in Figure 19, but also by the increase in  $\text{NO}_2$  seen when the lances were reinserted at 2000 hours. When the lances went in at this time, the  $\text{NO}_2$  emissions were essentially zero. After an hour, however, the  $\text{NO}_2$  emissions slowly began to increase, finally leveling out at approximately 8 ppm.

The above (Figures 16 through 19) demonstrate the synergistic benefits of the integrated process. The  $\text{NH}_3$  slip from the SNCR process suppresses the  $\text{NO}_2$  emissions associated with  $\text{NO}$  to  $\text{NO}_2$  oxidation by dry sodium injection. Concurrently, the sodium reduces the  $\text{NH}_3$  slip from the SNCR process. (Note: In the present case, the control system adjusts the urea injection rate to maintain a set  $\text{NH}_3$  slip level, and the tendency to reduce  $\text{NH}_3$  slip is manifested in a higher  $\text{N}/\text{NO}_x$  ratio for a given  $\text{NH}_3$  slip.)

### **Ammonia Absorption on the Fly Ash**

An issue that needs to be addressed with any post-combustion  $\text{NO}_x$  reduction technology with  $\text{NH}_3$  slip is the absorption of ammonia on the fly ash. This can have a number of impacts ranging from personnel safety while handling the ash, odor problems, or impacting the salability of the ash for future use as a cement aggregate. In the latter, a salable product becomes a disposal problem with an attached economic penalty. At the Arapahoe Station, the ash is not sold for use in cement. Thus, the only problems that have been encountered have been an occasional  $\text{NH}_3$  odor around the ash handling area and potential concern with worker safety should the concentrations become too high.

At Arapahoe Unit 4, ash is removed from the fabric filter hoppers with a vacuum system and transported dry to an ash silo. When loaded onto trucks for transport to the disposal site, the ash is wetted with about 20% water (by weight) in order to minimize fugitive dust emissions. Depending upon the specific ash characteristics, this wetting process can result in the release of  $\text{NH}_3$  vapors from the ash. Whether or not  $\text{NH}_3$  is released from the ash depends primarily on the pH of the aqueous phase on the surface of the ash particles. As the pH increases above a level of 9 to 9.5, there is an increased release of vapor-phase ammonia.

During the test program with urea injection alone, the ammonia concentration in the ash varied over the range of 100 to 200 ppm (measured on a weight basis). The ash ammonia content appeared to be primarily related to the  $\text{NH}_3$  slip levels from the SNCR system and, to some extent, the fabric filter cleaning cycles. During long-term testing with the SNCR system alone, and a 10 ppm  $\text{NH}_3$  slip limit at the stack, there were no incidents of excessive  $\text{NH}_3$  odors during the ash handling process.

Testing has shown that when the SNCR system is operated in conjunction with the dry sodium injection system, the urea injection rate could be increased substantially while maintaining a 10 ppm  $\text{NH}_3$  slip level at the stack. This is one of the synergistic benefits of the patented Integrated

Dry NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System discussed above. However, during these tests, the ammonia concentration in the ash increased to the range of 400 to 700 ppm (weight basis), and there were frequent occurrences of NH<sub>3</sub> odors at the ash silo. Reducing the NH<sub>3</sub> slip set point to the range of 4 to 5 ppm reduced the ammonia concentration of the fly ash down to the 100 to 200 ppm range (weight basis), but the odor problem persisted.

At first, it was thought that the odor problem was a result of the sodium changing the pH of the ash. The pH resulting from placing 0.5 gram of ash in 200 ml of distilled water was 9.3 for an ash sample without sodium injection. The same test run with an ash sample from a test with sodium injection resulted in a pH of 10.3. While the sodium did indeed increase the pH, which in turn would tend to release more NH<sub>3</sub> from the aqueous to the vapor phases, the pH difference did not appear significant enough to account for the ash handling problems encountered.

An interesting observation was made during the pH measurements. While the presence of sodium was found to slightly increase the final pH, it was also found to have a large effect on the rate at which the pH changed as the ash was wetted. Figure 20 shows the change in pH versus time after 0.5 gram of ash is placed into 200 ml of distilled water and stirred. With the coal ash alone, almost 30 minutes are required for the soluble components of the ash to dissolve and change the pH to a final value of 9.3. However, with sodium present in the ash sample, the pH develops almost instantaneously, presumably because of the higher solubility of the sodium compounds in the ash. This more rapid development of the high pH level can result in more rapid and localized release of the ammonia vapor, and may explain the odor problem encountered when concurrently operating the SNCR and sodium systems. Other than decreasing the level of NH<sub>3</sub> slip from the SNCR system, additional approaches to dealing with this issue have not been explored.

## CONCLUSIONS

Public Service Company of Colorado, in cooperation with the U.S. Department of Energy and the Electric Power Research Institute, has installed the Integrated Dry NO<sub>x</sub>/SO<sub>2</sub> Emissions Control System. The system has been in operation for over three years and preliminary conclusions are as follows:

- NO<sub>x</sub> reduction during baseload operation of the unit with low-NO<sub>x</sub> burners and overfire air ranges from 63 to 69% with no increase in unburned fly ash carbon or CO emissions.
- With the addition of retractable lances to the SNCR system, improved low load performance of the system urea-based SNCR injection allows an additional 30 to 52% NO<sub>x</sub> removal with an ammonia slip limit of 10 ppm at the fabric filter inlet. This increases total system NO<sub>x</sub> reduction to greater than 80% at full load, significantly exceeding the project goal of 70%.
- The ability to follow the temperature window by rotating the ARIL lances has been demonstrated and also proved to be an important feature in optimizing the performance of the SNCR system.

- SO<sub>2</sub> removal with calcium-based dry sorbent injection into the boiler at approximately 1000°F flue gas temperature was disappointing with less than 10% removal achieved.
- SO<sub>2</sub> removal with calcium-based dry sorbent injection into the fabric filter duct has been less than expected with a maximum short term removal rate approaching 40%.
- Sodium bicarbonate injection before the air heater has been very effective with short term SO<sub>2</sub> removals of over 80% possible. Longer term testing has demonstrated removal near 70% at an approximate NSR of 1.0.
- Sodium sesquicarbonate injection ahead of the fabric filter can achieve 70% removal on a long term basis, at an approximate NSR of 2.0.
- NO<sub>2</sub> emissions are generally higher when using sodium bicarbonate than when using sodium sesquicarbonate. The NO<sub>2</sub> generated during sodium-based injection is related to SO<sub>2</sub> removal and the cleaning cycle of the fabric filter, but all factors important to NO<sub>2</sub> generation are not fully understood.<sup>(11)</sup>
- Long term testing of the integrated system demonstrated the synergistic benefit of operation with SNCR and sodium-based dry sorbent injection (i.e., reduce NO<sub>2</sub> and NH<sub>3</sub> emissions).
- When the SNCR and dry sodium systems were operated concurrently, an NH<sub>3</sub> odor problem was encountered in the area around the unit 4 ash silo. This problem appears to be related to the rapid change in pH due to the presence of sodium in the ash.

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## **DISCLAIMER**

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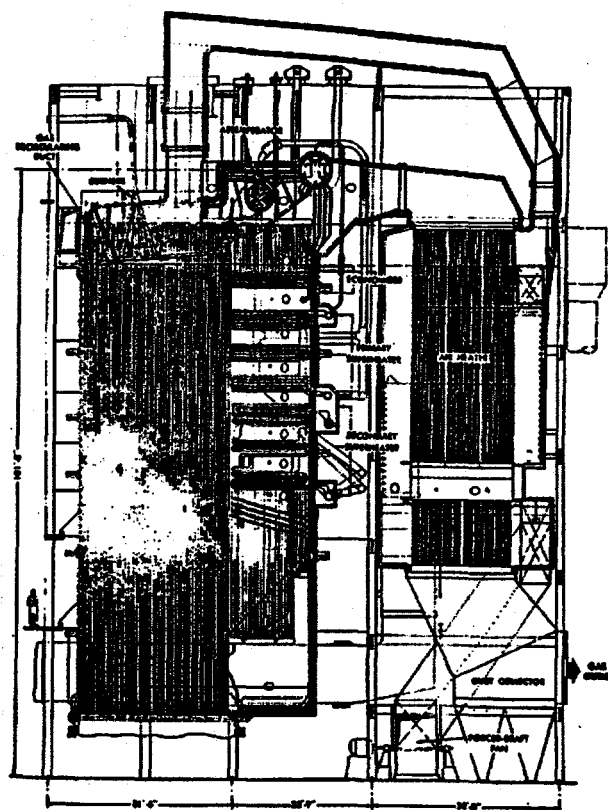


Figure 1. Boiler Elevation

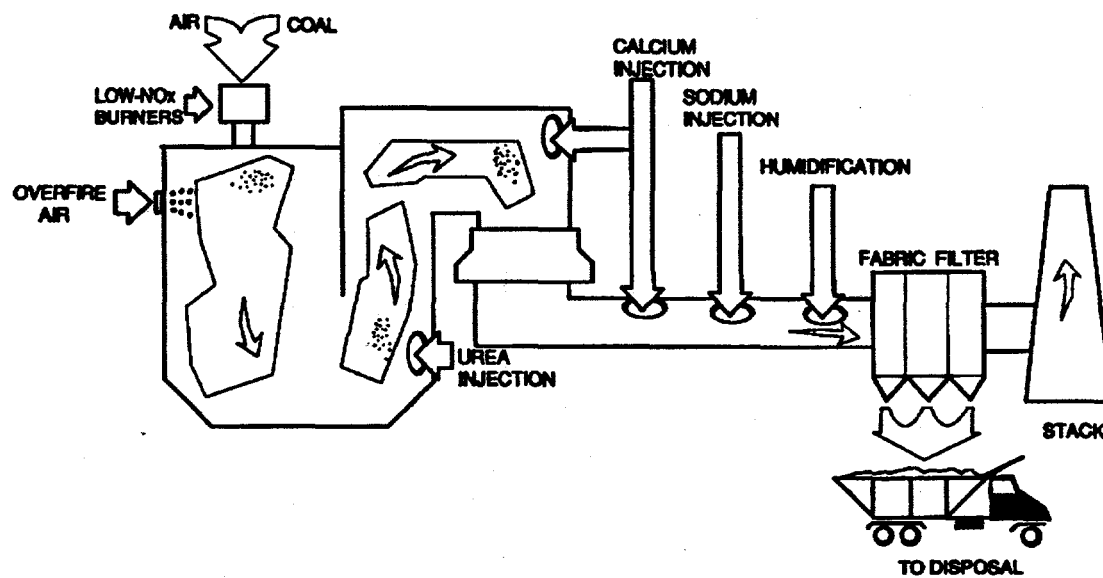


Figure 2. Process Flow Diagram of the Integrated NO<sub>x</sub>/SO<sub>2</sub> Emission Control System

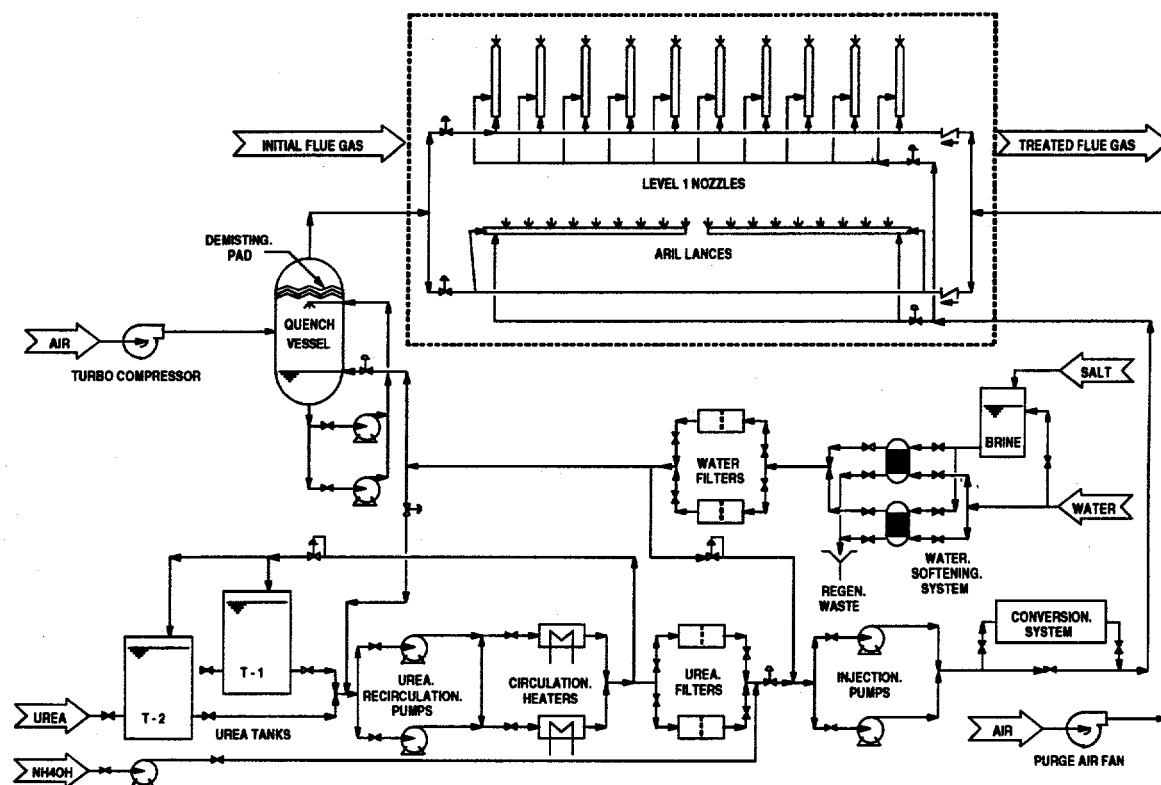


Figure 3. SNCR System Flow Diagram

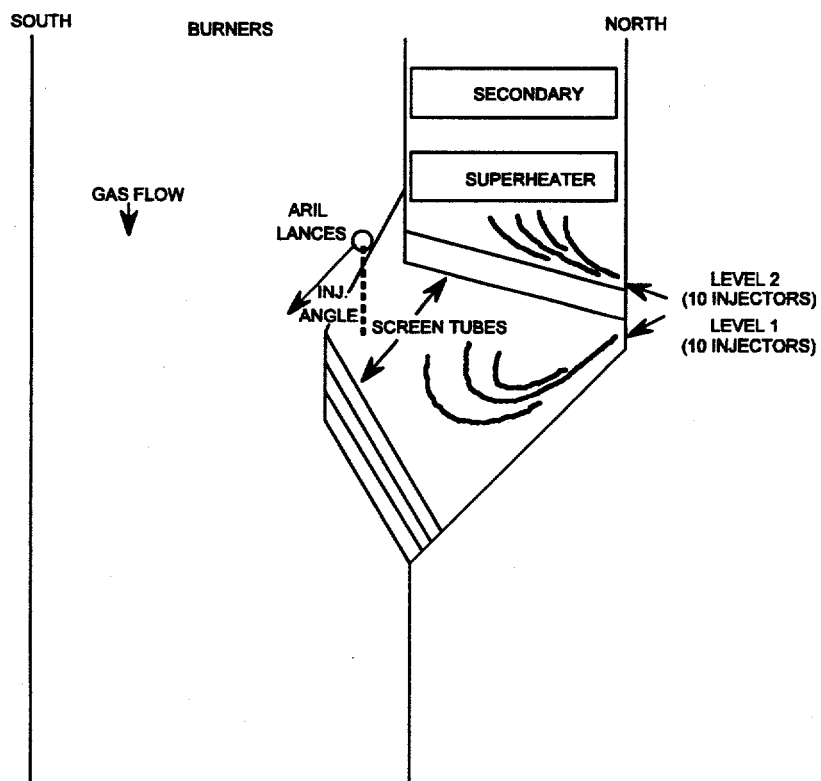


Figure 4. SNCR Injection Locations

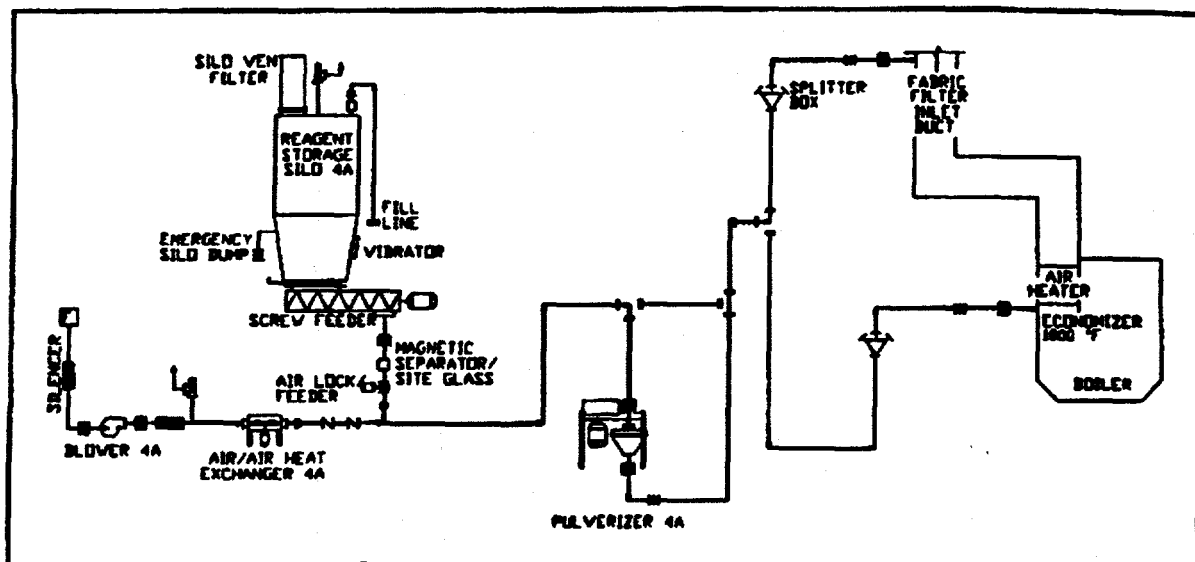


Figure 5. Dry Sorbent Injection Flow Diagram

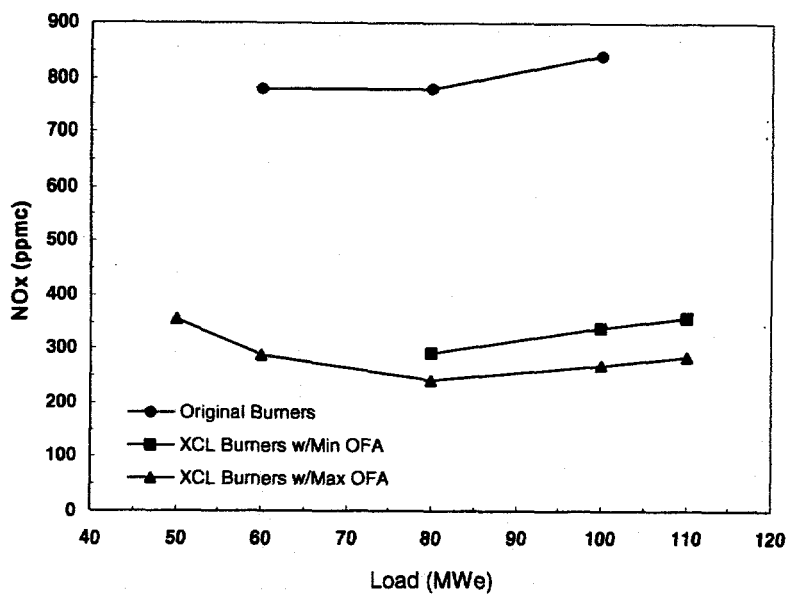


Figure 6. NO<sub>x</sub> Emissions Before and After Low-NO<sub>x</sub> Combustion System Retrofit

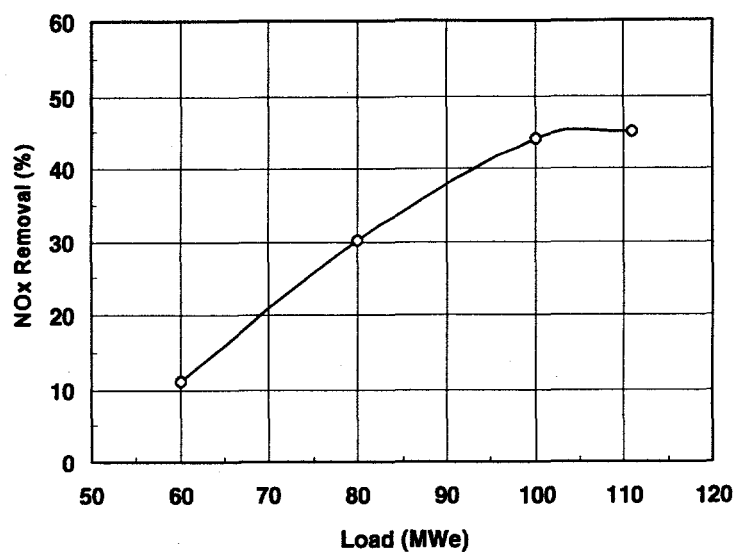


Figure 7. Performance of the Original SNCR System as a Function of Load (10 ppm  $\text{NH}_3$  slip limit)

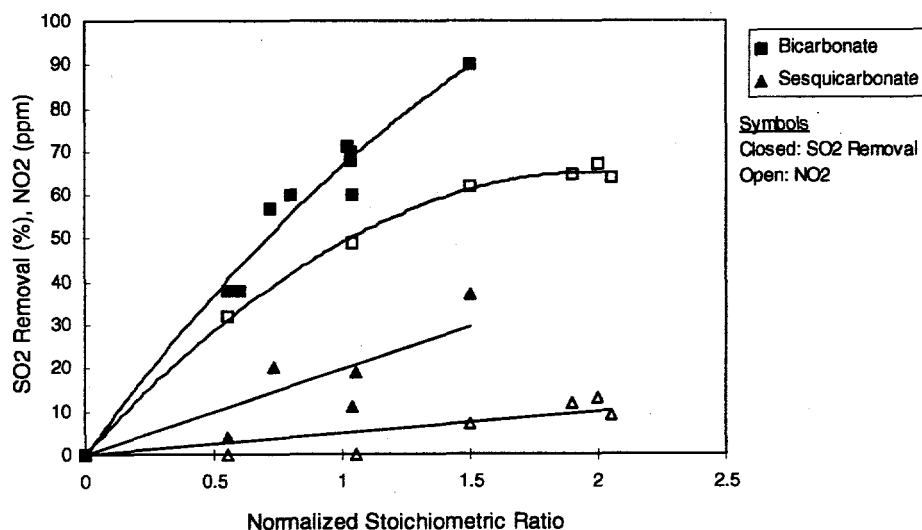
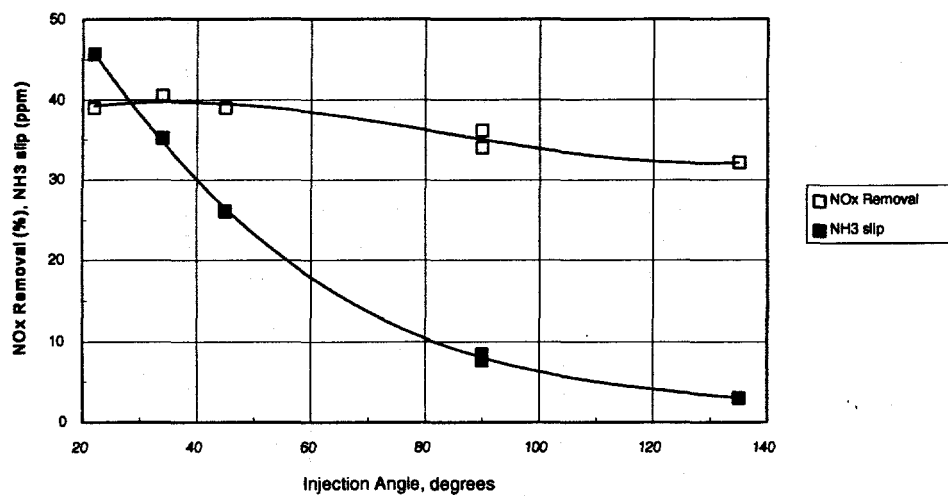
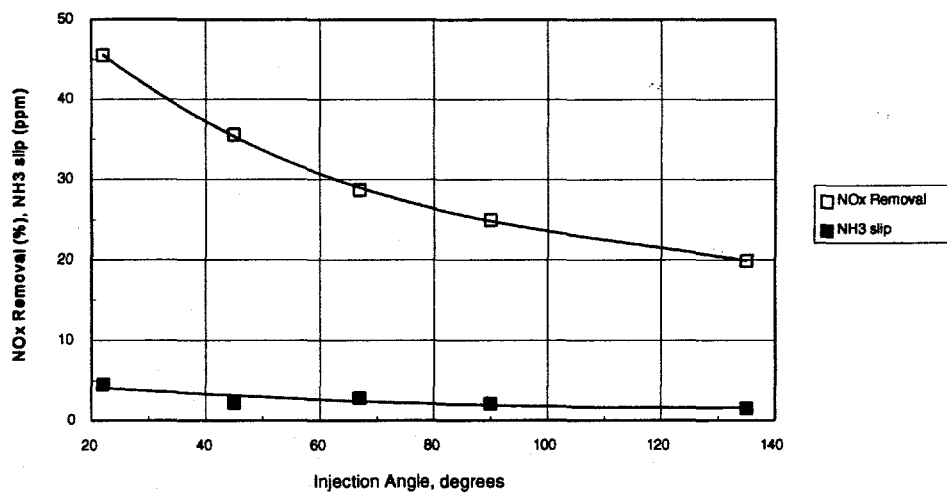


Figure 8. Performance of Sodium Dry Sorbent Injection (sodium sesquicarbonate injected ahead of the fabric filter; sodium bicarbonate injected ahead of the air preheater)

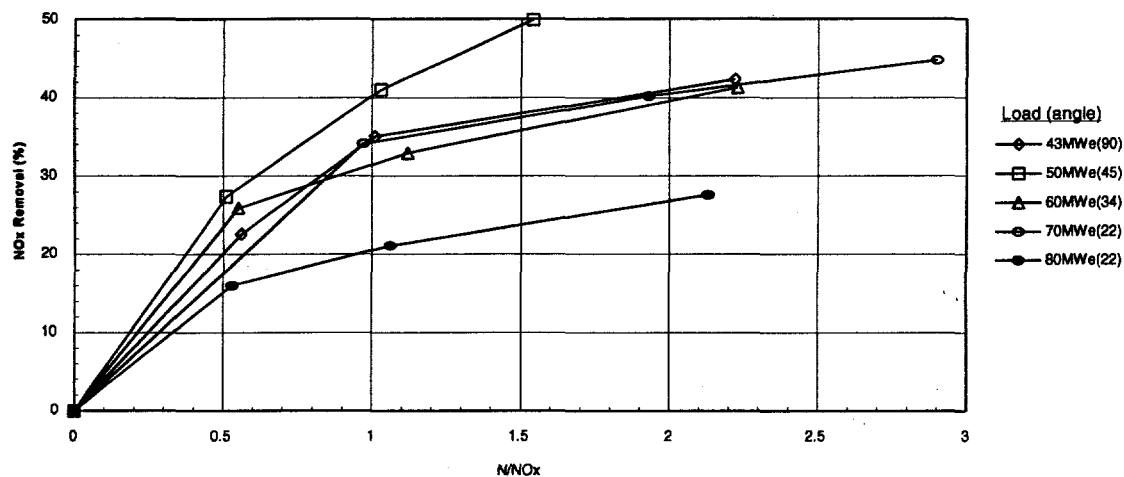


(a) 43 MWe

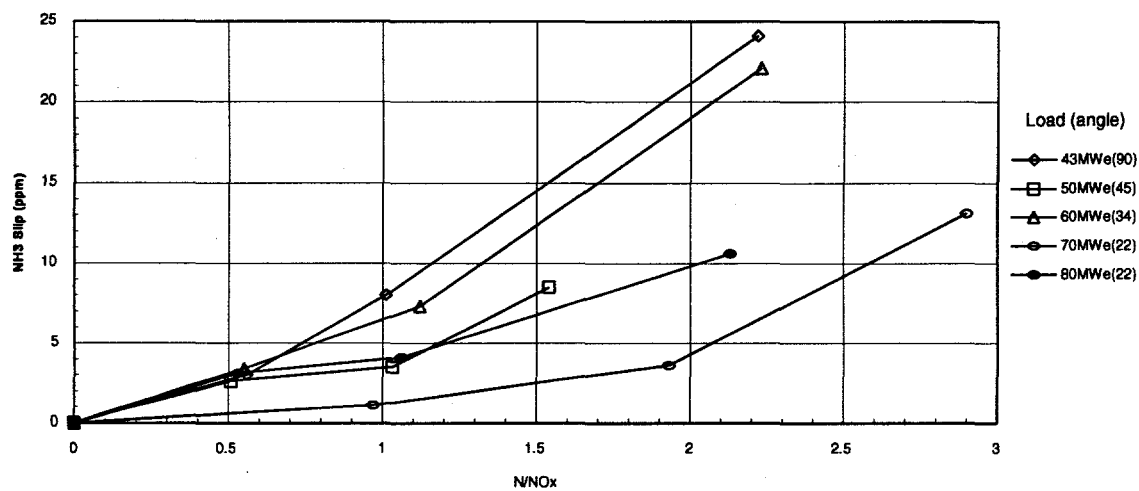


(b) 50 MWe

Figure 9. Effect of Injection Angle on NO<sub>x</sub> Removal and NH<sub>3</sub> Slip  
(Loads: 43 and 50 MWe, N/NO<sub>x</sub> = 1.0)



(a) NOx Removal



(b) NH3 slip

Figure 10. ARIL Lance Performance Over the Load Range: 43 to 80 MWe



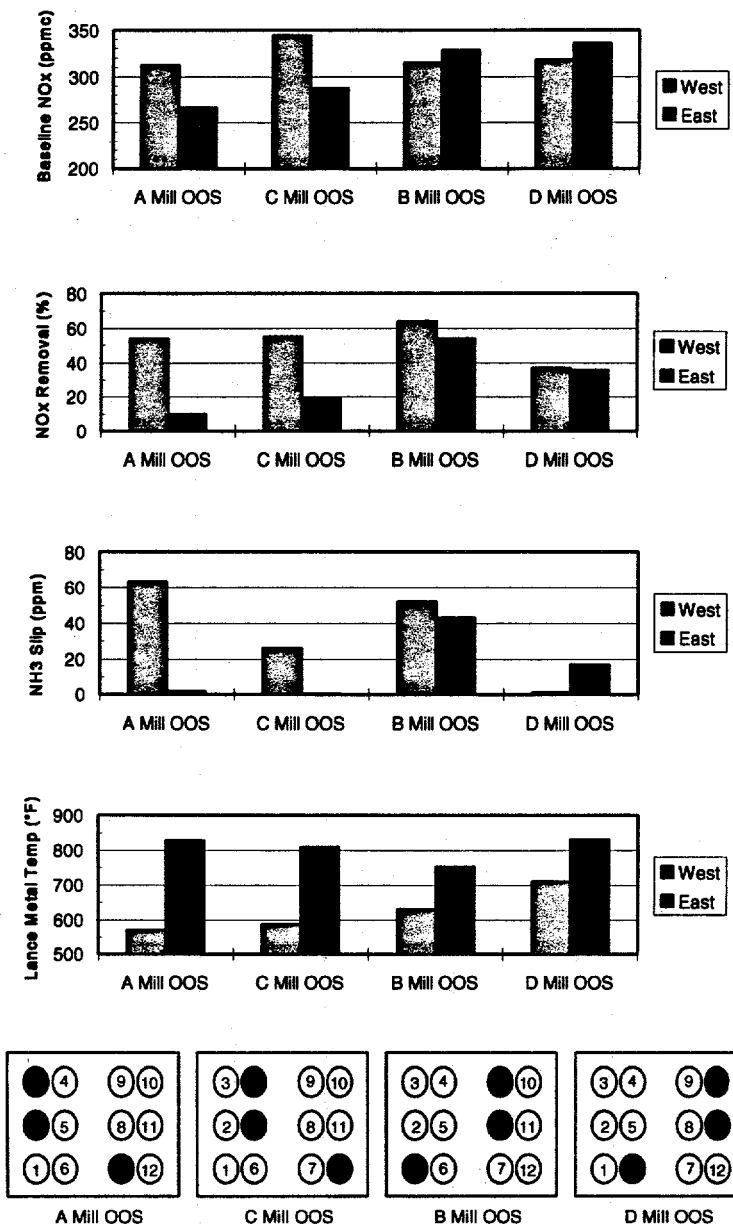


Figure 11. East-West Baseline NO<sub>x</sub> Emissions, NO<sub>x</sub> Removals, NH<sub>3</sub> Slip Levels and Lance Metal Temperature Distributions as a Function of Mill-in-Service Pattern at 60 MWe (N/NO<sub>x</sub> = 2.0, 22° Injection Angle)

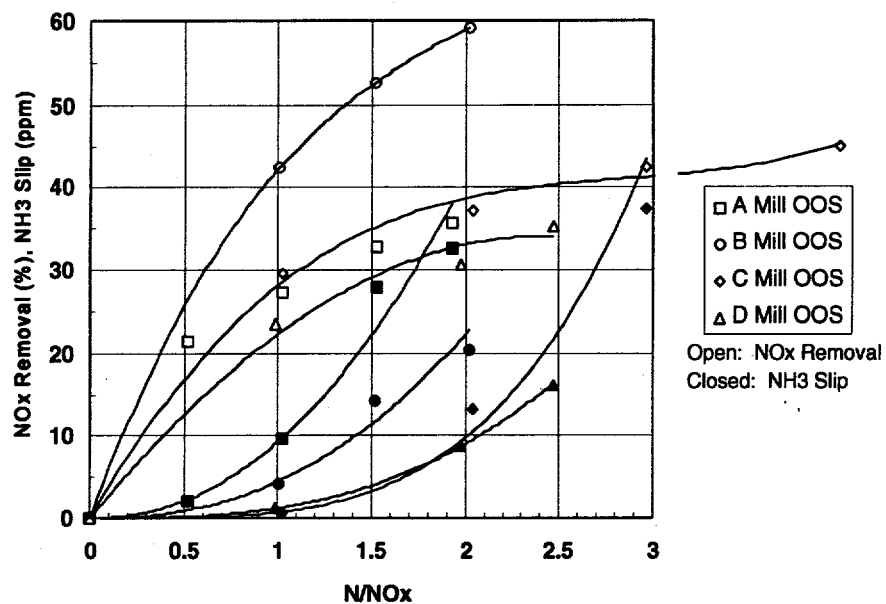


Figure 12. Effect of Mill-in-Service Pattern on ARIL Lance Performance at 60 MWe (22° Injection Angle)

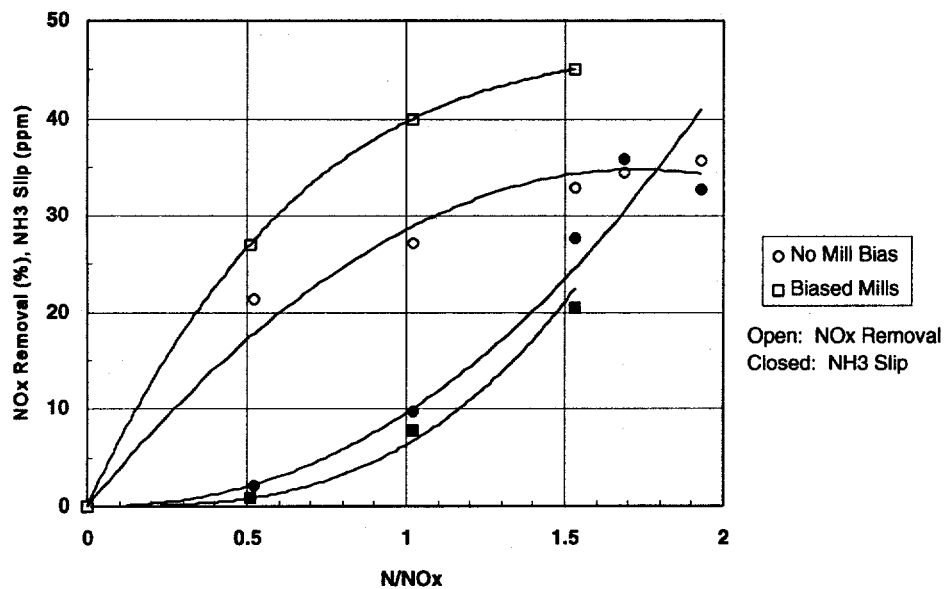


Figure 13. Effect of Coal Mill Bias on ARIL Lance Performance at 60 MWe (A Mill OOS, 22° Injection Angle)

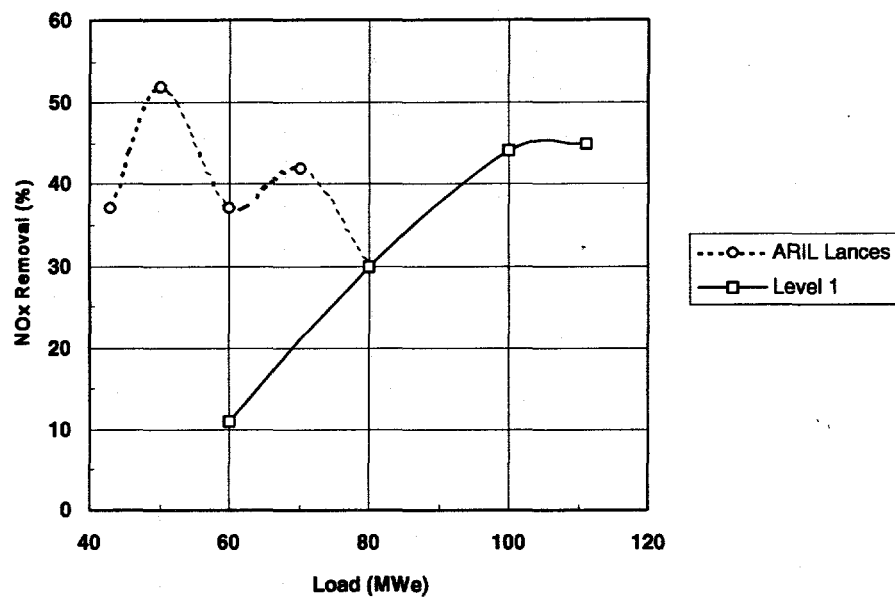


Figure 14. NO<sub>x</sub> Removal as a Function of Load for an NH<sub>3</sub> Slip Limit of 10 ppm

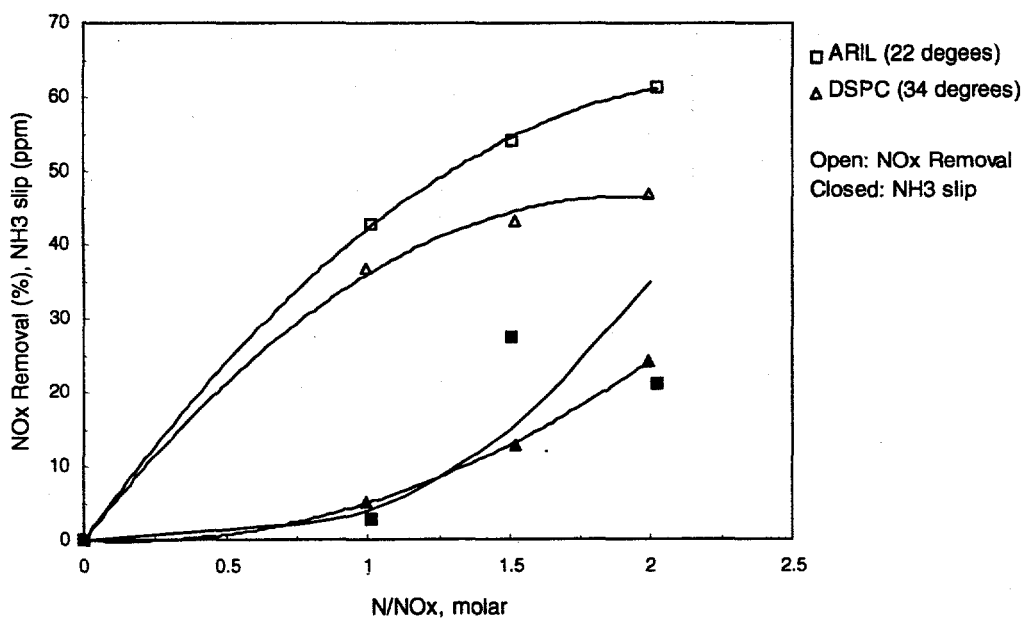


Figure 15. Comparison of the ARIL and DPSC Lance Performance

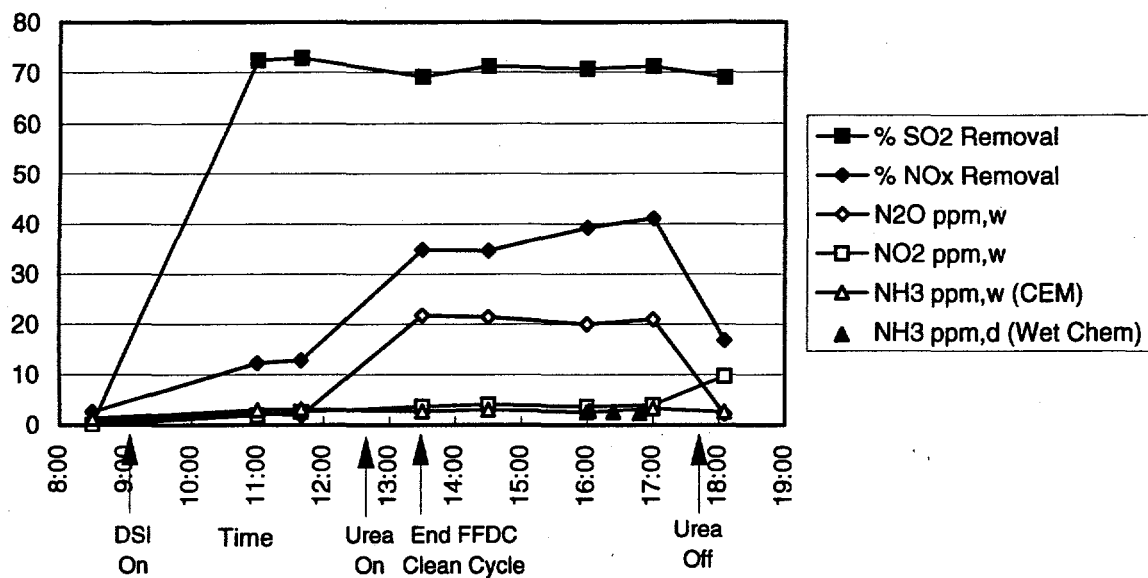


Figure 16. Time History of 100 MWe Integrated Test with Sodium Sesquicarbonate Injection ( $2\text{Na}/\text{S} = 2.0$ ,  $\text{N}/\text{NO}_x = 0.6$ , A Mill OOS)

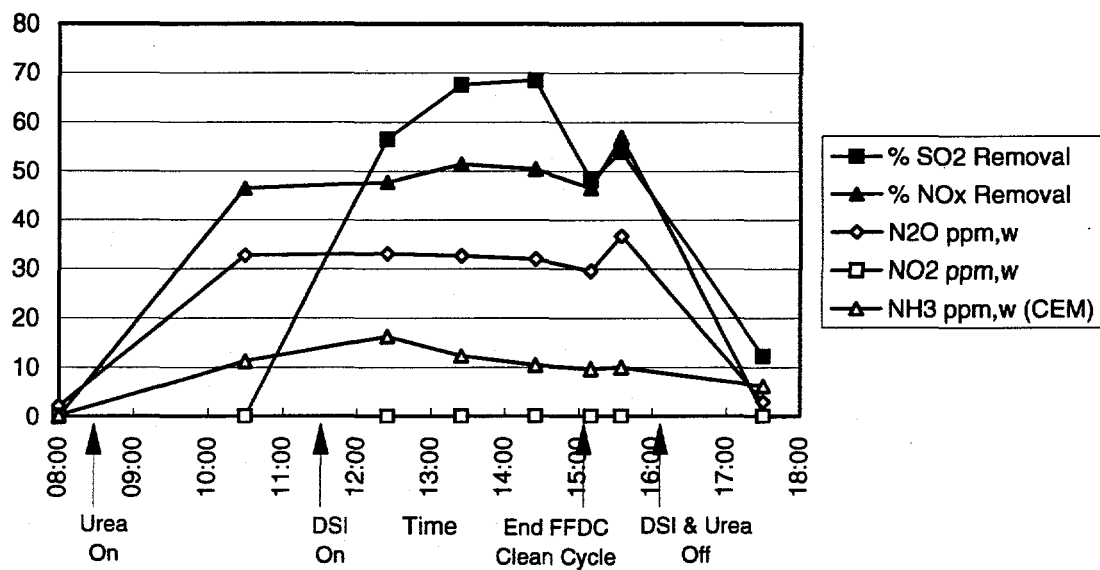


Figure 17. Time History of Integrated Test with Sodium Bicarbonate Injection ( $2\text{Na}/\text{S} = 1.1$ ,  $\text{N}/\text{NO}_x = 1.1$ , 100 MWe, 4 Mills in Service)

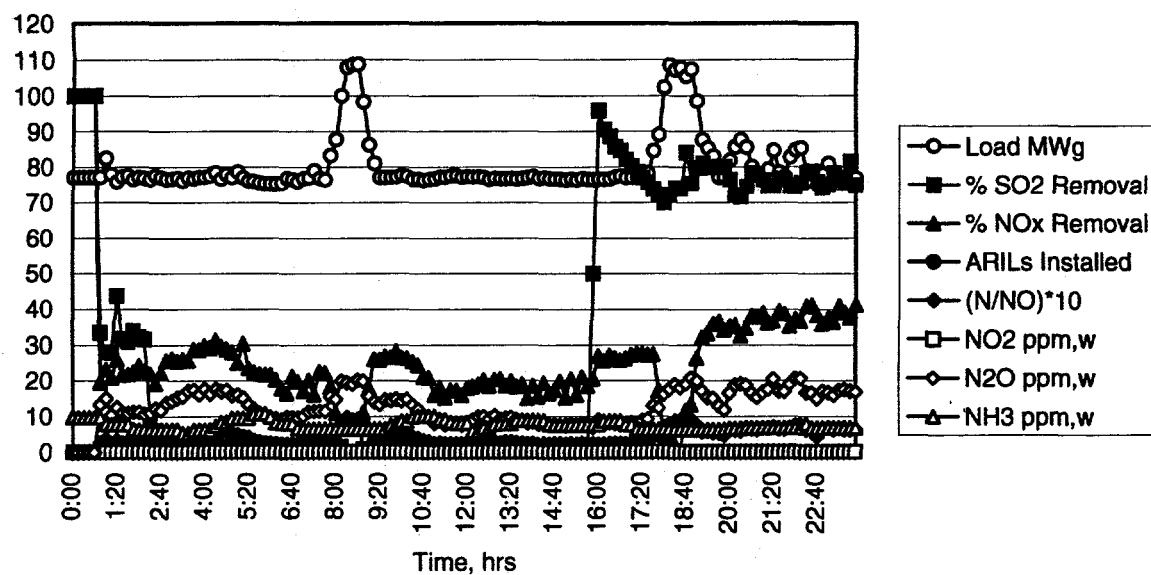


Figure 18. Long-Term Integrated Load-Following Test Results (February 29, 1996)

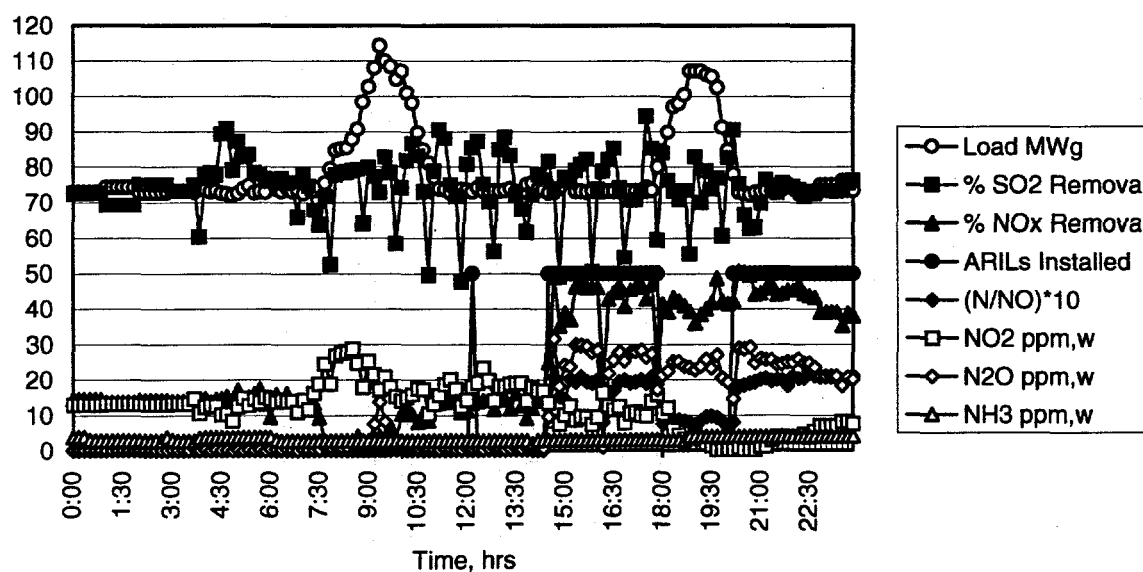


Figure 19. Long-Term Integrated Load Following Test Results (March 4, 1996)

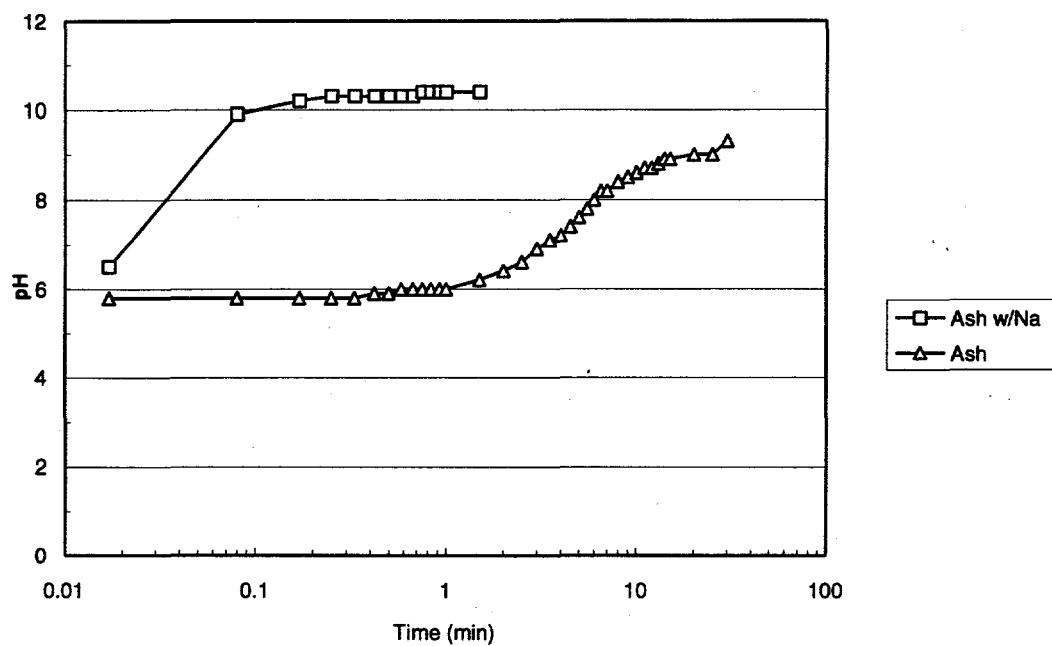
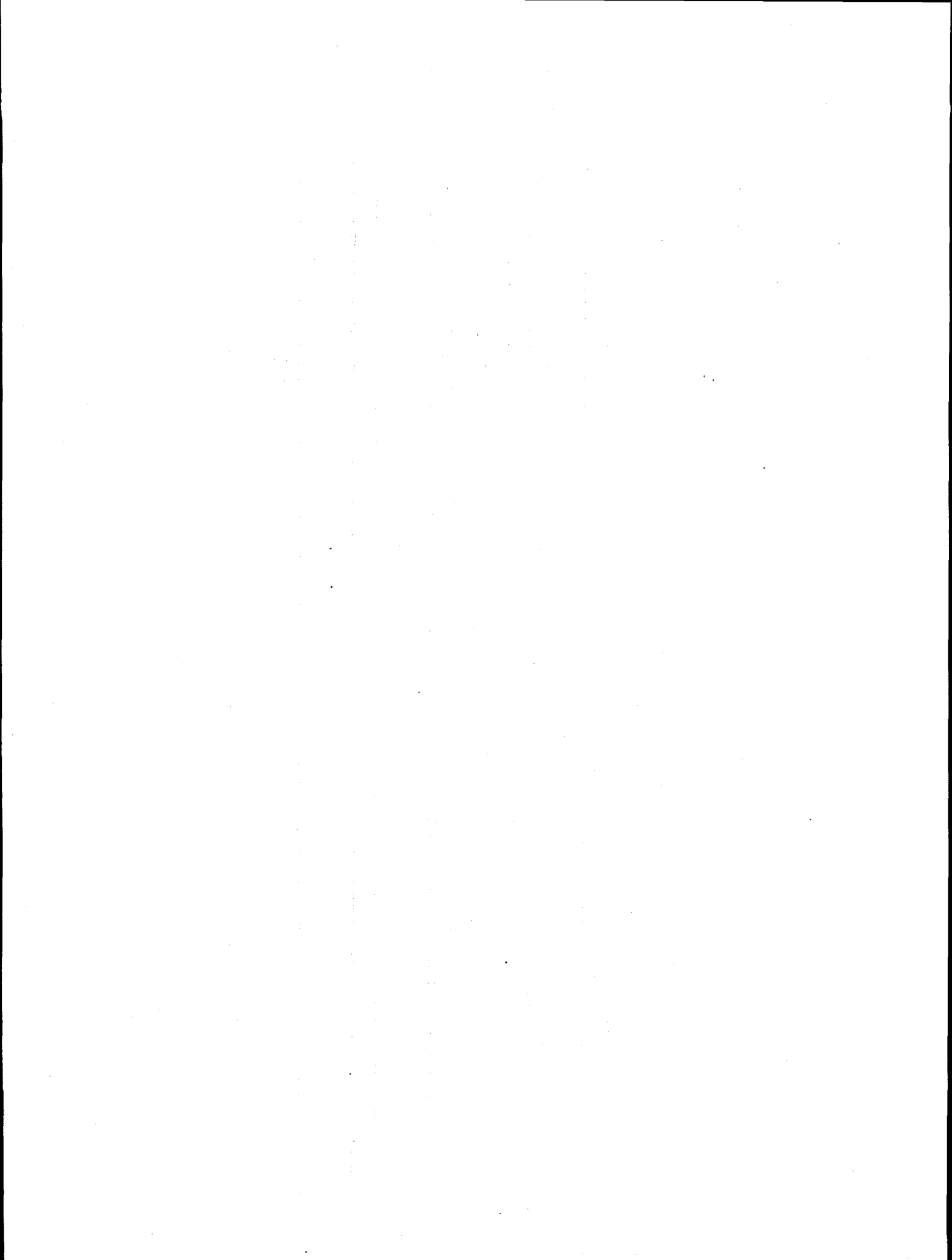


Figure 20. pH versus Time for Coal Ash and Coal Ash/Sodium Mixture  
(0.5 grams of ash in 200 ml of H<sub>2</sub>O)



# **PURE AIR'S BAILLY SCRUBBER A FOUR-YEAR RETROSPECTIVE**

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**The 5th Annual Clean Coal Technology Conference  
January 7 - 10, 1997  
Tampa, Florida**

## **ABSTRACT**

Pure Air's Advanced Flue Gas Desulfurization (AFGD) Clean Coal Project has completed four highly successful years of operation at NIPSCO's Bailly Station. As part of their program, Pure Air has concluded a six-part study of system performance. This paper will summarize the results of the demonstration program, including AFGD performance on coals ranging from 2.0 - 4.5% sulfur. The paper will highlight novel aspects of the Bailly facility, including pulverized limestone injection, air rotary sparger for oxidation, wastewater evaporation system and the production of PowerChip® gypsum. Operations and maintenance which have lead to the facility's notable 99.47% availability record will also be discussed. A project company, Pure Air on the Lake Limited Partnership, owns the AFGD facility. Pure Air was the turn key contractor and Air Products and Chemicals, Inc. is the operator of the AFGD system.



## Introduction

Pure Air's innovative AFGD project was selected in Round II of the Department of Energy's Clean Coal Technology Program. The project is located at Northern Indiana Public Service Company's (NIPSCO) Bailly Station in Chesterton, Indiana, about 40 miles southeast of Chicago, Illinois. The \$151.3 million project involves the retrofit construction and the first three years of operation of a single module AFGD system serving two coal-fired boilers.

On 2 June 1992, the AFGD system began to process flue gas, thus becoming the first commercial scrubber in the country meeting the requirements of the new U.S. Clean Air Act Amendments of 1990 and the first "own and operate" AFGD system in the world. In its first two years of operation, the facility has achieved an availability rate of 99.47% while having the ability to scrub more than 95% of the sulfur dioxide (SO<sub>2</sub>) emissions from two coal-fired boilers.

The specific technical objectives of the Pure Air's AFGD demonstration project are the following:

- Effectively demonstrate high reliability and SO<sub>2</sub> removal capability of a large single module absorber serving multiple boilers
- High velocity co-current absorber
- In-situ forced oxidation producing commercial grade gypsum
- Direct dry limestone injection system
- High efficiency air rotary sparger to enhance oxidation
- Wastewater evaporation system (WES)
- "Own and Operate" business arrangement

In addition to demonstrating the above objectives, a series of six testing periods will be performed by Pure Air and NIPSCO over a period of three years to demonstrate the operation of the facility using coal with a wide range of sulfur contents. Each demonstration test will last approximately 5 to 6 weeks and will test coals of specific sulfur content outlined below:

- 2.0 - 2.5% sulfur coal
- 2.5 - 3.0% sulfur coal
- 3.0 - 3.5% sulfur coal
- 3.5 - 4.0% sulfur coal
- 4.0 - 4.5% sulfur coal
- Optimum sulfur coal

The objective of the DOE demonstration test plans are to evaluate the effects of liquid-to-gas (L/G) ratio, calcium-to sulfur ratio and oxidation capability of the air rotary sparger on the overall system performance. This includes SO<sub>2</sub> removal efficiency, slurry composition and by-product gypsum quality.

## Plant Description

NIPSCO's Bailly Generating Station consists of two coal-fired boilers; Unit #7 rated at 183 MWG capacity began commercial operation in 1962 and Unit #8 rated at 345 MWG capacity began commercial operation in 1968. Both Units #7 and #8 burn Midwestern high sulfur coal. The AFGD facility includes one resin-lined absorber module to process the flue gas generated from Units #7 and #8 and the required ancillary systems. The absorber is a co-current grid-packed tower with two levels of slurry distribution and an integral reaction tank.

The flue gas generated from Units #7 and #8 is combined and sent to a single co-current absorber. Upon entering the absorber module, the flue gas is saturated by contacting a CO<sub>2</sub> enriched gypsum slurry as it passes through a 3 to 4 meter open-faced grid. The absorber grid provides the required surface area for the flue gas and slurry to react so that greater than 95% of incoming SO<sub>2</sub> is removed in the absorber section of the AFGD system. The cleaned flue gas then passes through a two-stage mist eliminator where liquid and solids droplets are removed prior to exiting the scrubber. (See Figure 1.)

The co-current design allows the flue gas and liquid slurry to flow in the same direction, and features a large gas-liquid disengagement zone above the absorber tank. This permits gas velocity through the absorber of up to 20 fps. The use of the high gas velocity co-current absorber allows Pure Air to design one single module for the 528 MWG Bailly Generating Station. There is no spare or back-up module. Instead, the scrubber is designed for a very high level of availability while removing 95% or more of the SO<sub>2</sub>, without the use of performance enhancing chemical additives. Therefore, a high degree of system reliability will be demonstrated. The Pure Air design for the Bailly AFGD system uses a non-pressurized slurry distribution system. This system requires approximately 30% less recirculation pump power than conventional counter-current spray towers. Also, since the fountain-like flow does not generate a fine mist, the mist eliminator loading is reduced by as much as 95% compared to counter-current systems.

An important aspect of Bailly's AFGD system is its in-situ forced oxidation using the air rotary sparger (ARS). The ARS has several advantages over the conventional stationary spargers such as higher oxygen utilization, overall reduction of oxidation air and agitation power, and reduced maintenance cost of the oxidation system.

The Bailly Station has very limited space available for the AFGD facility. In addition to the single module, another space-saving feature used in the Bailly AFGD design is a pulverized dry limestone injection system. The direct injection of pulverized limestone eliminates the need for on-site wet grinding systems, thereby reducing space requirements and capital costs. Unlike conventional wet ball mill systems, the dry limestone injection system does not require ball mills, tanks, pumps, and other associated equipment.

Another feature of the Bailly AFGD demonstration involves wastewater treatment. In addition to a conventional wastewater treatment system, Pure Air has provided a wastewater evaporation system (WES) for the Bailly FGD system. In the WES, a portion of the chloride purge stream is sprayed in the ductwork ahead of the Unit #8 precipitator. The purpose of this unique system is to evaporate the high chloride purge stream in the path of the hot flue gas prior to ESP. Calcium chloride solids remain after the water has evaporated and these solids, along with the fly ash, are collected and removed in the precipitator.

In addition to reducing emissions, this AFGD system converts the SO<sub>2</sub> into a high purity (on average between 96% and 97%) commercial grade of synthetic gypsum. This synthetic gypsum is used by United States Gypsum Company to produce wallboard at its East Chicago, Indiana plant. This makes East Chicago the only facility in North America to produce wallboard from 100% FGD gypsum and the sixth U.S. Gypsum plant to use synthetic gypsum on a sustained basis. The sale of this synthetic gypsum is covered under a long term agreement entered in June 1990 between NIPSCO and United States Gypsum Company. Since production began, U.S. Gypsum has taken NIPSCO's entire output of synthetic gypsum and is providing NIPSCO with a reliable alternative to landfill disposal.

At the Bailly Station, Pure Air is also demonstrating a unique gypsum agglomeration process known as PowerChip gypsum. There are many different FGD technologies that produce gypsum. Most make a product similar in consistency to wet sand. This type of gypsum is difficult and time consuming to unload from rail cars or trucks. The material handling problems with conventional FGD gypsum are numerous, as it has poor processing flexibility and contains agglomerates which must continuously be broken and dispersed. It requires an extensive drying process, using expensive, energy consuming drying equipment, before it can be used. For some users, the solution may be to purchase FGD gypsum that has already been dried. A technology from Pure Air is helping utilities produce PowerChip gypsum, a product with all the performance and purity benefits of FGD gypsum and all the handling benefits of natural gypsum. The particle size distribution of milled PowerChip gypsum much more closely resembles that of natural gypsum than does conventional unprocessed FGD gypsum. This makes PowerChip gypsum just as easy to use as natural rock gypsum. PowerChip gypsum is produced in relatively dry, consistently sized chips which will not freeze together in cold weather. It requires no special storage systems, nor expensive drying or conveying equipment.

The PowerChip agglomeration process utilizes a compression mill at an optimum compacting force with an exclusive curing time and temperature relationship that reformulates and modifies the physical structure of FGD by-product gypsum. This technology produces stable, semi-dry, agglomerated flakes of calcium sulfate dihydrate (gypsum) which range between 1/8" to 1/16" in thickness and 3/8" to 1-1/4" in length and width. The production rate of the PowerChip demonstration facility at the Bailly Station is 7 tons per hours. This material can be handled,

transported and stored with existing infrastructure designed for natural rock gypsum without costly modifications associated with handling other forms of FGD gypsum.

### **Bailly AFGD Performance**

#### **Maintenance**

The key components to achieving the 99.47% availability record over the three years of operation are the operating/maintenance philosophy of the Pure Air Bailly facility coupled with the incorporation of technological modifications. Pure Air embraces a modified reliability centered maintenance philosophy which began in the project design phase. Through critical mode analysis, equipment is identified that, if down, would have the largest impact on availability. In-line spares have been incorporated for these critical pieces of equipment. The operating technicians, who have both operating and maintenance responsibilities, are trained to focus on equipment important to availability, which has no installed spare. Operating technicians use predictive maintenance techniques to determine which equipment needs attention and to what extent. A team then determines the cost of repair versus replacement. Key equipment and spare parts identified and purchased in the design phase using operating history, are on hand locally at the facility. Pure Air also maintains a centrally located spare parts inventory which is available through a mainframe application within the parent company, Air Products and Chemicals. As such, virtually all critical parts and equipment are either on line or readily accessible. As part of the reliability centered maintenance system, a computerized maintenance system is utilized. This tool provides tracking on every piece of equipment and process system in terms of costs and maintenance history. Available to all personnel within the facility, this information is used to facilitate total impact operating and maintenance decisions.

#### **Material Construction**

Another major modification to the AFGD system was the implementation of new alloy technology. C-276 alloy over carbon steel clad material replaced existing alloy wallpaper construction within the absorber tower wet/dry interface. The installation of this technology advancement resulted in significant maintenance cost reductions.

#### **WES Modifications**

The original WES system consisted of a series of high pressure nozzles spraying wastewater in the path of flue gas ahead of the Unit #8 electrostatic precipitator. After a series of testing with the high pressure nozzles, it was determined that the high pressure nozzle did not perform satisfactorily and excessive amounts of solids accumulated in the duct work. It was decided to replace the high pressure nozzles with two-fluid nozzles, which provide better droplet size distribution and more control of liquid being evaporated. With the installation of a two-fluid nozzle, the WES has been running for the past eight months resulting in zero liquid discharged

for Unit #8. Inspection of the WES duct work-reveals no solids accumulation or corrosion in the duct work.

### **DOE Demonstration Test Results**

As stated earlier, the performance of the AFGD system will be evaluated over a three year period. A total of five demonstration tests will be conducted (Test I - Test V) during this period. The first three years of this demonstration have been completed and all the DOE tests have been completed. A summary of the completed tests in chronological order is provided below.

**DOE Demonstration Test III.** The first series of DOE demonstration tests started in August 1992 burning Indiana coal containing approximately 3.2% sulfur coal (see *Table 1*). This is the normal coal for NIPSCO's Bailly Generating Station. A total of 31 days of testing were conducted in a six week period. The AFGD system was tested at 100%, 67% and 33% of its design capacity. The primary variables tested were liquid-to-gas (L/G) ratio, calcium-to-sulfur (stoichiometric) ratio and ARS oxidation capability. The results of the 3.2% sulfur coal test have been previously issued (2).

**DOE Demonstration Test IV.** This DOE demonstration test started in May 1993 and lasted for approximately six weeks. The coal used for this test had a sulfur content of approximately 3.8% and heating value of 11,000 BTU/lb (*Table 1*).

A total of 48 tests were conducted during this demonstration. The variables studied during this demonstration were flue gas velocity (boiler load varied from 100% to 33% of its maximum capacity), liquid-to-gas (L/G) ratio, and calcium-to-sulfur (stoichiometric) ratio. The parametric study included the effect of the above variables on SO<sub>2</sub> removal and on gypsum quality.

**Stoichiometric Ratio (SR):** For wet limestone scrubbers, SR is defined as moles of total calcium (or calcium carbonate) fed to the FGD system per moles of SO<sub>2</sub> removed. To evaluate the effect of SR in the system performance, the absorber calcium carbonate level was varied from 50 to 130 mmole/l while maintaining 75% and 94% of design recirculation flow at 100% boiler load (*Figure 2*). As illustrated in this figure, the SO<sub>2</sub> removal efficiency increases from 94% to 97% by increasing SR from 1.03 to 1.08 moles/mole at 94% of design recirculation rate. The effect of SR on SO<sub>2</sub> removal efficiency was also tested at 67% and 33% boiler load. The results of the testing at 67% boiler load are presented in *Figure 3*. As illustrated in this figure, 98% SO<sub>2</sub> removal was achieved at SR of 1.065 while operating with only 92% of design L/G.

**Liquid-to-Gas (L/G) Ratio:** The L/G was fluctuated by varying the number of the operating recirculation pumps. At 100% load, the recirculation flow was varied from 75% to 94% of total liquid flow while maintaining relatively constant stoichiometric ratio of 1.03 and 1.08 moles/mole (moles of calcium per mole of SO<sub>2</sub> removed).

As expected, SO<sub>2</sub> removal increased with increasing recirculation flow rate (L/G). For example, at 100% boiler load and stoichiometric ratio of 1.03, SO<sub>2</sub> removal increased from about 90.5% to 94% by increasing absorber recirculation flow from 75% to 94% of its design value (*Figure 4*). The performance of the system is further improved from about 94% to 97% at higher SR (1.08) over the same range of L/G. Much better system performance was achieved under similar conditions at reduced (67%) boiler load (*Figure 5*).

**Gypsum Quality:** The gypsum purity during the 3.8% sulfur coal demonstration test period averaged 96.7%. All other gypsum parameters are presented in *Table 2*.

**DOE Demonstration Test V.** This test consisted of NIPSCO burning coal with greater than 4.5% sulfur for a period of at least 100 hours. During this period, the boiler was kept under full load (100%) conditions. The L/G and the SR were set so that the system would maintain at least 95% pure gypsum. This DOE demonstration test started in June 1994 and lasted for 16 days. The coal used for this test had a sulfur content of approximately 4.7% and a heating value of 12,700 BTU/lb (*Table 1*). The variables studied during this demonstration were calcium-to-sulfur (stoichiometric) ratio and liquid-to-gas (L/G) ratio.

**Stoichiometric Ratio (SR):** To evaluate the effect of SR on the system performance, the absorber calcium carbonate level was varied from 50 to 130 mmole/l while maintaining 75% and 93% of design recirculation flow (*Figure 6*). As indicated in this figure, 95% of SO<sub>2</sub> removal efficiency was achieved at SR of 1.05 (moles/mole) and 93% of design L/G.

**Liquid to Gas (L/G) Ratio:** During the 4.7% sulfur coal and 100% boiler load test, data was collected to evaluate the system performance at constant stoichiometric ratio (1.035 and 1.05) while varying the recirculation flow rate from 75% to 100% of its design rate. Again, as expected, SO<sub>2</sub> removal increased with increasing L/G. With 4.7% sulfur coal and a stoichiometric ratio of 1.05, SO<sub>2</sub> removal efficiency increased from 92.5% to 96% by increasing absorber recirculation flow from 75% to 100% of its design rate (*Figure 7*).

**Gypsum Quality:** Similar to 3.8% sulfur coal, very high gypsum purity was produced during the 4.7% sulfur coal testing (*Table 2*).

**DOE Demonstration Test I.** This demonstration test started in July 1994 and lasted for approximately 6 weeks. The coal used for this test had a sulfur content of approximately 2.25% and a heating value of 11,932 BTU/lb (*Table 1*).

A total of 42 tests were conducted during the demonstration, which consisted of 20 tests at full boiler load, 15 tests at 67% boiler load and 8 tests at 33% boiler load. The variables studied during this demonstration test were flue gas velocity, liquid-to-gas (L/G) ratio, calcium-to-sulfur ratio (stoichiometric) ratio and their effect on gypsum quality.

**Stoichiometric Ratio (SR):** To determine the effect of SR on SO<sub>2</sub> removal efficiency, the absorber calcium carbonate level was varied from 24 to 90 mmole/l (corresponding to SR of 1.015 to 1.056 moles of calcium carbonate fed per mole of SO<sub>2</sub> removed) while maintaining 70% and 85% of design recirculation flow rate at 100% boiler load (*Figure 8*). As illustrated in this figure, SO<sub>2</sub> removal efficiency increased with increasing absorber SR; in fact, the AFGD system could easily remove 97% of the inlet SO<sub>2</sub> at stoichiometric ratio of 1.055 with only 85% of design recirculation flow rate. 92% SO<sub>2</sub> removal efficiency is achievable with SR of only about 1.015 moles/mole at the same recirculation flow rate.

As expected, reducing the gas velocity from 100% to 67% of boiler load improves the system performance (*Figure 9*). For example, increasing the system SR from 1.014 to 1.03 at 84% of design L/G, resulted in increasing SO<sub>2</sub> removal from 94% to 97% at 67% boiler load.

**Liquid-to-Gas (L/G) Ratio:** The effect of the L/G on SO<sub>2</sub> removal was tested by varying the number of the operating recirculation pumps while maintaining constant stoichiometric ratio. With 2.25% sulfur coal, the recirculation flow rate was varied from 70% to 85% of the design flow rate at SR of 1.03 and 1.053 moles/mole at 100% boiler load (*Figure 10*) and at SR of 1.014 and 1.03 moles/mole at 67% boiler load (*Figure 11*). As presented in *Figures 10 and 11*, SO<sub>2</sub> removal efficiency increased with increasing L/G at constant SR. For example, increasing recirculation rate from 70% to 85% at the SR of 1.03 increased the SO<sub>2</sub> removal from 91% to 94% at 100% boiler load, and from 95% to 97% at 67% boiler load.

**Gypsum Quality:** The AFGD system produced very high quality gypsum (*Table 2*) during the 2.25% sulfur coal testing.

**DOE Demonstration Test II.** This demonstration test started in September 1994 and lasted for approximately four weeks. The coal used for this test was a blend of Captain coal containing 3.2% sulfur and Credero mined coal containing 2.25%. This blended coal had a sulfur content of approximately 2.75% (*Table 1*).

A total of 29 tests were conducted during this demonstration, which consisted of 14 tests at 100% boiler load, 8 tests at 67% boiler load, and 7 tests at 33% boiler load. The variables studied during this demonstration test were flue gas velocity, liquid-to-gas (L/G) ratio, calcium-to-sulfur (stoichiometric) ratio and their effect on gypsum quality.

**Stoichiometric Ratio (SR):** To evaluate the effect of SR on SO<sub>2</sub> removal efficiency, the absorber calcium carbonate level was varied from 25 to 100 mmole/l (corresponding to SR of 1.015 to 1.065 moles of calcium carbonate fed per mole of SO<sub>2</sub> removed) while maintaining 80%, 88% and 100% of design recirculation flow rate at 100% boiler load (*Figure 12*). As illustrated in this figure, SO<sub>2</sub> removal efficiency increases with increasing stoichiometric ratio. For example, at 100% of

design recirculation flow rate, increasing the absorber SR from 1.015 to 1.065 resulted in increased SO<sub>2</sub> removal from 94% to 96%.

As expected, reducing the gas velocity from 100% to 67% improves the system performance (*Figure 13*). For example, at 67% boiler load, increasing the SR from 1.02 to 1.04 at 88% of design L/G, resulted in increasing SO<sub>2</sub> removal from 96.5% to 97.5%.

**Liquid-to-Gas (L/G) Ratio:** The effect of the L/G on SO<sub>2</sub> removal was tested by varying the number of the operating recirculation pumps while maintaining constant stoichiometric ratio with 2.75% sulfur coal. The recirculation flow was varied from 80% to 100% of the design flow rate at SR of 1.015, 1.03, 1.04 and 1.055 moles/mole at 100% boiler load (*Figure 14*) and at SR of 1.025 and 1.04 moles/mole at 67% boiler load (*Figure 15*). Again as illustrated in *Figures 14 and 15*, SO<sub>2</sub> removal efficiency increased with increasing L/G at constant SR.

**Gypsum Quality:** The AFGD system produced very high quality gypsum (*Table 2*) during the 2.75% sulfur coal testing.

**DOE Demonstration Test VI.** This demonstration test started in March 1995 and continued until May 1995. The overall objective of this demonstration test was to evaluate the system performance while the boilers burning coal with an optimum level of sulfur for a period of sixty days. The optimum level of sulfur was identified as a result of the previous testing.

The coal used for this test had a sulfur content of approximately 3% sulfur and heating value of 10,970 Btu/lb (*Table 1*). A total of 52 tests were conducted during this demonstration and all of the tests were operated at full boiler load. The variables studied during this demonstration test were liquid-to-gas (L/G) ratio, calcium-to-sulfur (stoichiometric) ratio and their effect on gypsum quality.

**Stoichiometric Ratio (SR):** To evaluate the effect of SR on the system performance, the absorber calcium carbonate level was varied from 40 to 116 mmol/l (corresponding to SR of 1.025 to 1.078) while maintaining 84%, 93% and 96% of design recirculation flow rate at 100% boiler load (*Figure 16*). As illustrated in this figure, the SO<sub>2</sub> removal of efficiency increases from 94% to 96% by increasing SR from 1.03 to 1.06.

**Liquid-to-Gas (L/G) Ratio:** At 100% load, the recirculation flow was varied from 65% to 93% while maintaining relatively constant stoichiometric ratio of 1.03, 1.045 and 1.065 (moles of calcium per moles of SO<sub>2</sub> removed). As expected, SO<sub>2</sub> removal increased with increasing recirculation flow rate. For example, at 100% load and SR of 1.065, SO<sub>2</sub> removal increased from 94.5% to 96% by increasing absorber recirculation from 60% to 100% of its design value (*Figure 17*).



Gypsum Quality: The AFGD system produced very high quality gypsum (Table 2) during the optimum sulfur level coal testing.

### Summary of Results

The demonstration of the AFGD system at the Bailly Generating facility has established this technology as an efficient and reliable means of removing  $\text{SO}_2$ . The overall system  $\text{SO}_2$  removal efficiency results from the first four DOE demonstration tests have been analyzed and several relationships have been determined. First, at constant stoichiometric ratio,  $\text{SO}_2$  removal efficiency increases with the absorber recirculation rate (*Figure 18*).

The rate increase in  $\text{SO}_2$  removal efficiency is highest at low sulfur coal (2.25% sulfur) and decreases as coal sulfur content increases to 4.5%. Next, at moderately low calcium stoichiometric ratio, very high levels of  $\text{SO}_2$  removal are achieved at two thirds of design L/G ratio (*Figure 19*). As illustrated in this figure, the AFGD system is capable of achieving higher than 95% reduction in  $\text{SO}_2$  at 76% of design L/G for all sulfur coal tested.

To date, operations have gone well. The scrubber has already exceeded its target of demonstrating 95+%  $\text{SO}_2$  removal efficiency, while producing a commercial grade gypsum by-product and meeting all system operating and contractual requirements (*Tables 3 through 5*).

# FIGURE 1

## Advanced Flue Gas Desulfurization Process Flow Diagram

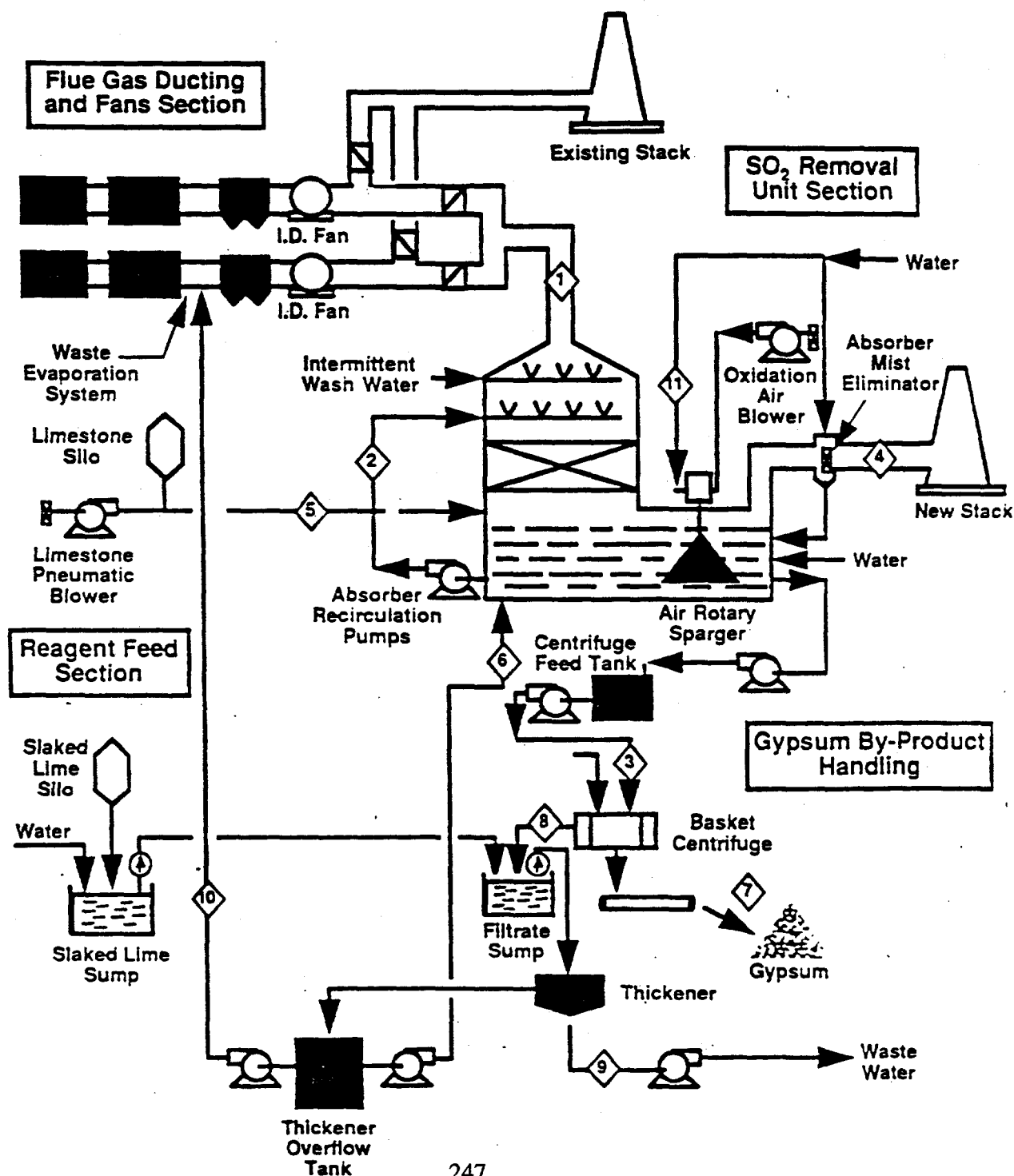


FIGURE 2

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

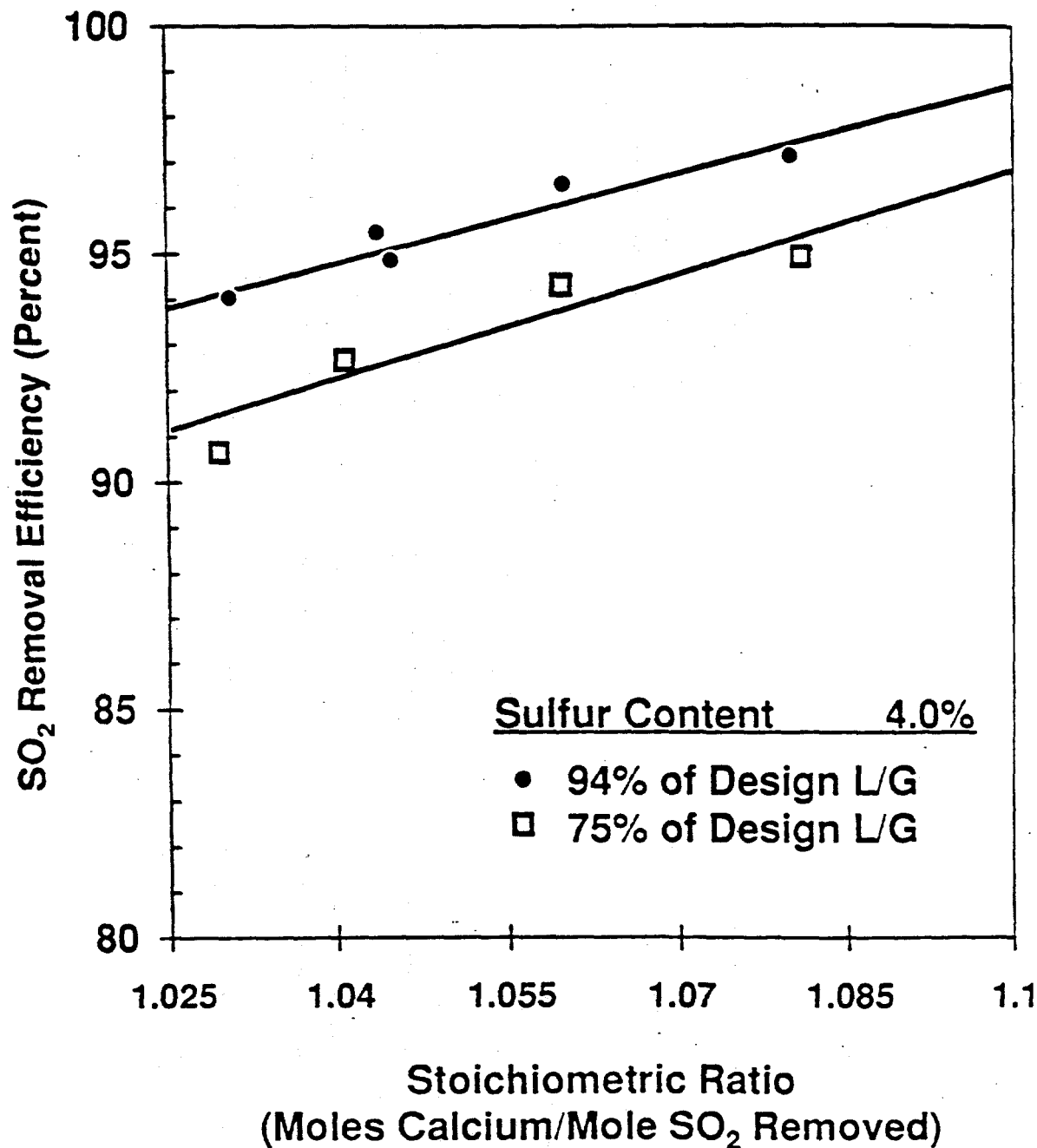


FIGURE 3

# SO<sub>2</sub> Removal Performance at Bailly AFGD (67 Percent Boiler Load)

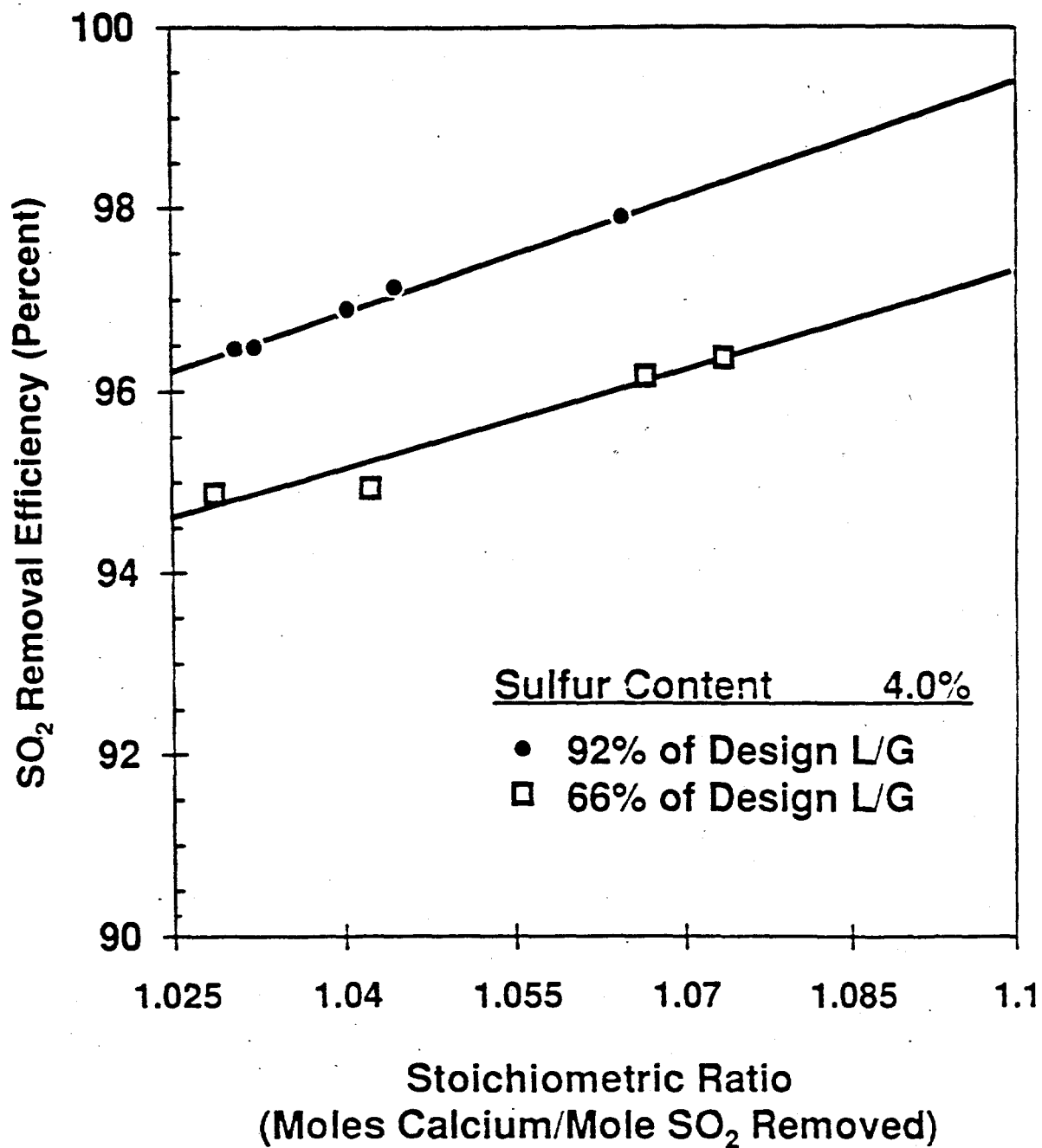


FIGURE 4

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

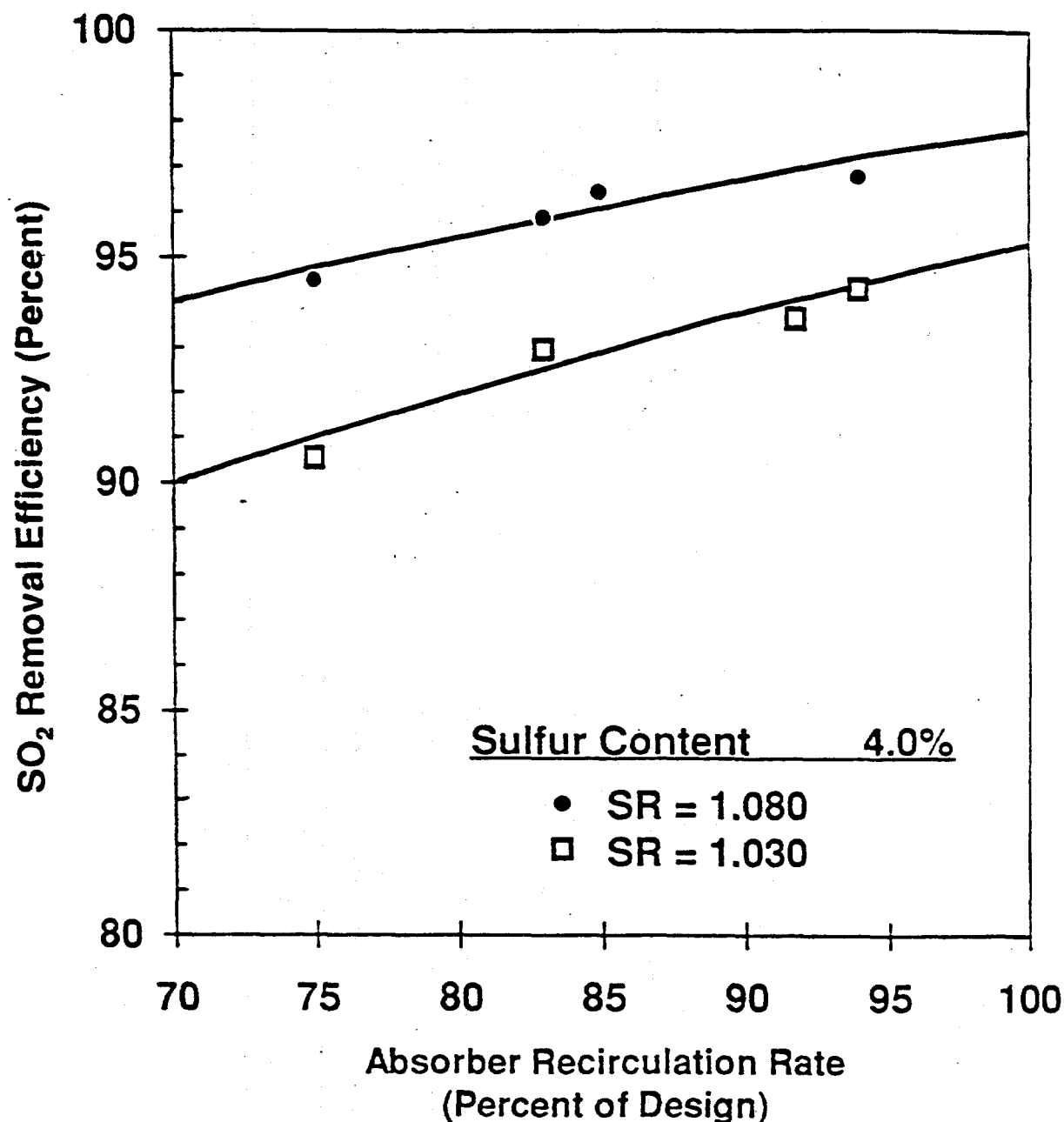


FIGURE 5

# SO<sub>2</sub> Removal Performance at Bailly AFGD (67 Percent Boiler Load)

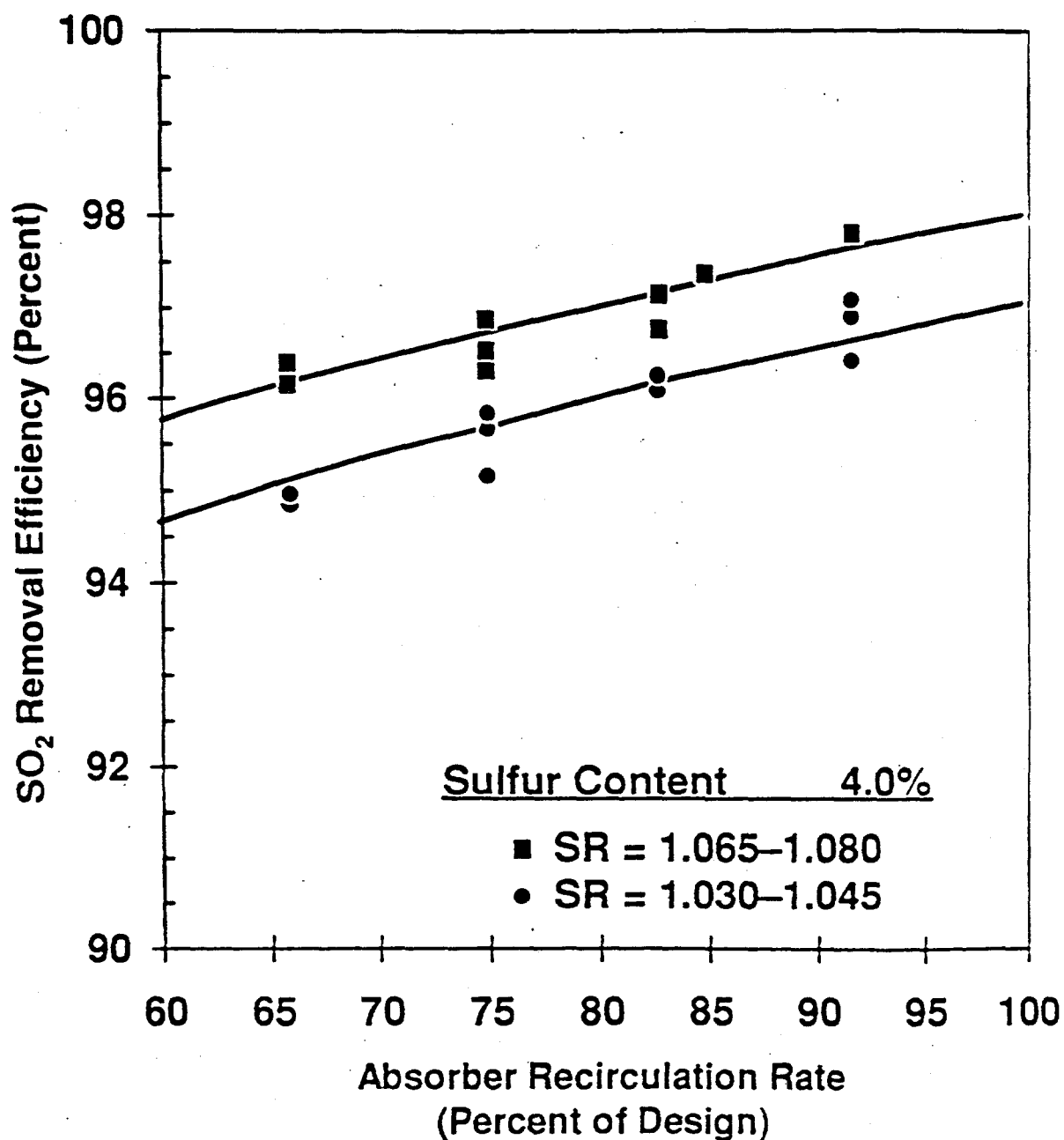


FIGURE 6

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

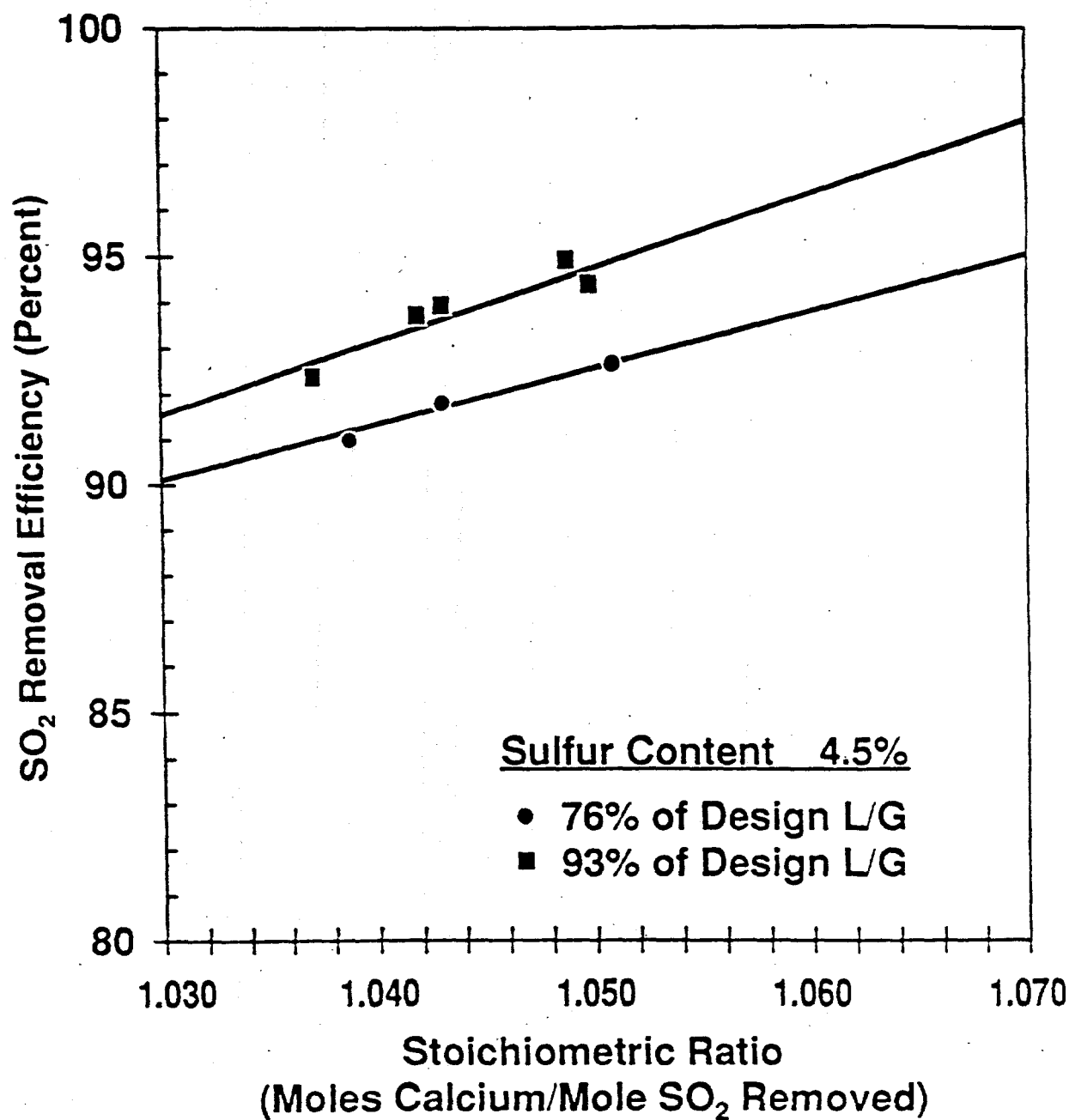


FIGURE 7

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

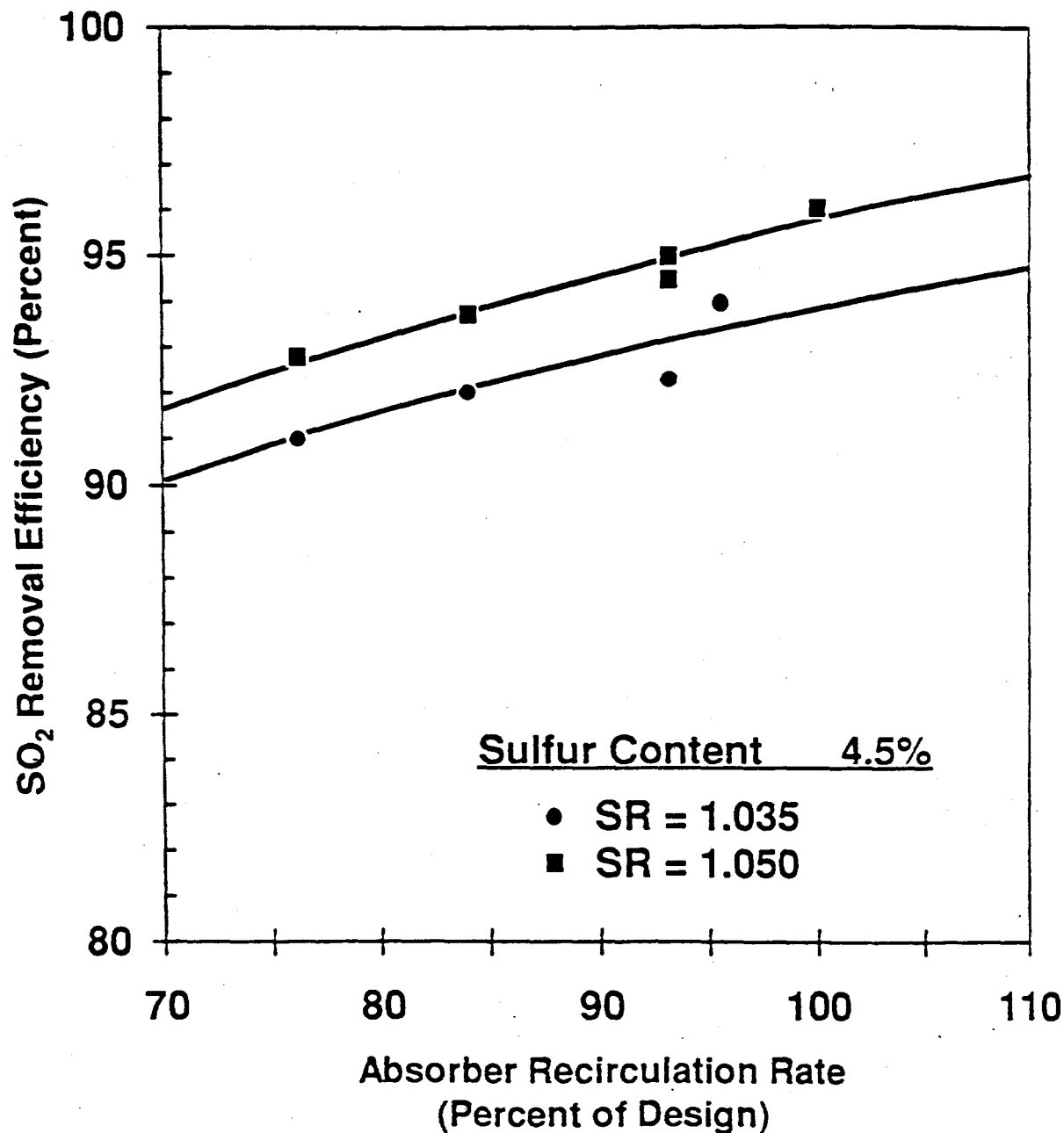




FIGURE 8

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

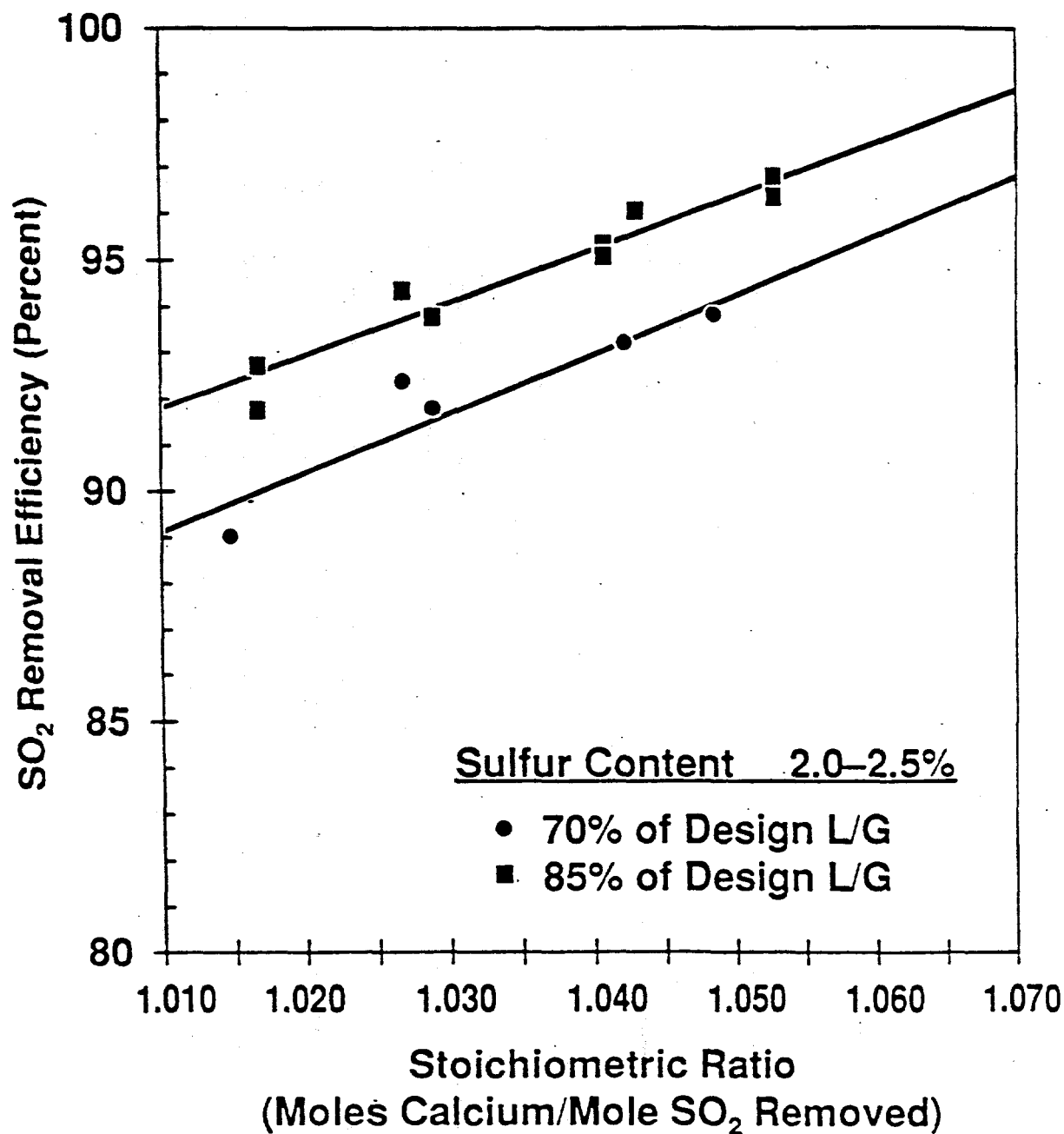


FIGURE 9

# SO<sub>2</sub> Removal Performance at Bailly AFGD (67 Percent Boiler Load)

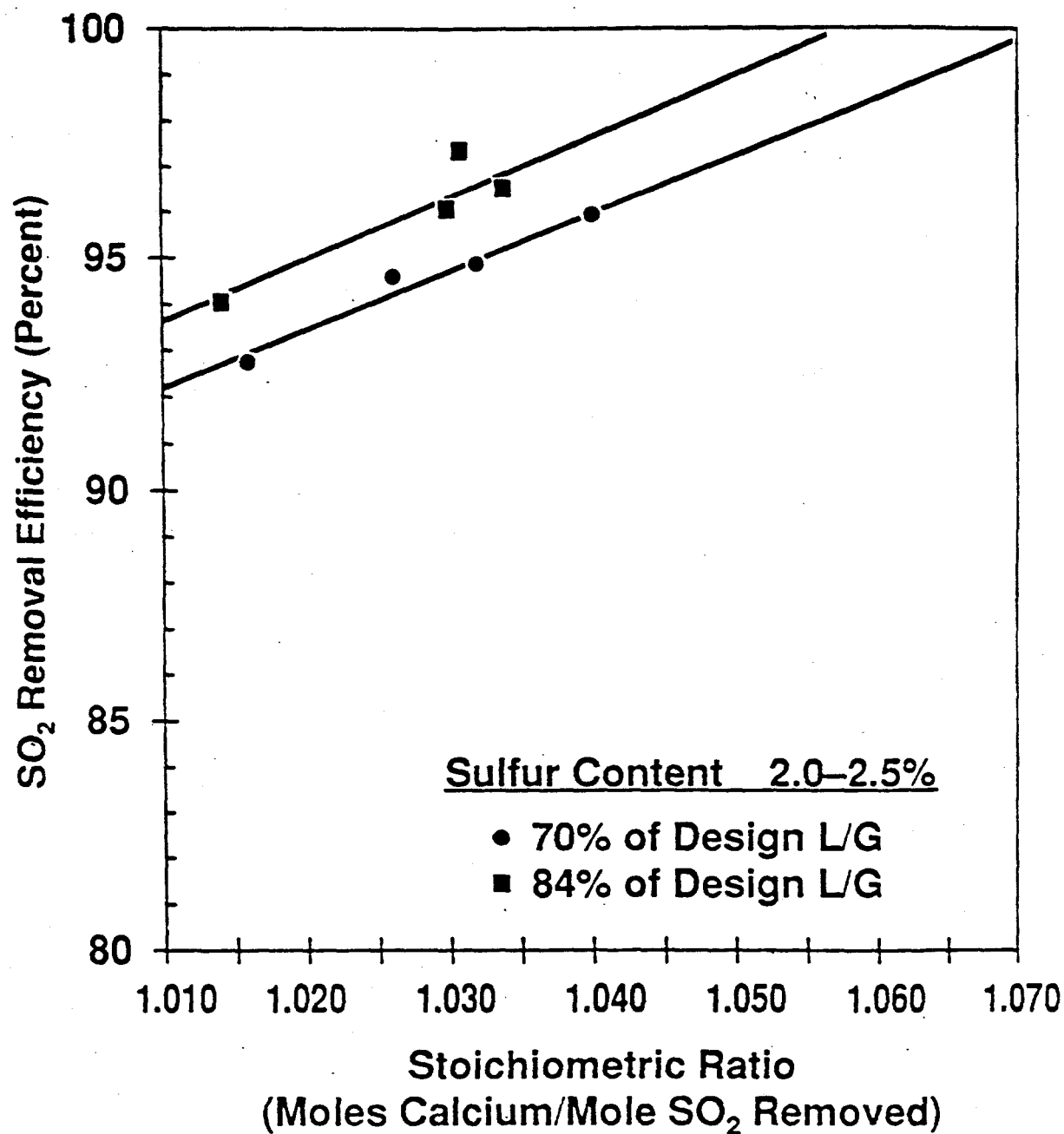


FIGURE 10

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

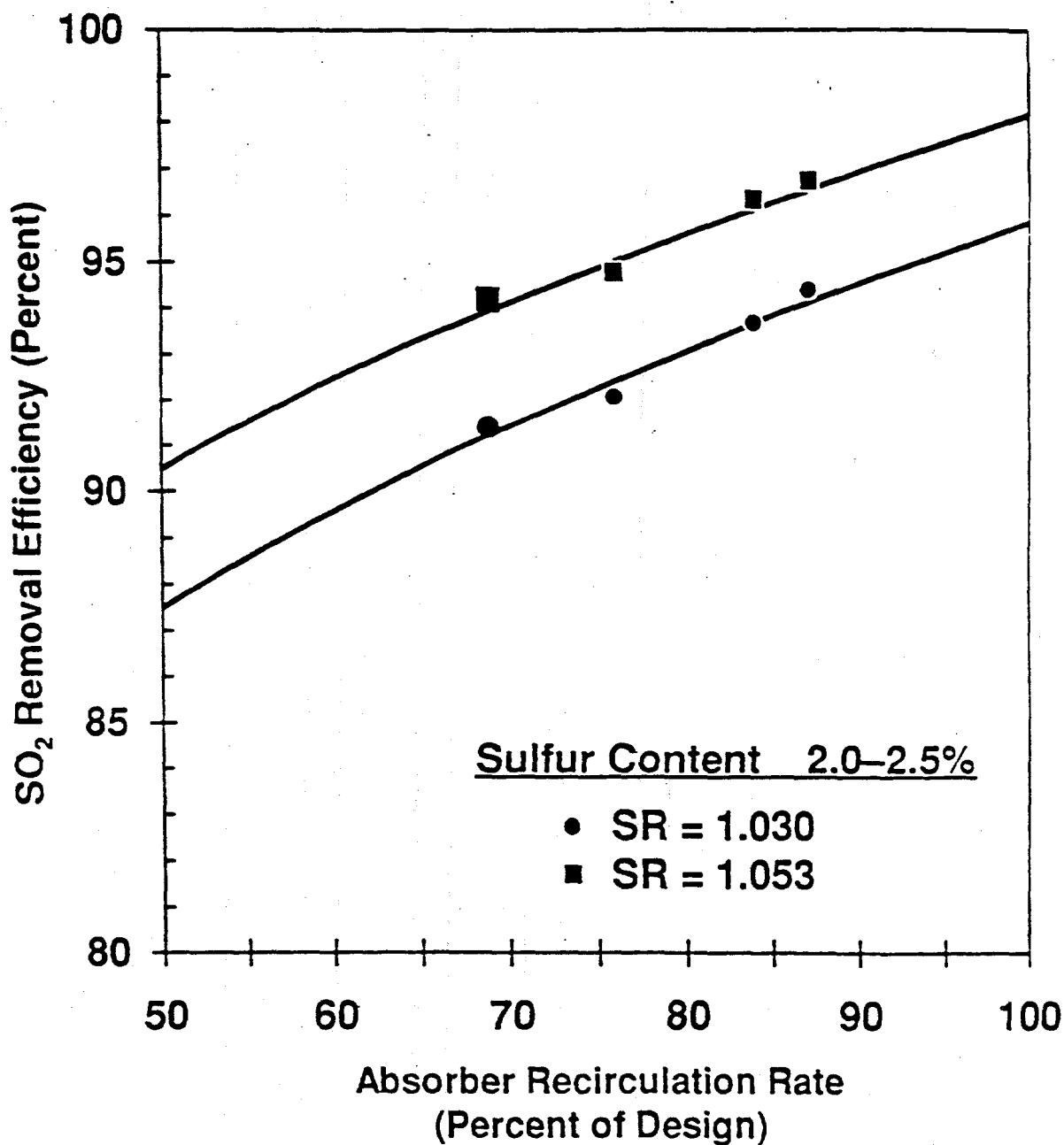


FIGURE 11

# SO<sub>2</sub> Removal Performance at Bailly AFGD (67 Percent Boiler Load)

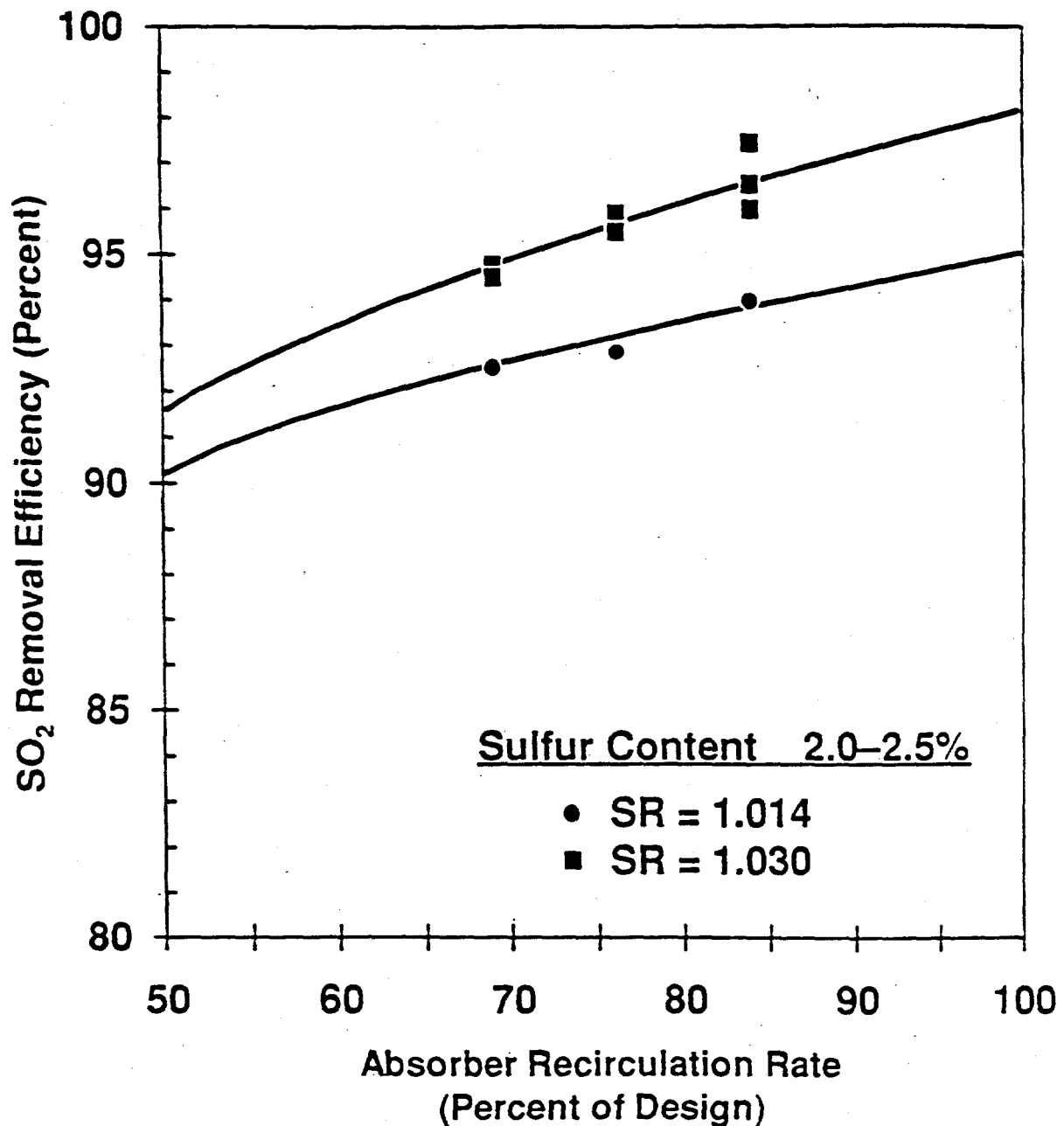


FIGURE 12

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

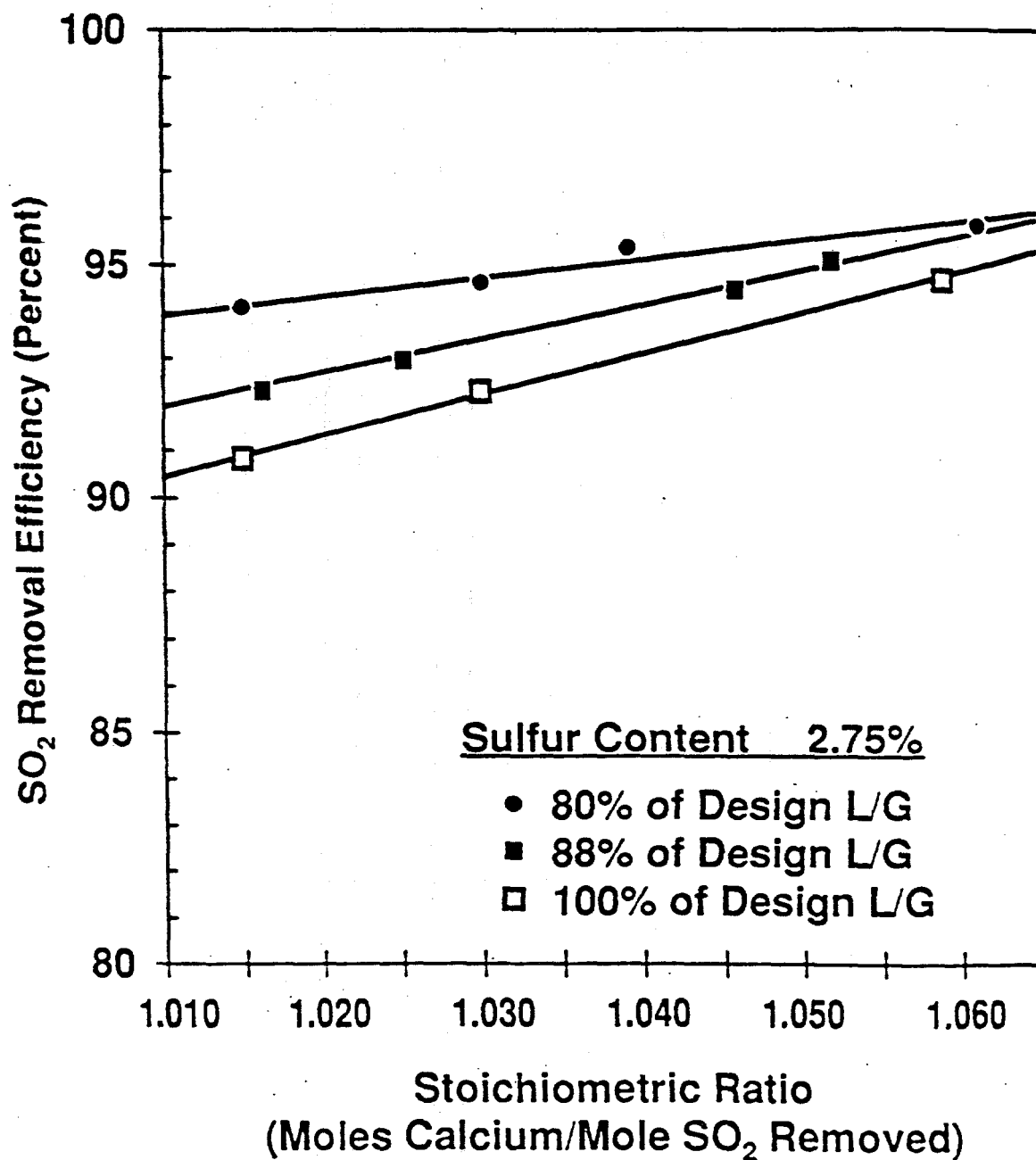


FIGURE 13

# SO<sub>2</sub> Removal Performance at Bailly AFGD (67 Percent Boiler Load)

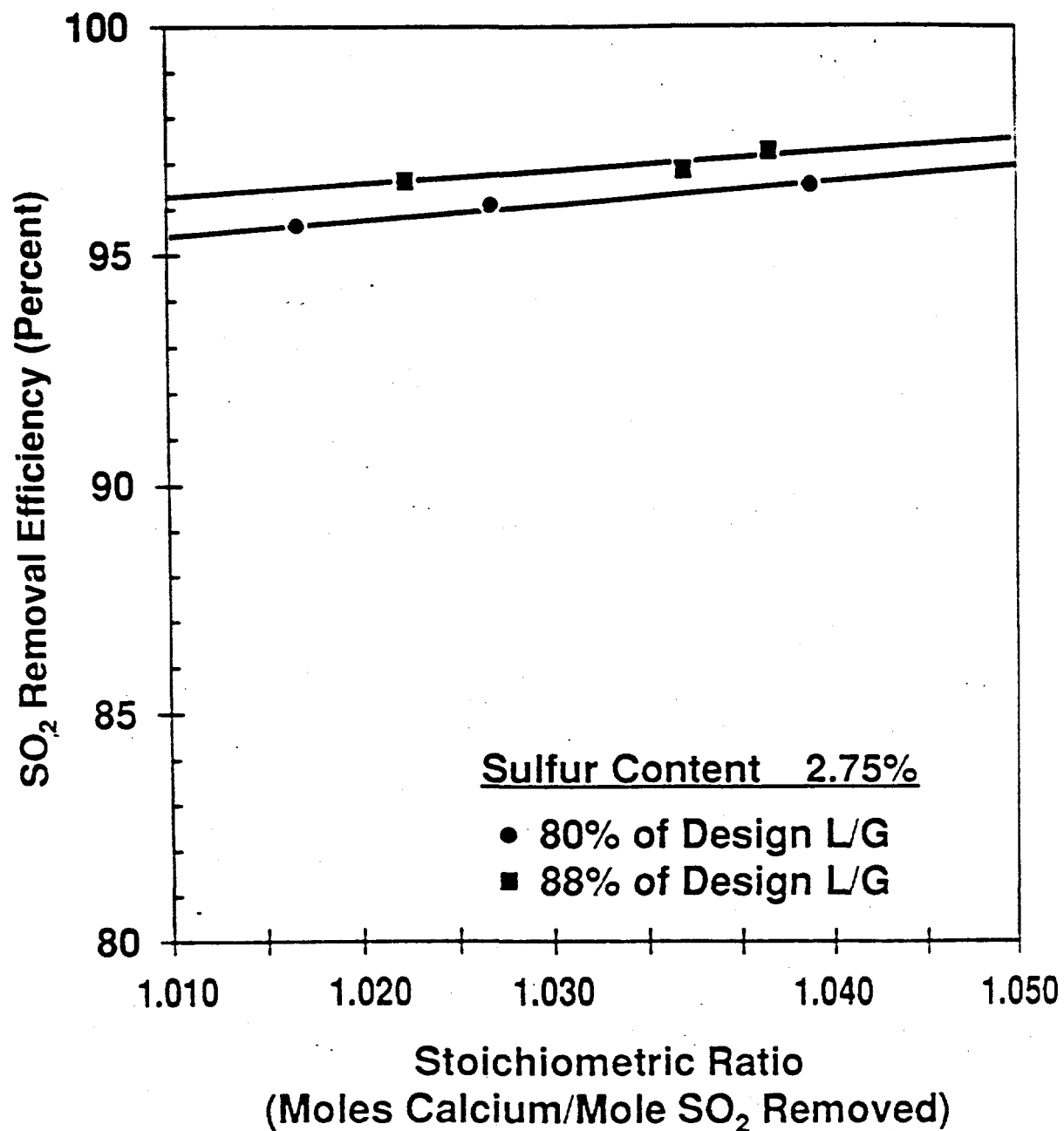


FIGURE 14

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

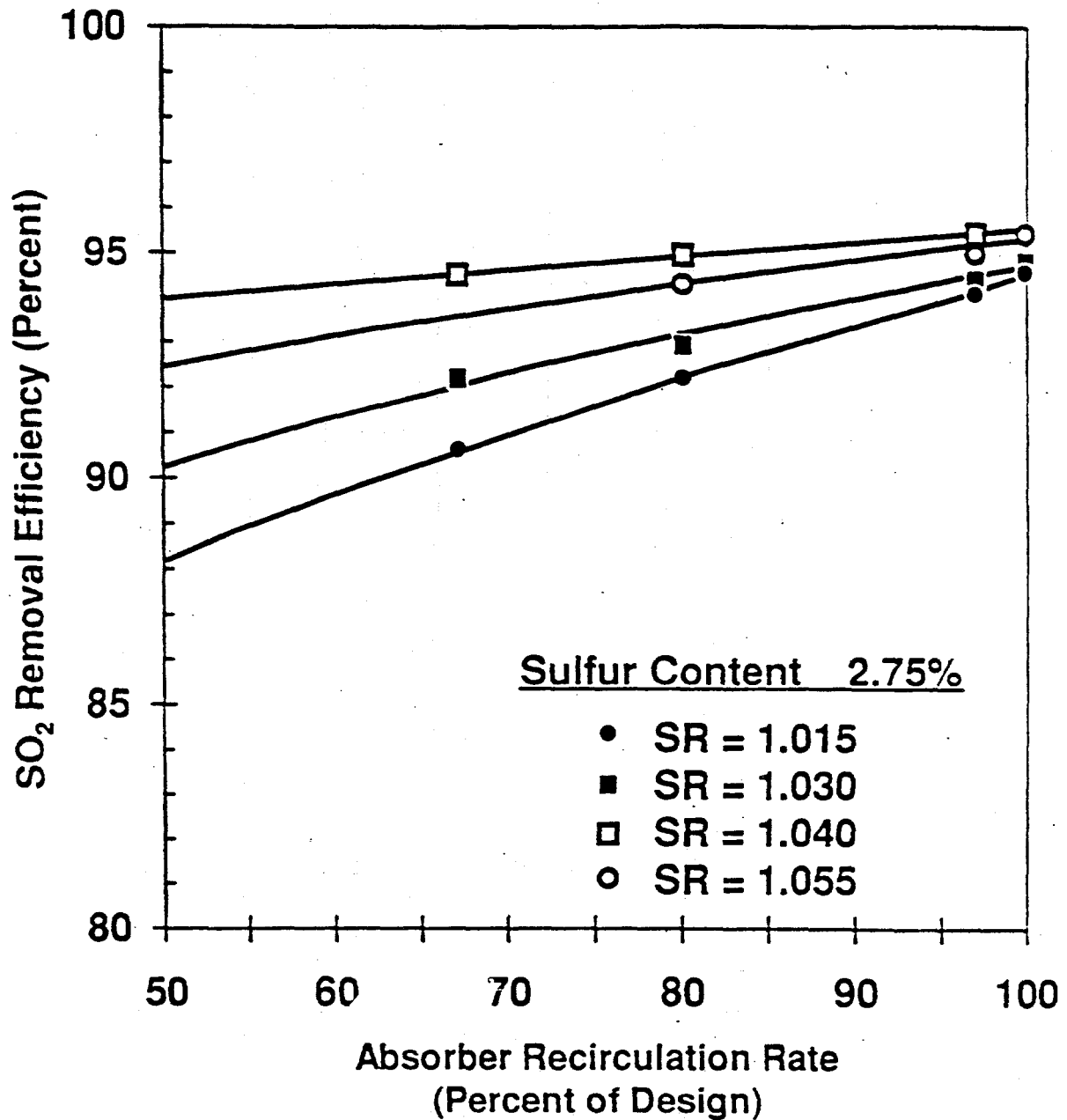


FIGURE 15

# SO<sub>2</sub> Removal Performance at Bailly AFGD (67 Percent Boiler Load)

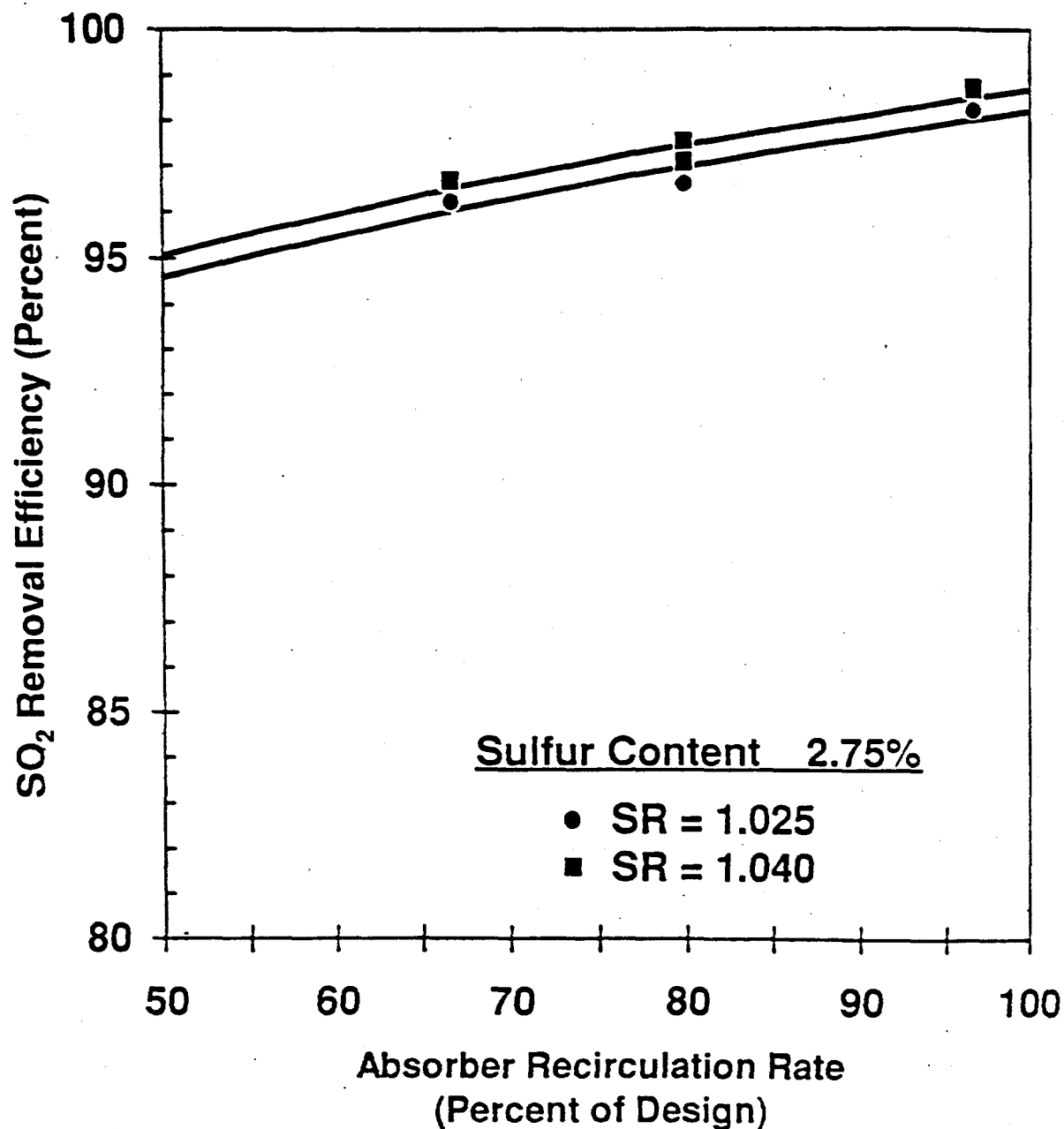




FIGURE 16

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

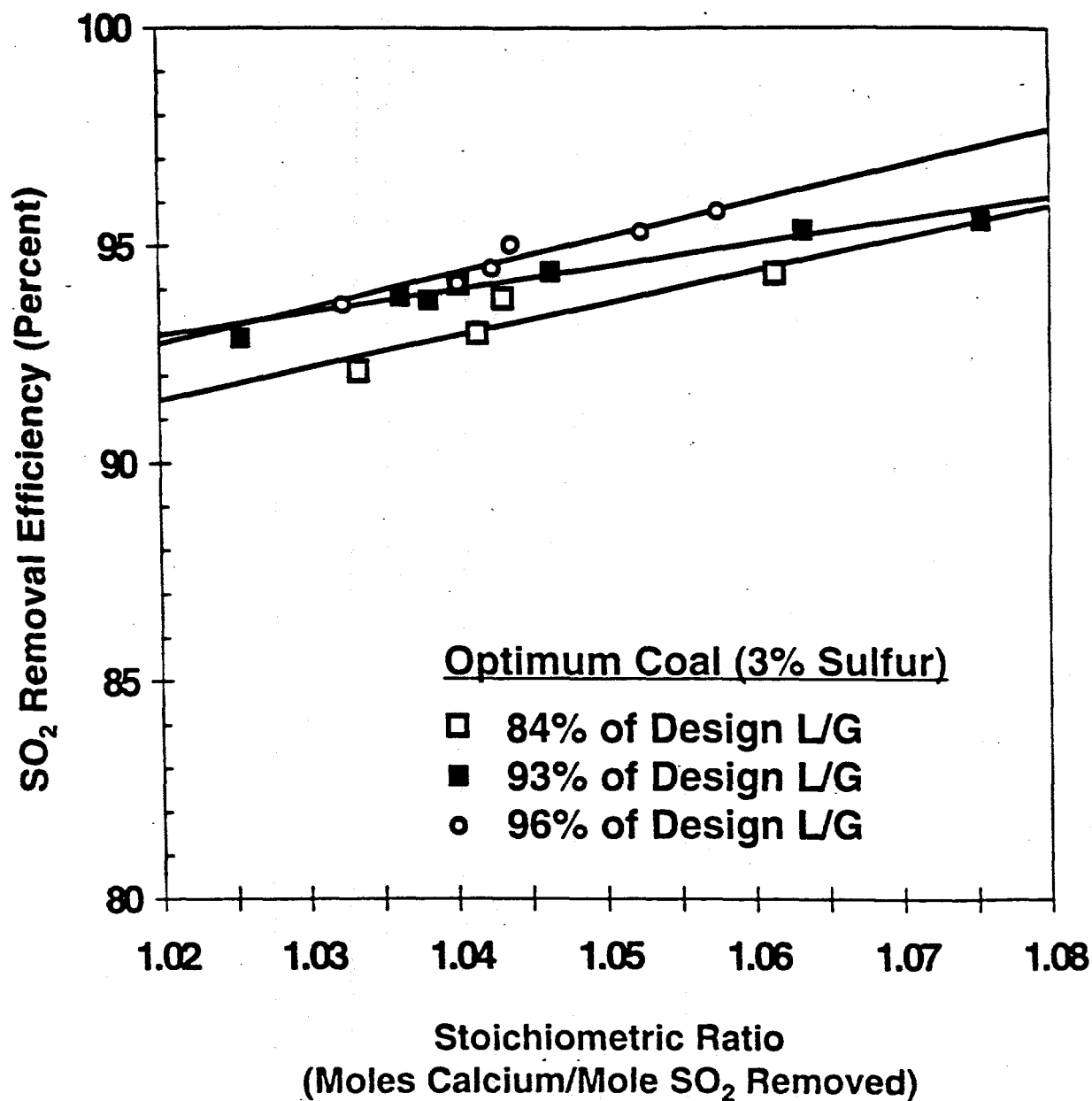


FIGURE 17

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

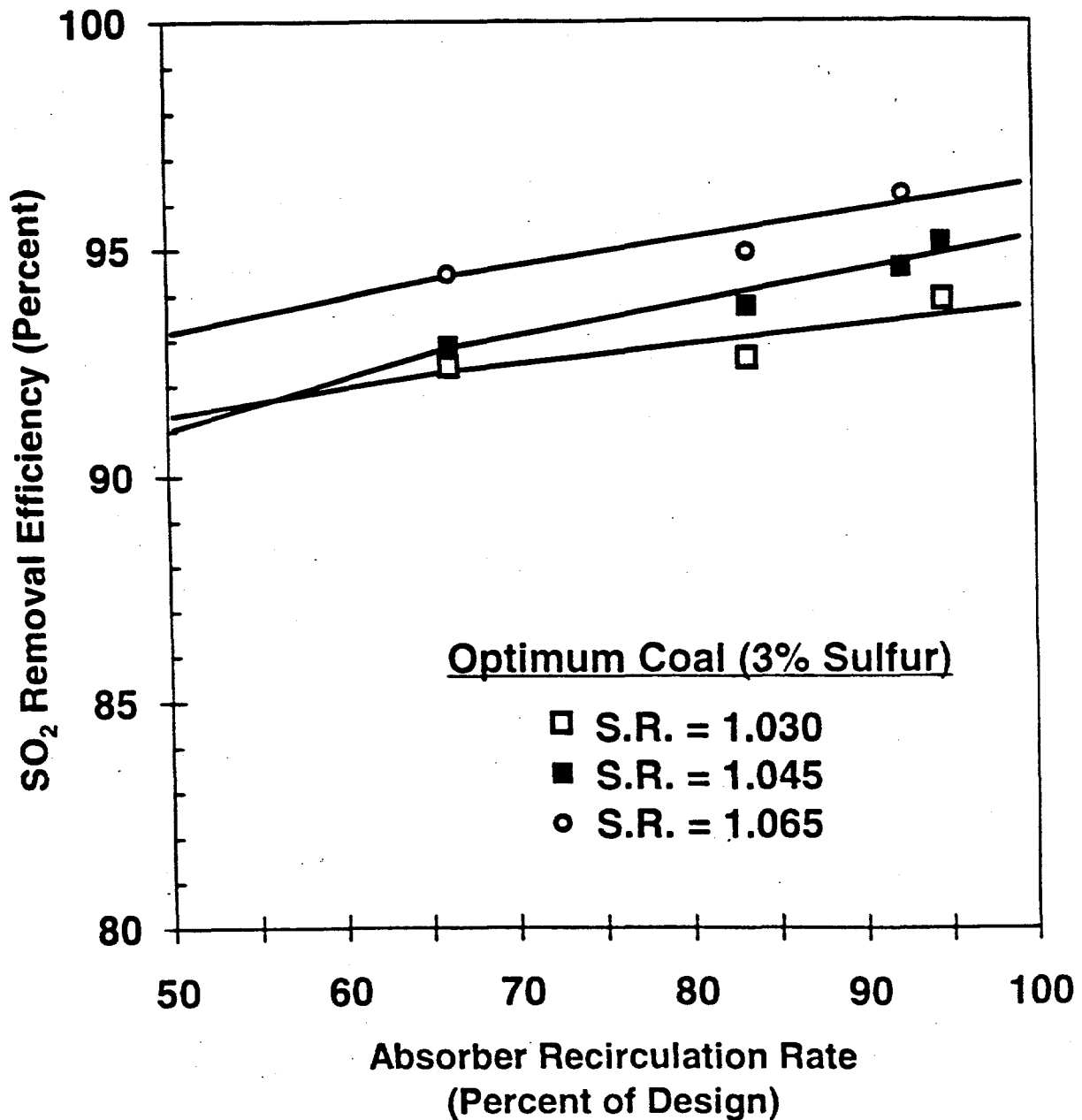


FIGURE 18

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)

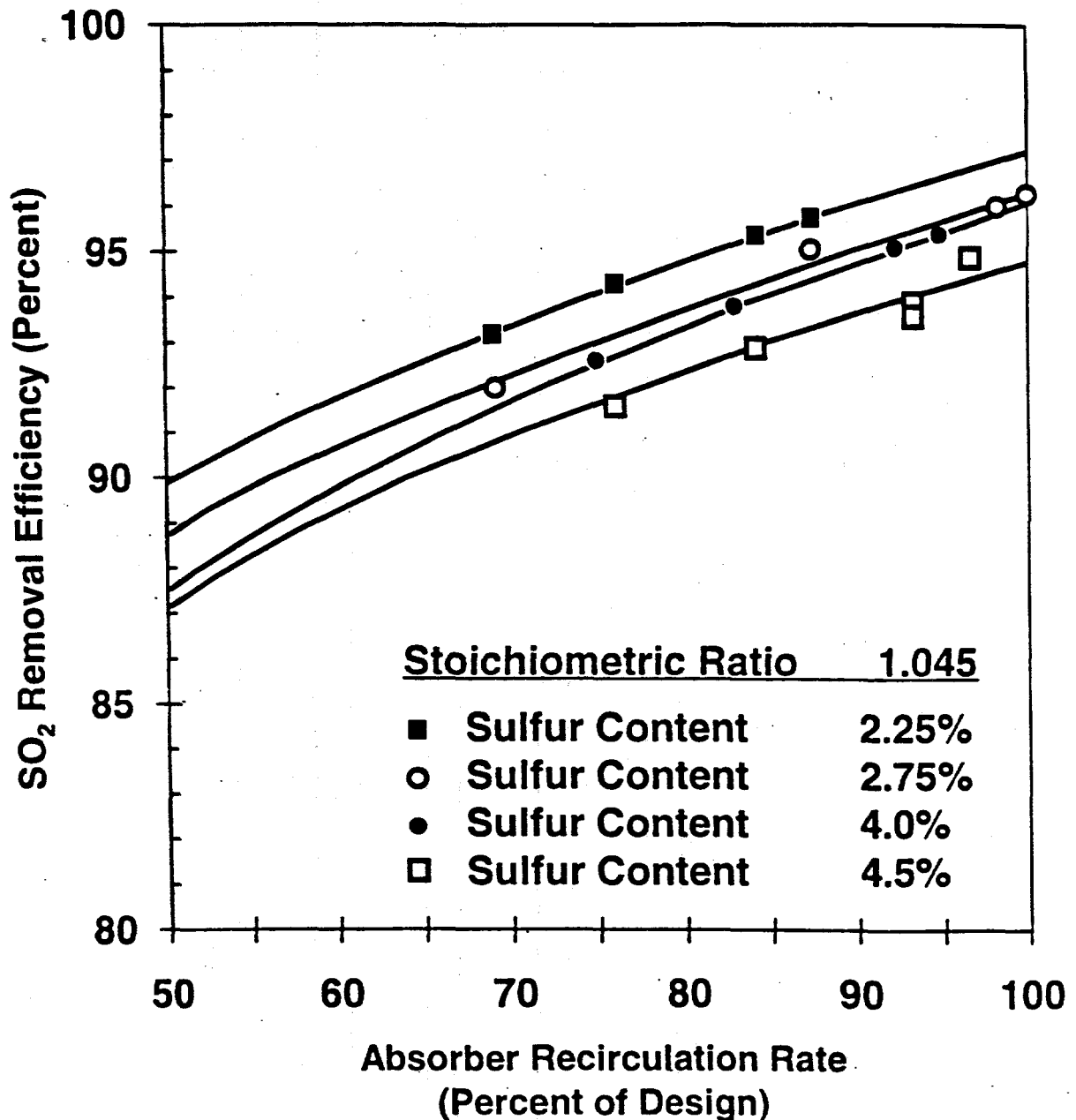
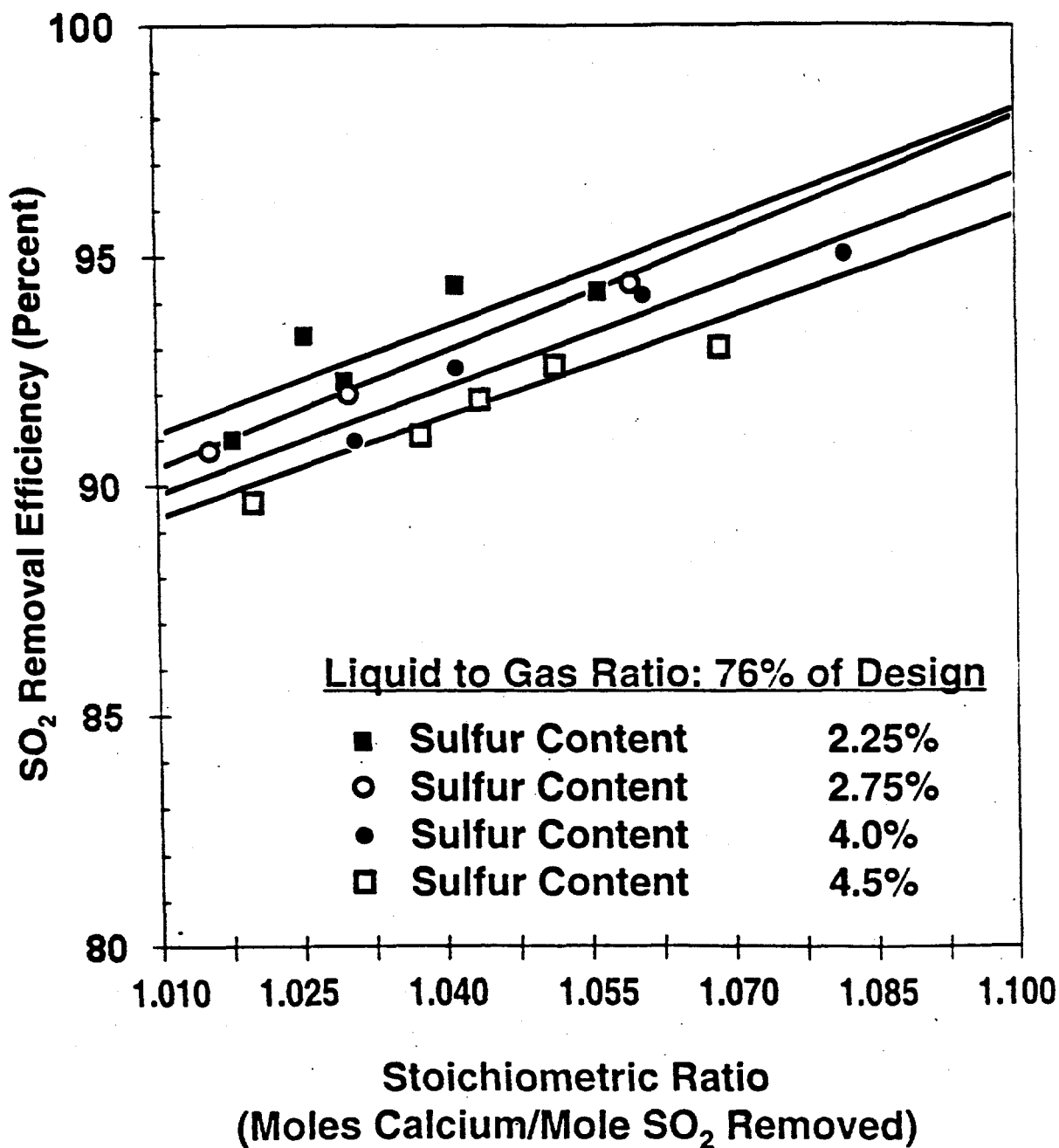


FIGURE 19

# SO<sub>2</sub> Removal Performance at Bailly AFGD (100 Percent Boiler Load)



**TABLE 1**  
**COAL ANALYSIS**

	DOE Demo I (S=2.25%)	DOE Demo II (S=2.75%)	DOE Demo III (S=3%)	DOE Demo IV (S=4%)	DOE Demo V (S=4.5%)	DOE Demo VI (S=3%)
Carbon	66.56	61.61	62.1	59.14	69.32	61.1
Hydrogen	4.5	4.39	4.09	4.37	4.94	4.32
Nitrogen	1.44	1.23	1.22	1.26	1.17	1.16
Sulfur	2.21	2.91	3.21	3.79	4.73	3.00
Oxygen	6.71	7.45	8.19	7.19	5.63	7.40
Chlorine	0.14	0.1	0.06	0.03	0.07	0.07
Moisture	8.6	12.92	11.14	13.72	4.74	13.05
Ash	9.53	9.63	10.1	10.7	9.3	9.93
Btu/lb	11,932	11,022	10,874	11,000	12,700	10,972

**TABLE 2**  
**GYPSUM ANALYSIS**

Ultimate Analysis (As Rec'd Weight %)

	DOE Demo I (S=2.25%)	DOE Demo II (S=2.75%)	DOE Demo III (S=3%)	DOE Demo IV (S=4%)	DOE Demo V (S=4.5%)	DOE Demo VI (S=3%)
Gypsum ( $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ , %)	96.7 - 99.7	96.3 - 99.4	94.6 - 98.8	93.5 - 97.3	95.6 - 99.7	97.4 - 99.1
Calcium Carbonate ( $\text{CaCO}_3$ , %)	0.7 - 2.8	0.4 - 2.8	1.5 - 3.7	0.4 - 4.5	1.6 - 2.9	1.4 - 3.5
Chloride (Cl, ppm)	<20 - 37	<20 - 38	<20 - 38	9 - 148	<20 - 37	20 - 107
Magnesium Oxide ( $\text{MgO}$ , %)	0.04 - 0.17	0.04 - 0.2	0.08 - 0.23	0.08 - 0.51	0.08 - .21	0.03 - 0.16
Free Moisture (%)	4.6 - 7.8	4.3 - 8.4	3.7 - 8.4	4.2 - 8.8	5.8 - 9.6	5.4 - 7.9

**TABLE 3**  
**OPERATIONS SUMMARY**

	<u>Expected</u>	<u>Achieved</u>
SO <sub>2</sub> Emissions	90% removal or 0.6 lb/MMBtu, whichever is less stringent	Averaged 94% (during DOE test up to 98+%, or 0.382 lb/MMBtu)
Power Consumption 24 hour average (kw)	<8,650	5,275
Facility Pressure Drop 24 hour average (IWC)	<13.5	3.23
Particulate Emissions (g/SCFD)	no net increase	0.04 inlet 0.0071 outlet
Availability (%)	95	99.47
Gypsum Moisture (%)	<10	6.64
Gypsum Chloride (ppm)	<120	33
Gypsum Purity (%)	93	97.2
Average Water Consumption (GPM)	3,000	1,560
Average Wastewater Flow (GPM)	275	81

**TABLE 4**  
**WALLBOARD-GRADE GYPSUM SPECIFICATIONS**

	<u>Expected</u>	<u>Four Year Average</u>
Gypsum Purity (wt. % dry)		
CaSO <sub>4</sub> · 2 H <sub>2</sub> O	>93.0	97.4
CaSO <sub>3</sub> · 1/2 H <sub>2</sub> O	<2.0	0.08
SiO <sub>2</sub>	<2.5	0.5
Fe <sub>2</sub> O <sub>3</sub>	<3.5	0.22
R <sub>2</sub> O <sub>3</sub>	—	0.3
Chlorides (ppm)	<120	30
Free H <sub>2</sub> O (wt %)	<10	6.5
Mean Particle Size (microns)	>20	50

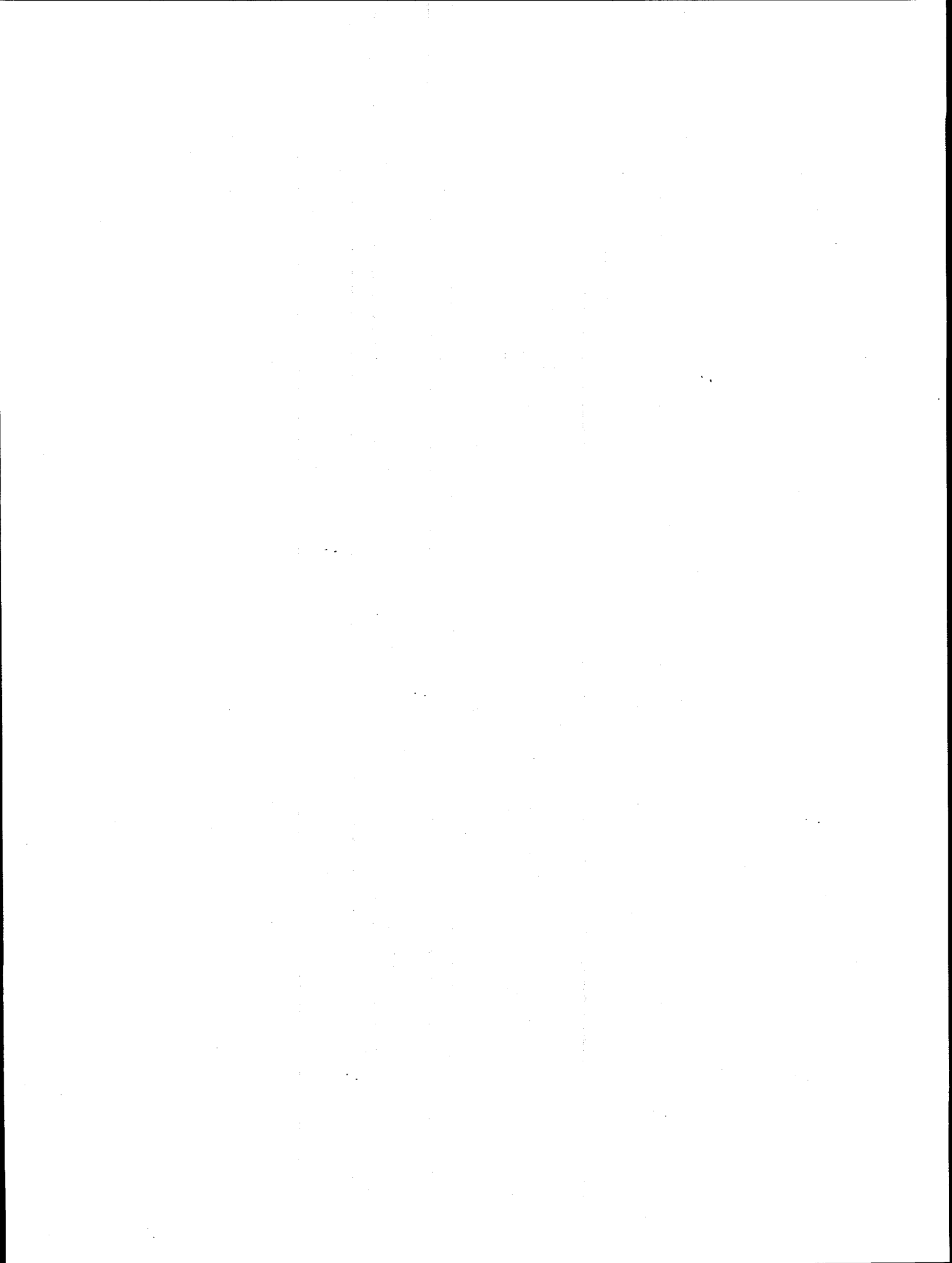


**TABLE 5**  
**WATER REQUIREMENTS**

	<u>Expected</u>	<u>Four Year Average</u>
Supply Water Flow (gpm)	<3,000	1,500
Wastewater pH	6.0 to 9.0	7.5 to 9.0
Wastewater Total Suspended Solids (ppm)	<30	<18
Wastewater Dissolved Solids (ppm)		
Chlorides (Cl)	<30,000	8,960
Sulfates ( $\text{SO}_4^{+2}$ )	<2,500	<2,500
Fluorides (F)	<1,100	23
Total Dissolved Solids	<100,000	15,000

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# DEMONSTRATION OF SCR TECHNOLOGY FOR THE CONTROL OF NO<sub>x</sub> EMISSIONS FROM HIGH-SULFUR COAL-FIRED UTILITY BOILERS

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## ABSTRACT

This paper describes the completed Innovative Clean Coal Technology project which demonstrated SCR technology for reduction of flue gas NO<sub>x</sub> emissions from a utility boiler burning U.S. high-sulfur coal. The project was sponsored by the U.S. Department of Energy, managed and co-funded by Southern Company Services, Inc. on behalf of the Southern Company, and also co-funded by the Electric Power Research Institute and Ontario Hydro. The project was located at Gulf Power Company's Plant Crist Unit 5 (a 75 MW tangentially-fired boiler burning U.S. coals that had a sulfur content ranging from 2.5 - 2.9%), near Pensacola, Florida. The test program was conducted for approximately two years to evaluate catalyst deactivation and other SCR operational effects. The SCR test facility had nine reactors: three 2.5 MW (5000 scfm), and six 0.2 MW (400 scfm). Eight reactors operated on high-dust flue gas, while the ninth reactor operated on low-dust flue gas. The reactors operated in parallel with commercially available SCR catalysts obtained from suppliers throughout the world. Long-term performance testing began in July 1993 and was completed in July 1995. A brief test facility description and the results of the project are presented in this paper.

## PROJECT GOALS

Although SCR is widely applied in Japan and western Europe, numerous technical uncertainties are associated with applying SCR to U.S. coals. These uncertainties include:

- (1) potential catalyst deactivation due to poisoning by trace metal species present in domestic coals but not present, or present at much lower concentrations, in fuels from other countries;
- (2) performance of the technology and effects on the balance-of-plant equipment in the presence of high amounts of SO<sub>2</sub> and SO<sub>3</sub> (e.g., plugging of downstream equipment with ammonia-sulfur compounds); and

- (3) performance of a wide variety of SCR catalyst compositions, geometries and manufacturing methods at typical U.S. high-sulfur coal-fired utility operating conditions.

These uncertainties were explored by constructing and operating a series of small-scale SCR reactors and simultaneously exposing different commercially available SCR catalysts to a common flue gas derived from the combustion of high-sulfur U.S. coal. First, long-term SCR catalyst performance was evaluated for two years under realistic operating conditions similar to those found in U.S. pulverized-coal-fired utility boilers. Deactivation rates for exposed catalysts were documented to determine life and associated process economics. Second, short-term parametric tests were performed during which SCR operating conditions were adjusted above and below design values to observe catalyst performance. The performance of air preheaters was also observed to evaluate the effects of SCR operating conditions on heat transfer, pressure drop and boiler efficiency. Third, honeycomb- and plate-type SCR catalysts of various commercial compositions from the U.S., Japan, and Europe were evaluated. Tests with these catalysts were used to expand knowledge of the performance of various catalyst types under U.S. utility operating conditions with high-sulfur coal.

### SCR TEST FACILITY DESCRIPTION

The SCR demonstration facility was located at Gulf Power Company's Plant Crist in Pensacola, Florida. The facility treated a flue gas slip-stream from Unit 5, a commercially operating 75-MW (nameplate) unit, firing U.S. coals with a sulfur content ranging from 2.5 - 2.9%. Unit 5 is a tangentially-fired, dry bottom boiler with hot- and cold-side electrostatic precipitators (ESPs) for particulate control. The SCR test facility consisted of nine reactors operating in parallel for comparisons of commercially available SCR catalysts obtained from suppliers throughout the world. With all reactors in operation, the amount of combustion flue gas that could be treated was 17,400 scfm or 12% of Unit 5's capacity (about 8.7 MWe). Table 1 shows the average test facility flue gas constituent concentrations measured over the life of the project. Also shown are representative particulate loadings.

Table 1: Test Facility Inlet Flue Gas Composition

Constituent	Average	Data Source
Unit Load (MW)	63	continuous data average over project life
NO <sub>x</sub> (ppmv at actual O <sub>2</sub> )	314	continuous data average over project life
O <sub>2</sub> (%)	4.8	continuous data average over project life
CO <sub>2</sub> (%)	14.7	continuous data average over project life
SO <sub>2</sub> (ppm)	1570	continuous data average over project life
SO <sub>3</sub> (ppm)	4	average of parametric test data
HCl (ppm)	104	start-up/commissioning test data
NH <sub>3</sub> (ppm)	< 0.4	start-up/commissioning test data
Particulate (gr/dscf)	3.1	estimated by interpolation using average unit load and high/low load start-up/commissioning data

The process flow diagram for the SCR test facility is shown in Figure 1. The facility consisted of three large SCR reactors (2.5 MW, 5000 scfm) and six smaller SCR reactors (0.2 MW, 400 scfm). Eight of the nine reactors were designed to treat flue gas containing full particulate loading (high dust) extracted from the inlet duct of the hot-side ESP, while one small reactor was designed to treat flue gas extracted from the hot-side ESP outlet (low dust). Only seven of the eight high-dust reactors were operated for catalyst long-term and parametric testing.

Each reactor train was equipped with an electric duct heater to independently control flue gas temperature and a venturi flow meter to measure the flue gas flow rate. An economizer bypass line maintained a minimum flue gas temperature of 620 °F to the high-dust reactors. Anhydrous ammonia was independently metered to a stream of heated dilution air and was injected via nozzles into the flue gas upstream of each SCR reactor. The flue gas and ammonia passed through the SCR reactors, which had the capacity to contain up to four catalyst layers.

For the large reactor trains, the flue gas exited the reactor and entered a specially modified pilot scale air preheater (APH). This modified APH was used to better simulate full scale APH's for improved extrapolation of results to commercial scale. The APHs were incorporated in the project to evaluate the effects of the SCR process on APH deposit formation and the effects of the deposits on APH performance and operation.

All reactor trains, except the low-dust train, had a cyclone downstream of the SCR reactor to protect the induced draft (ID) fans from particulates. The exhaust from all of the SCR reactors was combined into a single manifold and reinjected into the host boiler's flue gas stream ahead of the cold-side ESP. The preheated air from the APH on the large reactors was also combined into a single manifold and returned to the host boiler draft system at the existing host APH outlet. All particulate removed from the test facility was combined with ash from the host unit's ESP and sent to an ash disposal area.

The test facility examined the performance of eight SCR catalysts (one reactor was idled due to the withdrawal of a project participant), each with its own particular geometry and physical and chemical design. Each catalyst supplier was given great latitude in designing their particular catalyst offering for the project. Suppliers were required to meet the following general criteria.

- design catalyst baskets to match predetermined reactor dimensions
- provide a maximum of four catalyst layers
- insure a maximum base-line SO<sub>2</sub> oxidation rate of 0.75%
- insure a maximum base-line NH<sub>3</sub> slip at the reactor exit of 5 ppm
- maintain 80% NO<sub>x</sub> reduction while meeting NH<sub>3</sub> slip requirements
- design for 2-year life meeting performance criteria

The particular design characteristics of each catalyst follow in Table 2. The number of catalyst beds, general type of catalyst (honeycomb or plate) and reactor in which the catalyst was tested are provided. In addition, the project life flue gas exposure time for each of the catalysts is shown.

**Table 2. Catalyst Design Specifications**

Parameter	Grace Noxeram	NSKK	Siemens	Grace Synox	Cormetech High-Dust	Haldor Topsoe	Hitachi Zosen	Cormetech Low-dust
Reactor	A	B	C	D	E	F	G	J
Dust Level	High	High	High	High	High	High	High	Low
Composition	V-W/Ti	V-W/ Ti/Si	V/Ti	V/Ti/Si	V-W/Ti	NA <sup>6</sup>	V/Ti	V-W/Ti
Type <sup>5</sup>	HC	HC	Plate	HC	HC	Plate	Plate	HC
Pitch, mm (opening/wall thick.)	7.5 (6.1/1.4)	7.0	5.0	7.5 (6.1/1.4)	7.1 (6.0/1.1)	DNX-16 6.4 hyd.D.	5.5	3.7 (3.2/0.5)
Void Fraction, %	65	70	81	65	71	73	90	72
Density, lb/ft <sup>3</sup>	39±3.5	25	37 <sup>1</sup>	18±1.5	37	16	23 <sup>1</sup>	32
Geom. Surf. Area, m <sup>2</sup> /m <sup>3</sup>	430	470	383	430	470	455	420	910
Gas Flow, Nm <sup>3</sup> /hr	8500	8500	8500	680	680	680	680	680
Gas Flow, scfm	5000	5000	5000	400	400	400	400	400
Catalyst Volume, m <sup>3</sup>	3.1	3.026	2.30	0.19	0.245	0.189	0.27	0.097
GHSV @ 0°C, hr <sup>-1</sup>	2742	2809	3692	3579	2776	3600	2500	7033
Gas Flow Parametric Range (as % of design)								
Min.	63%	91%	60%	66%	60%	50%	NA	60%
Max.	126%	127%	150%	131%	150%	100%	NA	150%
Cross Sec. Area, m <sup>2</sup> <sup>2</sup>	1.080	1.08	1.106 (1.164)	0.09	0.081	0.094	0.09	0.081
Super.Lin. Veloc., Nm/s <sup>3</sup>								
Min.	2.186	2.186	2.135	2.10	2.34	2.00	2.1	2.34
Max.	2.76	2.5	3.203	2.76	3.0	2.5	2.5	3.5
No. of Beds	3	3	2	3	3	3	3	2
Temperature, °F	700	700	700	700	700	700	700	700
Temp. Parametric Range (°F)								
Min.	660	644	617	660	644	617	626	644
Max.	750	842	806	750	790	752	752	790
% SO <sub>2</sub> Oxidation	0.75	0.5	< 0.6	< 0.75	< 0.75	0.68	0.7	< 0.75
Inlet NO <sub>x</sub> , ppmv (wet)	400	400	417	400	400	400	400	400
NH <sub>3</sub> /NO <sub>x</sub>	0.813	0.811	0.817	0.813	0.81	0.81	0.85	0.81
NH <sub>3</sub> /NO <sub>x</sub> Parametric Range								
Min.	0.6	0.6	0.6	0.6	0.6	0.6	0.7	0.6
Max.	1.0	1.0	1.2	1.0	1.2	1.2	1.1	1.2
% deNO <sub>x</sub> Activity	80	80	80	80	80	80	80	80
NH <sub>3</sub> Slip, ppmv (wet)	< 5	5	5	< 5	< 5	< 5	< 5	< 5
Particulates, g/Nm <sup>3</sup>	6 - 8	6 - 8	6-8	6 - 8	6-8	6-8	6 - 8	NA
Pressure Drop, in. H <sub>2</sub> O	3.85 <sup>4</sup>	2.62	1.28	2.60 <sup>4</sup>	2.6	1.73	< 4	3.5
Actual Flue Gas Exposure Time (hrs.)	11012	11859	11632	10151	10151	10175	7293	5363

1. Includes basket, otherwise catalyst density only.

2. Catalyst only. Value in parentheses includes basket.

3. Velocity based on cross-sectional area of catalyst, not basket.

4. For Grace, ΔP includes baskets

5. HC = Honeycomb

6. NA = Not Available

## Fuel

One of the primary purposes of this project was to evaluate the performance of SCR technology on U.S. coals. To insure that an adequate database of coal composition was generated, daily as-burned coal samples were acquired from the host boiler. Monthly composites were then used to carefully track both primary and trace coal constituents. Where possible, neutron activation analysis (INAA) was used to augment other measurement techniques.

The coal supply during the project consisted of eastern bituminous coals with sulfur contents greater than or equal to 2.5%. When coal monitoring began (March 1993), Peabody coal having a 2.9% sulfur content was being burned. In September 1993, however, a coal miner's strike and buy-out of the Peabody contract caused a change in coal supply. During September and October, Kerr-McGee and Taft coals were purchased. The Kerr-McGee coal was a blend of Illinois #5 and #6 mines while the Taft coal was from Walker and Jefferson counties in Alabama. For November and December, coal from the Illinois basin in Western Kentucky was purchased from Jader to be burned. All of these coals had a sulfur content of approximately 2.5%. In January 1994, a new contract was set up to purchase high sulfur coal (2.7% sulfur) for the host unit from Old Ben Franklin and Kerr-McGee coal companies. The Old Ben coal from the Illinois #6 mine and Kerr-McGee from the Herring #6 mine continued to be the coal source throughout the remainder of the project.

A summary of the coal analyses performed over the project life follows in Tables 3 and 4. The analyses presented contain data measured by Alabama Power Company Laboratories (APCo) using atomic absorption and ion-coupled plasma techniques (primary constituents and some trace metals) and data measured by the University of Missouri-Columbia, Research Reactor Center using neutron activation analysis (INAA) to supplement trace metals testing. In cases where a constituent was measured by both laboratories, the data give a basis of comparison for the two analytical methods.

**Table 3. Proximate Coal Analysis Data**

Test	APCo Method	Units	APCo data
Moisture, Total	ASTM D 3302	% by Wt.	10.87
Ash	ASTM D 3180	% by Wt.	9.30
Gross Cal. Value	ASTM D 3180	Btu/lb	13268
Sulfur, Total	ASTM D 3180	% by Wt.	2.58
Sulfur lb/MMBtu	ASTM D 3180	lb/MMBtu	1.95
Carbon	ASTM D 3180	% by Wt.	74.82
Hydrogen	ASTM D 3180	% by Wt.	5.00
Nitrogen	ASTM D 3180	% by Wt.	1.58
Oxygen	ASTM D 3180	% by Wt.	6.73
Carbon, Fixed	ASTM D 3180	% by Wt.	52.83
Volatile Matter	ASTM D 3180	% by Wt.	37.88



**Table 4. Elemental Coal Analysis Data**

Element	Units	APCo data	INAA data	Element	Units	APCo data	INAA data	Element	Units	APCo data	INAA data
Aluminum	wt %	1.09	0.97	Hafnium	mg/kg		0.55	Scandium	mg/kg		2.61
Antimony	mg/kg	<1.0	0.74	Iron	wt %	1.08	1.04	Selenium	mg/kg	<2.0	1.91
Arsenic	mg/kg	3.2	7.6	Lanthanum	mg/kg		5.6	Silica	wt %	2.27	
Barium	mg/kg	40	58	Lead	mg/kg	12		Sodium	wt %	0.06	0.05
Beryllium	mg/kg	3		Lithium	mg/kg	9		Strontium	mg/kg		<38
Cadmium	mg/kg	< 1.0		Lutetium	mg/kg		0.12	Tantalum	mg/kg		0.13
Calcium	wt %	0.24	0.16	Magnesium	wt %	0.06		Terbium	mg/kg		0.14
Cerium	mg/kg		12.1	Manganese	mg/kg	24	24.6	Thorium	mg/kg		1.72
Cesium	mg/kg		0.97	Mercury	mg/kg	0.07	0.11	Titanium	wt %	0.06	0.06
Chlorine	mg/kg	1767		Molybdenum	mg/kg	7.8	9.3	Uranium	mg/kg		2.61
Chromium	mg/kg	19	16.7	Neodymium	mg/kg		6.2	Vanadium	mg/kg	41	39.5
Cobalt	mg/kg	7	3.61	Nickel	mg/kg	15	<28	Ytterbium	mg/kg		0.50
Copper	mg/kg	9		Phosphorus	wt %	0.02		Zinc	mg/kg	39	47
Dysprosium	mg/kg		0.92	Potassium	wt %	0.20	0.18	Zirconium	mg/kg		34
Europium	mg/kg		0.23	Rubidium	mg/kg		12.5				
Fluorine	mg/kg	56		Samarium	mg/kg		1.21				

## TEST PLAN

The project test plan was divided into five main categories as follows. Start-up and commissioning tests were performed during the initial start-up of the test facility beginning in early 1993 and continuing until the beginning of long term operations (July 1993). The remaining categories of tests were performed during the two year period of long term operations.

- |                                     |                        |
|-------------------------------------|------------------------|
| 1) Start-up and Commissioning Tests | 4) Air Preheater Tests |
| 2) Long-Term Performance Tests      | 5) Miscellaneous Tests |
| 3) Short-Term Parametric Tests      |                        |

### Start-Up and Commissioning Tests

The start-up and commissioning of the test facility was of extreme importance to the overall success of the project. Tests were performed to demonstrate that parameters such as ash loading, particle size distribution, flue gas flow, velocity distribution, and flue gas constituents (including trace metals), etc. were comparable between the reactors and were representative of the host unit flue gas parameters. Also, the start-up and commissioning tests showed that the specifications for the test facility design had been met. These basic design criteria follow in Table 5.

**Table 5. Test Facility Design Criteria**

Parameter	Design Specification
Flue Gas Flow Rate	Control to within $\pm 2\%$ for large reactors, $\pm 5\%$ for small reactors
NH <sub>3</sub> /NO <sub>x</sub> Ratio	Control to within 0.005 of desired NH <sub>3</sub> /NO <sub>x</sub>
Flue Gas Velocity Distribution	Not more than $\pm 10\%$ deviation in flow velocity across the cross-section of each individual reactor
NH <sub>3</sub> Distribution	Not more than $\pm 5\%$ deviation in ammonia distribution
Particulate Concentration	Not more than $\pm 5\%$ deviation between total mass loading between individual reactors, not more than $\pm 10\%$ deviation in particle size distribution

### **Long-Term Catalyst Performance Testing**

Long term catalyst performance testing focused on the performance of the catalysts/reactors at baseline conditions evaluating long-term performance parameters such as catalyst deactivation, ammonia slip, and reactor pressure drop. This evaluation depended on continuous or periodical measurement of various operational parameters to determine performance. The reactor baseline conditions under which these long-term tests were performed are identical to the baseline conditions shown for the short term parametric tests in [Table 6](#).

### **Catalyst Pressure Drop, Fouling, and Erosion**

Reactor pressure drops were measured continuously using the test facility's distributed control/data acquisition system. The data shown in [Figure 2](#) represent pressure drop in inches of water column across the catalyst beds in each of the reactors in the test facility. The pressure drop created by the dummy beds has been excluded to more closely represent actual catalyst pressure drop. In practice, full scale installations experience very little pressure drop across the flow straightening grid, unlike the test facility which required significantly higher dummy bed pressure drops to meet design specifications. To produce a meaningful plot, the data has been limited to periods of operation at or near baseline conditions. Periods where no data is shown represent periods when either the particular reactor or entire test facility was not in operation.

The general trend of the plot shows a fairly steady pressure drop over time for most of the test facility catalysts. This is encouraging because it indicates that sootblower operations were providing the cleaning necessary to prevent long term fouling of the catalysts. Sootblowing at the test facility was normally carried out on each of the catalyst beds and dummy beds every eight hours. The large reactors utilized a traversing rake steam sootblowing system. Steam was injected at approximately 250 psig at a distance of roughly 2 ft. above each of the catalyst layers and dummy layer. The small reactors were manually sootblown using a compressed air lance. Interestingly, due to the delicate nature of the Hitachi Zosen catalyst, this catalyst was not sootblown. This did not, however, adversely affect the fouling of this catalyst, as the extremely low facial area of the catalyst prevented fouling to a great degree.

Catalyst fouling and erosion was tracked through the analysis of reactor pressure drops and periodic visual inspections. In commercial SCRs, which have relatively infrequent visual inspections of the catalyst layers, the primary indication of catalyst fouling is by the monitoring of reactor pressure drop. In addition, monitoring of NO<sub>x</sub> and/or slip distributions at the reactor

outlet can give an indication of catalyst fouling. Erosion in commercial SCRs is tracked through visual inspection and by periodic sampling of the catalyst at which time erosion can be determined in the laboratory. Severe erosion may also be detected by an analysis of catalyst bed pressure drops.

Since catalyst sampling at the test facility required quarterly shut-down of the reactors, visual inspection of all catalyst beds was made at least on a quarterly basis. In practice, other outages were also used to make visual inspections, resulting in a quite frequent inspection schedule. These inspections usually indicated some fouling near the reactor walls. Problems due to this fouling were greatly mitigated on the large reactors by the design of the catalyst baskets. Catalyst suppliers had insured a smooth transition of flow from the reactor dimensions to the actual catalyst open face dimensions, thus preventing significant fouling due to wall effects. The small reactors exhibited more fouling near the reactor walls, primarily due to the small size of these reactors thus exacerbating the wall effects. In any case, all high-dust catalysts demonstrated an ability to handle the dust loadings without significant impacts to performance. In the case of the low-dust reactor, more fouling was noted. This was not a result of catalyst design, however, but was the result of test facility design. Long duct runs with increased heating requirements and a less than optimum take-off scoop all contributed to large particulate material being passed through the low-dust reactor system. The catalyst was not designed for these conditions and the fluctuation in this catalyst's pressure drop should not be construed as demonstrating the inapplicability of a low-dust configuration.

An important addition to each of the catalyst beds was a screen used to assist in the prevention of catalyst fouling. Screens placed several inches above the catalyst face with a mesh opening roughly equal to or slightly smaller than the catalyst channel openings proved to be extremely helpful in preventing channel blockage. The screens helped by breaking up large ash particles into small pieces which could pass through the catalyst rather than lying on the catalyst face, thereby blocking channels. Large ash particles that did not break up merely rested on the screen until they could be removed through periodic cleaning. Since the screens were placed several inches above the catalyst face, flow could redistribute around the large ash pieces, thus minimizing adverse effects.

The harsh environment in which catalysts in coal fired applications must operate make high physical strength and durability a necessity for catalysts. Erosion of the catalyst can be worsened by areas of high velocity or high particulate as well as by erosion from sootblowing operations. Severe erosion in certain areas of a particular catalysts is accelerated by the fact that eroded areas become the preferred flow path due to localized reduced pressure drop, thereby increasing the erosion potential in that particular area. This phenomenon seems to be more prevalent with honeycomb catalysts, however, leading edge hardening of the catalyst can help to mitigate facial erosion. In the case of some plate catalysts such as the Siemens, the catalysts support is steel, preventing significant erosion past the face of the catalyst. In the case of the Haldor Topsoe catalyst, a face hardening procedure appeared to enhance this catalyst's ability to withstand facial erosion. Little facial erosion was noted on the Hitachi Zosen catalyst likely due to the extremely small cross-sectional area of exposed catalyst face (due to extremely thin plates). The majority of

erosion on the honeycomb catalysts appeared to be due to aggressive sootblowing procedures (based on the erosion pattern). One exception to this finding however, was the Grace Synox catalyst. This special-design catalyst represents a significant deviation from other honeycomb catalysts currently marketed. The bimodal design resulted in a relatively soft material which showed significant erosion during the test program mainly indicated by the thinning of the catalyst walls. Also, some dummy bed material initially used in the test facility was considerably less durable than the catalysts. This material exhibited the phenomenon previously discussed where erosion problems were exacerbated by channeling.

### **Catalyst deNO<sub>x</sub> Activity and Deactivation**

As expected, the deNO<sub>x</sub> activity of the test catalysts varied greatly with respect to bulk volume. However, by varying bulk volume, all catalysts in the test program met the long-term performance criteria. DeNO<sub>x</sub> activity is an important parameter because the bulk volume (and weight) control to a large degree the number of catalysts beds and physical size and structure of an SCR reactor and the resulting capital cost. Generally, as NO<sub>x</sub> reduction increases to high levels (>90%) ammonia slip levels increase dramatically. The specific design of the reactor is important in this observation. A reactor could, of course, be designed for 90% NO<sub>x</sub> reduction while maintaining very low values of ammonia slip. This however, results in a high capital cost due to the disproportionately increased amount of catalyst required. High slip values at high NO<sub>x</sub> reduction are not only due to catalyst volume and design, but also to the physical design of the reactor. As NO<sub>x</sub> reduction nears 100%, non-idealities in the reactor design, such as NO<sub>x</sub> distribution, ammonia distribution, and velocity distribution become extremely important and eventually control the maximum NO<sub>x</sub> reduction that can be obtained for a specific reactor.

The catalyst suppliers were given a great deal of latitude in specifying the volume of catalyst for their respective reactors as previously mentioned. In some cases suppliers chose to minimize catalyst volume by designing a system which met, but did not exceed (improve upon) performance requirements. In other cases, catalyst suppliers chose to increase catalyst volume to insure performance margin and thus improve performance. Honeycomb catalysts are generally considered to have a high surface area when compared to plate-type catalysts (on a bulk volume basis) and would thus generally require less volume (although weight may be greater). However, offerings in the test program demonstrated the ability of plate-type catalysts to meet specifications with similar or less volume than honeycomb configurations. Thus, it appears that installed catalyst design margin can be as significant to overall catalyst volume as is the basic geometrical (honeycomb/plate) design.

One of the fundamental purposes of the test program was to address the potential catalyst deactivation associated with U.S. high-sulfur coal applications. Although a considerable experience base was present for foreign coals, little was known about potential catalyst poisons and the resulting catalyst deactivation of U.S. coals prior to this study. The primary method of tracking this deactivation was through laboratory tests performed by the catalyst suppliers. Catalyst samples were removed from the reactors on roughly a quarterly basis and sent to the respective laboratories for testing. Results were then reported directly to SCS. Data are presented in Figure 3 at the base condition for which the most data is available. This base

condition corresponds to a temperature of 700 °F, an area velocity of 50 Nm<sup>3</sup>/m<sup>2</sup>h, an inlet NO<sub>x</sub> concentration of 400 ppm, an NH<sub>3</sub>/NO<sub>x</sub> ratio of 1.0 and an SO<sub>2</sub> concentration of 2000 ppm. These deactivation trends are in keeping with expectations of the catalyst suppliers based on other experience, primarily foreign installations. Trace analysis of potential catalyst poisons such as arsenic and sodium were performed at the individual supplier's laboratories to correlate these concentrations to the measured deactivation rates. Interestingly, poison levels (especially arsenic) on the catalysts were higher than would be expected for the measured amount of deactivation.

### ***Long-Term Ammonia Slip***

The general result of catalyst deactivation is the increase in ammonia slip from the reactor over time assuming deNO<sub>x</sub> efficiency is held constant. Catalyst deactivation, however, is not always the source of increases in ammonia slip. Other factors such as maldistributions in NO<sub>x</sub> and ammonia, fouling, and erosion can all contribute to increases in ammonia slip over time. As in the test facility, most commercial SCRs operate at constant deNO<sub>x</sub> efficiency. This is necessary to meet emission regulations, but results in ammonia slip increases as reactor operation becomes less than optimum and the catalyst deactivates. Unfortunately, this ammonia slip is often poorly tracked in commercial SCRs, and thus the overall "health" of an SCR is not known. Consequently, the test facility monitoring program did closely track ammonia slip from the reactors at base-line conditions as well as other parametric conditions to adequately characterize the reactor/catalyst performance. As expected, ammonia slip increased over the life of the project. Figure 4 shows this general trend using the combination of all test facility catalysts. Data has been limited to operation at or very near base-line conditions. It should be noted that the depicted increases in ammonia slip are due not only to catalyst deactivation, but also due to degradation in reactor-specific parameters such as NH<sub>3</sub>/NO<sub>x</sub> distribution, velocity distribution, etc. Careful monitoring of these parameters, however, helped to minimize their contribution to the overall performance degradation of the reactor. The plot shows very low ammonia slip at the start of the project (<2 ppm for all catalysts with an average of less than 1ppm). As exposure time increased, however, ammonia slip increased, up to an average of roughly 3 ppm, with some catalysts at or near the design limit of 5 ppm. This demonstrates maturity in catalyst design and also shows that deactivation beyond that normally expected did not occur (based on prior world-wide experience). Figure 5 shows average ammonia slip for each of the catalysts on an individual basis using data taken over the life of the project at or near the base-line operating condition. It should be cautioned that ammonia slip is a direct function of catalyst volume and economics must be addressed in determining the most appropriate catalyst for a particular installation. It should also be noted that catalyst exposure times differ significantly in some cases (see Table 2). The data presented represent average ammonia slip over the life of the project at conditions very near baseline. High and low measurement variability is shown. This is a measure of repeatability over the project life. The high values shown should not be construed as end-of-project ammonia slip values and therefore do not indicate a failure of the particular catalyst to meet design specifications.

### ***Intermediate NO<sub>x</sub> Reduction***

Intermediate NO<sub>x</sub> reduction was tracked throughout the life of the project using NO<sub>x</sub> and ammonia measurements taken upstream and downstream of the first catalyst bed at various

parametric conditions during the five parametric test sequences. In general, deactivation based on  $\text{NO}_x$  measurements is difficult to see in a small scale facility due to the relatively slight decline in activity, especially considering measurement accuracy. However, first bed  $\text{NO}_x$  reduction measurements are most likely to exhibit an effect from catalyst deactivation than are other  $\text{NO}_x$  measurements such as over-all  $\text{NO}_x$  reduction. Since intermediate  $\text{NO}_x$  reduction is a strong function of the  $\text{NH}_3/\text{NO}_x$  ratio, it can be difficult to discern trends since in practice the actual  $\text{NH}_3/\text{NO}_x$  ratio is difficult to hold constant. However, a large number of points at or near the same  $\text{NH}_3/\text{NO}_x$  ratio can be used to track decreases in catalyst activity. Also shown in Figure 4 is the intermediate  $\text{NO}_x$  reduction versus catalyst exposure time at base-line conditions. An average value of all reactors has been used to increase the total amount of data available. The plot exhibits a very slight decrease in  $\text{NO}_x$  reduction over the nearly 12,000 hours of catalyst exposure time. This indicates that on average, the catalysts were not deactivating rapidly as is supported by the laboratory activity tests previously discussed.

### Short-Term Parametric Testing

Five sequences of parametric tests were performed on the test facility catalysts during the project life. These parametric tests were designed to examine ammonia slip,  $\text{deNO}_x$  efficiency (intermediate ammonia),  $\text{SO}_2$  oxidation,  $\text{N}_2\text{O}$  formation,  $\text{NO}_x$  and ammonia distributions, fly ash composition and loading, velocity distributions, and particulate distributions at varying conditions. Table 6 shows the general variations in temperature, flow rate, and  $\text{NH}_3/\text{NO}_x$  ratio that were tested. These parametric tests are most important in helping to define the applicable operating boundaries of the SCR reactors and their responses to changes in the various parameters.

Table 6. Parametric Conditions

Parameter	Minimum	Base-Line	Maximum
Temperature, ( $^{\circ}\text{F}$ )	620	700	750
$\text{NH}_3/\text{NO}_x$ molar ratio	0.6	0.8	1.0
Space velocity, % of design flow	60	100	150
Flow rate (scfm) -large reactor	3000	5000	7500
-small reactor	240	400	600

### Ammonia Slip

In general, ammonia slip, not  $\text{NO}_x$  reduction is the controlling factor in the long term operation of a commercial SCR. Unless extremely severe deactivation has occurred, required  $\text{NO}_x$  reduction can almost always be achieved if no restriction is placed on ammonia slip. In practice, this is not acceptable due to balance-of-plant impacts and, consequently, limits of ammonia slip are placed on commercial applications, usually  $\leq 5$  ppm. The specifications for the test facility catalysts required the base-line ammonia slip not exceed 5 ppm through the end of the two year test period.

In addition to the ammonia slip depending on catalyst exposure time, it is a strong function of other parameters such as flow rate, temperature,  $\text{NH}_3/\text{NO}_x$  distribution, and especially  $\text{NH}_3/\text{NO}_x$  ratio ( $\text{NO}_x$  reduction). The catalyst design margin partly controls the ability of a particular installation to handle changes in these parameters without exceeding desired ammonia slip levels.

In addition, the catalyst design plays a role in its ability to handle parametric changes and consequently some catalysts appear to be more susceptible to increases in ammonia slip due to changes in parametric conditions than do others.

#### Flow Rate Effects

Theoretically, from a kinetic standpoint,  $\text{NO}_x$  reduction should be inversely proportional to changes in flow rate. In practice, improvements in mass transfer tend to mitigate some of the effects of increased flow rate and in general flows could be increased to 150% of design without the catalyst exceeding the 5 ppm limit at 80%  $\text{NO}_x$  reduction and design temperature. This ability depends on installed margin and design, although sensitivity to this parameter is not excessive.

#### Temperature Effects

Temperature increases would be expected to significantly improve reactivity based on a kinetic model of the catalysts. However, in practice, mass transfer limitations become controlling in commercial SCRs and improvements in test facility catalyst activity were not very significant above 700 °F. Most catalysts exhibited fairly significant improvements in overall performance as temperatures were increased from 620 °F to 700 °F, but relatively little improvement was noted as the temperature was increased from 700 °F to 750 °F. This shows that the benefits of high temperature operation probably do not outweigh the heat rate penalties involved in operating the SCR at the higher temperature. The reader should be cautioned that these performance conclusions are based on catalyst designed for a 700 °F base-line operating temperature. Catalyst suppliers are quite mature in their design capabilities and would likely be able to take advantage of a higher temperature operating environment in a specialized design and thus realize more significant improvements in performance.

#### $\text{NH}_3/\text{NO}_x$ Ratio Effects

Changes in the  $\text{NH}_3/\text{NO}_x$  ratio and consequently  $\text{NO}_x$  reduction generally produce the most significant changes in ammonia slip. Decreases in  $\text{NO}_x$  reduction below 80%, down to values of roughly 60% were examined in the test facility. In most cases the ammonia slip at the low  $\text{NO}_x$  reduction was at or near the ammonia slip detection limit of 1 ppm. As  $\text{NO}_x$  reduction was increased above 80%, ammonia slip also increased. Slip values were fairly reasonable up to  $\text{NO}_x$  reduction levels of approximately 90%. As  $\text{NO}_x$  reduction increased past 90%, however, ammonia slip levels increased dramatically.

#### $\text{SO}_2$ Oxidation

Unfortunately, the catalytically active species that result in de $\text{NO}_x$  activity often contribute to  $\text{SO}_2$  oxidation activity. Since increased  $\text{SO}_3$  is detrimental to equipment downstream of the SCR, these two reactions tend to bound the catalyst design. In general, as requirements to minimize  $\text{SO}_2$  oxidation relax, de $\text{NO}_x$  activity per volume of catalyst can be increased. The upper bound for  $\text{SO}_2$  oxidation for the test facility catalysts was set at 0.75% at base-line conditions. The oxidation of  $\text{SO}_2$  was measured in the test facility at various operating conditions in addition to the base-line measurements. The average  $\text{SO}_2$  oxidation rates for each of the test facility catalysts are shown in [Figure 6](#). These data reflect base-line conditions over the life of the project. All of the catalysts were within the design limits, with most exhibiting oxidation rates well below the

specified limit. One notable exception is the Siemens catalyst which had a high activity with minimum catalyst volume, resulting in an SO<sub>2</sub> oxidation rate very close to the design value. In practice, some benefits may be realized by decreasing SO<sub>2</sub> oxidation past some arbitrarily chosen point, but these decreases are generally made at the expense of increasing catalyst volume. As with deNO<sub>x</sub> activity, the SO<sub>2</sub> oxidation margin chosen by the individual catalyst suppliers likely contributed significantly to the overall catalyst volume. It is difficult to determine the point of diminishing return for decreasing SO<sub>2</sub> oxidation, as little full scale long-term operational data is available for U.S. boilers. Although differences in SO<sub>2</sub> oxidation characteristics were very apparent between the catalyst suppliers, these differences do not necessarily denote a difference in catalyst design expertise between the suppliers, but more likely reflect a philosophical decision as to the catalyst offering made. In practice, all suppliers would likely be able to meet an applications specific SO<sub>2</sub> oxidation requirements (assuming flue gas constituents do not contribute to the oxidation characteristics). Unlike ammonia slip (or catalyst deNO<sub>x</sub> activity), the SO<sub>2</sub> oxidation rate of a catalyst is not expected to change significantly with time. This was the general finding based on measurements at the test facility as well as with the catalyst supplier laboratory tests. Since SO<sub>2</sub> oxidation remains relatively constant over time for a given catalyst volume, balance-of-plant effects must be evaluated in terms of the maximum SO<sub>2</sub> oxidation that may occur in later years with the addition of catalyst layers.

#### Flow Rate Effects

Due to the reaction order and the relatively constant SO<sub>2</sub> concentration, SO<sub>2</sub> oxidation should be inversely proportional to reactor flow rate (i.e. space velocity, linear velocity). However, most of the test facility catalysts exhibited a fairly constant SO<sub>2</sub> oxidation rate with respect to flow. The Siemens catalyst, with a somewhat higher SO<sub>2</sub> oxidation rate did exhibit a more linear response to flow rate. It is believed that physical phenomenon in the test facility such as SO<sub>3</sub> deposition within the reactor may have masked some of the effects on SO<sub>2</sub> oxidation from flow rate changes.

#### Temperature Effects

The oxidation of SO<sub>2</sub> is normally a much stronger function of temperature than of flow rate. Theoretically the catalysts should exhibit an exponential relationship of SO<sub>2</sub> oxidation to temperature. However, the measurements in the test facility showed that this relationship was more linear than exponential. Little difference was noted in SO<sub>2</sub> oxidation between 620 and 700 °F. However, SO<sub>2</sub> oxidation did increase more significantly between 700 and 750 °F. Figure 7 shows average SO<sub>2</sub> oxidation for the test facility catalysts at high temperature, with fairly significant increases in oxidation rate over base-line values previously shown.

#### NH<sub>3</sub>/NO<sub>x</sub> Ratio Effects

The rate of SO<sub>2</sub> oxidation is not expected to be a direct function of the NH<sub>3</sub>/NO<sub>x</sub> ratio, since ammonia does not play a direct role in the SO<sub>2</sub> oxidation reaction. However, other extraneous factors can create apparent changes in SO<sub>2</sub> oxidation rate with changes in NH<sub>3</sub>/NO<sub>x</sub> ratio. This is mainly due to physical effects such as precipitation of ammonium bisulfate. Increases in the NH<sub>3</sub>/NO<sub>x</sub> ratio change the ammonium bisulfate formation patterns between the measurement points and can result in an apparent change in SO<sub>2</sub> oxidation rate. For this reason, most SO<sub>2</sub> oxidation tests were made at 80% NO<sub>x</sub> reduction.



### **Other Parametric Tests**

Other tests were performed as part of the parametric sequences. These included evaluation of  $N_2O$  formation,  $NO_x$  and ammonia distribution analysis, fly ash composition and loading, velocity distributions and particulate distributions. The majority of these tests were performed at base-line conditions and were performed as part of an ongoing assessment of reactor operating conditions and performance. The results of these tests showed no significant changes in  $N_2O$  levels across the SCR reactors and that parameters such as particulate and velocity distributions remained within the original design specifications.

### **Air Preheater Tests**

The three test facility air preheaters consisted of a two layer (APH A) and a three layer (APH B) Ljungstrom<sup>®</sup> air preheater and a heat-pipe or Q-pipe<sup>®</sup> (APH C) all provided by ABB Air Preheater, Inc. (ABB). The performance of the air preheaters was evaluated using manual tests conducted during the parametric test sequences, long-term continuous tests, and evaluations made by ABB utilizing laboratory analyses of deposit samples, basket materials, wash water, and by visual inspections. The following results were reported by ABB.

The thermal performance of the three air preheaters was measured by the drop in number of transfer units (Ntu) from the initial values to the final values. The thermal performance dropped by about 11%, 7%, and 25% for APHs A, B, and C respectively. The fact that APH B seemed to deteriorate less than the others is attributed to the fact that this air preheater received significantly less ammonia slip as an average over the project life (partially due to differences in reactor operating conditions) than did the other two air preheaters. It is believed that APH C deteriorated most because it was a recuperator, and was therefore more sensitive to a given amount of fouling than a regenerator.

As one would expect, the gas side pressure drops were more sensitive to the degree of fouling and plugging than were the Ntu values. In general, all three air preheaters showed steady increases in gas-side pressure drop ( $\Delta P_g$ ) during the test period, punctuated by occasional spikes which may have been caused by system upsets such as sootblower failure. In general, the high  $\Delta P$ 's could be reduced by aggressive cleaning methods, including sootblowing at 4 hour intervals, thorough water washing, and occasional increases in the gas outlet temperature. It was not possible, however, to maintain the original, clean  $\Delta P$  of any of the air preheaters. The air and gas Euler (Eu) numbers (defined as  $g_c(-dp)/\rho V^2$ ) for air preheater A increased by 145% and 115%, respectively, from the beginning of the test to the end. For air preheater B, the increases in both Eu numbers were in the 50-55% range.

Although the 3-layer air preheater appeared to perform better than the 2-layer air preheater, it cannot be concluded that the 3-layer design is superior to the 2-layer design. This is because the 2-layer air preheater received much more ammonia slip than the 3-layer air preheater -- possibly as much as four to five times more. Given this significant difference in operating condition, the 2-layer preheater performed remarkably well, and might logically have done better than the 3-layer design if the concentration of ammonia into the two regenerators had been equal.

The Q-pipe<sup>®</sup> air preheater seemed to steadily lose thermal performance with time during most of the test period. It is possible that some part of this performance loss may have been due to the loss of some of the air heater pipes in the unit as a result of the sootblower eroding the tube wall and the consequent loss of heat transfer fluid from those pipes. However, the majority of the drop in Ntu is believed to be attributable to fouling of the finned tubes on the gas side.

Corrosion tests were performed on various heat transfer surface materials used in APHs A and B. It was determined that enameled heat transfer surfaces should be used for Ljungstrom<sup>®</sup> air heaters when ammonia and sulfur compounds are both present in the gas stream. Other conclusions from the study were as follows.

- 1) Ammonium bisulfate or its corrosion products were shown to be a major constituent in the Ljungstrom<sup>®</sup> air preheater deposits at temperatures less than the ammonium bisulfate formation point.
- 2) The magnitude of the ammonia slip had a significant effect on the corrosion losses and deposit accumulations in the units.
- 3) The enameled heat transfer surface was an order of magnitude superior to the low-alloy corrosion resistant and carbon steel materials with regard to corrosion losses and deposit accumulations.
- 4) The enameled heat transfer surface exhibited superior cleanability compared to the low-alloy corrosion resistant and carbon steel materials.

Table 7 summarizes the changes in Ntu, air-side Euler number ( $Eu_a$ ), and gas-side Euler number ( $Eu_g$ ) for the three air preheaters. Worst case increases in  $Eu_g$  varied from 185% up to 320%. Worst case increases in air-side pressure drop ( $\Delta P_a$ ) varied from 245% up to 345%. Final air and gas Eu numbers were about 115-145% higher than initial values on air preheater A, and about 50-55% higher on air preheater B. This is roughly consistent with the declines in Ntu (final vs. initial) of 11% for air preheater A and only 7% for air preheater B, since the air preheater with the high increase in Eu (air preheater A) also had the greater loss of Ntu.

**Table 7. Air Preheater Performance Data**

Air Heater	Initial Value (I)	Worst Case Value (WC)	Final Case Value (F)	% Change WC vs. I	% Change F vs. I
<i>Ntu values</i>					
A	3.75	3.1	3.3	-17%	-12%
B	3.50	2.8	3.3	-20%	-7%
C	2.40	1.4	1.8	-42%	-25%
<i>Gas-side Eu Number</i>					
A	35	140	75	+300%	+115%
B	28	118	43	+320%	+55%
C	63	180	135	+185%	+115%
<i>Air-side Eu Number</i>					
A	40	137	98	+245%	+145%
B	35	155	53	+345%	+50%

\* Excludes questionable data from months 25-26

## Miscellaneous tests

### ***Waste Stream Impacts***

A concern associated with the implementation of SCR technology is the effect that it may have on plant waste streams. This is thought to be primarily due to the adsorption of slip ammonia on fly ash. A special study was conducted to examine ammonia volatilization, ammonia extraction, and the effect of ammonia on metals extraction from fly ash.

### Ammonia Extractability and Volatilization

Almost no ammonia volatilizes from SCR fly ash until a significant amount of water has been absorbed by the ash. A plausible mechanism for the apparent volatilization is that enough water must be gained by the ash to form a moist layer with a pH high enough to evolve gas-phase ammonia from the ammonium compounds on the ash. In closed-container experiments, nearly all of the ammonia on the ash evolved to the gas phase when wet. Ammonia concentrations in enclosed spaces depend on the ammonia concentration of the ash, the volume of air surrounding the ash and the presence of a humid atmosphere

Tests were performed to examine the extractability of ammonia from fly ash using aqueous solutions of various pH. It was determined that the extraction does depend upon solution pH. All or nearly all of the ammonia present was extracted in buffered solutions at pH 4.7, and pH 6.2, but not all was recovered in alkaline unbuffered extracts. In the pH 6.2 buffer, however, the completeness of extraction seemed to fall off somewhat as the ratio of ash to buffer increased. At 3g of ash per 50 ml of pH 6.2 buffer, the amount of ammonia extracted was about 200 µg/g whereas at 1 g per 50 ml, the amount was near 250 µg/g.

Tests were also performed to examine the particle size dependency of fly ash on ammonia adsorption using a series cyclone in-situ ash separation method. Ammonia concentration in the

ash was much higher for the smaller particle sizes, but most of the total ammonia was found to reside with larger particles simply because these comprise the vast majority of the ash mass. The implication is that little slip ammonia will exit the process in the gas phase when high efficiency particulate emission controls are in place. This is due to much of the ammonia being in the solid phase at the air heater exit ( $\geq 50\%$ ) and to most of the ammonia being associated with the larger particle sizes which are most readily collected. In addition, the cooler temperatures downstream of the air preheater allow for easy deposition of ammonia by-products throughout the plant duct-work and in other pollution control processes (if any) prior to the stack.

#### **Metal Extractability from SCR Fly Ash**

Test facility ash samples were subjected to extraction with water, and extracts were analyzed for each of 28 metals. This was done to ascertain whether exposure of the fly ash to ammonia vapor caused an enhancement of the metal's extractabilities under conditions resembling those that might exist in an ash pond. The tested metals were Ag, Al, As, B, Ba, Be, Ca, Cd, Ce, Co, Cr, Cu, Fe, Hg, Mg, Mn, Mo, Ni, P, Pb, Rb, Sb, Se, Sn, Sr, Tl, V, and Zn. Of these 28 metals included in the study, only 17 could be detected in the fly-ash extracts (Ag, Be, Ce, Cu, Hg, Mn, Ni, Pb, Sn, Tl, and Fe were not detected). Of the 17 detectable metals, only barium underwent an increase in extractability following exposure to ammonia. The magnitude of the increases was found to depend directly on the magnitude of the  $\text{NH}_3/\text{NO}_x$  ratio in the SCR unit (ammonia slip), however, the increase was slight for all  $\text{NH}_3/\text{NO}_x$  ratios tested. Of the 16 additional metals that could be detected in the fly ash extracts, none displayed what were considered to be genuine enhancements in extractability, and several exhibited decreases in extractability as a result of exposure of the fly ash to ammonia.

#### ***TCLP Analysis***

Toxicity Characteristics Leaching Procedure (TCLP) analyses were performed on fly ash samples from the test facility at various times throughout the project life. These samples were taken as grab samples from the cyclone ash hoppers from each of the high-dust SCR reactors. The reactors were operated at base-line conditions at least 12 hours prior to sampling. All hoppers were cleaned of ash prior to the period of base-line operation to insure that all ash acquired in the sampling was produced during reactor base-line operation. In general the TCLP analyses showed non-detectable amounts of metal species. In those cases where detectable amounts were present, little or no change was noted between the SCR hopper samples and the host unit reference sample. It was concluded from this study that the SCR process does not significantly affect the TCLP results of fly ash.

#### ***Ammonia Partitioning***

As part of the three series of air preheater tests, the gas/solid phase partitioning of ammonia was determined. These tests showed that a large portion of the gas-phase ammonia present at the reactor exit adsorbed onto the fly ash as the flue gas was cooled through the air preheaters. In general, roughly one half of the total ammonia present was adsorbed onto the fly ash on a mass basis. Ash ammonia concentrations are normally reported on a ppm by weight ( $\mu\text{g/g}$ ) basis as compared to gas-phase concentration reported on a ppm by volume basis. Using these conflicting units, the ash phase concentration was roughly 100 times that of the gas phase. In other words, an

ammonia slip value of 1 ppm by volume gave roughly 100  $\mu\text{g/g}$  (or ppm by weight) of ammonia on the fly ash downstream of the air preheater (this finding compares well with similar foreign experience). It should be cautioned that these results are highly variable and are strongly dependant on temperature regimes, fly ash mass loading, and fly ash composition as well as the ammonia slip level. However, solid-phase ammonia concentrations can be correlated to ammonia slip levels if large amounts of data are used to help prevent skewing of the results due process variations other than ammonia slip. This data is useful only in the determination of long term trends and would not be adequate for the evaluation of short-term slip such as during a parametric or compliance test.

## ECONOMIC EVALUATION

The estimated capital and O&M costs from an economic evaluation of applying SCR technology at full scale to a new facility (coal-fired boiler utilizing high-sulfur, U.S. indigenous coal) are reported in this section. This information is not applicable to retrofit situations due to site-specific costs associated with space limitations, fan or air preheater upgrades, or major ductwork redesign. In addition, this economic analysis is not meant to supplant the need to perform site-specific financial analyses when evaluating SCR technology to a specific new facility. There will be project-specific constraints, sensitivity analyses, and market forces which no generalized economic analysis will be able to capture. Rather, the information reflects a macro-economic analysis of SCR technology based on historical data measured at the test facility, empirical laboratory data generated as a result of the test program, and consolidation of operational lessons learned, tempered with the knowledge of the current commercial market trends. This analysis is in a draft form and dollar values presented are likely to change during review and subsequent modification.

### Base Case

The economic estimates for the base case evaluation are founded upon the application of a high-dust, hot-side SCR (i.e., located between the boiler economizer outlet and the air preheater inlet) to a new coal fired installation utilizing high-sulfur domestic coal. The technical design premises used to prepare the economic analysis were selected to be representative of actual or anticipated plant configurations and  $\text{NO}_x$  control requirements currently being permitted or likely to be permitted on coal-fired boilers in the United States. Thus, defining assumptions were selected in an effort to have broad utility applicability.

The base case represents a new, base-load 250 MW pulverized-coal power plant typical of the majority of new coal-fired projects currently under development, construction, or recently declared in commercial operation. The unit size of 250 MW is consistent with future trends of new domestic power plants. The plant is located in a rural area with minimal space limitations. The fuel is a high-sulfur bituminous Illinois No. 6 coal.

The plant will utilize a single, balanced-draft, pulverized-coal fired boiler complete with all required auxiliary equipment. The boiler will be designed to produce approximately 1,610,000 lb/hr of main steam at turbine inlet conditions of 2400 psig and 1000 °F. Utilizing current generation low  $\text{NO}_x$  combustion systems, the boiler is assumed to produce a  $\text{NO}_x$  emission rate of

0.35 lb/MBTU. For purposes of this study, it is assumed that tangentially-fired boilers and wall-fired boilers are interchangeable with respect to all thermal performance and flue gas constituents.

The flue gas exits the boiler and enters a single, hot-side SCR. Flue gas flow is vertically downward through the reactor. The SCR is located directly above the air preheater and is designed as a "universal" reactor able to accept either (or both) plate type catalyst and honeycomb type catalyst. Ammonia injection will utilize stand-alone dilution air fans rather than combustion air.

A single, trisector, Ljungstrom® regenerative air preheater will be utilized to reclaim heat from the flue gas stream and transfer that heat to the primary and secondary air. Physical features of the air preheater are typical of what is commercially offered as a "deNO<sub>x</sub>" air preheater. The heater transfer surface arrangement will include hot, intermediate, and cold sections.

Sulfur dioxide will be removed using a lime spray dryer FGD system. A reverse gas, fabric filter baghouse will be used which will collect the dried reaction products from the spray dryer as well as the fly ash produced in the boiler by the combustion of coal.

The SCR reactor for this analysis includes three catalyst support layers plus a flow straightener (dummy bed). At time zero, two of the three catalyst support layers are loaded with catalyst. To optimize catalyst life, a spare (empty) catalyst support elevation inside the reactor is provided. The spare layer allows catalyst suppliers to develop optimized catalyst management plans which increase catalyst utilization. Thus, a fresh catalyst layer can be added to the reactor after the guarantee period when the ammonia slip begins to exceed the guaranteed limit. The activity of the new catalyst combined with the residual activity of the existing catalyst restores the performance of the SCR and extends the next addition/replacement outage beyond the initial guarantee interval.

The catalyst management plan developed for the economic analysis is based on composite empirical laboratory data developed on the test facility catalysts and uses a 16,000 hour (2 year) catalyst life guarantee period. After the initial guarantee period of 2 years, a new layer of catalyst is added to the reactor, thus, taking advantage of the residual activity in the initial layers to "boost" the performance of the SCR. The next addition of catalyst is required in project year 6 when one of the initial layers is replaced. Thus, catalyst is added and/or replaced after 2, 6, 9, 12, 15, 18, 21, 24, and 27 years for a total of 9 times during the 30 year evaluation period. It should be noted that catalyst life projections of 30 years are speculative at best due to uncertainty of catalyst structural properties with long-term exposure to flue gas and thermal cycling.

The base case design criteria are shown below in Table 8. Also included in the table are charge and levelization factors as well as fixed and variable O & M assumptions and unit costs.

**Table 8. Base Case Design Criteria**

Parameter	Specification
Type of SCR	Hot-Side
Number of SCR Reactors	One
Reactor Configuration	3 catalyst support layers + 1 dummy layer
Initial Catalyst Load	2 of 3 layers loaded, 1 spare layer
Required Range of Operation	35% to 100% boiler load
NO <sub>x</sub> concentration at Inlet	0.35 lb/MBtu
Design NO <sub>x</sub> reduction	60%
Flue Gas Temperature at SCR Inlet	700 °F
Flue Gas Pressure at SCR Inlet	-5 in. W.G.
Design Ammonia Slip	5 ppm
Guaranteed Catalyst Life	2 years (16,000 hours)
SO <sub>2</sub> Oxidation	0.75% (initial catalyst load)
Maximum Pressure Drop	6 in. W.G. (fully loaded reactor)
Velocity Distribution	$\Delta V/V_{\text{mean}} < 10\%$ over 90% of reactor area $\Delta V/V_{\text{mean}} < 20\%$ over remaining 10% of area
Ammonia Distribution	$\Delta C/C_{\text{mean}} < 10\%$
Temperature Distribution	$\Delta T < 10\text{ }^{\circ}\text{C}$ max. deviation from mean
Anhydrous Ammonia Cost	\$ 250/ton
SCR Catalyst Cost	\$ 400/ft <sup>3</sup>
SCR Catalyst Guarantee Period	2 years
SCR Catalyst Escalation	3.0%
Power Cost	30 mills/kWh
ID Fan Efficiency	75%
SCR Draft Loss (fully loaded reactor)	3.0 in. W.G.
Ductwork Draft Loss	0.75 in. W.G.
Ammonia Injection Grid Draft Loss	0.75 in. W.G.
Unrecoverable Air Preheater Draft Loss	1.0 in. W.G.
Operating Labor Man-Hour Rate	\$ 23/hr
Maintenance Factor (% of total process capital)	2.0%
<b>Current Dollar Analysis (1996 dollars):</b>	
Capital Charge Factor	0.150
O & M Cost Levelization Factor	1.362
<b>Constant Dollar Analysis</b>	
Capital Charge Factor	0.116
O & M Cost Levelization Factor	1.000

### Capital, O&M, and Levelized Cost vs. Unit Size

In order to examine the change in SCR costs vs. unit size, additional capital and O&M estimates were repeated for a 125 MW unit and 700 MW unit. To maintain consistency with the 250 MW

base case unit, an SCR removal efficiency of 60% NO<sub>x</sub> reduction was assumed. Where possible, consistent (or identical) assumptions were made with regard to the 125 MW and 700 MW units.

Results are shown in Table 9 for the base case (250 MW), 125 MW, and 700 MW unit sizes for an SCR with a NO<sub>x</sub> removal efficiency of 60%. On a \$/kw vs. unit size basis, the total capital requirement of the SCR system shows a trend of decreasing unit cost (\$/kw) with increasing unit size indicating significant economies of scale. Total capital requirement ranges from \$61/kw for the 125 MW unit to \$45/kw for the 700 MW unit.

**Table 9. Capital, O&M, and Levelized Cost vs. Unit Size**  
(60 % NO<sub>x</sub> Removal)

Unit Size	125 MW	250 MW	700 MW
Total capital requirement (\$)	7,602,000	13,415,000	31,327,000
Total capital requirement (\$/kw)	61	54	45
First year fixed operating cost (\$)	213,000	312,000	614,000
First year variable operating cost (\$)	367,000	733,000	2,053,000
<b>Current Dollar Analysis</b>			
Levelized Cost (mills/kWh)	2.89	2.57	2.22
Levelized Cost (\$/ton)	2,811	2,500	2,165
<b>Constant Dollar Analysis</b>			
Levelized Cost (mills/kWh)	2.09	1.85	1.59
Levelized Cost (\$/ton)	2,037	1,802	1,547

#### **Capital, O&M, and Levelized Cost vs. NO<sub>x</sub> Removal Efficiency**

In addition to the 250 MW base case NO<sub>x</sub> removal efficiency of 60%, two additional NO<sub>x</sub> reduction cases at 40% and 80% were calculated to examine the impact on cost. Tabular results of this analysis are shown below in Table 10.

**Table 10. Capital, O&M, and Levelized Cost vs. NO<sub>x</sub> Removal Efficiency**  
(250 MW Unit Size)

NO <sub>x</sub> Reduction	40%	60%	80%
Total capital requirement (\$)	12,974,000	13,415,000	14,142,000
Total capital requirement (\$/kw)	52	54	57
First year fixed operating cost (\$)	305,000	312,000	324,000
First year variable operating cost (\$)	621,000	733,000	857,000
<b>Current Dollar Analysis</b>			
Levelized Cost (mills/kWh)	2.39	2.57	2.79
Levelized Cost (\$/ton)	3,502	2,500	2,036
<b>Constant Dollar Analysis</b>			
Levelized Cost (mills/kWh)	1.74	1.85	2.00
Levelized Cost (\$/ton)	2,536	1,802	1,460



### O&M Cost vs. Inlet NO<sub>x</sub> Concentration

Many new boiler installations face difficult decisions on how to best optimize overall NO<sub>x</sub> reduction requirements using a combination of low NO<sub>x</sub> burners and SCR. While maximizing combustion NO<sub>x</sub> reductions can allow lower SCR variable O&M costs, it typically comes at the expense of increased LOI in the fly ash, and hence, lower plant cycle efficiency. Optimizing the burners to minimize LOI often leads to higher NO<sub>x</sub> concentrations entering the SCR and, hence, higher variable O&M costs to achieve a permitted outlet NO<sub>x</sub> emission limit. Results showing leveled cost vs. SCR inlet NO<sub>x</sub> concentration for a 250 MW unit operating at 60% NO<sub>x</sub> removal are summarized in Table 11.

**Table 11. Levelized Cost vs. SCR Inlet NO<sub>x</sub> Concentration**  
(250 MW @ 60% NO<sub>x</sub> Removal)

Inlet NO <sub>x</sub> (lb/MMBtu)	.45	.40	.35	.30	.25
<b>Current Dollar Analysis</b>					
Levelized Cost (mills/kWh)	2.61	2.59	2.57	2.55	2.53
Levelized Cost (\$/ton)	1,977	2,205	2,500	2,894	3,446
<b>Constant Dollar Analysis</b>					
Levelized Cost (mills/kWh)	1.88	1.87	1.85	1.84	1.82
Levelized Cost (\$/ton)	1,425	1,590	1,802	2,086	2,483

## CONCLUSIONS

### SCR Catalysts

All of the catalysts in the test program proved to be acceptable for the application and met design specifications. Significant differences, however, were noted in catalyst parameters such as volume, weight, activity (both deNO<sub>x</sub> and SO<sub>2</sub> oxidation), fouling, and pressure drop. No one catalyst can be considered superior since each may possess advantages particular to the characteristics of a proposed installation. For instance, pressure drop may not be a critical factor in a new-plant installation, but may be critical in the catalyst selection for a retrofit situation where a fan retrofit lies in the balance. Similarly, weight may be the controlling factor for an in-duct installation, while other installations may be particularly concerned with SO<sub>2</sub> oxidation. In short, careful examination of catalyst characteristics are necessary to choose the optimum catalyst for a particular installation.

### Deactivation

The deactivation trends of the catalysts were within expected ranges based on operating experience in European and Japanese installations. Catalyst poison concentrations were somewhat higher than prior experience. However, no unusual acceleration in deactivation was noted and it appears that at least for the coals tested during this project, the resulting deactivation is similar in significance to other world-wide installations.

### ***Volume/DeNO<sub>x</sub> Activity***

Total catalyst volume and weight are strong functions not only of catalyst design (surface area, activity, etc.), but of installed margin. In some cases catalyst suppliers chose to minimize catalyst volume by designing a system which met, but did not exceed performance requirements. In other cases, catalyst suppliers chose to increase catalyst volume to insure performance margin and thus improve performance. Honeycomb catalysts are generally considered to have higher surface area when compare to plate-type catalysts (on a bulk volume basis) and would thus generally require less volume (although weight may be greater). However, offerings in the program have demonstrated the ability of plate-type catalyst to meet specifications with similar or less bulk volume than honeycomb configurations. Thus, it appears that installed catalyst design margin can be as significant to bulk catalyst volume as is the basic geometrical (honeycomb/plate) design.

### ***SO<sub>2</sub> Oxidation***

While some catalysts essentially met the SO<sub>2</sub> oxidation requirement, others improved upon it greatly. In practice some benefits may be realized by decreasing SO<sub>2</sub> oxidation past some arbitrarily chosen point, but those decreases are generally made at the expense of increasing catalyst volume. As with deNO<sub>x</sub> activity, the SO<sub>2</sub> oxidation design margin chosen by the individual catalyst suppliers likely contributed to the overall catalyst volume. It is difficult to determine the point of diminishing return for decreasing SO<sub>2</sub> oxidation as little full scale long-term operational data is available for U.S. boilers. Differences in SO<sub>2</sub> oxidation characteristics were very apparent between the catalysts. However, these differences do not necessarily denote a difference in catalyst design expertise between suppliers, but more likely reflect a philosophical decision as to the catalyst offering made. In practice, all suppliers would likely be able to meet a particular application's SO<sub>2</sub> oxidation requirements.

### ***Pressure Drop***

Catalyst pressure drops are generally a function of catalyst geometry and volume. Honeycomb catalysts commonly have less open area than plate catalysts, resulting in increased pressure drop per unit length of catalyst. In some cases, this is mitigated partially by less volume of honeycomb catalyst required compared to plate catalysts. In addition, honeycomb pressure drops can be modified by adjusting the wall thickness of the catalyst. In the case of the Siemens plate, the pressure drop was considerably lower than the other catalysts especially considering the fact that Siemens utilized only two catalyst beds to meet the program requirements without substantial margin. On a per volume basis, the Haldor and Hitachi catalysts also had low pressure drop, but the increased volume, compared to the Siemens catalyst, resulted in an overall pressure drop more similar to the honeycomb catalysts in the program. Thus, the overall reactor pressure drop must be considered when evaluating catalysts and neither the basic geometry nor the catalyst volume are exclusively controlling.

### ***Fouling and Erosion***

The fouling characteristics of SCR catalysts are important to the successful long-term operation of the reactor. In practice, the pressure drop across a catalyst increases slightly from new conditions once the catalyst is placed in service. The pressure loss, however, should remain

relatively consistent after the initial start-up of the reactor. Steadily increasing pressure drops over time indicate that a catalyst is either improperly designed for the application or that appropriate sootblowing steps are not being taken. Small reactors similar to those present in the test facility represent worst case scenarios for catalyst fouling due to the wall effects and space constraints preventing optimum sootblowing. However, the test facility pressure drop data show relatively level pressure drop over time for the catalysts. This indicates that the sootblowing procedures used in the test facility were effective. The plate-type configurations exhibited somewhat less fouling potential than did the honeycomb configurations (based on visual inspection) although quantification is difficult, but both configurations were acceptable for the application.

Erosion of the test facility catalysts was not considered to be a significant problem. Most of the erosion in the test facility is thought to be due at least in part to the aggressive sootblowing operations. The primary conclusion is that erosion is likely not the controlling factor in catalyst life. However, catalyst design does play a major role in erosion susceptibility. In both honeycomb and plate catalysts, the substrate support material as well as the primary catalytically active material contribute to the overall erosion resistance of the particular offering.

### **Air Preheater Performance**

As expected, the study concluded that the SCR process exacerbates performance degradation of air preheaters mainly due to ammonia slip and subsequent by-product formation. It is recommended that enameled heat transfer surfaces be used to aid in corrosion resistance and cleaning. Air preheater performance is likely a direct function of the magnitude of ammonia slip, however, it is undetermined if deleterious effects have an ammonia slip threshold value. A comparison of regenerator versus recuperator type air preheaters showed that regenerators tend to outperform recuperators in SCR applications in terms of both thermal performance and fouling.

### **Low/High Dust Configuration**

The installation of the reactor J low-dust configuration allowed a comparison of low-dust to high-dust applications. The test results were somewhat inconclusive due to fouling problems associated with the design of the test facility low-dust reactor and extraction scoop, biasing the physical testing. The results did show that catalyst parameters such as deactivation,  $\text{NO}_x$  reduction capability, and  $\text{SO}_2$  oxidation were similar to the high-dust configuration. The results also showed that comparable  $\text{NO}_x$  reduction could be accomplished with significantly less volume of catalyst (primarily due to the higher specific surface area of the low-dust catalyst). Physical characteristics such as fouling and pressure drop, however, were much more difficult to determine. The results showed that a low-dust configuration could be very sensitive to upsets in boiler/ESP operation. The small channel openings in the low-dust configuration present a particular problem during upset conditions where large amounts of particulate are forced through the catalyst. Since it appears that relatively short periods of operation at such upset conditions could be catastrophic to the catalyst, process design must unfortunately focus on mitigating the effects of upset conditions. It is unclear as to the degree that upsets would be experienced on full

scale installations as compared to the test facility experience. It is also unclear how these full-scale upsets would compare in severity to the test facility upsets. The study basically concludes that the low-dust configuration worked well for the conditions for which it was designed, but to the degree that excursions from these design conditions occur, the catalyst could be adversely impacted.

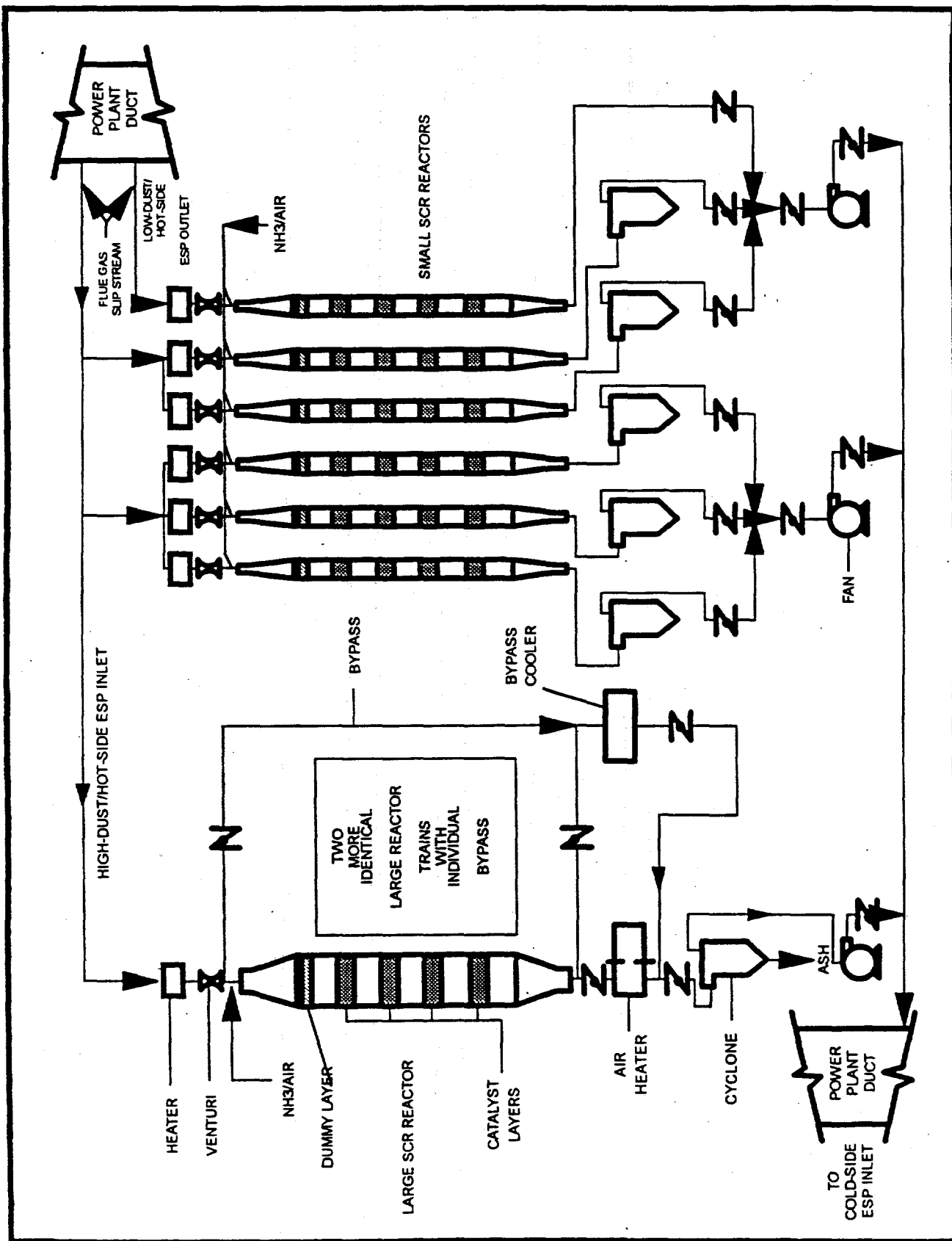
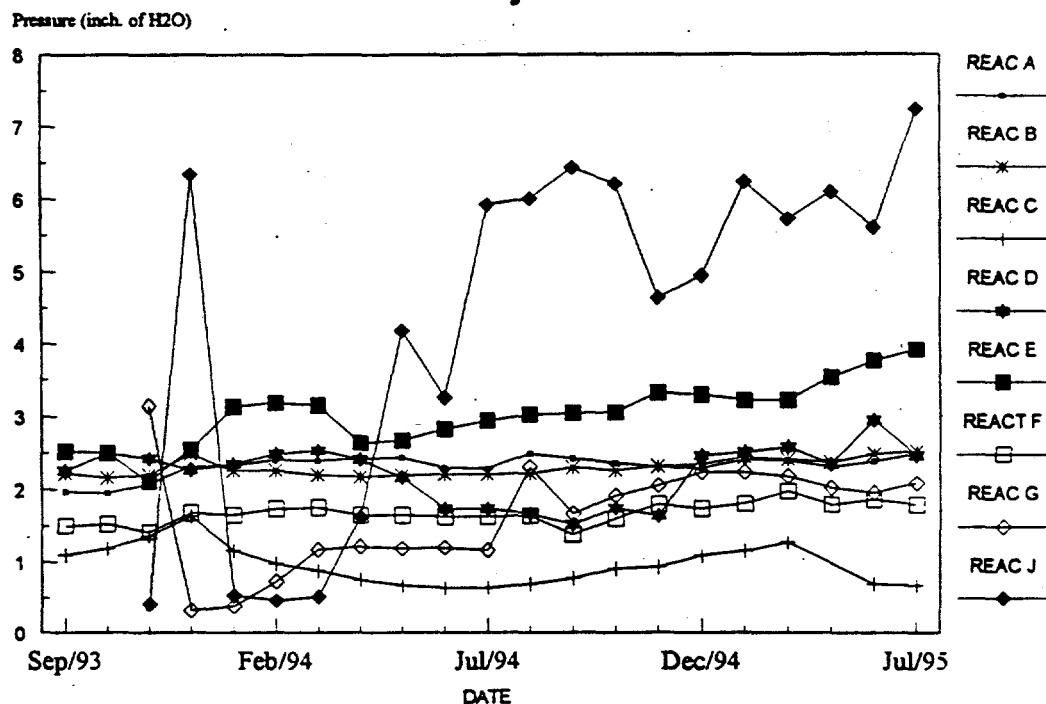
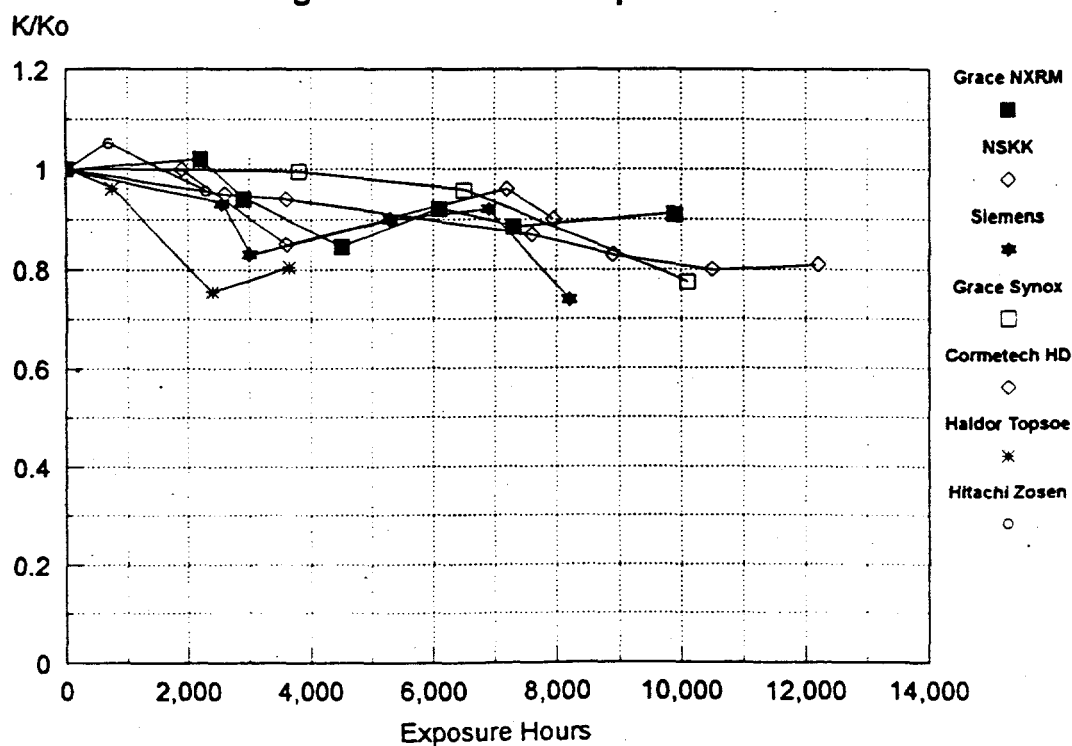


Figure 1. SCR Demonstration Facility-Process Flow Diagram

**Figure 2. Catalyst Bed Pressure Drop  
vs. Project Date**

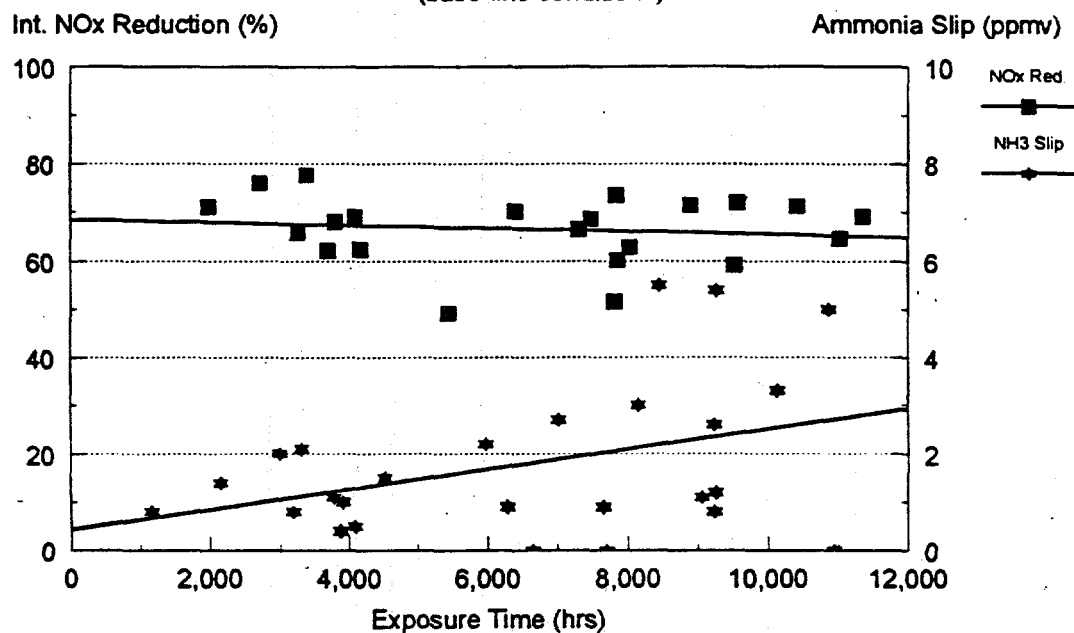


**Figure 3. K/Ko vs. Exposure Time**

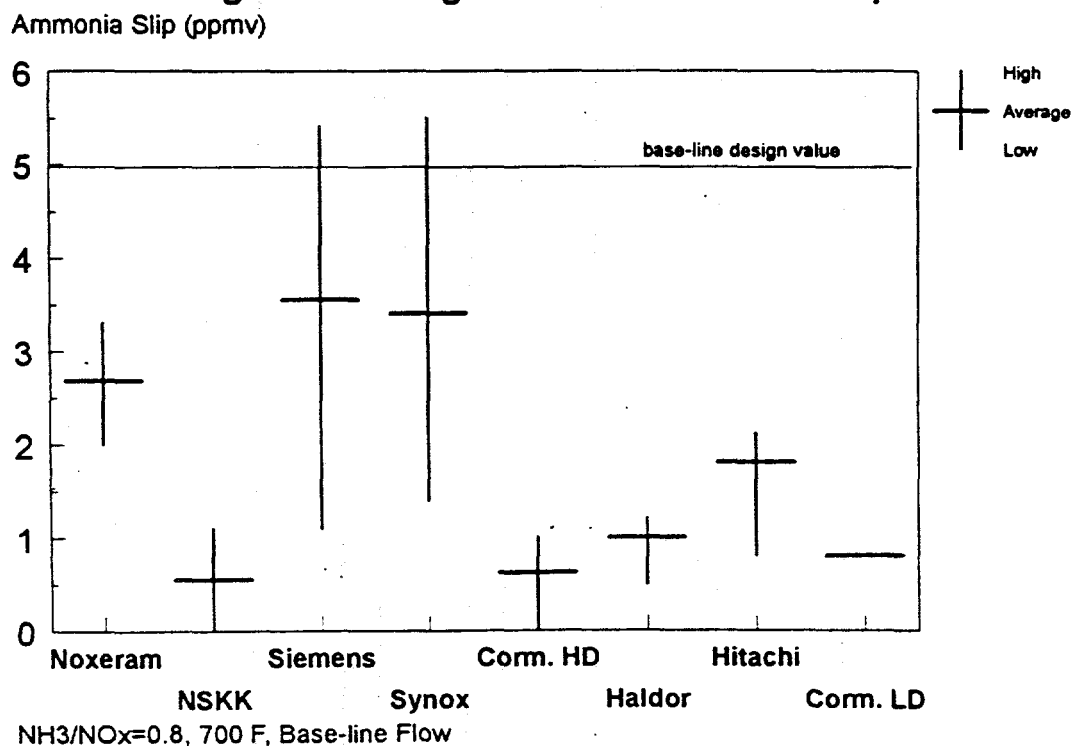


**Figure 4. Intermediate NO<sub>x</sub> Reduction and Ammonia Slip vs. Time**

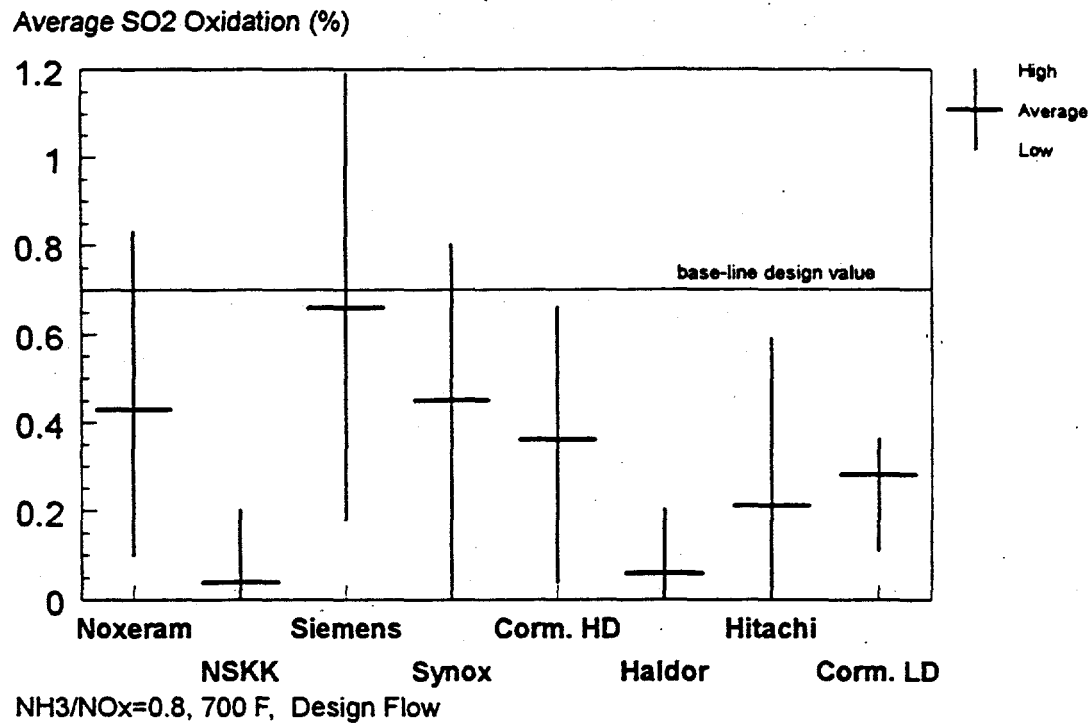
(base-line conditions)



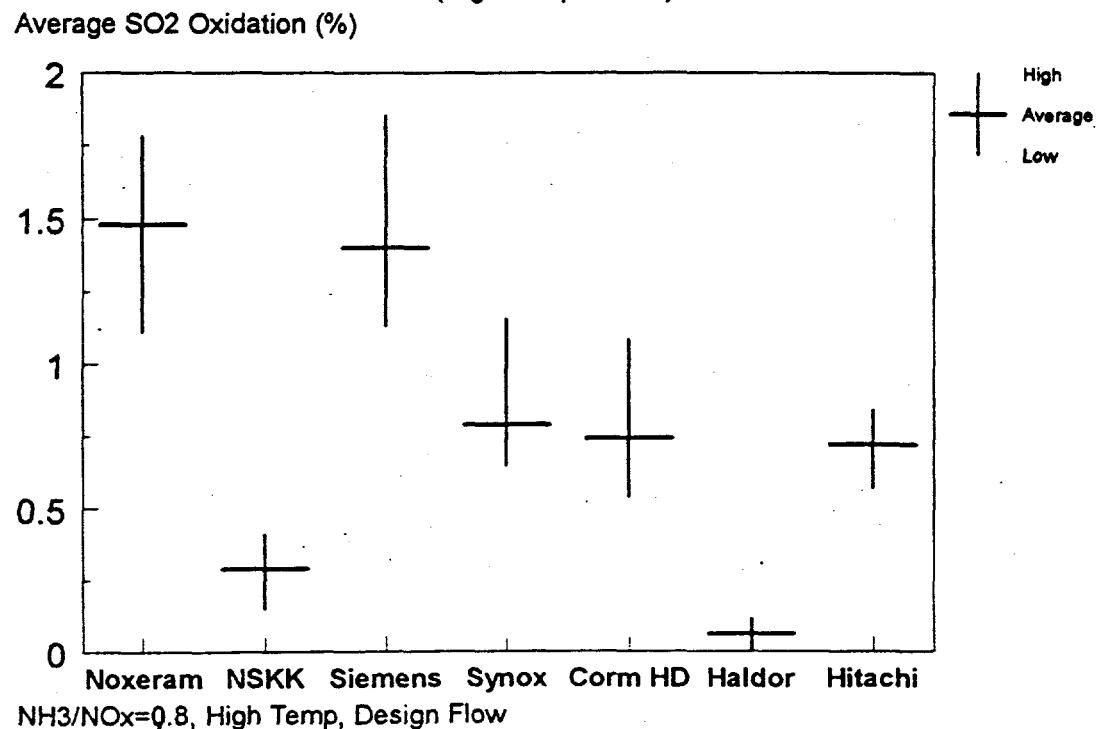
**Figure 5. Average Base-line Ammonia Slip**



**Figure 6. Average SO<sub>2</sub> Oxidation Rate**  
(Base-Line)



**Figure 7. Average SO<sub>2</sub> Oxidation Rate**  
(High Temperature)





## **ACKNOWLEDGMENT**

This project has been a collective effort on the part of many individuals and organizations too numerous to name individually. In lieu of a detailed listing, we would like to recognize the following organizations and their employees for their contribution to the success of this project; U.S. Department of Energy, Southern Company Services, Inc., Electric Power Research Institute, Ontario Hydro, Gulf Power Company, Southern Research Institute, ABB Air Preheater, Inc., Spectrum Systems, Inc., ICS, Inc., Cormetech, Inc., Haldor Topsoe A/S, Hitachi Zosen, Nippon Shokobai, Siemens, and W.R. Grace & Co.

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**500 MW DEMONSTRATION OF ADVANCED WALL-FIRED COMBUSTION  
TECHNIQUES FOR THE REDUCTION OF NITROGEN OXIDE EMISSIONS FROM  
COAL-FIRED BOILERS**

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**ABSTRACT**

This paper discusses the technical progress of a U. S. Department of Energy Innovative Clean Coal Technology project demonstrating advanced wall-fired combustion techniques for the reduction of nitrogen oxide (NO<sub>x</sub>) emissions from coal-fired boilers. The primary objective of the demonstration is to determine the long-term NO<sub>x</sub> reduction performance of advanced overfire air (AOFA), low NO<sub>x</sub> burners (LNB), and advanced digital control/optimization methodologies applied in a stepwise fashion to a 500 MW boiler. The focus of this paper is to report (1) on the installation of three on-line carbon-in-ash monitors and (2) the design and results to date from the advanced digital control/optimization phase of the project.

## INTRODUCTION

This paper discusses the technical progress of one of the U. S. Department of Energy's Innovative Clean Coal Technology (ICCT) projects demonstrating advanced combustion techniques for the reduction of nitrogen oxide (NO<sub>x</sub>) emissions from wall-fired boilers. The demonstration is being conducted at Georgia Power Company's Plant Hammond Unit 4, a 500 MW, pre-NSPS (New Source Performance Standards), wall-fired boiler. Plant Hammond is located near Rome, Georgia, northwest of Atlanta.

The Hammond project is being managed by Southern Company Services, Inc. (SCS) on behalf of the project co-funders: the Southern Company, the U. S. Department of Energy (DOE), and the Electric Power Research Institute (EPRI). In addition to SCS, Southern includes the five electric operating companies: Alabama Power, Georgia Power, Gulf Power, Mississippi Power, and Savannah Electric and Power. SCS provides engineering and research services to the Southern electric system. The ICCT program is a jointly funded effort between DOE and industry to move the most promising advanced coal-based technologies to the commercial marketplace. The goal of ICCT projects is the demonstration of commercially feasible, advanced coal-based technologies that have already reached the "proof-of-concept" stage. The ICCT projects are jointly funded endeavors between the government and the private sector in which the industrial participant contributes at least 50 percent of the total project cost. The DOE is participating through the Office of Clean Coal Technology at the Pittsburgh Energy Technology Center (PETC).

The primary objective of the demonstration is to determine the long-term NO<sub>x</sub> reduction performance of advanced overfire air (AOFA), low NO<sub>x</sub> burners (LNB), and advanced digital control/optimization methodologies applied in a stepwise fashion to a 500 MW boiler. Short-term tests of each technology are also being performed to provide engineering information about emissions and performance trends [1,2,3,4].

Following a brief unit and technology review, this paper focuses on the design and results to date from the advanced digital control/optimization phase of the project.

## UNIT AND TECHNOLOGY REVIEW

Georgia Power Company's Plant Hammond Unit 4 is a Foster Wheeler Energy Corporation (FWEC) opposed wall-fired boiler, rated at 500 MW gross, with design steam conditions of 2500 psig and 1000/1000°F superheat/reheat temperatures, respectively. The unit was placed into commercial operation on December 14, 1970. Prior to the LNB retrofit in 1991, six FWEC Planetary Roller and Table type mills provided pulverized eastern bituminous coal (12,900 Btu/lb, 33% VM, 53% FC, 72% C, 1.7% S, 1.4% N, 10% ash) to 24 pre-NSPS, Intervane burners. The burners are arranged in a matrix of 12 burners (4W x 3H) on opposing walls with each mill supplying coal to four burners per elevation (Figure 1).

During a spring 1991 unit outage, the Intervane burners were replaced with FWEC Controlled Flow/Split Flame (CF/SF) burners. In the CF/SF burner, secondary combustion air is divided between inner and outer flow cylinders. A sliding sleeve damper regulates the total secondary air

flow entering the burner and is used to balance the burner air flow distribution. An adjustable outer register assembly divides the burner's secondary air into two concentric paths and also imparts some swirl to the air streams. The secondary air that traverses the inner path, flows across an adjustable inner register assembly that, by providing a variable pressure drop, apportions the flow between the inner and outer flow paths. The inner register also controls the degree of additional swirl imparted to the coal/air

mixture in the near throat region. The outer air flow enters the furnace axially, providing the remaining air necessary to complete combustion. An axially movable inner sleeve tip provides a means for varying the primary air velocity while maintaining a constant primary flow. The split flame nozzle segregates the coal/air mixture into four concentrated streams, each of which forms an individual flame when entering the furnace. This segregation minimizes mixing between the coal and the primary air, assisting in the staged combustion process.

As part of this demonstration project, the unit was also retrofit with an Advanced Overfire Air (AOFA) system. The FWEC design diverts air from the secondary air ductwork and incorporates four flow control dampers at the corners of the overfire air windbox and four overfire air ports on both the front and rear furnace walls. Due to budgetary and physical constraints, FWEC designed an eight port AOFA system more suitable to the project and unit than the twelve port system originally proposed.

The Unit 4 boiler was designed for pressurized furnace operation but was converted to balanced draft operation in 1977. The unit is equipped with a coldside ESP and utilizes two regenerative secondary air preheaters and two regenerative primary air heaters. During the course of the ICCT demonstration, the unit was retrofitted with six Babcock & Wilcox MPS 75 mills (two each during the spring 1991, spring 1992, and fall 1993 outages).

## REVIEW OF PRIOR TESTING

Baseline, AOFA, LNB, and LNB+AOFA test phases have been completed (Table 1). Short-term and long-term baseline testing was conducted in an "as-found" condition from November 1989 through March 1990. Following retrofit of the AOFA system during a four-week outage in spring 1990, the AOFA configuration was tested from August 1990 through March 1991. The

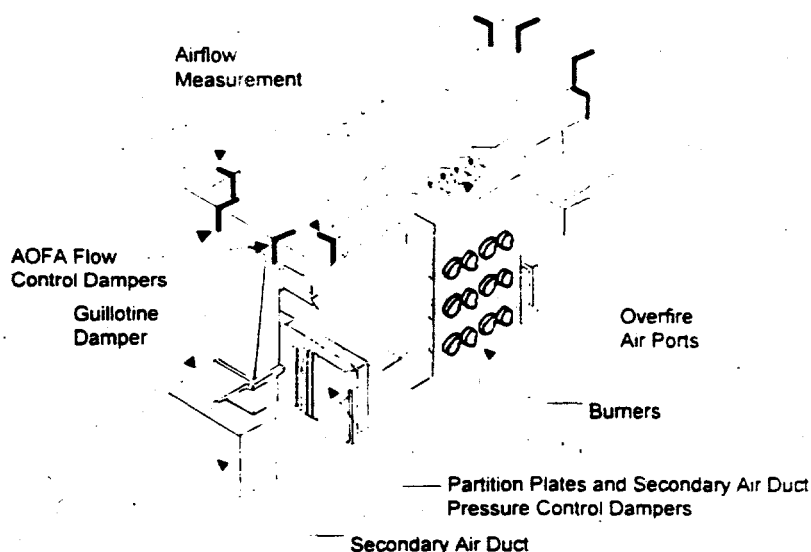


Figure 1. Hammond Unit 4 Furnace Layout

FWEC CF/SF low NO<sub>x</sub> burners were then installed during a seven week outage starting on March 8, 1991 and continuing to May 5, 1991. Following optimization of the LNBs and ancillary combustion equipment by FWEC personnel, LNB testing was commenced during July 1991 and continued until January 1992. Testing in the LNB+AOFA configuration was completed during August 1993. During both the LNB and LNB+AOFA, there were significant increases (when compared to baseline) in precipitator fly ash loading and gas flow rate and also, increases in fly ash LOI which adversely impacted stack particulate emissions and forced the unit to be load limited [5].

Table 1. Project Schedule

Phase	Description	Date	Status
0	Pre-Award Negotiations		
1	Baseline Characterization	8/89 - 4/90	Completed
2	Advanced Overfire Air Retrofit (AOFA) & Characterization	4/90 - 3/91	Completed
3A	Low NO <sub>x</sub> Burner Retrofit (LNB) & Characterization	3/91 - 1/92	Completed
3B	LNB+AOFA Characterization	1/92 - 8/93	Completed
4	Digital Controls/Optimization Retrofit & Characterization	8/93 - 12/96*	In Progress
5	Final Reporting and Disposition	9/95 - 12/96*	In Progress

\* Indicates change from original work breakdown structure. Final schedule dependent upon availability of unit.

A summary of the baseline, AOFA, LNB, and LNB+AOFA long-term NO<sub>x</sub> emissions data for Hammond Unit 4 are shown in Figure 2. Baseline testing was performed in an "as-found" condition. For the AOFA, LNB, and LNB+AOFA test phases, following optimization of the unit by FWEC personnel, the unit was operated according to FWEC instructions provided in the design manuals. As shown, the AOFA, LNBs, and LNB+AOFA provide a long-term, *full load*, NO<sub>x</sub> reduction of 24, 48, and 68 percent, respectively. The *load-weighted* average of NO<sub>x</sub> emissions reductions was 14, 48, and 63 percent, respectively, for AOFA, LNBs, and LNB+AOFA test phases. Although the LNB plus AOFA NO<sub>x</sub> level represents a 67 percent reduction from baseline levels, a substantial portion of the incremental change in NO<sub>x</sub> emissions between the LNB and LNB+AOFA configurations is the result of operational changes and is not the result of the AOFA system [6].

The *time-weighted* average of NO<sub>x</sub> emissions for the baseline, AOFA, LNB, LNB+AOFA test phases are shown in Table 2. Since NO<sub>x</sub> emissions are generally dependent on unit load, the NO<sub>x</sub> values shown in this table are influenced by the load dispatch of the unit during the corresponding test frame. Also shown in this table are the 30 day and annual achievable emission limits as determined during these test periods. The 30-day rolling average achievable emission limit is defined as the value that will be exceeded, on average, no more than one time per ten years. For the annual average, a compliance level of 95 percent was used in the calculation.

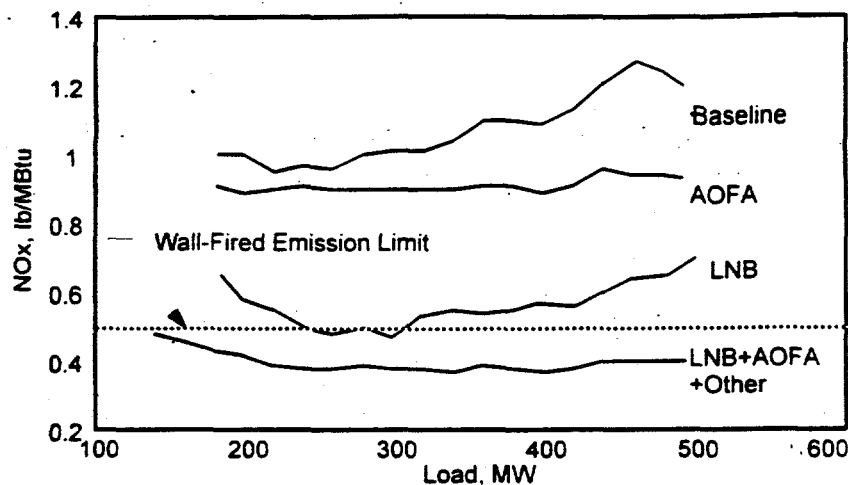


Figure 2. Long-Term NOx Emissions vs. Load Characteristic

Table 2. Long-Term NOx Emissions

Unit Configuration →	Baseline		AOFA		LNB		LNB-AOFA	
Parameter ↓	Mean	RSD. %	Mean	RSD. %	Mean	RSD. %	Mean	RSD. %
Number of Daily Avg. Values	52	-	86	-	94	-	63	-
Load (MW)	407	9.4	386	17.9	305	17.7	293	23.9
NOx Emissions (lb/MBtu)	1.12	9.5	0.92	8.6	0.53	13.7	0.41	12.9
O <sub>2</sub> Level (percent at stack)	5.8	11.7	7.3	12.6	8.4	7.7	8.73	16.3
NOx 30 Day AEL (lb/MBtu)	1.24	-	1.03	-	0.64	-	0.51	-
NOx Annual AEL (lb MBtu)	1.13	-	0.93	-	0.55	-	0.42	-

AEL = Achievable Emission Limit; RSD = Relative Standard Deviation

## EVALUATION OF ON-LINE CARBON-IN-ASH ANALYZERS

A subsidiary goal of the Wall-Fired project is the evaluation of advanced instrumentation as applied to combustion control. Based on this goal, several on-line carbon-in-ash monitors are being evaluated as to their:

- Reliability and maintenance.
- Accuracy and repeatability, and
- Suitability for use in the control strategies being demonstrated at Hammond Unit 4.

This instrumentation has the potential to allow faster response to changes in boiler conditions, resulting in benefits to boiler efficiency and ash monitoring.

Three units are currently installed at this site: (1) Applied Synergistics FOCUS, (2) CAMRAC Corporation CAM, and (3) Clyde-Sturtevant SEKAM. The SEKAM unit samples from two locations at the economizer outlet while the CAM unit samples from a single location at the

precipitator inlet. The FOCUS unit is a non-extractive system that utilizes two cameras located above the nose of the furnace. The following paragraphs briefly describe these devices.

### CAMRAC

The CAM (Carbon-in-Ash-Monitor) was developed by GAI Consultants during the 1980's for the CAMRAC company. Financial support was provided by several utilities throughout the United States as well as the Electric Power Research Institute. This instrument offers automatic monitoring of unburned carbon in combustion products. As of June 1996, a total of 5 units have been installed at various locations. The system has been tested on Georgia Power Company's Plant Hammond Unit 4 as well as Alabama Power Company's Plant Gaston Unit 4. Other locations outside the Southern Company include Pennsylvania Electric Company's Conemaugh Station, Allegheny Power Systems' Harrison Station, Philadelphia Electric's Eddystone Station and Duquesne Light's Cheswick Station.

The CAM's operating principle is based on the microwave absorptive properties of carbon particles in ash. Approximately 5 grams of fly ash are automatically extracted isokinetically from the duct and placed in a small collection cell. Microwaves at a frequency of 2450 MHz are passed through the collection cell. A power level of less than 150 milliwatts is used. Relative microwave absorbance between carbon and carbon-free ash is used to determine the carbon content of the sample. In other words, the power into the collection cell minus the transmitted and reflected power is equivalent to the power absorbed by carbon in the ash sample. An internal calibration curve is used to convert this absorbance to percent carbon. According to the manufacturer, the measured carbon level is independent of the coal being burned. The system sends results of the sample analysis to the plant control room for combustion performance optimization. Following analysis of the sample, the sample is returned to the combustion duct.

Although the unit can accommodate extraction locations on the cold gas side of the air heaters, CAMRAC's preferred sampling location is between the economizer and the air heaters. An adaptive sampler is operated by CAM so that isokinetic sampling conditions are maintained at all load levels. For single point sampling, the collection cell is purged and a new ash sample is collected for analysis every five to ten minutes. Multipoint designs with up to ten samplers are available for additional accuracy in duct LOI characterization, although sample cycle time is greater for multipoint sampling.

Figure 3 shows the CAM unit as installed at Hammond 4. A schematic of a typical CAM arrangement is shown as Figure 4. Table 3 provides other aspects of the system including size, accuracy and cost. These specifications were taken from CAM material and system dimensions.

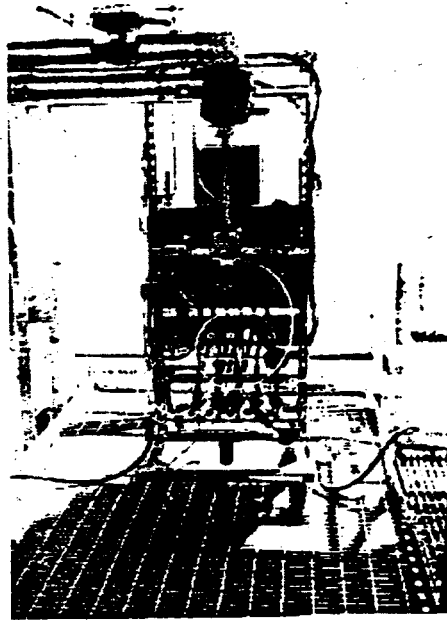
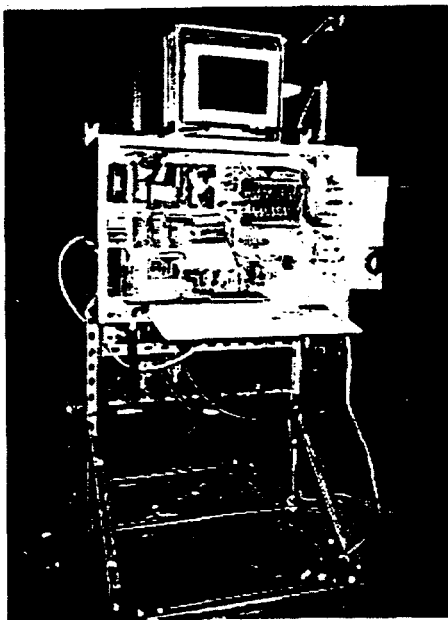


Figure 3. CAM On-Line Analyzer

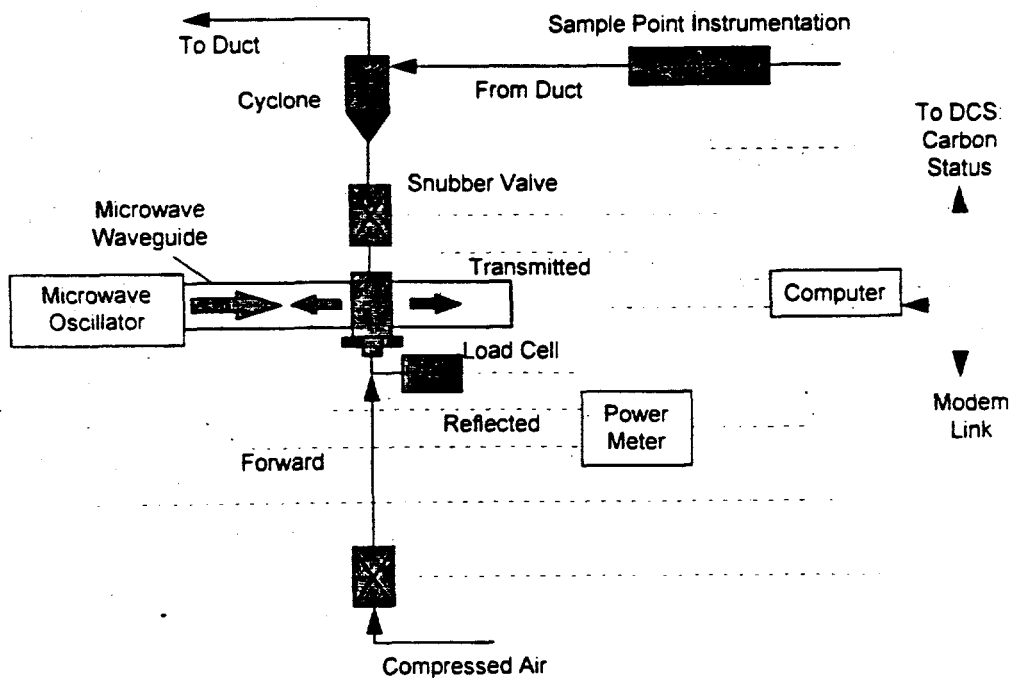


Figure 4. CAM General Arrangement



Table 3. CAM Performance Parameters

Carbon-in-Ash-Monitor (CAM)	
Operating Principle	Absorbance of microwave energy
Instrument Size (WxDxH)	8 ft x 1 ft x 6 ft
Mobility	instrumentation: medium sampling device: medium
Sample Size	~5 grams
Quoted Accuracy	$\pm 0.5\%$ (absolute) below 5% carbon $\pm 10\%$ (relative) above 5% carbon
Analysis Display	actual % carbon of collected sample
Response Time	~5 minutes
Normal Maintenance	calibrate pressure cells - 6 months calibrate load cell - 6 months replace air filters @ intervals based on site air quality
Cost	\$50-100 K depending on options
Contact	Mr. Anthony DiGioia or Mr. Phil Glogowski CAMRAC Company, Inc. 570 Beatty Road Monroeville, Pennsylvania 15146 (412) 856-3200 phone (412) 856-4970 fax

### CLYDE-STURTEVANT

The SEKAM unit was developed by the United Kingdom's Central Electric Generating Board (CEGB) in conjunction with Sturtevant Engineering Systems. Ownership was later transferred to Clyde-Sturtevant Engineering. Commercial production of the current monitor began in late 1990. As of June 1996, total installations worldwide were 40 instruments. In the United States, the system has been tested at four locations including Georgia Power Company's Plant Hammond Unit 4, Carolina Power and Light Roxboro Station, PEPCO in Alexandria, Virginia and on a fluidized bed combustion unit at AES Thames in Connecticut.

The operating principle associated with SEKAM involves trapping fly ash in a glass cell and measuring its capacitance which varies inversely with carbon content. Ash is collected super-isokinetically from the flue gas stream using multiple probes positioned in the flue gas stream. The ash is separated from the gas by a cyclone and is then deposited into a rectangular glass chamber, known as a Kajaani cell; that is located between two capacitance sensors. Ash passes through the vertical glass chamber on a plug flow basis rather than a batch basis. Here the sample capacitance is measured and converted to percent unburned carbon based upon a correlation curve of carbon vs. capacitance. Upon completion of analysis, a portion of the cell is purged to allow a small amount of newly-collected ash to enter the system. In this way, the system displays percent unburned carbon as a rolling average. The total cycle time varies with ash loading. Cycle times of 15 minutes (full load) to greater than two hours (low loads) have been observed.

Photos of the unit as installed at Hammond are shown in Figure 5. A schematic of a general SEKAM arrangement is shown in Figure 6. Information provided in system specifications is shown in Table 4. Some aspects mentioned include accuracy, mobility and cost as provided in marketing literature.

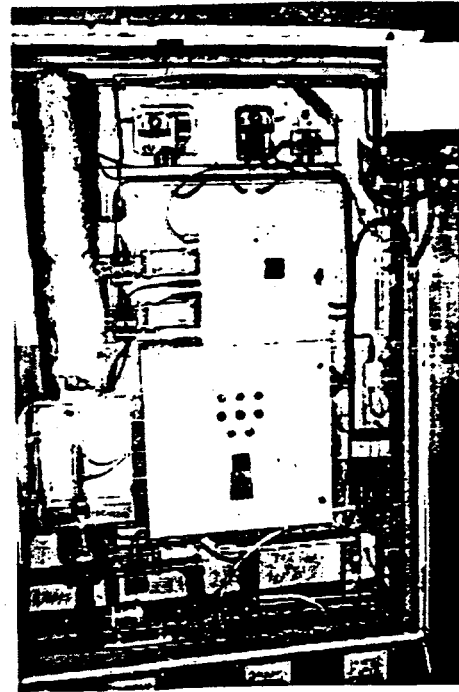
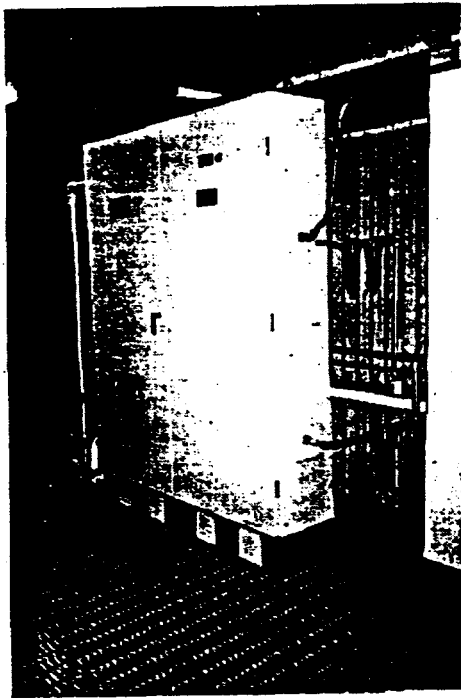


Figure 5. SEKAM On-Line Analyzer

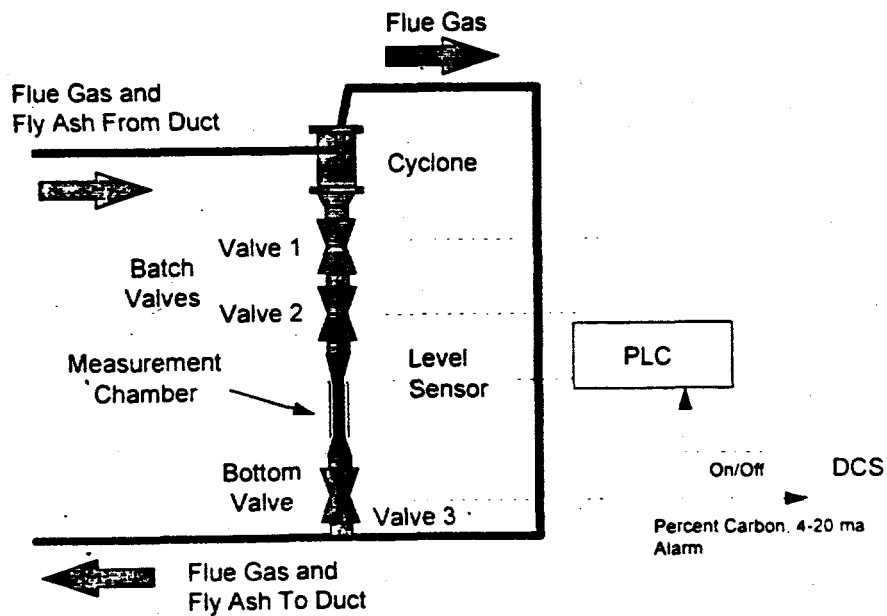


Figure 6. SEKAM General Arrangement

Table 4. SEKAM Performance Parameters

SEKAM	
Operating Principle	Capacitance
Instrument Size (WxDxH)	6 ft x 6 ft x 8.5 ft
Mobility	instrumentation: low sampling device: low
Sample Size	~ 375 grams
Quoted Accuracy	± 1.2%
Analysis Display	% carbon or LOI
Response Time	~ 15 minutes (full load)
Normal Maintenance	replace seals - 6 months
Cost	\$45-50 K
Contact	Mr. Peter Wilson 5732 Rebel Drive Charlotte, North Carolina 28210 (704) 556-1555 phone (704) 556-0136 fax

### APPLIED SYNERGISTICS

Applied Synergistics' FOCUS (Furnace On-line CombUstion System) Unburned Carbon Module is a non-intrusive device that provides a continuous real-time indication of on-line unburned carbon levels in fly ash. Presently, there are cameras installed at six locations worldwide. Testing on the system has been conducted at Georgia Power Company's Plant Hammond Unit 4, Dairyland Power Cooperative's Genoa Unit 3, Baltimore Gas & Electric Company's Brandon Shores Station, and Potomac Electric Power Company's Morgantown Station.

The FOCUS operating principle is based on the premise that unburned carbonaceous material exiting the furnace will be hotter than the surrounding gases and carbon-free ash. Therefore, the carbon-laden particles will emit higher levels of radiant energy in the infrared range. Infrared video cameras installed along the wall of the furnace will record these hotter particles as white spots. These images are then processed to determine the number of traverses in counts per minute. Site-specific equations enable the processor to predict LOI as a function of counts per minute, load, and excess O<sub>2</sub>.

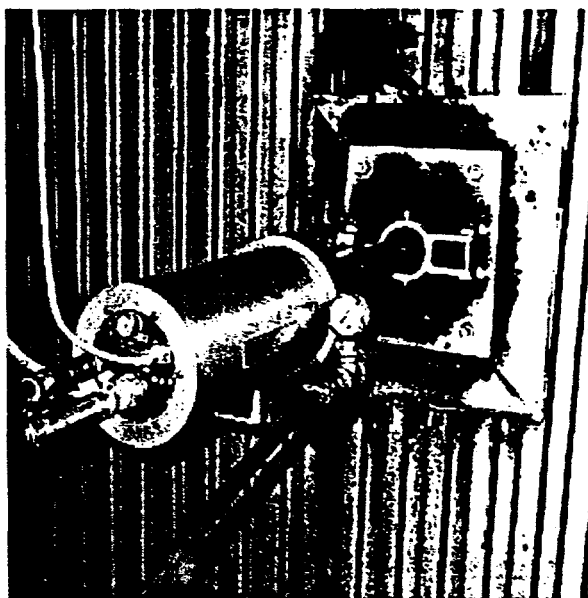


Figure 7. FOCUS On-Line Analyzer

Figure 7 is a photograph of a camera from the FOCUS set-up at Hammond. A schematic of a typical FOCUS arrangement showing major elements of the system is shown in Figure 8. Table 5 presents various aspects of the system including accuracy, instrument size, and cost as provided in company material and system specifications.

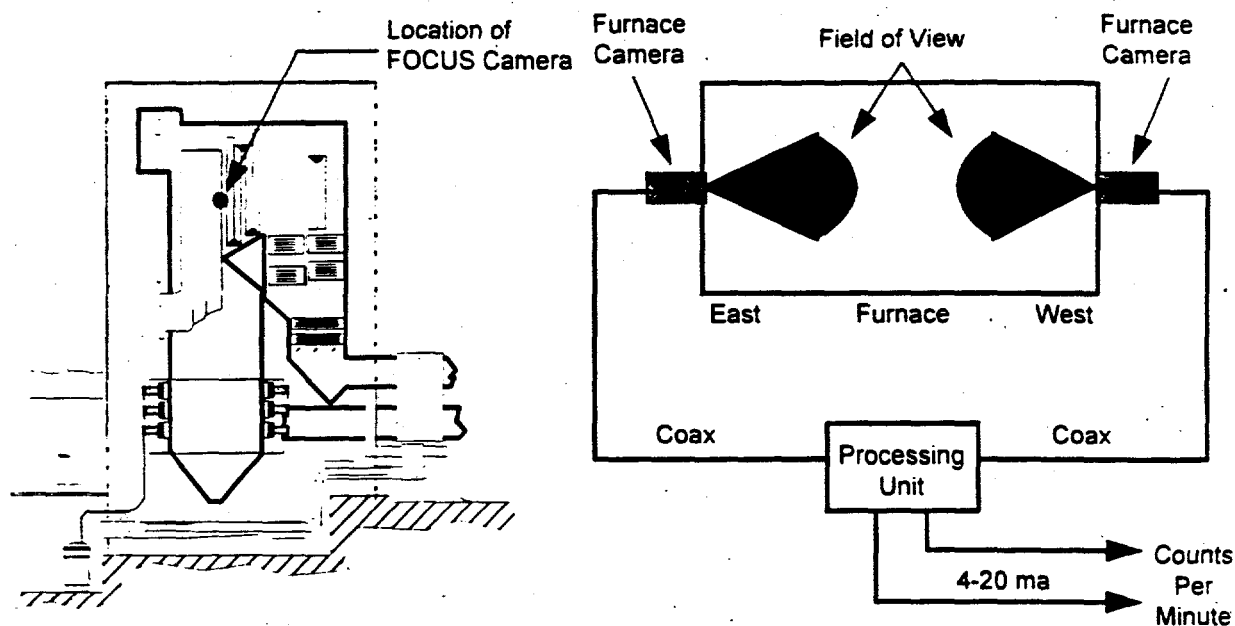


Figure 8. FOCUS General Arrangement

Table 5. FOCUS Performance Parameters

Furnace On-line CombUstion System	
Operating Principle	infrared detection
Instrument Size (WxDxH)	1.0' x 8.9' x 1.0' (one camera)
Mobility	instrumentation: medium monitoring device: medium
Sample Size	none collected
Quoted Accuracy	1.1% standard error of LOI
Analysis Display	bar graph of counts for 24 hour period
Response Time	less than one minute
Normal Maintenance	replace compressed air filters @ intervals dependent on site air quality
Cost	\$40-55 K depending on # of cameras installed
Contact	Mr. Randy Carter Applied Synergistics, Inc. 3831 Old Forest Road, Suite 6 Lynchburg, Virginia 24501 (804) 385-6102 phone (804) 385-0714 fax

Goals of the equipment evaluation at Plant Hammond were as follows: to provide data for additional calibration of instruments demonstrated at this site (CAM, FOCUS, and SEKAM); to compare accuracy of instrument readings versus laboratory determinations of ash samples (except FOCUS); to determine the response time of analyzers to changes in boiler conditions; to compare isokinetic duct conditions to instrument readings and ESP hopper samples; and to estimate availability and durability of instruments using current information on equipment problems (type and duration).

### Equipment Set-Up

CAM, SEKAM and FOCUS on-line LOI analyzers were installed at Georgia Power Company's Plant Hammond Unit 4. Each was placed in a different location as described below. CAM was set up for single point isokinetic sampling with the ash samples being extracted from the "B" side duct between the air heater and the cold side ESP. SEKAM collects samples from the ductwork between the economizer and the air heater. Two sample probes are used to simultaneously extract ash from both the "A" and "B" sides. FOCUS has two cameras positioned in the superheat region of the boiler. One camera provides counts from the east wall ("A" side) and the other from the west wall ("B" side) of the furnace. During the test program, data from the instruments was archived on the unit's digital control system and later retrieved for analysis.

### Test Conditions

A series of tests were conducted in July 1995 (Test 150 & 151) and in February 1996 (Test 152 & 153) in an attempt to evaluate the performance of the CAM, SEKAM, and FOCUS units. Composite duct samples were collected on the "A" and "B" side at the precipitator inlet during each test with each composite sample consisting of ash extracted from 15 locations per side. A duct traverse was conducted at low, normal, and high oxygen levels while the unit was running at nominal 300, 400 and 500 MW loads. In addition to the composite duct samples collected during the duct traverse, ESP hopper samples were collected from the front row of hoppers on A and B sides during each test. SEKAM, CAM, and FOCUS unit readings were also recorded via the digital control system.

### Accuracy

Instrument accuracies were compared in two ways. First, composite duct samples for each test were compared to average unit readings taken during the same testing period. The second method involved placing ash samples from other plants into the units for evaluation, then comparing the instrument and lab values. Since the FOCUS unit does not collect a sample, the latter method was used for the CAM and SEKAM analyzers only.

Test data in Figures 9 through 11 gives an indication of the accuracy of each instrument when compared to the appropriate isokinetic duct composite samples. Since the ash samples collected by the instruments were not analyzed, measurement errors include that resulting from non-representative ash sampling and inherent instrument inaccuracies. It should also be noted that the isokinetic results are themselves just estimates of the actual carbon levels in the duct.

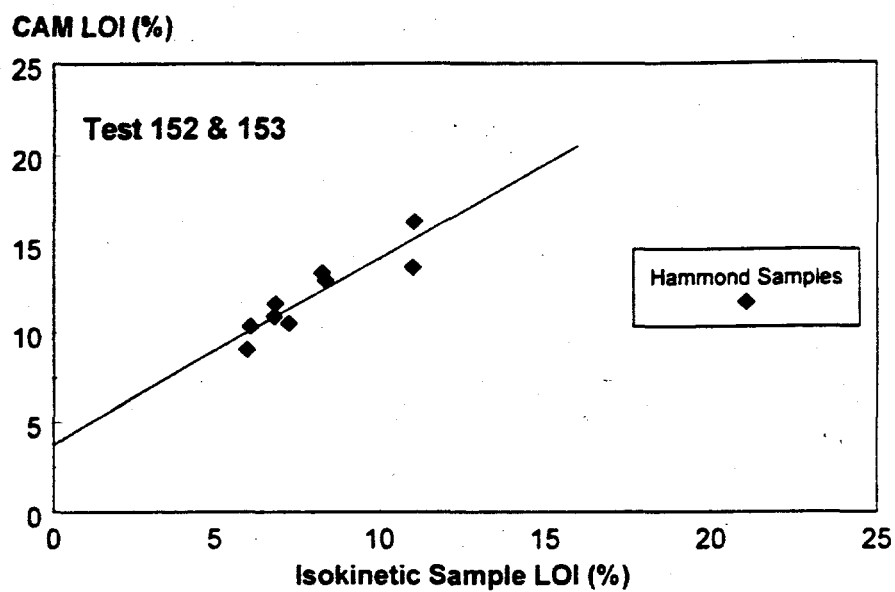


Figure 9. CAM versus Isokinetic LOI

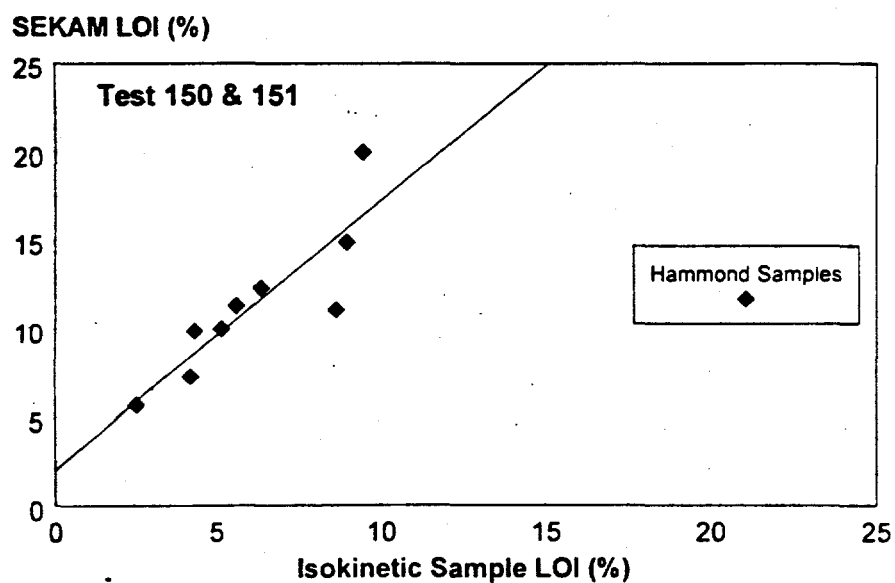


Figure 10. SEKAM versus Isokinetic LOI

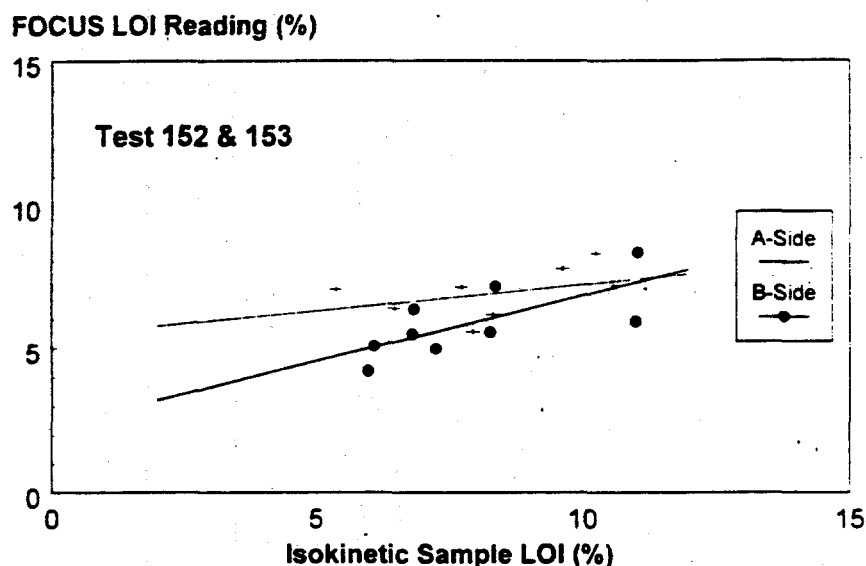


Figure 11. FOCUS versus Isokinetic LOI

As shown, all instruments provide minimal scatter around the line of best fit through the data, indicating adequate precision or repeatability of readings. In terms of accuracy, CAM and SEKAM provide best results, although the degree of accuracy was not always consistent. While the A-side and B-side curves for FOCUS readings deviated further from an ideal 45 degree prediction line (than CAM or SEKAM), the curves possess positive slopes. LOI values predicted using counts from the B-side camera appear to be more accurate than corresponding results from the A-side data.

Accuracy of CAM and SEKAM instruments was evaluated further by directly placing ash samples with known LOI into each unit's sample collection cell for analysis. An advantage of this procedure is the removal of concerns about collecting representative samples to compare with duct composites. In addition, it presented an opportunity to select ash sources which would intentionally provide a larger range of LOI values over which to evaluate accuracies. Figures 12 and 13 show CAM with a slight advantage in accuracy and consistency in these tests. Possible explanations for the differences in instrument and lab LOI include small amounts of moisture which could have been absorbed from the atmosphere during storage prior to use with the instruments. With either analyzer, it is expected that moisture would result in a higher reading.

#### Response Time

The time required for each unit to recognize a change in excess oxygen level was also considered in the evaluation. As mentioned earlier, the test series consisted of sampling at three loads and three oxygen levels at each load. To monitor the response of each instrument, the load and oxygen levels were plotted along with the LOI readings for each unit over a period of time. From Figure 14 it can be seen that, regardless of accuracy, the CAM and FOCUS units respond promptly to changes in boiler conditions. The SEKAM was much slower to respond due to its

sampling procedure primarily as the result of the instrument requiring a relatively large ash sample to perform its analysis.

### Equipment Problems

In addition to performance testing, a log was kept to reflect the availability of each unit and the problems encountered during operation. A summary for each unit is provided below.

SEKAM was installed in November 1994 and has reflected a high availability. Some of the problems that have been encountered and handled are listed as follows:

- Unit not providing readings; A/D converter card replaced.
- Sample valve cycled on and off; valve replaced.
- Extremely low LOI readings; instrument calibrated.
- Small leak in sample cell; valve seals replaced.
- Samples not collected; small holes in sample line patched.

CAM was installed in March 1995 and has demonstrated low to moderate availability, with noticeable improvement in the latter portion of the operating period. Some of the problems associated with this instrument included the following:

- Probe flanges too short; spacer inserted.
- Probes plugged; cleaned out probes.
- Unit shutoff due to a locked monitor; instrument restarted.
- Faulty heat tracing line; line replaced.
- Transmitter not working properly; transmitter replaced.
- Unit not responding during sample collection; weigh cell replaced.
- Moisture in plant air; additional filters installed.

FOCUS was installed in July 1995 and has shown a high level of availability. Maintenance items included:

- An East camera count error occurred due to slag screen movement. A lens filter was installed and the camera was repositioned.
- The automatic iris arrangement on the East side was also changed to a fixed aperture.



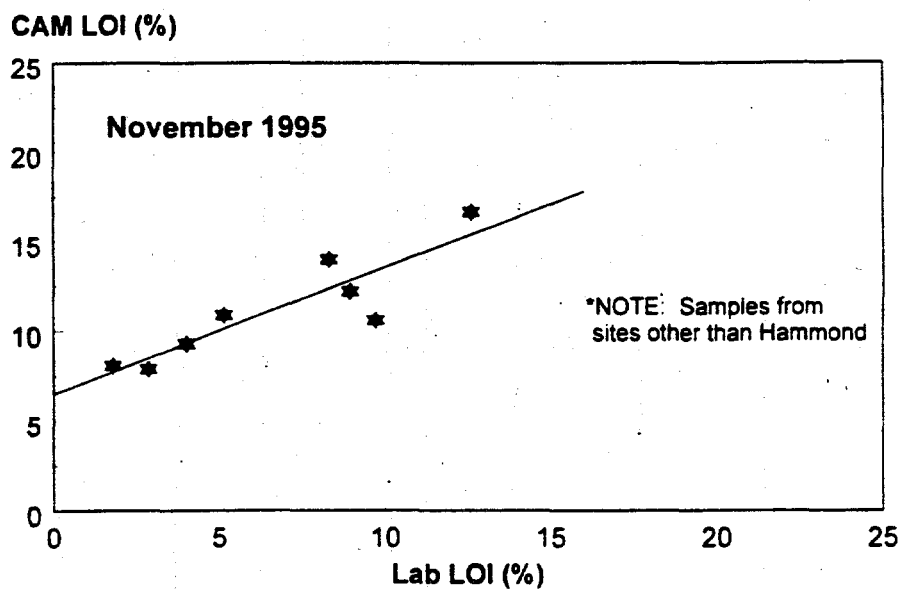


Figure 12. CAM Accuracy

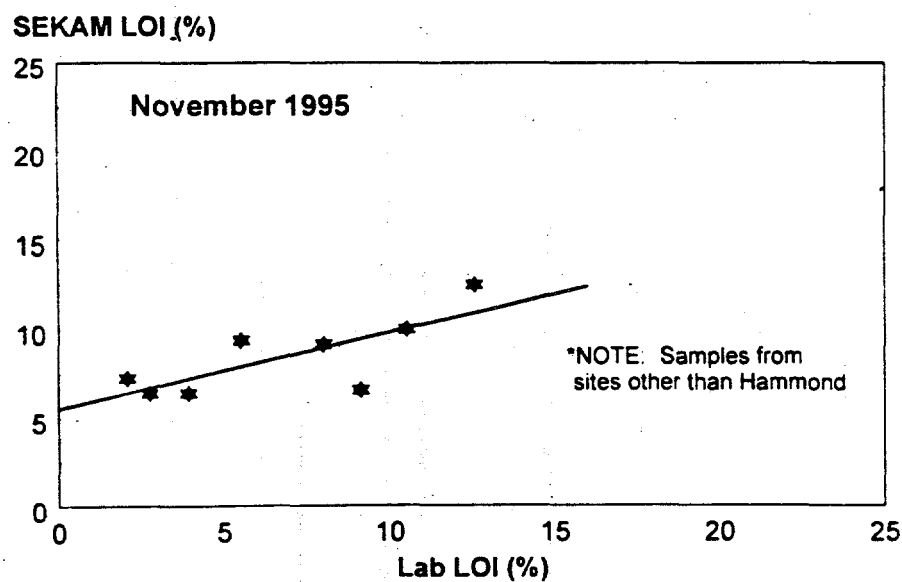


Figure 13. SEKAM Accuracy

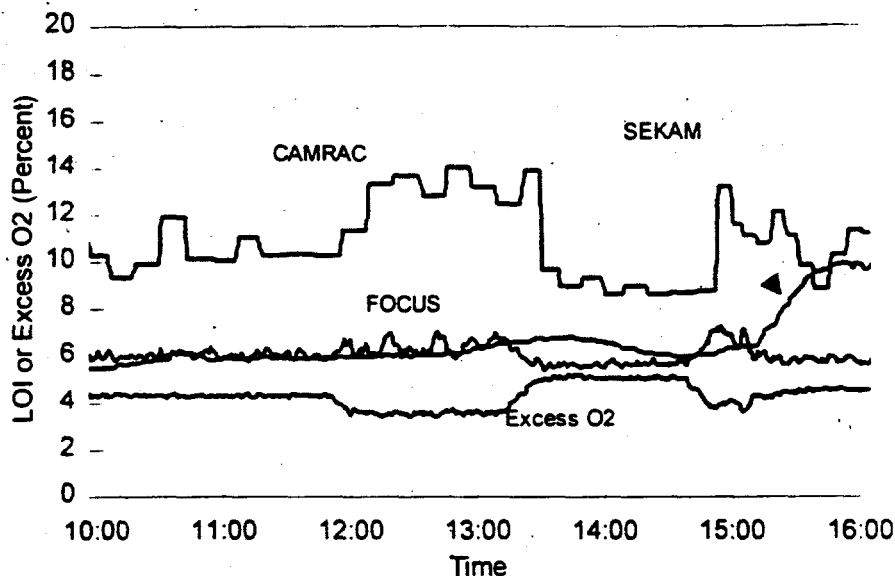


Figure 14. Analyzer Time Response

#### PHASE 4 - ADVANCED CONTROLS / OPTIMIZATION

Phase 4 of the project was the installation and demonstration of an advanced on-line optimization technology -- specifically, GNOCIS. GNOCIS (Generic NO<sub>x</sub> Control Intelligent System) is an enhancement to digital control systems (DCS) targeted at improving utility boiler efficiency and reducing emissions. GNOCIS is designed to operate on units burning gas, oil, or coal and is available for all combustion firing geometries. GNOCIS utilizes a neural-network model of the combustion characteristics of the boiler that reflects both short-term and longer-term trends in boiler characteristics. A constrained-nonlinear optimizing procedure is applied to identify the best set points for the plant. These recommended set points can be implemented automatically without operator intervention (closed-loop), or, at the plant's discretion, conveyed to the plant operators for implementation (open-loop). The software is designed for continuous on-line use. The major elements of GNOCIS are shown in Figure 15.

Alabama Power Company's Gaston Unit 4, a 270 MW wall-fired unit, and PowerGen's Kingsnorth Unit 1, a 500 MW tangentially-fired unit served as developmental sites for GNOCIS [7].

GNOCIS development was funded by a consortium consisting of the Electric Power Research Institute, PowerGen, Southern

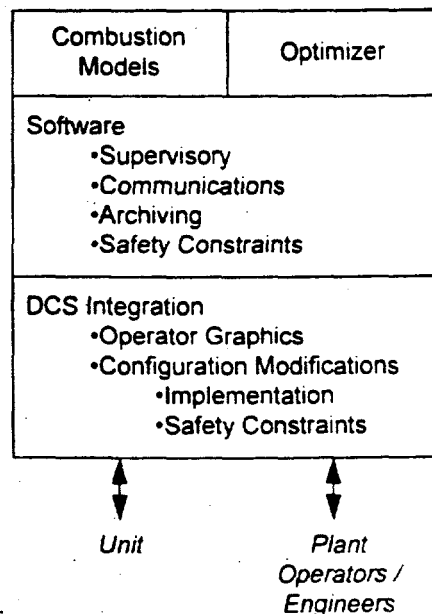


Figure 15. Major Elements of GNOCIS

Company, Radian International, U.K. Department of Trade and Industry, and U.S. Department of Energy.

### GNOCIS Implementation

From project inception, the goal of the GNOCIS installation at Hammond has been to implement a closed-loop, supervisory system. The Foxboro DCS, installed in 1994, included configuration enhancements which facilitated incorporation of GNOCIS into the overall control strategy. As at Gaston, all operator interaction with GNOCIS is through the DCS operator displays. The GNOCIS host platform at this site is a Sun Sparc 5 running the Solaris 2.3 operating system. This platform was chosen here since the Foxboro system also uses the Sparc architecture. The Sun interfaces to the DCS using local area network connection and TCP/IP.

### Model Development

As at Gaston and Kingsnorth, data collected through the DCS are used to create the combustion models with modeling efforts concentrating on the most recent long-term data. As necessary, tests are run at off-design conditions to augment data available from normal operation and thereby expand the range over which the combustion model could make estimates.

### Trial Results

Following the completion of installation, preliminary testing of GNOCIS at Hammond 4 began during February 1996 with tests being conducted at loads of 500 MW, 400 MW, and 300 MW. Various combinations of objectives were tested including minimizing NO<sub>x</sub> emissions, minimizing carbon-in-ash, and maximizing efficiency in both open- and closed-loop modes. Implementation of the GNOCIS recommendations were greatly facilitated as a result of enhancements made to the DCS. Results from these early tests suggested that further modifications be made to the system with the most important modification being the substitution of the overfire flow control damper positions for the corresponding overfire air flows. This change was necessary since these flow measurements had, to a large degree, become unreliable. On February 24, the unit went off-line for a scheduled maintenance outage.

During May 1996, testing of GNOCIS in both open- and closed-loop modes resumed with 22 tests being conducted. As before, various objectives were tested. Although relatively narrow limits were placed on the recommendations that GNOCIS could provide, preliminary analysis of the results are encouraging. On May 17, the unit came off-line as a result of turbine problems and has just recently returned to service. Several of the tests conducted prior to the outage are discussed below.

**Test Day 158.** Test 158 was conducted on May 7 with the unit off economic dispatch and at 480 MW. The purpose of the test was to evaluate the performance of GNOCIS in regards to boiler efficiency improvements as GNOCIS was made sequentially less constrained (Table 6). The tests were conducted in open-loop mode. Boiler efficiency and a subset of the independent control variables during the course of the test period are shown in Figure 16. As shown, nominal boiler efficiency was near 87.5 percent at the beginning of the testing and with sequential

application of the GNOCIS recommendations, an efficiency of approximately 88.3 percent was attained. As can be seen in the figure, recommendations for excess oxygen, AOFA damper, and mill loading were implemented at approximately 11:15, 12:10, and 12:45, respectively. Also note that the recommended damper position is dependent on whether the mills are included in the optimization mix.

**Test Day 161.** Test 161 was conducted on May 15, 1996 at full load and in closed-loop mode. Results and control actions taken are shown in Figures 17 and 18 respectively. During the test day several objectives were tested including minimizing NO<sub>x</sub> and LOI and maximizing boiler efficiency. As with previous closed-loop tests at this site, recommendations were intentionally made narrow until further confidence was gained in the stability of GNOCIS recommendations (Table 6). When NO<sub>x</sub> minimization was the goal (Test 161-1), NO<sub>x</sub> emissions were reduced by approximately 10 percent from baseline. Similarly, when efficiency and LOI were goals (161-2 and 161-3), improvements of near 0.7 percent and 2 percent, respectively, were obtained. Also, in Test 161-2, simultaneous improvements in NO<sub>x</sub>, LOI, and efficiency were obtained. As can be seen in Figure 18, GNOCIS did not adversely affect the stability of the control actions.

**Test Day 162.** Test day 162 (Figures 19 and 20), conducted on May 16, 1996 was also at full load and GNOCIS was again operating in closed-loop mode. In Test 162-1, minimize LOI was the goal and as shown, a reduction of approximately 2.2 percent was obtained. As expected, NO<sub>x</sub> emissions increased with increasing O<sub>2</sub> levels. The goal was then changed to minimize NO<sub>x</sub> with O<sub>2</sub> clamped to the current levels. As shown, at least for the conditions present for this test, GNOCIS estimated that the other independent control variables (AOFA dampers and mill loadings) would have minimal impact on NO<sub>x</sub> emissions and therefore no control action was taken (Figure 20). The final test (162-3) freed up excess O<sub>2</sub> and the control action was taken resulting in a NO<sub>x</sub> reduction of approximately 10 percent. As with the prior days testing, there was no apparent adverse impact on the stability of the unit (Figure 20).

Table 6. Hammond / Short-Term Tests

Test	Goals			Limits		
	NO <sub>x</sub>	LOI	Efficiency	Excess O <sub>2</sub>	AOFA Dmpr.	Mill Flows
158-1	-	-	Max	±0.2	Clamped	Clamped
158-2	-	-	Max	±0.2	±5	Clamped
158-3	-	-	Max	±0.2	±5	±5
161-1	Min	-	-	±0.2	±5	±5
161-2	-	-	Max	±0.2	±5	±5
161-3	-	Min	-	±0.2	±5	±5
162-1	-	Min	-	±0.4	±5	±5
162-2	Min	-	-	Clamped	±5	±5
162-3	Min	-	-	±0.4	±5	±5

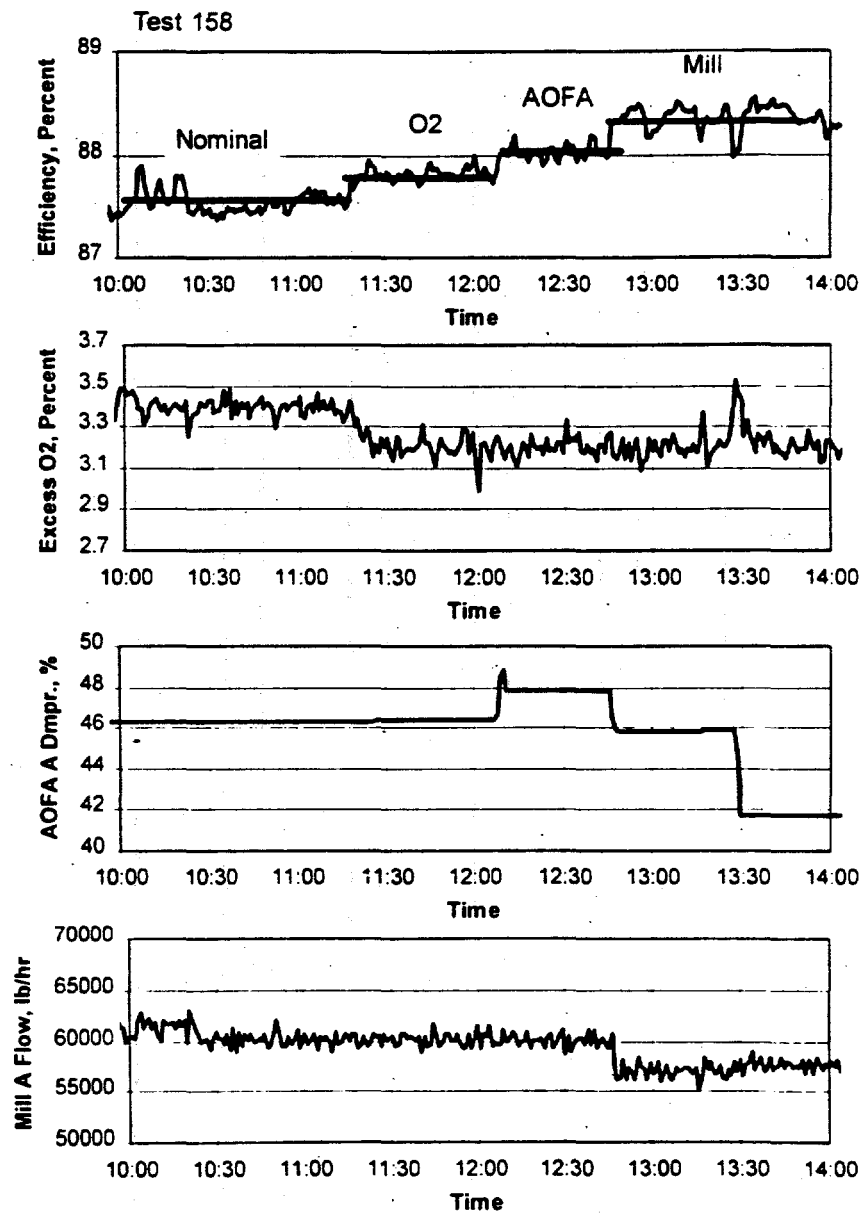


Figure 16. Hammond / Results of May 7, 1996 Testing

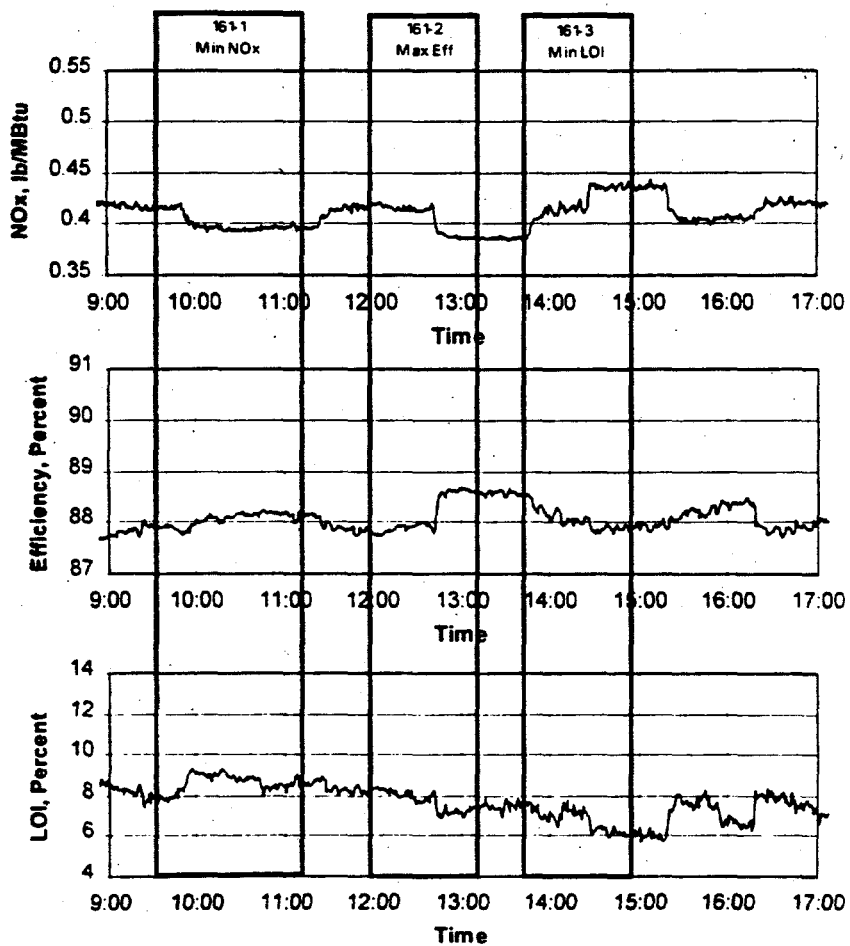


Figure 17. Hammond / Results of May 15, 1996 Testing

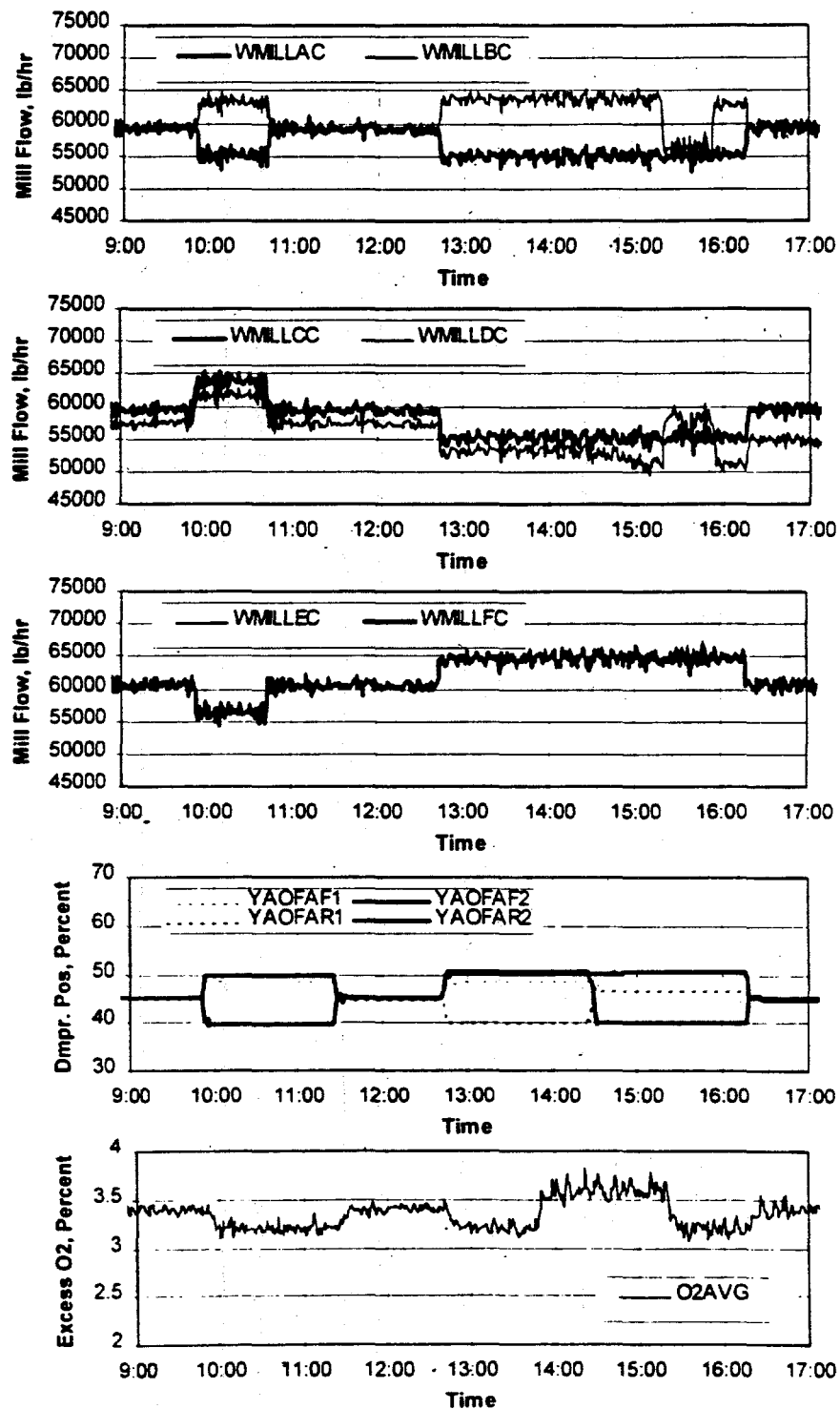


Figure 18. Hammond / Control Actions During May 15, 1996 Testing

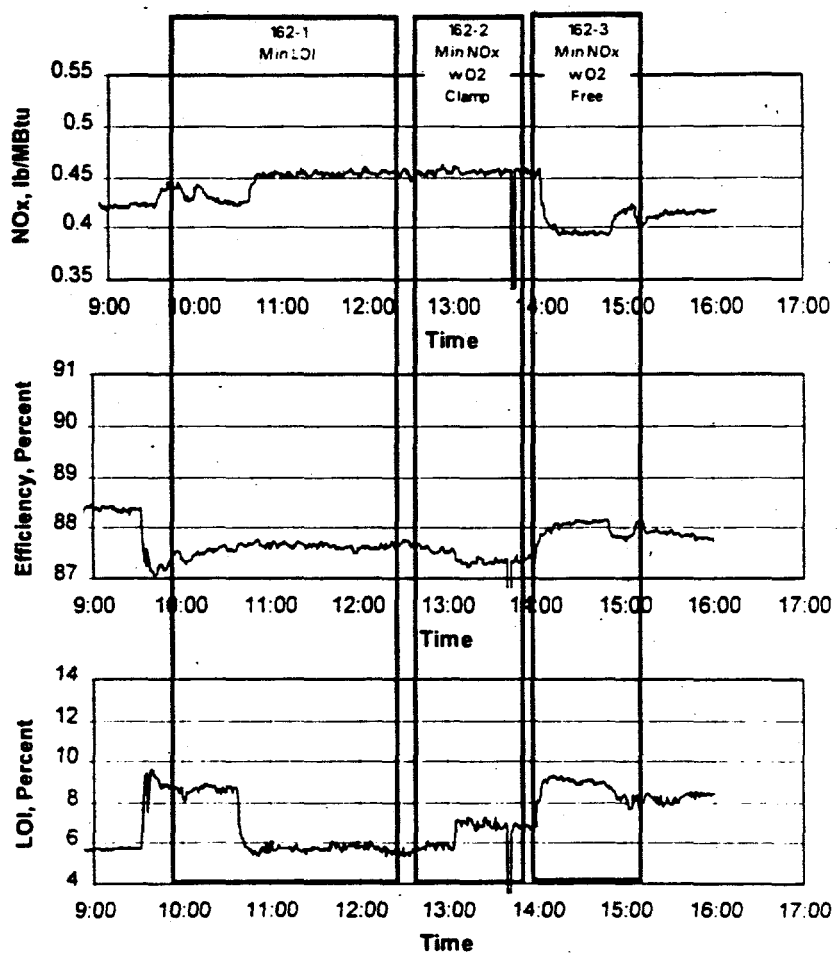


Figure 19. Hammond / Results of May 16, 1996 Testing



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3. *500 MW Demonstration of Advanced Wall-Fired Combustion Techniques for the Reduction of Nitrogen Oxide Emissions from Wall-Fired Boilers - Phase 3A Low NO<sub>x</sub> Burner Tests.* Southern Company Services, Birmingham, AL: 1994.
  4. *500 MW Demonstration of Advanced Wall-Fired Combustion Techniques for the Reduction of Nitrogen Oxide Emissions from Wall-Fired Boilers - Phase 3B Low NO<sub>x</sub> Burner plus Advanced Overfire Air Tests.* Southern Company Services, Birmingham. AL: 1995.
  5. *500 MW Demonstration of Advanced Wall-Fired Combustion Techniques for the Reduction of Nitrogen Oxide (NO<sub>x</sub>) Emissions from Coal Fired Boilers - Technical Progress Report - Third Quarter 1991.* Southern Company Services Inc.. Birmingham. AL: 1992.
  6. Sorge, J., Wilson, S., "500 MW Demonstration of Advanced Wall-Fired Combustion Techniques for the Reduction of Nitrogen Oxide (NO<sub>x</sub>) Emissions from Coal Fired Boilers," Third Annual Clean Coal Technology Conference, September 6-8, 1994. Chicago, Illinois.
  7. Sorge, J., Squires, R., Menzies, W., Stallings, J., "GNOCIS An Update on the Generic NO<sub>x</sub> Control Intelligent System," EPRI NO<sub>x</sub> Controls for Utility Boilers, August 6-8, 1996, Cincinnati, OH.

**Technical Session IV**  
**Advanced Power Generation**  
**Systems**



# **Wabash River Coal Gasification Repowering Project - First Year Operation Experience**

**E. J. (Chip) Troxclair**  
Destec Energy, Inc.  
Houston, Texas

**Jack Stultz**  
PSI Energy, Inc.  
West Terre Haute, Indiana

## **ABSTRACT**

The Wabash River Coal Gasification Repowering Project (WRCGRP), a joint venture between Destec Energy, Inc. and PSI Energy, Inc., began commercial operation in November of 1995. The Project, selected by the United States Department of Energy (DOE) under the Clean Coal Program (Round IV) represents the largest operating coal gasification combined cycle plant in the world. This Demonstration Project has allowed PSI Energy to repower a 1950's vintage steam turbine and install a new syngas fired combustion turbine to provide 262 MW (net) of electricity in a clean, efficient manner in a commercial utility setting while utilizing locally mined high sulfur Indiana bituminous coal. In doing so, the Project is also demonstrating some novel technology while advancing the commercialization of integrated coal gasification combined cycle technology. This paper will discuss the first year operation experience of the Wabash Project, focusing on the progress towards achievement of the demonstration objectives.

## **Acknowledgements**

DOE Project Manager: Gary Nelkin  
Participant Joint Venture Manager: Phil Amick, Destec Energy, Inc.  
Demonstration Period: December, 1995 - November, 1998

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Research sponsored by the U.S. Department of Energy's Morgantown Energy Technology Center, under contract DE-FC21-92MC29310 with Wabash River Coal Gasification Repowering Project Joint Venture, 2500 CityWest Blvd., Suite 150, Houston, Texas 77042: telefax: (713) 735-4837.

## **Introduction**

When the Wabash River Coal Gasification Repowering Project Joint Venture (the JV) signed the Cooperative Agreement with the U.S. Department of Energy (the DOE) in July 1992, this marked the beginning of a truly beneficial alignment amongst the entities involved. PSI needed a clean, low cost, energy efficient baseload capacity addition that would function as a substantial element of their plan to comply with the requirements of the Clean Air Act. Also important was this projects' ability to process locally-mined (Indiana) high sulfur coal. Finally, PSI needed a project that would pass the approval of the Indiana Utility Regulatory Commission as the low cost option for baseload capacity addition.

Encouraged by the data and experience gained at its Louisiana Gasification Technology, Inc. plant (LGTI) and by the DOE Clean Coal Technology Program, Destec was interested in advancing its gasification technology to the next generation to enhance the competitive position of gasification technology for future IGCC projects.

The DOE, through its Clean Coal Round IV Program, wanted a commercial demonstration of a clean coal technology to abate the barriers to commercialization of clean coal technologies and gain data to enable power generators to make informed decisions concerning utilization of clean coal technologies.

Through the Wabash River Coal Gasification Repowering Project (the Project), the needs of the participants and the DOE are being met with this 262 MW commercial power plant. This Project is demonstrating a clean, highly efficient technology that meets today's energy demand and tomorrow's (year 2000) clean air requirements.

## **Overview**

The Project Participants, Destec Energy, Inc. (Destec) of Houston, Texas and PSI Energy, Inc., (PSI) of Plainfield, Indiana, formed the JV to participate in DOE's Clean Coal Technology (CCT) program to demonstrate the coal gasification repowering of an existing generating unit affected by the Clean Air Act. The Participants jointly developed, but separately designed, constructed, own, and are now operating an integrated coal gasification combined cycle power plant, using Destec's coal gasification technology to repower the oldest of the six units at PSI's Wabash River Generating Station in West Terre Haute, Indiana. Destec's gasification process is integrated with a new GE 7 FA combustion turbine generator and heat recovery stream generator in repowering of a 1950's - vintage steam turbine generator using pre-existing coal handling facilities, interconnects, and other auxiliaries.

The Project has completed the first year of a three year Demonstration Period under the DOE CCT program. The early operation of the Project, which is now the world's largest single-train coal gasification combined cycle plant operating commercially, has demonstrated the ability to run at full load capability while meeting the environmental requirements for sulfur and NO<sub>x</sub> emissions. CINergy, PSI's post-merger organization, dispatches the Project second behind their hydro facilities

on the basis of environmental emissions and efficiency, with a demonstrated heat rate of approximately 9,000 Btu/KWh (HHV).

## **Background**

### **Destec Gasification Technology Evolution**

Destec's parent Company, the Dow Chemical Company (Dow), began the development of the Destec Gasification process in the early 1970's. Dow wanted to diversify its fuel base from natural gas to lignite and coal for its power intensive chlor-alkali processes and began to develop the gasification process through basic R&D and pilot plants. Dow's first commercial gasification plant followed, the Louisiana Gasification Technology, Inc. (LGTI) facility in Plaquemine, La. This project operated from the second quarter 1987 until the third quarter 1995 under subsidy from the Synthetic Fuels Corporation and later the Treasury Department. When Destec was formed in 1989 the gasification technology was transferred from Dow to Destec.

### **Wabash Project Development**

Destec approached PSI in early 1990 to initiate discussions concerning the DOE Clean Coal Technology Round IV program solicitation. Through the Wabash River Coal Gasification Repowering Project Joint Venture, the project submittal was made. In September 1991, the Project was among nine projects selected from 33 proposals. The Project was selected to demonstrate the integration of Destec's gasification process with a new GE 7FA combustion turbine generator and HRSG in the repowering of an aged steam turbine generator to achieve improved efficiency and reduced emissions.

### **Goals of Participants**

- PSI wants to demonstrate an alternative technology for new units and repowering of existing units. Also PSI is incorporating this IGCC power plant into their system and wants to demonstrate this as a reliable and cost-effective element of their baseload generation capability.
- Destec is demonstrating the operability, cost effectiveness and economic viability of its gasification technology in a commercial utility setting.
- Destec wants to further enhance its gasification technology's competitive position by demonstrating new techniques and process enhancements as well as substantiate performance expectations and capital and operating costs.

- The DOE wants to abate the barriers to commercializing clean coal technologies, particularly gasification and repowering applications, and otherwise enable power generators to make informed commercial decisions concerning the utilization of clean coal technology.

### **Project Organization, Commercial Structure, and Costs**

There are two major agreements which establish the basis of the Project. First, the Joint Venture Agreement was created between PSI and Destec to form the Wabash River Coal Gasification Repowering Project Joint Venture in order to administer the Project under the DOE Cooperative Agreement. Second, the Gasification Services Agreement (GSA) was developed between PSI and Destec and contains the commercial terms under which the Project was developed and is now operated.

#### **PSI Responsibilities:**

- build power generation facility to an agreed schedule
- own & operate the power generation facility
- furnish Destec with a site, coal, electric power, stormwater and wastewater facilities, and other utilities and services.

#### **Destec Responsibilities:**

- build gasification facility to agreed schedule
- own and operate the gasification facility
- guarantee operating performance of coal gasification facility including product & by-product quality
- deliver syngas and steam to the power generation facility

### **Project Costs**

The overall combined cost of the gasification and power generation facilities was \$417 million at completion. This cost includes the costs of engineering and environmental studies, equipment procurement, construction, pre-operations management (including operator training), and start-up. This figure includes escalation during the project. The start-up costs include the costs of construction and operations, excluding coal and power, up to the date of commercial operation in November 1995. Soft costs such as legal and financing fees and interest during construction are not included in this figure.

A savings of \$30-40 million was realized by the repowering of the existing PSI facility, re-using the steam turbine and auxiliaries and coal handling equipment. This probably also reduced the project schedule by as much as a year, because of the simplified permitting effort versus a greenfield project.

Two areas of significant impact that increased the cost of the project were unanticipated construction problems and start-up delays. The construction effort was plagued by weather

problems in the first nine months of the schedule, and later by labor shortages and construction contractor problems, that led to massive acceleration in the last 25% of the two year construction schedule. During the combined start-up of the gasification and power generation facilities, certain delays contributed to extension of the project fixed costs that also contributed to the final cost.

Project participants anticipate the costs of future units to be reduced dramatically, to the \$1200/kw range for dual train facilities. Advances in turbine technology should bring the installed cost to under \$1000 / kw for greenfield installations by the year 2000.

### **Project Schedule**

The schedule for this project spans the time from selection in September, 1991 by the DOE during Clean Coal Round IV awards, to the end of the three year demonstration period in November 1998. The major project activities and corresponding milestones are as follows:

DOE Selection in Round IV	September	1991
Cooperative Agreement Finalized	August	1992
Environmental Assessment Complete	May	1993
State Air Permits Complete	May	1993
Indiana Utility Regulatory Approval Complete	May	1993
Began Construction	September	1993
Completed Construction	July	1995
First Coal Operation	August	1995
Began Commercial Operation	November	1995
Began Demonstration Period	December	1995
Complete Demonstration Period	November	1998
Final Report	February	1999

This aggressive schedule was possible by overlapping of activities between the development and engineering periods as well as the engineering and construction periods.

### **Review of Technology**

#### **General Design and Process Flow**

The Destec coal gasification process features an oxygen-blown, continuous-slugging, two-stage, entrained-flow gasifier which uses natural gas for startup. Coal is milled with water in a rodmill to



form a slurry. The slurry is combined with oxygen in mixer nozzles and injected into the first stage of the gasifier, which operates at 2600 F and 400 psig. Oxygen of 95% purity is supplied by a turnkey, 2060-ton/day low-pressure cryogenic distillation facility which Destec owns and operates.

In the first stage, coal slurry undergoes a partial oxidation reaction at temperatures high enough to bring the coal's ash above its melting point. The fluid ash falls through a taphole at the bottom of the first stage into a water quench, forming an inert vitreous slag. The syngas then flows to the second stage, where additional coal slurry is injected. This coal is pyrolyzed in an endothermic reaction with the hot syngas to enhance syngas heating value.

The syngas then flows to the High Temperature Heat Recovery Unit (the HTHRU), essentially a firetube steam generator, to produce high pressure saturated steam. After cooling in the HTHRU, particulates in the syngas are removed in a hot/dry filter and recycled to the gasifier where the carbon in the char is converted into syngas. Filter-element construction is a proprietary design proven at full scale at LGTI. The syngas is further cooled in a series of heat exchangers and passed through a catalyst which hydrolyzes carbonyl sulfide into hydrogen sulfide. Hydrogen sulfide is removed using MDEA-based stripper columns. The "sweet" syngas is then moisturized, preheated, and piped over to the power block.

The key elements of the power block are the General Electric MS 7001 FA high-temperature combustion turbine/generator, the heat recovery steam generator (the HRSG), and the repowered steam turbine.

The GE 7FA is a dual-fuel machine (syngas for operations and No. 2 fuel oil for startup) capable of a nominal 192MW when firing syngas, which is attributed to the increased mass flows associated with syngas. Steam injection is used for NO<sub>x</sub> control, but the steam flow requirement is minimal compared to conventional systems because the syngas is moisturized at the gasification facility, making use of low-level heat in the process. The water consumed in this process is continuously made up at the power block by water treatment systems which clarify and treat river water.

The HRSG for this project is a single-drum design capable of superheating 754,000 lb/hr of high-pressure steam at 1010 F, and 600,820 lb/hr of reheat steam at 1010 F when operating on design-basis syngas. The HRSG configuration was specifically optimized to utilize both the gas-turbine exhaust energy and the heat energy made available in the gasification process. The nature of the gasification process in combination with the need for strict temperature and pressure control of the steam turbine led to a great deal of creative integration between the HRSG and the gasification facility.

The repowered unit, originally installed in 1952, consisted of a conventional coal-fired boiler feeding a Westinghouse reheat steam turbine rated at 99MW but derated in recent years to 90MW for environmental dispatch. Repowering involved refurbishing the steam turbine to both extend its life and withstand the increased steam flows and pressures associated with the combined cycle operation.

The repowered steam turbine produces 104MW which combines with the combustion turbine generator's 192MW and the system's auxiliary load of approximately 34MW to yield 262MW (net) to the CINergy grid.

At the design point, the Air Separation Unit (ASU) provides oxygen and nitrogen for use in the gasification process but is not an integral part of the plant thermal balance. The ASU uses services such as cooling water and steam from the gasification facilities and is operated from the gasification plant control room.

The gasification facility produces two commercial byproducts during operation. Sulfur removed as 99.9 percent pure elemental sulfur is marketed to sulfur users. Slag will be sold as aggregate in asphalt roads and as structural fill in various types of construction applications.

### **Technical Advances**

Using integrated coal gasification combined cycle technology to repower a 1950's-vintage coal-fired power generating unit essentially demonstrates a technical advance in and of itself.

More specifically, high energy efficiency and superior environmental performance while using high sulfur bituminous coal is the result of several improvements to Destec's gasification technology, including:

- Hot/Dry Particulate Removal, applied at full commercial scale with no provision for bypass.
- Syngas Recycle, which provides fuel and process flexibility while maintaining high efficiency.
- A High Pressure Boiler, which cools the hot, raw gas by producing steam at a pressure of 1,600 psia.
- A Dedicated Oxygen Plant, which produces 95% pure oxygen for use by the Project. Use of 95% purity increases overall efficiency of the Project by lowering the power required for production of oxygen.
- Integration of the Gasification Facility with the Heat Recovery Steam Generator to optimize both efficiency and operating costs.
- The Carbonyl Sulfide Hydrolysis system, which allows such a high percentage of sulfur removal.
- The Slag Fines Recycle system, which recovers carbon remaining in the slag byproduct stream and recycles it back for enhanced carbon conversion. This also results in a higher quality byproduct slag.
- Fuel Gas Moisturization, which uses low-level heat to reduce steam injection required for NO<sub>x</sub> control.
- Sour water treatment and Tail Gas Recycling, which allow more complete recycling of combustible elements, thereby increasing efficiency and reducing waste water and air emissions.

The Project's superior energy efficiency is also attributable to the power generation facilities

included in the Project. These facilities incorporate the latest advancements in combined cycle system design while accommodating design constraints necessary to repower the steam turbine, including:

- The Project is the first application of Advanced Gas Turbine technology for syngas fuel, incorporating redesigned compressor and turbine stages, higher firing temperatures and higher pressure ratios, specially modified for syngas combustion.
- Repowering of the Existing Steam Turbine involved upgrading the unit in order to accept increased steam flows generated by the HRSG. In this manner, the cycle efficiency is maximized because more of the available energy in the cycle is utilized.

## **Operations Experience**

The Project completed the commissioning phase in August of 1995 and began the start-up process. By late August, the gasifier was ready for coal feed. The Project was in the start-up and testing mode through mid November at which time the start-up tests were complete and the Project was ready for the commercial operation and demonstration phase to begin. Significant in the start-up phase was the successful demonstration of the thermal integration of the combined operations. There were no substantial problems integrating the steam and water systems, although some early feedwater control problems contributed to early operation interruptions that carried over to the commercial operating period. These problems have since been resolved. The startup phase also demonstrated product (syngas) and by-product (slag & sulfur) quality and environmental performance.

### **Demonstration Period Test Plan**

With this project being a full scale commercial unit in a utility environment, the Test Plan for the Demonstration Period focuses on successful operation of the plant as a base-load unit in the PSI system. Specifically, the goals of the participants for the Demonstration Test Plan primarily address continuous improvement in plant availability, operating and maintenance costs, maintaining dispatch, and improvement in overall performance while fulfilling the reporting requirements for environmental performance and equipment/system performance. Towards these goals, the next section will address the first year of performance under the three year demonstration period.

### **Operations Statistics/Milestones**

The early commercial operation of the WRCGRP saw the plant build on the success of the start-up period with primary focus on attaining maximum sustained capacity for the purpose of final performance testing for the Air Separation Unit (ASU) Facility and Gasification Plant. The ASU Performance Testing was completed in February 1996 during an operating campaign that lasted over 300 hours. In March 1996 just four months into the operating period, the gasification plant demonstrated extended operation at 100% of rated design by running over 100 hours at or above

gasifier design capacity. During these February and March operating campaigns the combustion turbine ran smoothly on syngas and had periods of operation at the 192 MW maximum rated capacity on syngas.

As the Project accumulated the early run time, evaluation of the technical advances noted previously showed that most of the new unit operations performed very well, however two of the areas contributed problems which affected run time. The primary problem area has been the reliability of the particulate removal system, primarily due to breakage of ceramic candle filters. Further testing and modifications to the particulate removal system are underway to minimize element breakage. Another problem area was chloride concentrations in both the COS hydrolysis catalyst beds and downstream heat exchangers in the syngas cooler line-up. Unexpected localized high chloride concentrations contributed to catalyst poisoning and chloride stress corrosion cracking in the low temperature syngas heat exchangers. A scrubber system has been installed to remove the chlorides from the syngas prior to the COS hydrolysis beds and syngas heat exchangers. These modifications are in place as the plant moves into the second operating year.

On the Power Block side the new Advanced Gas Turbine has performed very well on syngas. The turbine's operation has been more stable on syngas than on oil, with blade temperatures more evenly distributed and less temperature spiking. NO<sub>x</sub> is controlled with steam injection to meet air permit requirements. The turbine experienced three problem areas after the acceptance of syngas. The first was in the syngas module and the piping from the module to the gas turbine. Expansion bellows required redesign and replacement to eliminate mechanical cracking in the flow sleeves. This problem was corrected by GE efforts in early syngas runs. The second problem has been the syngas purge control. These problems were primarily related to field devices such as solenoid valves and flow measuring devices. The solenoids have been redesigned and replaced and GE continues to work on flow measuring devices. The third area was the GE required row 2-3 spacer modifications, a fleet problem unrelated to syngas utilization.

Table I shows the production statistics for both the Gasification plant and combined cycle plant through October 1996.

<b><u>Gasification Plant Production Statistics</u></b>	
First Coal Gasified	August 17, 1995
Total Gasifier Hours on Coal*	2035
Total Syngas produced*	2,814,066 MMBtu (Dry)
Total Coal Processed*	189,233 Tons
Highest Capacity Demonstrated (% Nameplate)	103% (1825 MMBtu/hr, HHV)
Longest Continuous Coal Run* (Hours)	253
Cold Gas Efficiency (%)	>74%

<b><u>Combined Cycle Plant Production Statistics</u></b>	
First Syngas to Combustion Turbine (C.T.)	October 3, 1995
Total C.T. Hours*	2872
Total C.T. Hours on Syngas*	1340
MWH'S produced on Syngas*	333,486
Highest C.T. Capacity Demonstrated (% Nameplate)	100% (192 MW)
Longest Continuous Syngas Operation* (Hours)	151

\*(All Production Statistics through October 1996)

**TABLE I**

Following is an operations summary of each major operating area, including the areas mentioned above, with a discussion of the process modifications incorporated to address the early problems encountered.

### **Area Operations Summaries**

#### **Coal Slurry Preparation**

Coal is ground into a slurry in a rodmill, using recycled water from the gasification process. Wet milling reduces potential fugitive particulate emissions and minimizes water consumption and effluent waste water volume. The slurry is stored in an agitated tank large enough to supply the gasifier during rodmill forced outages.

The slurry preparation area has now processed (189,233) tons of coal with no significant problems. Typical problems handling coal during low ambient temperature conditions and heavy snowfall were experienced, primarily with the automatic sampling equipment, but the slurry has consistently met target solids concentration. The slurry storage and feed systems have also performed very well since the beginning. Typical Coal properties are shown in Table II.

COAL PROPERTIES	
Moisture	5-15%
Ash	5-15%
Sulfur (dry)	2.3 - 5.9%
Ash fusion temperature	2000-2500 F
Heating Value (MAF)	Over 13,500 Btu/lb (HHV)

TABLE II

## Oxygen/Nitrogen Generation and Supply

The Air Separation Unit (ASU), supplied by Liquid Air Engineering Co. (LAEC), produces 2060 t/d oxygen at 95% purity as well as high purity nitrogen and dry process air for use in the gasification process. The process involves air compression, purification, cryogenic distillation, oxygen compression, and a nitrogen storage and handling system. After modifications to improve nitrogen production the ASU has reliably supplied products to the gasifier island at specified quantities and quality.

## Gasification and Slag Handling

The two stage Destec gasifier operates with a slagging first stage and an entrained flow second stage. Coal slurry and oxygen are fed to the first stage as well as recycled char from the particulate removal system. This stage operates at 2600 F, producing syngas which exits to the second stage. Molten slag exits the first stage through a taphole and is quenched in a water bath prior to removal through Destec's continuous slag removal system. The second stage of the gasifier uses additional coal slurry and recycled syngas to lower the temperature to 1900 F. Raw syngas exits the gasifier enroute to the syngas cooler.

The gasification and slag handling areas have performed very well thus far. Slag removal has been essentially trouble free since the beginning. The gasifier has consistently processed the coal into high quality syngas.

## Syngas Cooling, Particulate Removal, and COS Hydrolysis

Syngas containing entrained particulates exit the gasifier and is cooled in a firetube heat recovery boiler system, producing 1600 psig saturated steam. Cooled raw gas leaving the boiler passes through a barrier filter unit to remove particulates (char) for recycle to the first stage of the gasifier. The particulate free gas is further cooled prior to entering the COS hydrolysis unit where COS in the raw gas is converted to H<sub>2</sub>S for removal in the Acid Gas Removal system. This area of the

gasification plant has experienced problems which can be summarized into three areas: (1) Ash accumulation at the inlet to the firetube boiler, (2) particulate breakthrough from the barrier filter system, and (3) poisoning of the COS catalyst due to chlorides and trace amounts of arsenic in the syngas.

Ash deposition has not been a major contributor to overall downtime, but has limited runtime somewhat due to ash accumulation at the inlet to the boiler tubes. Improvements have been incorporated to reduce and manage this ash, and more improvements are planned.

Particulate breakthrough has been primarily due to movement and breakage of the ceramic candle filter elements. Substantial downtime is associated with entry into the particulate filter vessels, therefore there has been significant emphasis on improvements to this system. These improvements will be implemented during the third quarter and fourth quarter of 1996.

Poisoning of the COS catalyst due to chlorides and trace arsenic led to early replacement of the catalyst. To address this concern as well as metallurgy concerns with chlorides further downstream in the process, a scrubber system has been installed. The scrubber has satisfactorily resolved these problems.

### **Low Temperature Heat Recovery and Syngas Moisturization**

After exiting the COS hydrolysis unit, low level heat is removed from the syngas in a series of shell-and-tube heat exchangers prior to Acid Gas Removal. This low level heat is used for syngas moisturization, stripping of the acid gases in the Acid Gas Removal system, and preheating condensate. This section of the process has performed well in terms of providing the moisturization for the syngas and providing heat transfer as designed. However, localized chloride stress corrosion cracking to some of these exchangers necessitated replacement with alternate metallurgy. The scrubber mentioned earlier in addition to protecting the COS catalyst, has eliminated metallurgy concerns in this section of the process.

### **Acid Gas Removal and Sulfur Recovery**

The Acid Gas Removal system consists primarily of an  $H_2S$  absorber column and an  $H_2S$  stripper column.  $H_2S$  is removed from the syngas in the absorber using a solvent (MDEA) and the syngas is then routed to the moisturizer column mentioned previously. The  $H_2S$  absorbed is stripped and routed to the Claus process where it is converted to elemental sulfur. The remaining small amount of unconverted  $H_2S$  in the acid gas is compressed for recycle to the gasifier. During process upsets, the spent acid gas is sent to an incinerator, which is one of the permitted air emissions sources. The Acid Gas Removal process has effectively demonstrated removal of over 99% of the sulfur in the syngas. The typical product syngas composition from the plant is shown in Table III.

TYPICAL PRODUCT SYNGAS COMPOSITION	
Component	Volume Percent
Hydrogen (H <sub>2</sub> )	28
Carbon Monoxide (CO)	38
Carbon Dioxide (CO <sub>2</sub> )	10
Methane (CH <sub>4</sub> )	1
Nitrogen (N <sub>2</sub> )	1
Water (H <sub>2</sub> O)	22
Sulfur Compounds	<50 ppmV
Heating Value (dry)	285 Btu/scf (HHV)

TABLE III

### **Environmental Performance**

Total sulfur dioxide emissions from the three permitted emissions points (HRSG stack, gasification flare stack, and tail gas incinerator stack) have demonstrated the ability of the gasification process to successfully operate below 0.2 lbs/MMBtu of coal input. To date, emission rates of less than 0.1 lbs/MMBtu have been attained. This represents a 94% reduction in SO<sub>2</sub> emissions from the decommissioned Unit 1 boiler at Wabash River. The 0.2 lbs/MMBtu is significantly below Acid Rain limits for the year 2000, which are set at 1.2 lbs/MMBtu under the Clean Air Act.

### **Sour Water Treatment**

Sour water is condensed from the syngas in the low temperature heat recovery section of the gasification plant. This water is primarily used for recycle to the slurry preparation plant. The recycled water is stripped of all dissolved gases except ammonia, which remains in the recycled water. Excess water is stripped of all dissolved gases and discharged through a permitted outfall. The sour water treatment system has performed well.

### **Combustion Turbine**

The combustion turbine has operated in excess of (2800) fired hours on syngas and No 2 fuel oil. The turbine has operated in the designed baseload configuration and as a liquid fuel fired combined cycle peak service generator. Both modes of operation have proven to be stable and viable options for the operation of the generator on the bulk power system. The combustion turbine control system (Mark V) has proven, after initial startup tuning, to be reliable and maintainable by on-site PSI technicians. This system does require formal training for the technicians to develop the



necessary skills for long term maintenance. Technicians were trained to maintain Gas Turbine Controls (Mark V), the excitation system (EX2000) and the Gas Turbine cranking system, (LCI). On site control maintenance capability is critical to establishing an available and reliable Gas Turbine.

## **Steam Turbine**

The steam turbine is an early 1950's vintage Westinghouse reheat turbine. The original nameplate for the steam turbine was 99MW, but the repowered rating is 104MW due to the removal of the steam extractions. Throttle pressure has been maintained at the original 1450 psig and throttle temperature is 1005 F. The steam turbine and turbine auxiliaries are located approximately 1600 feet from the gas turbine power block and consequently required extensive piping and drains installations. Although the steam turbine is remotely located with respect to the new power block, the steam turbine operation interface is in the new control room with the new power block controls, Westinghouse WDPF.

Additional modifications were required to the repowered steam turbine as follows. The condensate and feedwater heating extractions were removed and capped. The cold reheat extraction was inspected and maintained for the repowered operation. One row of blading was replaced in the low pressure turbine as a result of the repowering. The generator was rewound and the generator rotor was replaced. A new static excitation system was installed to improve the reliability. The hydraulic turbine controls were replaced with the Westinghouse DEH control system. Existing Turbine Supervisory Instrumentation (TSI) was left in place and remains functional.

The turbine experienced a control shaft failure during the early operation due to an improperly sized cold reheat orifice causing the rotor to thrust, resulting in the failure. Otherwise, the steam turbine has operated very well in the new configuration.

## **Water Treatment**

Water treatment was designed to meet the needs of both the power block and the gasification island. Surface water is drawn from the Wabash River and clarified with a CBI Claricone, filtered then metered to various demands at both operating blocks of the project. Some filtered water is treated in two parallel 480 gpm demineralizers. There is 750,000 gallons of demineralized water storage capability. This water is the supply for the steam cycles of the power block and the gasification island. The control of the water facility is also included in the scope of the Westinghouse WDPF system and can be operated from the central control room. Operation of the water facility has been reliable and cost effective.

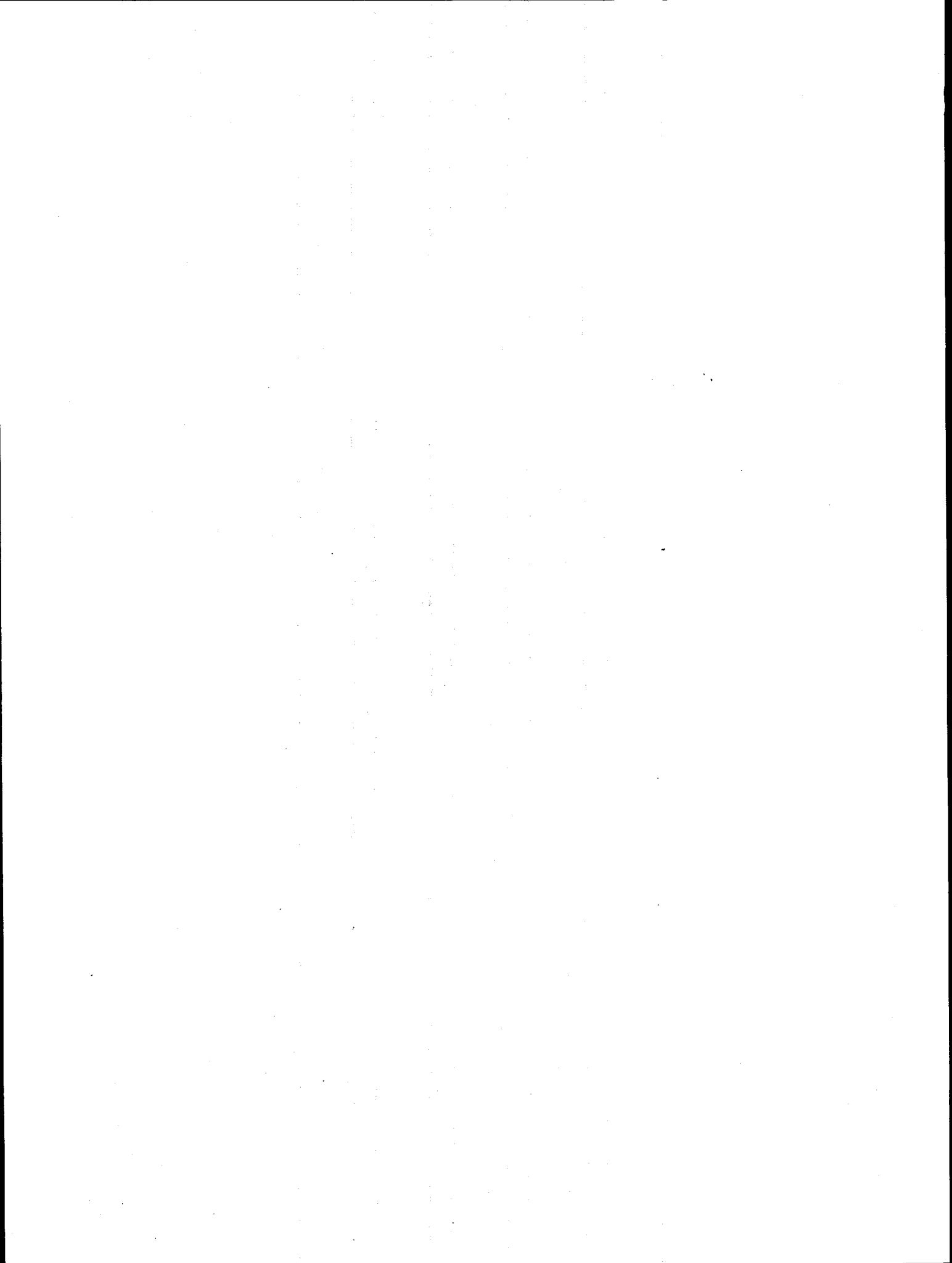
## **OUTLOOK/SUMMARY**

Through the first year of the demonstration period, the Wabash River Coal Gasification Repowering Project has made good strides towards achieving the Project Goals. Both the Gasification and Combined Cycle Plants have demonstrated the ability to run at capacity and within environmental compliance while using locally mined coal. The technology advancements which made this a DOE demonstration project have, for the most part, operated well. Modifications were made to address those problem areas identified through the early operation experience, modifications which have improved plant operation and will further allow demonstration of the Project Goals as the project moves into the second year of the demonstration period.

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# **Tampa Electric Company Polk Power Station IGCC Project Project Status**

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Fifth Annual Clean Coal Technology Conference  
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## **ABSTRACT**

The Tampa Electric Company Polk Power Station is a nominal 250 MW (net) Integrated Gasification Combined Cycle (IGCC) power plant located to the southeast of Tampa, Florida in Polk County, Florida. This project is being partially funded under the Department of Energy's Clean Coal Technology Program pursuant to a Round III award. The Polk Power Station uses oxygen-blown, entrained-flow IGCC technology licensed from Texaco Development Corporation to demonstrate significant reductions of SO<sub>2</sub> and NO<sub>x</sub> emissions when compared to existing and future conventional coal-fired power plants. In addition, this project demonstrates the technical feasibility of commercial scale IGCC and Hot Gas Clean Up (HGCU) technology.

The Polk Power Station achieved "first fire" of the gasification system on schedule in mid-July, 1996. Since that time, significant advances have occurred in the operation of the entire IGCC train. This paper addresses the operating experiences which occurred in the start-up and shakedown phase of the plant. Also, with the plant being declared in commercial operation as of September 30, 1996, the paper will discuss the challenges encountered in the early phases of commercial operation. Finally, the future plans for improving the reliability and efficiency of the Unit in the first quarter of 1997 and beyond, as well as plans for future alternate fuel test burns, are detailed.

The presentation will feature an up-to-the-minute update on actual performance parameters achieved by the Polk Power Station. These parameters include overall Unit capacity, heat rate, and availability. In addition, the current status of the start-up activities for the HGCU portion of the plant will be discussed.

## **INTRODUCTION**

Over the last seven years, Tampa Electric Company has taken the Polk Power Station from a concept to a reality. We have previously reported on the permitting, engineering, construction, contracting and staffing status of the project. We would like to concentrate in this paper on our recent checkout and startup experience and to discuss our operating history to date. We will also review our plans for 1997. In order to view our operations results in the proper perspective, it will be helpful to first briefly discuss some background of the Polk Power Station Project.

## **BACKGROUND**

### **PARTICIPANTS**

Tampa Electric Company (TEC) is an investor-owned electric utility, headquartered in Tampa, Florida. It is the principal, wholly-owned subsidiary of TECO Energy, Inc., an energy related holding company heavily involved in coal mining, transportation, and utilization. TEC has about 3650 MW of generating capacity. Over 97 percent of TEC's power is produced from coal. TEC serves over 500,000 customers in an area of about 2,000 square miles in west-central Florida, primarily in and around Tampa, Florida.

TECO Power Services (TPS) is a subsidiary of TECO Energy, Inc., and an affiliate of TEC. This company was formed in the late 1980's to take advantage of the opportunities in the non-regulated utility generation market. TPS currently owns and operates a 295 MW natural gas-fired combined cycle power plant in Hardee County, Florida. Seminole Electric Cooperative and TEC are purchasing the output of this plant under a twenty-year power sales agreement. In addition, TPS owns and operates a 78 MW plant in Guatemala.

TPS is responsible for the overall project management for the DOE portion of this IGCC project. TPS is also concentrating on commercialization of this IGCC technology as part of the Cooperative Agreement with the U.S. Department of Energy.

The project is partially funded by the U.S. Department of Energy (DOE) under Round III of its Clean Coal Technology Program. Use of a new hot gas clean-up system (HGCU) on a 10% slip stream of syngas will highlight this demonstration of IGCC technology.



## OBJECTIVES

The Polk unit is an integral part of TEC's generation expansion plan. TEC's original objective was to build a coal-based generating unit providing reliable, low-cost electric power. IGCC technology will meet those requirements.

Demonstration of the oxygen-blown entrained-flow IGCC technology is expected to show that such a plant can achieve significant reductions of SO<sub>2</sub> and NO<sub>x</sub> emissions when compared to existing and future conventional coal-fired power plants. In addition, this project is expected to demonstrate the technical feasibility of a commercial scale IGCC and HGCU technology. With the exception of the HGCU, only commercially available equipment has been used for this project. The approach supported by DOE is the highly integrated arrangement of these commercially available pieces of hardware and systems, in a new arrangement which is intended to optimize cycle performance, costs, and marketability at a commercially acceptable size of nominally 250 MW (net). Use of the HGCU will provide additional system efficiencies by demonstrating the technical improvements realized from cleaning syngas at a temperature of about 1000°F rather than utilizing more traditional cold gas clean-up (CGCU) methods: cooling the gas to about 100°F before the sulfur removal process. This low temperature process has the disadvantage of the irreversible cooling losses and associated reheating before admitting the syngas to the combustion turbine (CT).

## SITING

The plant site is a 4300-acre tract about 11 miles west of Fort Meade and 11 miles south of Mulberry in Polk County, Florida. The process through which this site was selected is one of the many success stories of the project.

In late 1989, TEC formed an independent citizen's task force made up of 17 people representing environmental and community leaders, educators, and economists to help guide the site search. Some of the various groups who had members on the task force were: The National Audubon Society, Florida Audubon Society, 1000 Friends of Florida, Sierra Club, The Hillsborough Environmental Coalition, University of South Florida, and others. We made sure that at least half of the group was comprised of members of the environmental community. We knew that protecting the environment would be a very high priority in selecting the plant's technology and site.

The task force conducted a year-long study of more than 35 sites in six counties with the assistance of a professional environmental consulting firm.

The task force ultimately decided - after much debate - that it was better to recommend sites that had already been touched by industry. In their final analysis, they recommended three former phosphate tracts in southwest Polk County. They believed it was best, from both an environmental and economic standpoint, to place previously mined phosphate land back into productive use.

With that recommendation in hand, we began negotiations with the land owners. That is how we came to select the site we have today.

This proactive approach to siting has been very successful for us. We have established strong support for our project and are maintaining a high level of interaction with the community so that we can maintain that support.

We have employed a process of open and regular communications with the local community, our customers, and the media demonstrating that, even in today's environmental climate, we can successfully site and build coal-fired generation.

In a recent survey, three out of four of our customers agreed that we need to build this facility. Two out of three think we made the right decision to use coal. Many of you know that these results are virtually the opposite of current national trends in public opinion. We will continue with our communications-based approach to this project, just as we have with all of our operations within Tampa Electric.

## **CAPITAL COST**

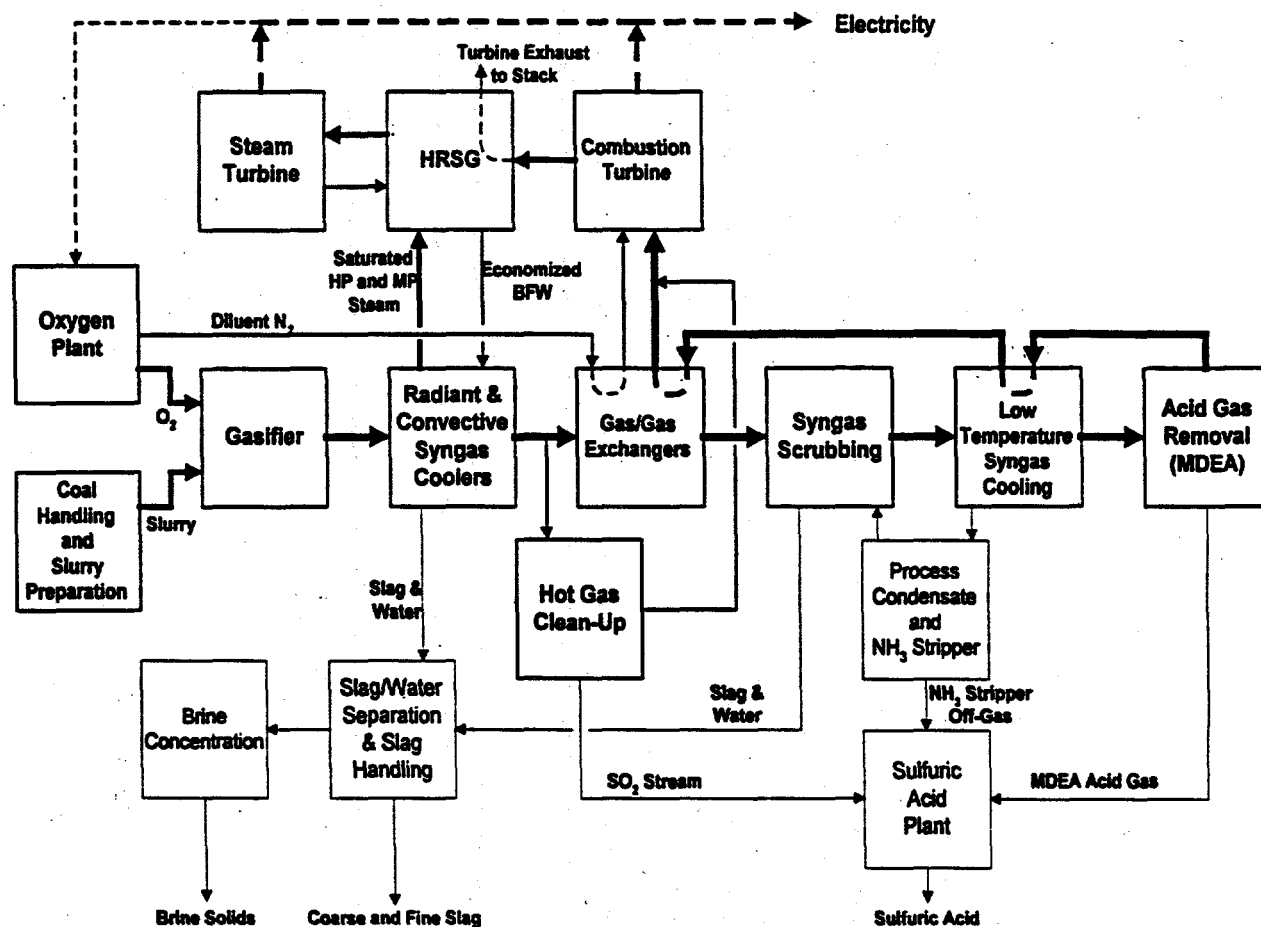
The total project capital cost was approximately \$510 million, which includes DOE's \$122 million cost share. At about \$2,000 / kW, this seems high in comparison to the commercial offerings of other technologies. However, we must consider three mitigating factors:

- Polk Power is a first-of-a-kind design. The next similar plant should be able to build on Polk's experience base to significantly reduce costs in several areas.
- Polk's capital costs include expenses for development and reclamation of the entire 4300 acre site up to its permitted capacity of 1150 MW. The Polk site should satisfy TEC's plant site needs for the next 10 to 20 years.
- Polk has two parallel gas clean-up systems.
- Polk Power is a very clean plant utilizing our most abundant indigenous fuel resource, coal.

Considering these factors, we expect the next generation of IGCC plant to cost between \$1200 and \$1500 when compared on a consistent basis to other technologies. Given the trend in environmental costs for new plants and the likely long term cost and availability of coal, IGCC appears quite attractive.

## TECHNICAL DESCRIPTION

A general flow diagram of the entire process is shown in Figure 1.



**FIGURE 1**  
**Polk Unit #1 IGCC Block Flow Diagram**

This unit utilizes commercially available oxygen-blown entrained-flow coal gasification (CG) technology licensed by Texaco Development Corporation. In this arrangement, coal is ground to specification and slurried in water to the desired concentration (60-70 percent solids) in rod mills. The unit is designed to utilize about 2200 tons per day of coal (dry basis). This coal slurry and an oxidant (95 percent pure oxygen) are then mixed in the gasifier feed injector. This produces syngas with a heat content of about 250 BTU/SCF (LHV). The oxygen is produced in an air separation unit (ASU). The gasifier is designed to achieve greater than 95 percent carbon conversion in a single pass. The gasifier is a single vessel feeding into one radiant syngas cooler which was designed to reduce the gas temperature to 1400°F.

After the radiant cooler, the gas is split into two (2) parallel convective coolers, where the temperature is further reduced to less than 900°F. A 10% slip stream goes to the HGCU system and the remainder is processed in a traditional CGCU system.

The CGCU system is a traditional amine scrubber type. Sulfur removed in the HGCU and CGCU systems is recovered in the form of sulfuric acid. This product has a ready market in the phosphate industry in the central Florida area. It is expected that the annual production of 45,000 tons of sulfuric acid produced by this 250 MW (net) IGCC unit will have minimal impact on the price and availability of sulfuric acid in the phosphate industry.

Most of the ungasified material in the coal exits the bottom of the radiant syngas cooler into the slag lockhopper where it is mixed with water. These solids generally consist of slag and uncombusted coal products. As they exit the slag lockhopper, these non-leachable products are saleable for blasting grit, roofing tiles, and construction building products. TEC has been marketing slag from its existing units for such uses for over 25 years.

The water in the slag lockhoppers requires treatment before it can be reused. All of the water from the gasification process is cleaned and recycled, thereby creating no requirement for discharging process water from the gasification system.

The ASU uses ambient air to produce oxygen for use in the gasification system and sulfur recovery unit, and nitrogen which is sent to the advanced CT. The addition of nitrogen in the CT combustion chamber has dual benefits. First, this additional mass flow has the advantage of producing higher CT power output. Second, the nitrogen acts to control potential  $\text{NO}_x$  emissions by reducing the combustor flame temperature which, in turn, reduces the formation of thermal  $\text{NO}_x$  in the fuel combustion process.

The ASU is sized to produce about 2100 tons per day of 95 percent pure oxygen and 6300 tons per day of nitrogen. The ASU was provided by Air Products.

The HGCU system is being developed by General Electric Environmental Services, Inc. (GEESI). Instead of having to cool the gas prior to sulfur removal, the HGCU will accept gas at 900-1000°F. The successful demonstration of this technology will provide for higher efficiency IGCC systems.

A regeneration system for the HGCU will produce a concentrated (about 13 percent)  $\text{SO}_2$  stream. This will feed a sulfuric acid plant, for production of a saleable acid byproduct.

Other support processes will also be tested in conjunction with HGCU:

- In addition to the high efficiency cyclones upstream of the HGCU system, a high temperature barrier filter is installed downstream of the HGCU to protect the CT.
- Sodium bicarbonate,  $\text{NaHCO}_3$ , will be used upstream of the HGCU sorbent bed for removal of chloride and fluoride species. The resulting stable solids sodium chloride and sodium fluoride will be disposed of with other plant solid byproduct streams.

The key components of the combined cycle are the advanced combustion turbine (CT), heat recovery steam generator (HRSG), steam turbine (ST), and electric generators. The power block is provided by General Electric.

The HRSG is installed in the CT exhaust to complete the traditional combined cycle arrangement and provide steam to the 130 MW ST. No auxiliary firing is done in the HRSG system. Hot exhaust from the CT is channeled through the HRSG to recover the CT exhaust heat energy. The HRSG high pressure steam production is augmented by high pressure steam production from the coal gasification plant. All high pressure steam is superheated in the HRSG before delivery to the high pressure ST.

The ST is a double-flow reheat turbine with low pressure crossover extraction. The ST and associated generator are designed specifically for highly efficient combined cycle operation with nominal turbine inlet throttle steam conditions of approximately 1450 psig and 1000°F with 1000°F reheat inlet temperature.

The operation of the combined cycle power plant is coordinated and integrated with the operation of the CG process plant. The initial startup of the power plant is carried out on low sulfur No. 2 fuel oil. Transfer to syngas occurs upon establishment of fuel production from the CG plant.

Under normal operation, syngas and nitrogen from the ASU are provided to the CT. The syngas/nitrogen mix at the CT combustion chamber is regulated by the CT control system to control the NO<sub>x</sub> emission levels from the unit.

Cold reheat steam from the high pressure turbine exhaust and HRSG intermediate pressure steam are combined before reheating in the HRSG and subsequent admission to the intermediate pressure ST. Some intermediate pressure steam is also supplied to the HRSG from the sulfur recovery unit.

The heart of the overall project is the integration of the various pieces of hardware and systems. Maximum usage of heat and process flow streams can increase overall cycle effectiveness and efficiency. In our arrangement, benefits are derived from using the experience of other IGCC projects, such as the Cool Water Coal Gasification Program, to optimize the flows from different subsystems. For example, low pressure steam from the HRSG is produced to supply heat to the CG facilities for process use. The HRSG also receives steam energy from the CG syngas coolers to supplement the steam cycle power output. Additional low energy integration occurs between the HRSG and the CG plant. Condensate from the ST condenser is returned to the HRSG/integral deaerator by way of the CG facilities, where some condensate preheating occurs by recovering low level heat.

Probably the most novel integration concept in this project is our use of the ASU. This system provides oxygen to the gasifier in the traditional arrangement, while simultaneously using what is normally excess or wasted nitrogen to increase power output and improve cycle efficiency and also lower NO<sub>x</sub> formation.

The primary source of emissions from the IGCC unit is combustion of syngas in the advanced CT (GE 7F). The exhaust gas from the CT leaves the system via the HRSG stack. Emissions from the HRSG stack are primarily NO<sub>x</sub> and SO<sub>2</sub> with lesser quantities of CO, VOC, and particulate matter (PM). The

CGCU and HGCU systems are designed to remove at least 96 percent of the sulfur present in the coal. The emission control capabilities of the HGCU system are yet to be fully demonstrated. Therefore, some emission estimates are higher compared to estimated emissions from the CGCU system. After the completion of the two-year, phase-one demonstration period, the lower emission rates from the CGCU system must be achieved to meet permit requirements.

The advanced CT in the IGCC unit uses nitrogen addition to control  $\text{NO}_x$  emissions during syngas firing. Nitrogen acts as a diluent to lower peak flame temperatures and reduce  $\text{NO}_x$  formation without the water consumption and treatment/disposal requirements associated with water or steam injection  $\text{NO}_x$  control methods. Maximum nitrogen diluent is injected to minimize  $\text{NO}_x$  exhaust concentrations consistent with safe and stable operation of the CT. Water injection is employed to control  $\text{NO}_x$  emissions when backup distillate fuel oil is used.

Part of our cooperative agreement with DOE is a four-year demonstration phase. During the first two years of this period, it is planned that four different types of coals will be tested in the operating IGCC power plant. The results of these tests will compare this unit's efficiency, operability and costs, and report on each of these test coals specified against the design basis coal, Pittsburgh #8. These results should provide a menu of operating parameters and costs which can be used by utilities in the future as they make their selection on methods for satisfying their generation needs, in compliance with environmental regulations.

## EARLY OPERATING HISTORY

### Third Quarter, 1996: Start-up and Commissioning:

First syngas was produced on July 19, 1996. The first gasifier run lasted 21.5 hours which set the longevity record for first fire on a solid fuel Texaco gasifier. Ten gasifier runs totaling 174 hours were completed during the third quarter. All plant systems had been successfully commissioned by the end of this period, so Polk Power Station Unit #1 was placed in commercial operation at the end of the third quarter on September 30, 1996. The major accomplishments and shutdown causes of the first ten runs are summarized in Table 1.

**TABLE 1**  
**Gasifier Runs, Major Accomplishments, Shutdown Causes**  
**Commissioning Phase (Third Quarter, 1996)**

Run Number	Duration (Hours)	Major Accomplishments	Shutdown Cause
1	21.5	First Syngas	Process Water Plugging - Clarifier
2	5.6		O <sub>2</sub> Flow Set Point Entry
3	29.5	Lined Out Process Water System	Loss of BFW from Power Block
4	10.3	First Time Through Low Temperature Gas Cooling	MDEA Foaming and Carry-Over
5	4.1		Raw Gas Flare Valve I/P Failure
6	3.7		False Indication of Cooling Water Loss
7	6.7		Lockhopper Problems
8	67.3	First Steady MDEA Operation	Process Water and Convective SGC Plugging
9	2.4		HP BFW Valve Failure
10	22.4	100% Gasifier Load, First Syngas to CT, First H <sub>2</sub> SO <sub>4</sub> , First Brine Crystals	CT Fuel Oil Leak, Convective SGC Plugging

#### Fourth Quarter, 1996: Initial Commercial Operation:

Ten gasifier runs totaling 701 hours were made in October and November, 1996, prior to a planned outage which began December 5 for routine maintenance, inspections, and some minor improvements. In the 30 days preceding the outage, the gasifier was on-line 67% of the time and the gas turbine was on 100% syngas fuel 59% of the time. This was a major accomplishment which exceeded our target expectations for this period. The longest continuous gasifier run was 7.5 days, and the combustion turbine was on syngas fuel continuously for 7.3 days during this run. The last four gasifier runs were shut down by transmission system voltage swings external to the plant. The protections systems have been reconfigured so even minor external disturbances such as these will no longer trip the unit. These runs are summarized in Table 2.

**TABLE 2**  
**Gasifier Runs, Shutdown Causes**  
**Early Commercial Operation (October and November, 1996)**

Run Number	Duration (Hours)	Turbine On Syngas (Hours)	Shutdown Cause
11	31.2	2.8	Convective SGC Plugging
12	101.6	16.8	Combustion Turbine Vibration-Rotor Bolt
13	81.9	51.4	Steam Turbine Trip (Excitation) Caused BFW Loss
14	4.5	0.0	Lockhopper Problems
15	54.2	40.4	Steam Turbine Trip (Excitation) Caused SGC Drum Upset
16	17.9	0.0	Main Air Compressor Trip - Execution of DCS Change
17	153.8	149.0	Oxygen Compressor Trip - Transmission System Voltage Swing
18	4.5	0.0	Slurry Feed Pump Trip - Transmission System Voltage Swing
19	71.0	64.6	Main Air Compressor Trip - Transmission System Voltage Swing
20	180.1	174.7	Slurry Feed Pump Trip - Transmission System Voltage Swing

Specific operational experiences and challenges during the commissioning and initial commercial operational phases are detailed below.



## **AIR SEPARATION**

The oxygen plant has operated essentially trouble free through both the commissioning and initial commercial operational phases. Early high vibration of the main air compressor motor has been reduced to normal levels. Three recent gasifier trips have resulted from oxygen plant trips due to problems external to the oxygen plant itself. Polk Power Station does not have a backup liquid oxygen supply system which could have saved these gasifier runs. When backup systems were being evaluated in the design phase, their cost could not be justified based on the expected incremental availability they were expected to provide. This is probably still the case, but TEC will continue to monitor the frequency of ASU plant trips.

The process performance of the oxygen plant has been exceptionally good. It comfortably met its rated production under hot ambient conditions with all product purities better than design and with capacity still available on the columns, exchangers, and compressors. Power consumption appears to be generally consistent with our expectations, but because of the number of variables involved, it must still be checked during a detailed performance test.

The advanced controls handle minor perturbations around steady state well, and we now always operate with them engaged. They adjust the feed air flow and internal flows based on the demand for the various products.

## **SLURRY PREPARATION**

Slurry preparation performed extremely well during the commissioning phase. For three months, we produced stable, pumpable slurries up to 64% concentration without the use of additives with virtually no operational problems. Early high vibration of the rod mills was quickly eliminated by reinforcing the foundations. However, some severe problems did develop beginning early in the fourth quarter of 1996. Specifically, we observed:

- Settling and partial plugging in many horizontal piping runs (reduced pumping capacity and caused instrumentation problems)
- Severe liner wear on the slurry transfer pumps (reduced pumping capacity)
- Overflowing of the slurry screens (operational problems)
- pH swings in the product slurry (corrosion of tanks and piping)
- Failure of the purge water filter (operational problems)

Factors which may have contributed to some or all of these problems are as follows:

- Variations in feed coal properties have been observed. Distinct property variations in the off-site coal pile have been documented, apparently due to weathering and/or aging; and aging in the on-site coal storage silos is also likely. This seems to be linked to the pH swings.
- The installed slurry screens are finer than Texaco had specified and the pump manufacturer required. This contributes to the problem of overflowing screens.
- A low dose rate of viscosity reducing additive has been used occasionally. This temporarily facilitates pumping, but may contribute to the line plugging in the long term.
- Rod loading in the mills has been adjusted several times to try to fine tune the particle size distribution.

We are addressing these problems through a series of steps such as installing appropriately sized slurry screens, restoring the initial rod loading of the rod mills, and more carefully monitoring and controlling the slurry pH with ammonia injection. Some additional modifications may be required once these easier changes are completed.

## **GASIFIER**

The gasifier itself is quite simple and it has performed reliably throughout the commissioning and early commercial operational phases.

The gasifier safety system performance has been excellent to date. We have had no nuisance shutdowns - all automatic shutdowns have been the result of problems in other parts of the plant which properly tripped the unit. The gasifier feed controls have also been excellent. These adjust the overall gasifier load as well as the ratio of oxygen to slurry to control gasifier temperature.

Thermocouple life had been a problem in the commissioning phase. However, early in the operational phase, we began running at lower temperatures which prolonged thermocouple life. Also, the on-line analyzers were proven sufficiently reliable and, in parallel, useful correlations between gasifier temperature and the syngas composition were developed. Consequently, although thermocouples are still necessary at times and they must still be replaced more often than we would prefer, concern and expense in this area has been significantly reduced. Additional development work is underway to further increase thermocouple life and reduce cost.

During the commissioning phase, we observed the performance of the gasifier at various temperatures, loads, and slurry concentrations. Some minor feed injector design changes were made as a result. We believe the operating conditions are now near optimum for this feed injector design and refractory liner.

The following Table 3 shows that some aspects of the gasifier's performance at current operating conditions that do not yet meet "Design" or "Commercially Expected" values.

**TABLE 3**  
**Slag Characteristics and Refractory Liner Life**

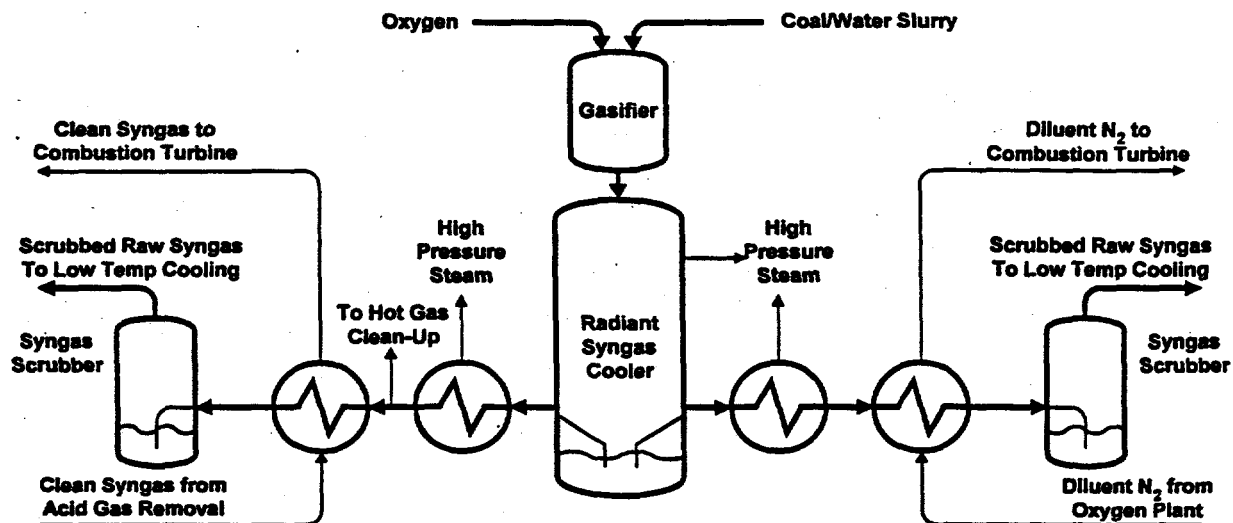
	Current Full Load Operation	Design or Commercial Expectation
Slag Carbon Content (Weight % Dry Basis)	34	14 to 28
Slag Quantity (Dry Tons/Day)	250	185 to 215
Heating Value Lost To Slag (MMBTU/Hr HHV)	70	20 to 50
Refractory Liner Life (Years at 85% On-Stream Factor)	½	2

Carbon conversion can be increased at the expense of refractory liner life, and vice-versa, by adjusting gasifier temperature. However, as can be seen from the table, there is little available to sacrifice on either parameter. The higher than expected carbon content of the slag creates handling problems and makes it a less desirable byproduct for many applications. It also increases the mass and volume of the material we must handle. Furthermore, the heating value of the carbon lost with the slag increases net plant heat rate by 75 to 200 BTU/KWH. The current "startup" gasifier refractory liner is less expensive with reduced slag resistance compared to the material we expect to use long-term. Our first liner replacement is scheduled for the spring of 1997. It will be a more slag-resistant material, so at current operating conditions, it may approach our commercial expectations of a 2 year liner life. However, some additional feed injector adjustments to improve carbon conversion at less severe reactor conditions are still required for us to realize our commercial expectations for liner life, heat rate, and slag quantity/quality. Texaco has an excellent team on-site and at other Texaco engineering and development centers working with us on these issues.

## HIGH TEMPERATURE SYNGAS COOLING

High Temperature Syngas Cooling consists of a Radiant Syngas Cooler (RSC) followed by Convective Syngas Coolers (CSC).

**FIGURE 2**  
**Syngas Cooler System**



Raw syngas from the gasifier first passes downward through an RSC where high pressure steam is generated. The CSC System consists of two wings, each of which handles 50% of the RSC outlet gas. Each wing consists of a fire-tube convective heat exchanger producing high pressure steam, followed by a two stage gas/gas heat exchanger where the raw syngas (tube side) heats either the clean syngas or diluent nitrogen to the combustion turbine.

The RSC and its associated steam systems have been trouble-free through both the commissioning and early operational phases. Fouling factors have been only  $\frac{1}{3}$  of the design value with Pittsburgh #8 coal, so no soot-blowing has been required. Fouling has been so low that we may need to elevate the RSC outlet temperature by covering part of the RSC surface with insulating refractory to meet the HGU minimum inlet temperature requirement. Soot blowing may be required as we achieve longer run times and gasify other coals, but all indications are that we will have no difficulty achieving target heat transfer. There have been no hints of plugging in the RSC.

As with the RSC, heat transfer in the CSC exchangers has recently been excellent. Fouling factors have been 30% or less of design values where we could measure them.

One of the greater challenges during the commissioning phase was pluggage within the CSC system. Many of the gas/gas exchanger tubes plugged with ash deposits during several of the commissioning phase runs. This increased the pressure drop above the allowable level, so it was necessary to cool,

open, enter, and clean this equipment often. The deposits absorbed moisture during this downtime, some from the ambient air and some from other sources. This produced rapid downtime corrosion. Pits penetrated through up to 60% of the tube wall thickness in some places. Fortunately, very early in the operating phase, we learned how to eliminate this plugging by controlling temperatures and velocities in the equipment. Also, we have been more careful in our shutdown and startup practices to minimize conditions leading to downtime corrosion. The inspection during the December, 1996, planned outage revealed no plugged tubes and no increased corrosion.

## **LOW TEMPERATURE GAS COOLING (LTGC)**

Immediately downstream of the CSC's are the Syngas Scrubbers where particulates and chloride are removed from the raw syngas in a water wash. The raw syngas is water saturated as it leaves the scrubbers at about 300°F. The LTGC system cools the syngas to near ambient temperature for the acid gas removal system. As the gas cools, most of the water vapor condenses and becomes what is referred to as process condensate. The LTGC system consists of three partially condensing heat exchangers and associated knock-out drums, the process condensate return system, and an ammonia stripper to rid the system of the ammonia which condenses from the syngas with the process condensate.

The system has generally performed well to date. Some minor modifications were made to accommodate the somewhat different than expected flow rates of process condensate from some of the exchangers. The greatest difficulties have been in the ammonia stripper overhead piping. Ammonia combines with carbon dioxide to form solid salts which plug the piping if the temperature falls below about 160°F. Heat tracing was inadequate in some line segments and it was completely overlooked in others. Furthermore, the piping and control valves were inadequately sized, and this has prevented us from feeding this entire stream to the Sulfuric Acid Plant where the ammonia is to be converted to nitrogen and water vapor. These problems have been corrected and we expect no further difficulty with the LTGC system.

## **ACID GAS REMOVAL**

A tertiary amine (MDEA) system is being used in the Polk plant for removing hydrogen sulfide ( $H_2S$ ) from the raw syngas in the cold gas clean-up (CGCU) system.

We experienced a significant amount of foaming when we first introduced syngas to the MDEA absorber during Gasifier Run #4 early in the Commissioning Phase. Foaming is a known problem with all amine based acid gas removal systems. We quickly brought this foaming under control with filtration and anti-foam agents and have experienced no foaming during subsequent runs. However, some amine contamination of other plant systems persists, probably through a slight amount of carry-over with the clean syngas during startup. This amine finds its way into the grey water system, and ultimately into the brine concentration unit where it causes foaming in the falling film evaporator. This foaming must also be controlled with anti-foam agents.

Tuning of the MDEA system operation continued through the remainder of the commissioning phase, and clean gas within Polk's environmental requirements was consistently being produced by the beginning of the fourth quarter of 1996. 95% overall sulfur removal is achieved. The MDEA now routinely removes 99% or more of the  $H_2S$ . The remainder of the sulfur emissions are derived from carbonyl sulfide (COS), a compound which our plant configuration and MDEA solvent are not designed to remove. The gasifier produces more COS than was expected, and we are hoping the high COS production rate observed to date is peculiar to the Pittsburgh #8 coal we are now running. If it is not, we may have to change solvents, adjust operating conditions, and possibly make other modifications to run higher sulfur coals within our current permit limits.

While the MDEA does remove virtually all the  $H_2S$ , it typically only removes about 12% of the carbon dioxide ( $CO_2$ ) from the syngas. The plant design assumed 20% of the  $CO_2$  would be removed. This extra  $CO_2$  in the syngas improves overall plant efficiency by increasing the "free" mass flow to the turbine and reducing the steam required to regenerate the solvent.

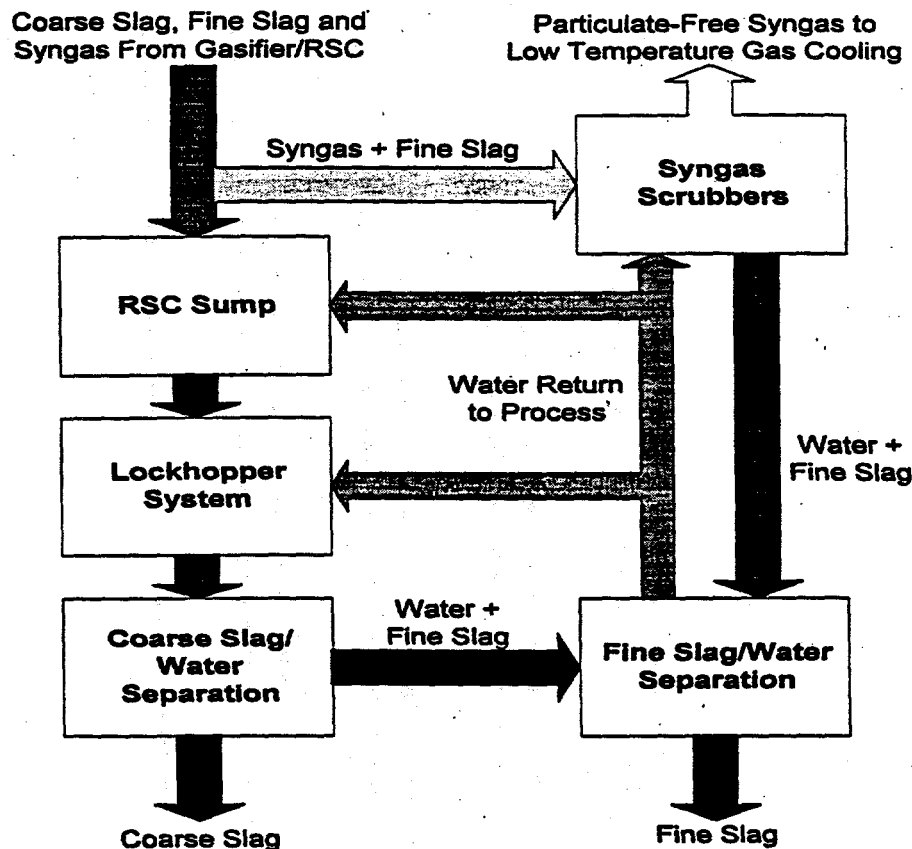
A steady rise in the concentration of degradation products has occurred in the MDEA solvent but not at an unexpected rate. A water wash column is installed upstream of the absorber to remove trace compounds to minimize formation of these MDEA degradation products. We have not yet built sufficient operating history to evaluate its effectiveness.

## **SULFURIC ACID PLANT**

The Acid Gas Removal system produces the main feed stream for the Sulfuric Acid Plant, an acid gas stream consisting of 20% to 30%  $H_2S$  and most of the remainder  $CO_2$ . The other main feed stream is the Ammonia Stripper off-gas. The Sulfuric Acid Plant has performed very well once steady, efficient operation of the Acid Gas Removal system was achieved early in the fourth quarter. The plant has tripped four times due to pressure fluctuations of the feed streams. These were not related to the Acid Plant itself.

The Pittsburgh #8 coal we are currently gasifying has a sulfur content of less than 2.5%, compared to a design concentration of 3.5%. As a result, 1 to 2 MMBTU/Hr of supplemental fuel is sometimes required as expected to maintain temperature in the catalytic reactors.

## SLAG HANDLING, FINES REMOVAL, AND PROCESS WATER SYSTEMS



**FIGURE 3**  
**Process Water Systems, Fine and Coarse Slag Handling**

Coarse slag and some of the fine slag from the gasifier falls through the RSC into a water pool at the bottom. This pool is referred to as the RSC Sump. From there, the slag is removed via a lockhopper system which cycles approximately twice per hour. With each cycle, the water and slag mixture from the lockhopper dumps into a concrete holding area where it is separated (Coarse Slag/Water Separation). The coarse slag is hauled to the slag holding area. The water, containing some fine slag, is pumped to the Fine Slag/Water Separation System.

The fine slag which does not fall into the RSC sump passes through the CSC system with the syngas and is removed in the syngas scrubbers. Fine slag and water are continuously blown down from the scrubbers. This stream is also routed to the Fine Slag/Water Separation system.

Fine Slag/Water Separation consists first of a settler where the fines are concentrated. The fines in the settler bottoms are then removed in a rotary drum vacuum filter and are also hauled to the slag holding area. The water is returned to the process.

The fines removal system has performed beyond expectations. It typically handles much more water and fines than design. During the commissioning phase, upsets of the settler did occur due to excessive traffic and/or loss of polymer feeds. This led to solids carryover from the gravity settler, resulting in plugged process piping. These problems have been largely eliminated in early commercial operation with operating experience.

Likewise, the lockhopper, RSC sump, and syngas scrubbers also experienced some plugging in the Commissioning Phase during periods of excessive solids traffic, but these problems also have been resolved with experience and some minor piping modifications. Erosion has been encountered in some control stations during early commercial operation. This was not unexpected, and it is being addressed with materials and configuration changes.

The Coarse Slag/Water Separation system has been a challenge. The water was expected to easily separate from the slag in the concrete holding area after each lockhopper dump. However, the fine slag stayed in suspension. These fines plugged the local sump and increased the loading on the gravity settler. Barrier walls were installed in the slag holding area and the water is now pumped off in batches after settling. This added settling time greatly reduces the fine slag in the water. The system is now operable, but still very labor intensive. Significant configuration changes are being considered for the long term.

## **BRINE CONCENTRATION**

The Polk Power station is permitted as a zero process water discharge facility requiring that all of the process water is recovered and reused. Through recycling, the chlorides removed from the syngas in the Syngas Scrubber would build to unacceptably high levels for affordable metallurgy. Therefore, a brine concentration system was incorporated into the plant design. It consists of a falling film evaporator, followed by a forced circulation evaporator feeding a crystallization and centrifuge separation step.

During the third quarter, the falling film unit was commissioned with excellent results. As previously mentioned, foaming problems resulting from amine in the feed has resulted in the greatest operational problems such as sump level control and carryover. Anti-foam agents have been effective, and a permanent anti-foaming injection system is being pursued.

The forced circulation evaporator has been the greatest challenge in operating the brine concentration unit in the early commercial phase. Corrosion has been excessive. Using corrosion coupon tests, coupled with laboratory tests, the corrosion mechanism is being understood which will lead to metallurgical and process modifications in this system.

Control of the centrifuge has been difficult, resulting in crystals of variable quality. We believe this is due in large part to erratic flows causing periodic line pluggage. The control scheme is being modified.



## COMBINED CYCLE

The key components of the combined cycle are the advanced combustion turbine (CT), heat recovery steam generator (HRSG), steam turbine (ST), and electric generators. The combined cycle power plant was provided by General Electric.

The CT is a modified Frame 7F capable of producing 150 MW (gross) from #2 fuel oil (the startup and backup fuel) and 192 MW (gross) from syngas fuel. When firing syngas fuel, nitrogen from the ASU provides both NO<sub>x</sub> abatement and power augmentation.

Hot exhaust from the CT is channeled through the HRSG to recover energy. The HRSG performs most of the plant's economizing and all of the superheating, while most of the high pressure steam is generated in the syngas coolers when the gasifier is on line. The HRSG also produces much of the low pressure steam consumed by the gasification plant. Consequently, Polk's HRSG contains significantly more superheater, economizer, and low pressure evaporator surface compared to HRSGs in conventional combined cycles.

The 130 MW ST is a double-flow reheat turbine. Nominal turbine inlet steam conditions are 1450 psig and 1000°F with 1000°F reheat temperature. Low pressure extraction provides the remainder of the low pressure steam for the gasification plant.

The combined cycle was commissioned on May 4, 1996. Ever since, it has been dispatched as a normal Tampa Electric generation resource. It has produced approximately 150,000 megawatt hours on distillate fuel and 200,000 megawatt hours on syngas fuel through the end of 1996.

The combustion turbine was first operated on 100 percent syngas fuel for 4.1 hours during Gasifier Run #10 in mid September. It reached a maximum load of 161 MW on syngas, generating 520 megawatt hours over this period. Combined cycle output reached 210 MW. However, this first period of operation on syngas revealed a design problem with the fuel nozzles which led to some local overheating. The combined cycle was out of service for the remainder of September for repairs and modifications. This problem has not recurred.

A brief period of operation on syngas fuel occurred during a short gasifier run on October 1. During the next gasifier run, Run 12, the CT reached full syngas load (192 MW gross) on October 13. This run was highly successful, but it did identify two additional problems:

- 1) Performance data during this run showed that the diluent nitrogen control valve was undersized for the design flow. Diluent N<sub>2</sub> is used for NO<sub>x</sub> abatement, and sufficient N<sub>2</sub> could only be provided to keep NO<sub>x</sub> emissions within permit limits with a CT output of 185 MW (gross). A larger valve is due in February, 1997.
- 2) GE observed high CT vibration on October 16. Their on-line diagnostics showed this was caused by a crack in a large turbine rotor bolt. GE replaced all these bolts in the subsequent 11 day outage. The CT has had no further high vibration problems.

ST excitation system failures caused ST trips on October 31 and November 6. These led to gasifier trips due to an incorrect valve lineup in the Hot Gas Cleanup System which had not yet been commissioned. These trips clearly demonstrate some of the drawbacks of integration: problems in one process unit can create even greater problems in another. The valve lineup was quickly corrected once it was found, and subsequent ST trips have not caused gasifier trips. We believe we have also finally found and corrected the cause of ST excitation system failures.

The best combined cycle performance prior to the December planned outage occurred during gasifier Run 20 from November 26 to December 4. The gasifier was on line continuously for 180 hours and the CT was on syngas fuel continuously for 175 hours. The average gross power production for the entire period was 300 MW, 184 MW from the CT and 116 MW from the ST.

At the time of printing of this paper, Gasifier Run #22 was still in progress. Through January 5, the gasifier had been on-line continuously for 306 hours and the CT had been on syngas fuel for a total of 296 hours and continuously for 267 hours.

Starting reliability of the CT on distillate fuel has been good, but fuel transfers to syngas fuel have been inconsistent. The first attempts to transfer the combustion turbine to syngas fuel in August were unsuccessful. Corrections were made, and the next attempt in mid-September went smoothly. But new problems appeared and there were 3 failed transfer attempts during gasifier Run 16 in early November. The CT never successfully transferred to syngas fuel in 18 hours of gasifier operation during that run. However, in mid and late November, transfers were smooth and routine. The purge system and the CT control system were modified in the December outage. Despite the changes, fuel transfers again were problematic in late December. We are hoping for a speedy resolution once all purge system modifications are completed and the controls are retuned.

## CONTROL SYSTEM

The plant's main control system is a Bailey Infi-90 Distributed Control System (DCS). The DCS communicates directly with 3 other plant control systems: the CT GE Mark V, the ST GE Mark V, and the Triconex Gasifier Safety System. There are about 7200 direct Input/Output variables. Over 500 process control graphics available on any of 14 CRT screens provide the operator interface.

The DCS has performed well. No gasifier or plant trips have been caused by DCS module or I/O failures. The overall DCS availability in fourth quarter of 1996 was 100.0 %.

Two systems associated with the DCS have also been highly successful: 1) the data storage, and 2) retrieval system and the operator training simulator. It is not an exaggeration to say that the Polk plant would not be running as well as it is today without these systems.

- Data storage and retrieval is done by a product called Plant Information Systems (PI) from Oil Systems Inc. Data storage has been almost 100% reliable, and retrieval is easy in several different formats (graphs, tables, spreadsheets).

- The operator training simulator was furnished by Bailey and TRAX, Inc. A copy of the actual plant control system (DCS and Triconex hardware and software) interacts with process plant models running on seven PC's. This simulator enabled plant personnel to become familiar with plant operation before startup and correct control system and procedural errors before they occurred in the real plant.

Although the DCS has performed well, the required level of technical support has been higher than expected to achieve these results. A full-time team of seven with some supplemental help worked throughout most of 1996 to address the following issues:

- DCS module infant mortality was fairly high in the Commissioning Phase, but failure rates have declined dramatically. All failed modules were replaced under warranty.
- Initially there were over 8000 possible alarms, and at times during the Commissioning Phase over 1000 of these were simultaneously active. Such information overload causes alarms to be ignored. A separate "alarm team", formed late in the Commissioning Phase, reduced the number of alarms to about 4000. Further reduction in the number of possible alarms and prioritization of the remaining alarms is still in progress.
- Conveying information which can be quickly and easily interpreted for split-second decision making is always a challenge. To meet this challenge, it has been necessary to improve plant diagnostics by adding more "first out" indications, dedicated displays, and ready lists. Graphic displays have also been modified to be more concise and easily readable. These efforts will undoubtedly continue into the foreseeable future.
- The data links between the DCS and both CT and ST Mark V control systems have been troublesome. Making changes is particularly hard. (In contrast, the data link between the DCS and the Triconex Gasifier Safety System has worked very well.) Also, working on the Mark V and GE's user interface is difficult. We must still rely more heavily on GE than we would prefer at this stage of operation. It would have been preferable to have done as many of the turbine control functions as possible directly in the DCS.
- Almost all logic and configuration errors have been eliminated, initial tuning has been done on all control loops, and some optimization has been done. However, initial operation and tuning efforts have shown that new or modified control logic will be necessary for several plant areas such as:
  - Overall plant load control
  - Combustion Turbine fuel transfers,
  - pH control in water treatment,
  - Grey Water inventory control
  - Centrifuge control in Brine Concentration.

## PLANS FOR FUTURE OPERATION

During the fourth quarter of 1996, most of our efforts were geared toward keeping the unit on line as much as possible to obtain operating experience. We are now moving into a period where we will look to improve performance and "fine tune" plant operation. Most of our efforts for 1997 will be focused primarily in four areas:

- Equipment and/or operational modifications to increase availability and reduce operating costs,
- Operational modifications to improve overall Unit heat rate,
- Continued start-up efforts for the Hot Gas Clean-Up (HGCU) system, and
- Preparation for alternate fuel test burns (DOE demonstration test burns).

## INCREASE AVAILABILITY AND REDUCE OPERATING COSTS

Previous sections of this paper have touched on steps that Polk is taking to reduce some operating costs, e.g., by improving the slurry preparation and slag/water separation areas which are currently labor-intensive, and reducing corrosion in the brine concentration area. However, even greater reductions in Polk's operating costs can be realized by increasing IGCC availability. High availability is an important attribute of any type of power plant. For IGCC plants, high availability is even more important than for most since there is a stronger link between high IGCC availability and low operating costs for the reasons discussed below. Polk's IGCC availability has recently exceeded our expectations for this period, but the plant has still not reached our projections for "mature" operation and there is much room for improvement.

Reduced IGCC availability usually occurs under two conditions:

- The gasifier is off-line
- The gasifier is on-line but the CT is not on syngas fuel

Whenever the gasifier is off-line, a considerable amount of energy must be consumed for heating the gasifier, heat maintenance in other parts of the plant (steam system, Sulfuric Acid Plant) and starting or maintaining operation of the ASU. The time to return the gasifier to service after a trip can be significant. This is primarily due to various heat up and cool down rate restrictions and strict sequencing requirements in various parts of the plant to prepare for a start-up. And although the power block can usually continue operation on its back-up fuel when the gasifier is down, this is often not an attractive operating mode for IGCC plants. IGCC combined cycles are optimized for integrated operation with syngas fuel so they are likely to be less efficient than other generation options when on the back-up fuel, and back-up fuels (at least in Polk's case, distillate) can be expensive.

- The operator training simulator was furnished by Bailey and TRAX, Inc. A copy of the actual plant control system (DCS and Triconex hardware and software) interacts with process plant models running on seven PC's. This simulator enabled plant personnel to become familiar with plant operation before startup and correct control system and procedural errors before they occurred in the real plant.

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  - Overall plant load control
  - Combustion Turbine fuel transfers,
  - pH control in water treatment,
  - Grey Water inventory control
  - Centrifuge control in Brine Concentration.

For each gasifier shutdown to date, to the extent that it was economically justified, the shutdown cause has been identified and corrected.

For example, Table 4 shows that the most serious factor affecting unit availability has been gas side plugging in the fire-tube heat exchangers. A significant effort was expended by Tampa Electric, Texaco, Bechtel and the equipment manufacturer, L. C. Steinmuller, to better understand the physical and chemical phenomenon which caused the plugging. Fortunately, in mid-October, TEC was able to develop empirical temperature and velocity correlations which enabled us to avoid operating in the plugging regime. There have been no shutdowns due to plugging since then. The correlations and operating procedures continue to be improved to optimize unit operation while still avoiding gas side pluggage. The phenomena which caused the plugging in the first place have not yet been positively identified, so TEC has requested Texaco, Steinmuller, and Bechtel to continue these investigations.

As another example, one of the shutdowns attributed to a valve failure was actually due to the failure of an I/P transducer. These devices inevitably fail, and providing redundancy to preclude all such failures for important valves could not be economically justified. However, we did replace the failed transducer with one from a different manufacturer whose transducers have demonstrated higher reliability in the TEC system.

The effort to identify and eliminate direct shutdown causes will continue in earnest as an extremely high priority in 1997, and probably throughout the life of the plant.

Some of the other steps which will be taken in 1997 to improve availability and reduce operating costs in the other areas are:

- To help survive operational upsets:
  - DCS operator displays, ready lists, etc., will be improved
  - The DCS "alarm team" will complete its work to eliminate nuisance alarms and prioritize the remaining ones.
  - Alternate routes for process water streams will be installed.
- To reduce gasifier turn-around and restart time:
  - Establish "hot-restart" procedures (restarting the gasifier immediately following a trip without having to preheat).
- To reduce the time between gasifier startup and when specification fuel is available to the CT:
  - Streamline procedures for placing the acid gas removal system in service
- To reduce/eliminate CT fuel transfer problems:
  - Complete modifications and tune the nitrogen purge system.

With these and other changes, we expect to improve on Polk Power's already excellent record of reliability growth.

## IMPROVE UNIT HEAT RATE

To improve unit heat rate, we are primarily considering operational changes that can be made with the existing plant equipment. Due to the complex and integrated nature of the process, there are still several areas of the plant which require "shakedown" or "fine tuning". It is difficult to make multiple changes within the process and accurately track the impact on unit performance. Therefore, we have instituted a systematic plan to analyze and improve plant heat rate.

The plant engineering staff, along with the plant operators (IGCC Process Specialists), have formed various process improvement teams. These teams look at specific areas of the plant where potential improvements are available. Recommendations are evaluated and implemented by the teams and results identified. Due to the high amount of integration in the process, it is important to understand overall relationships so that improvements in one area do not have an adverse effect in another area. Table 5 summarizes the areas currently under evaluation for plant heat rate improvements.

**TABLE 5**  
**Heat Rate Improvements**

Plant Area	Estimated Improvement Available
Air Separation Unit - Overall pressure balance and optimization	1.5 to 2.5 MW or approx. 75 BTU/KWH
"Other" Plant internal load	1.0 to 1.5 MW or approx. 50 BTU/KWH
Gasification - Heat Balance Optimization - Slurry Concentration - O/C Ratio - Operating Temperature - Carbon Conversion	Combined 300 to 500 BTU/KWH
Power Block and Steam Cycle Optimization	Combined 100 to 200 BTU/KWH

Current Heat Rate (BTU/KWH) = 9300 to 9500

Improvements Identified (BTU/KWH)\* = 500 to 800

Projected Heat Rate(BTU/KWH) = 8500 to 9000 \*\*

\* These are improvements identified based on current equipment in basically original configuration.

\*\* Initial projections of heat rate at ISO conditions were in the 8600 BTU/KWH range and this remains our target. The numbers presented above indicate that if all the improvements identified for the current equipment are realized, we will meet or slightly beat the heat rate goal for the unit.

## **START-UP OF THE HGCU SYSTEM**

During 1996, the HGCU system was operated without syngas to check out all the mechanical equipment and controls and to complete a cold flow attrition test with the sorbent Z-sorb III. In 1997, it is anticipated that the HGCU system will be run with the major objective of achieving steady state operation and optimizing the HGCU system using the first commercially available HGCU sorbent. During the period 1997-2000, the HGCU system will be tested using four different types of coals. Both sulfur and chlorine content will be varied by switching coals.

It was intended to have operated the HGCU system during the original check-out and start-up phase of the plant. However, the syngas temperature at the take-off to the HGCU system is significantly lower than designed. This is primarily due to the fact that the Radiant and Convective Syngas Coolers are removing more heat from the gas than anticipated. Several options are currently being evaluated to provide a suitable gas temperature to the HGCU system. Once this requirement is resolved, the HGCU test plan will resume.

The primary parameters that will be monitored are:

- a) H<sub>2</sub>S removal efficiency
- b) Ammonia inlet/outlet concentrations
- c) Availability
- d) Sorbent attrition rate
- e) Consumption of power, water, air, etc.
- f) Chloride removal efficiency
- g) Barrier filter performance
- h) Regeneration efficiency
- i) IGCC cycle efficiency
- j) Off-gas SO<sub>2</sub> purity
- k) Effects of impurities on sorbent performance
- l) Flyash slag removal efficiency (primary cyclone)

## **PREPARATION FOR ALTERNATE FUEL TEST BURNS (DOE DEMONSTRATION TEST BURNS)**

Polk Power Station has initiated a 4½ year Demonstration Test Burn period designed to cover overall unit and individual subsystem/component performance parameters. The unit has been designed to utilize eastern (U.S.) caking coals. Tampa Electric and the Department of Energy have agreed on a basic test plan that will evaluate the unit performance on four distinct coals. Data collection will provide valuable information on how the IGCC process, including the HGCU system, is started up, operated in stable and load changing modes, and then shut down in a safe manner. Also, this test program will help Tampa Electric identify the least "overall" cost fuel for continued operation of Polk Power Station.

Part of the test plan will also evaluate maintenance issues. Since much of the equipment being developed for this project is the first and/or largest of its kind, maintainability and reliability are as



important as operability. Issues such as corrosion rates, materials of construction, and accessibility for repair will be studied.

The initial start-up and non-test burn periods will use a Pittsburgh #8 coal. Performance tests for the major components and systems are specified in the respective contracts to be done using Pittsburgh #8 coal, and occasionally a "modified" Pittsburgh #8 coal with higher than standard sulfur. The plans call for a series of short term test burns on the various alternate fuels to be evaluated. This period will be followed by a period of time available to make any modifications required to the system in order to perform a long term test burn. Next, a long term test burn will be performed.

The planned sequence for the test burns is as follows:

- a) Run the unit on Pittsburgh #8 Coal
- b) Introduce the test fuel for 100 hours and monitor performance
- c) Return to Pittsburgh #8 coal
- d) Repeat a, b, and c above for each of the selected fuels
- e) Evaluate short term performance of the test fuel and determine any modifications required to perform a long term test burn
- f) Institute required changes
- g) Introduce the test fuel for the long term test burn
- h) Evaluate the long term performance
- i) Return to Pittsburgh #8 coal

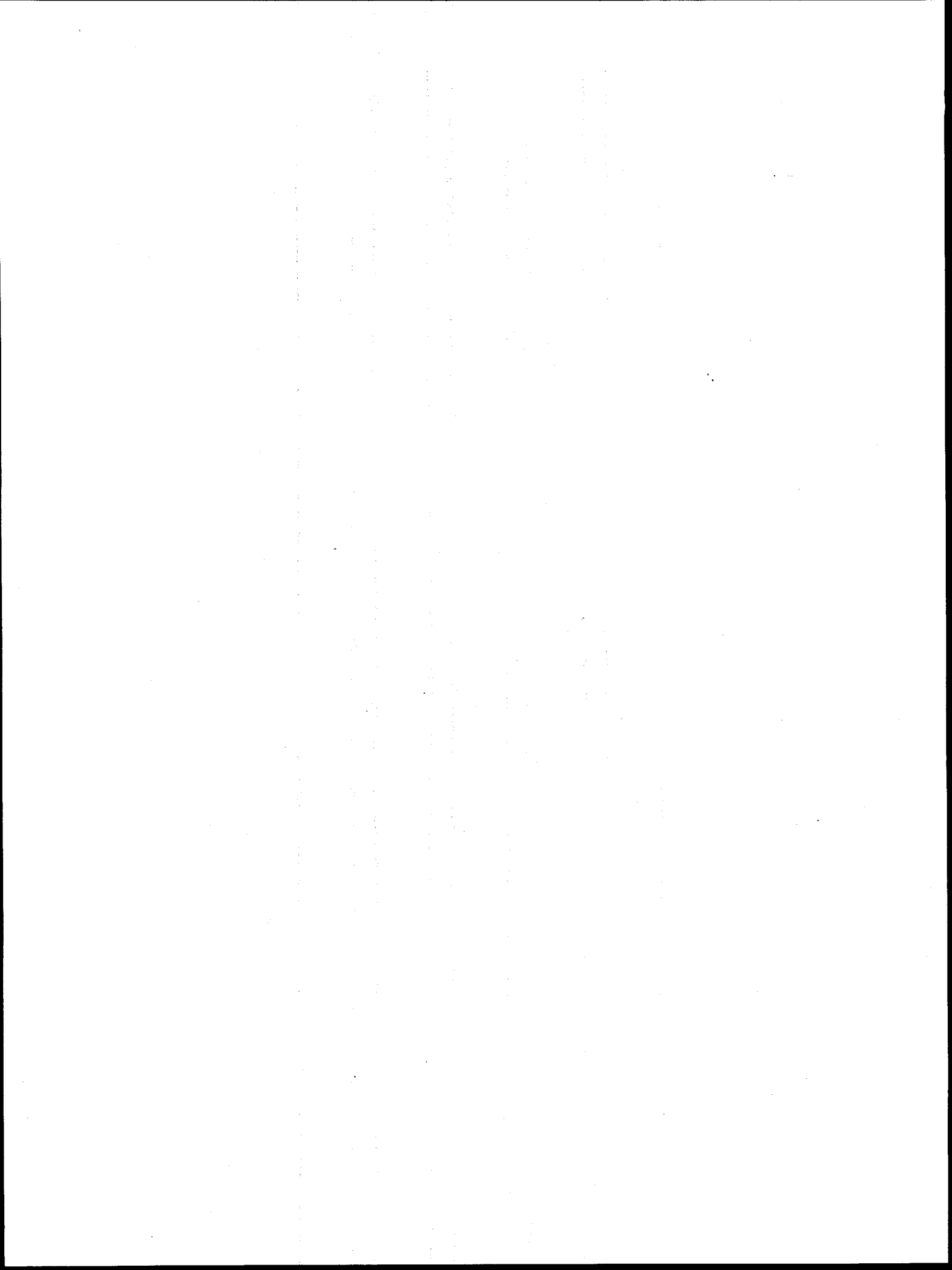
Each short term test burn will last approximately 100 hours (4 days). During the short term test burn, all performance related test parameters will be collected. This data will be used for two purposes. First, it will be used to report on IGCC performance using this fuel. Second, the data will be used to ascertain the type and extent of any modifications necessary to allow the system to operate on that fuel for a long term test along with the economics associated with the modifications identified.

It is anticipated that all four coals evaluated during the short term test burns will be selected for long term tests. Each of the long term test burns will last for approximately one month. Performance parameters and operating characteristics will be monitored. In addition, evaluations of refractory wear rate, material corrosion rates and other specific areas of concern will be performed.

## CONCLUSION

Polk Power Station was completed and commissioned on schedule and on budget. Recent operation has been excellent. We are very pleased with the plant's performance to date and are optimistic about the future of this clean coal technology in general and the Polk Power Plant in particular. We are looking forward to 1997 when we expect to further improve Polk's reliability, reduce operating costs, improve heat rate, commission the HGCU unit, and begin alternate coal testing.

TEC gratefully acknowledges the financial and technical support provided by the U.S. Department of Energy through their Clean Coal Technology Program. Without DOE participation, Polk Power would not have been possible. With DOE participation, we have been able to prove that IGCC can reduce NO<sub>x</sub> and SO<sub>x</sub> emissions in a utility scale application and advance the commercialization of this important clean coal technology.



## **Piñon Pine Power Project Nears Start-up**

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Houston, Texas

### **ABSTRACT**

The IGCC facility being built by Sierra Pacific Power Company (SPPCo) at their Tracy Station in Nevada is one of three IGCC facilities being cost-shared by the U.S. Department of Energy (DOE) under their Clean Coal Technology Program. The specific technology to be demonstrated in SPPCo's Round Four Project, known as the Piñon Pine IGCC Project, includes the KRW air blown pressurized fluidized bed gasification process with hot gas cleanup coupled with a combined cycle facility based on a new GE 6FA gas turbine. Construction of the 100 MW IGCC facility began in February 1995 and the first firing of the gas turbine occurred as scheduled on August 15, 1996 with natural gas. Mechanical completion of the gasifier and other outstanding work is due in January 1997. Following the startup of the plant, the project will enter a 42 month operating and testing period during which low sulfur western and high sulfur eastern or midwestern coals will be processed.

### **Acknowledgements**

The following are acknowledged for their ongoing, valuable support to SPPCo on this project: Douglas M. Jewell of Morgantown Energy Technology Center, who has been the DOE Project Manager since the project began with the signing of the Cooperative Agreement in August

1992; Foster Wheeler USA Corporation (FWUSA) of Clinton, New Jersey, subcontractor to SPPCo for project management, engineering, and construction management for the overall facility; and The M W Kellogg Company of Houston, Texas, subcontractor to FWUSA for design engineering, equipment procurement, and other services for the coal gasification section of the plant.

## **Introduction**

The Piñon Pine Power Project was one of the successful proposals in response to the U.S. Department of Energy's (DOE) Clean Coal Technology, Round Four solicitation which invited submissions for cost-shared projects to demonstrate technologies capable of replacing, retrofitting or repowering existing coal based facilities. The Program Opportunity Notice (PON) for the fourth round called for projects to demonstrate innovative, clean and energy efficient technologies with particular emphasis on achieving significant reduction in emissions of sulfur dioxide and/or nitrogen oxides from existing facilities and/or providing for future energy needs in an environmentally acceptable manner.

In the Piñon Pine IGCC Project, Sierra Pacific Power Co., (SPPCo) aims to demonstrate the use of advanced coal technologies to produce clean and low cost power to meet their growing customer needs. The facility is being built at SPPCo's Tracy Station some 20 miles east of Reno, Nevada and includes the design, engineering, procurement, construction and testing of a nominal 100 MW coal fueled integrated gasification combined cycle (IGCC) plant. The KRW air blown pressurized fluidized bed coal gasifier with hot gas cleanup will produce high temperature coal gas to be burned in a GE frame 6FA combustion turbine to generate about 60% of the plant power output. The rest of the power will be produced in a steam turbine generator driven by steam produced primarily from the combustion turbine exhaust gases. Foster Wheeler USA Corporation (FWUSA) is providing the design, engineering, procurement, and construction management of the overall facility with The M W Kellogg Company (MWK) subcontracted to supply the design, engineering, equipment procurement, and other services relating to the gasification island.

The GE Frame 6FA combustion turbine at the Piñon Pine Project is the first of its kind in the world and was successfully fired for the first time on August 15, 1996 using natural gas. The combined cycle part of the plant began commercial operation on natural gas in November 1996. Work is proceeding to complete the coal gasification island and support facilities in January 1997. The operation and testing phase of the project will last for 42 months and will include operation with the design coal, Southern Utah bituminous, as well as tests on high sulfur eastern or midwestern coal. The cost of building, commissioning and demonstrating the overall facility will be about \$335 million to be shared equally between DOE and SPPCo.

## Project Goals

First and foremost in SPPCo's decision to proceed with the project was the objective to generate low cost, base load power using coal in a clean environmentally acceptable manner.

To this end the aims of the Piñon Pine Power Project include:

- Demonstrating air blown, pressurized fluidized bed IGCC technology incorporating hot gas cleanup.
- Evaluating a combustion turbine firing low Btu coal gas.
- Assessing efficiency and the long term reliability, maintainability and environmental performance of the complete facility.
- Providing SPPCo with increased fuel flexibility in their generation system.

At the average ambient conditions for the Reno area, the plant is expected to perform in accordance with Table 1.

**TABLE 1**

Coal Feed (T/D)	880.6
Gas Turbine Power (Mwe)	60.99
Steam Turbine Power (Mwe)	46.23
Gross Power (Mwe)	107.22
Auxiliary Power (Mwe)	7.51
Net Power (Mwe)	99.71
Net Heat Rate (Btu (LHV)/kWh)	8096
Net Heat Rate (Btu (HHV)/kWh)	8390
Thermal Efficiency (LHV)%	42.1
Thermal Efficiency (HHV)%	40.6

## Key Technology Features

The KRW process (licensed by The M W Kellogg Technology Company) to produce clean high temperature coal gas improves upon the first generation IGCC technologies in several aspects which can be summarized as follows:

- Air Blown Gasification

Using air in place of oxygen as the oxidant in the gasification process leads to a simpler plant configuration and lower capital cost. In the air blown process 15 to 20% of the gas turbine compressor air is extracted for use as oxidant in the gasifier.

The gasifier is capable of operating with a wide variety of coals. This fuel flexibility is a major advantage of the process. During the testing period the design coal, a low sulfur western U.S. coal, will be the predominant fuel with campaign test runs of eastern or midwestern high sulfur coal.

Addition of limestone (or dolomite) to the gasifier serves several purposes. Dolomite offers economic, as well as processing, advantages for sulfation and ultimate ash disposal. Limestone (or dolomite) captures a large percentage of the sulfur released from the coal in the gasification process. Furthermore, previous test work indicates that the presence of limestone reduces the amount of ammonia produced, the latter being a contributor to NO<sub>x</sub> generation in the gas turbine combustor.

- **Hot Gas Particulate Removal**

Filtering of the gas at high temperature enables the sensible heat to be maintained resulting in higher plant efficiency.

- **Hot Gas Desulfurization**

Sulfur contained in the coal is removed in two steps. Some of the hydrogen sulfide produced in the reducing environment of the gasifier is captured as calcium sulfide by limestone fed to the gasifier with the coal. Chemical equilibrium considerations limit the capture to about 50% with low sulfur coal, but with high sulfur coals this can approach 90%. Sulfur, primarily in the form of hydrogen sulfide not captured by the limestone or retained by the ash exits the gasifier in the product gas stream, and is removed by the zinc based sorbent in the external hot gas transport desulfurization system.

- **Sulfation**

Coal ash with spent limestone (LASH) containing calcium sulfide and unconverted carbon is treated in the sulfator system which oxidizes the sulfide to calcium sulfate, combusts unconverted carbon and absorbs sulfur dioxide from the external transport desulfurization system regeneration gas.

## **Plant Configuration**

Block flow diagram **Figure 1**, indicates the main parts of this IGCC facility. Crushed coal ¼" x 0 and limestone (or dolomite) 16 x 200 mesh are pneumatically fed via lock hoppers into the gasifier, along with additional air from the booster compressor, through the concentric central

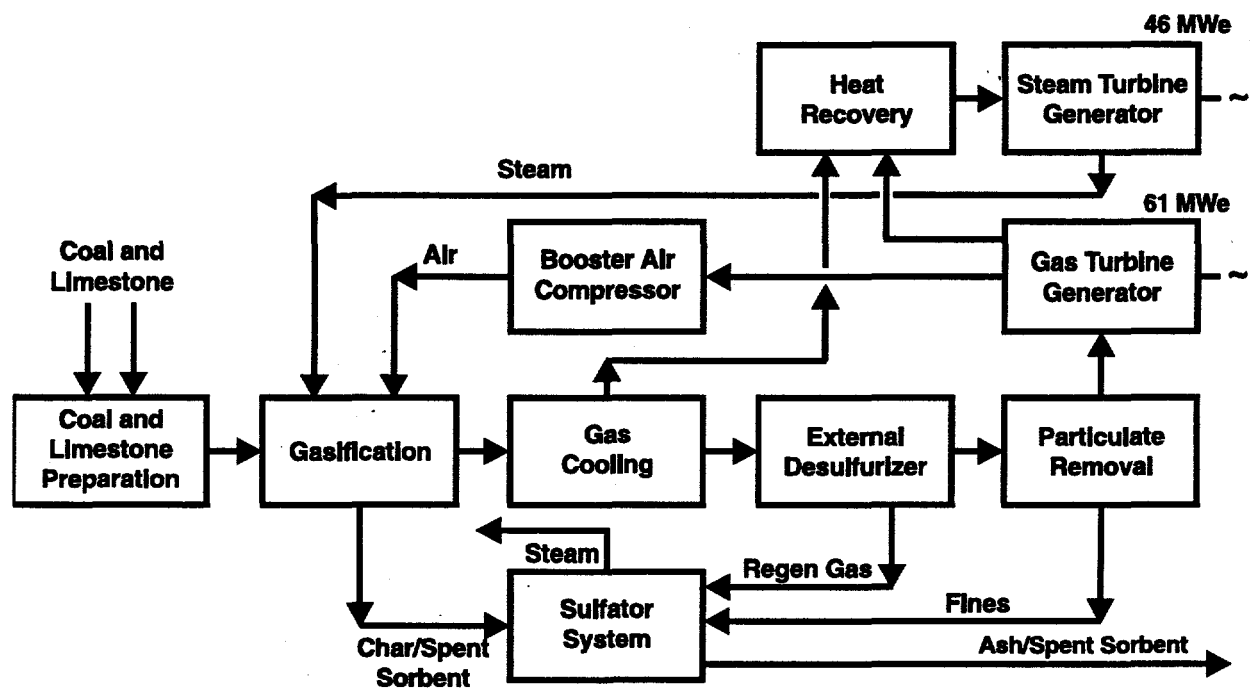
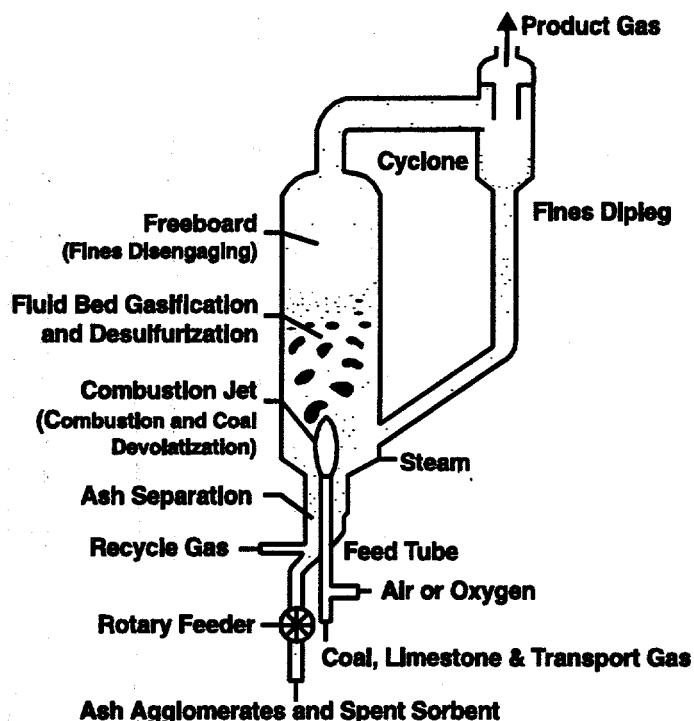


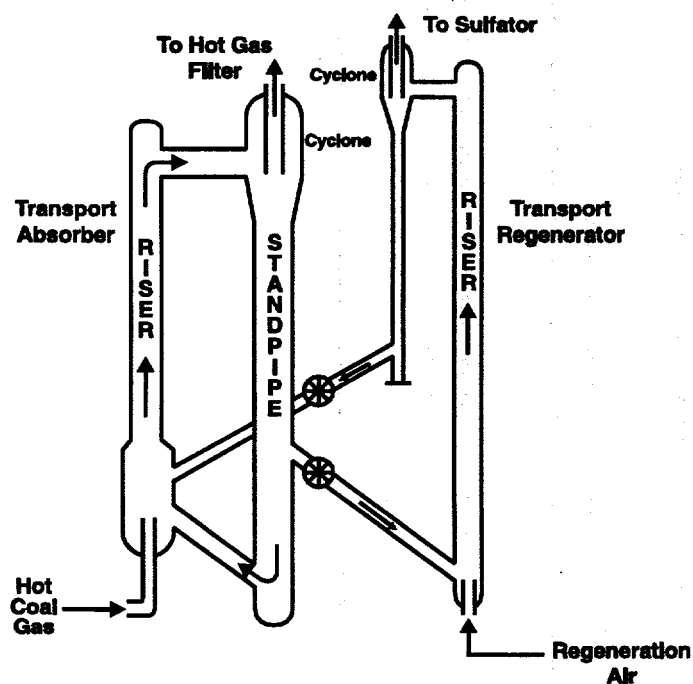
Figure 1: Block Flow Diagram



feed tube forming a high velocity jet (See **Figure 2**). Once in the gasifier the coal is quickly devolatilized. Partial combustion of char and gas occurring within the jet provides the heat necessary for the endothermic devolatilization, gasification, and desulfurization reactions. Extraction steam from the steam turbine is injected through the gasifier grid to aid in fluidization and drive the gasification reactions. LASH particles which separate from the bed particles due to their higher density are cooled and removed from the bottom of the gasifier. Recycle gas is used for fluidization and for cooling the LASH. The coal gas leaving the top of the gasifier contains significant quantities of entrained solids consisting of char, ash, and sorbent. A cyclone removes most of the entrained solids which are returned to the gasifier bed via the dipleg. The product gas is cooled from about 1800°F to 1000°F in a series of exchangers with the heat recovered as high pressure steam. The cooled coal gas is treated in the hot gas cleanup system to meet the specification required for fuel to the combustion turbine.



**Figure 2: KRW Gasifier**



**Figure 3: Transport Desulfurizer**

In the transport desulfurizer system shown in **Figure 3**, a zinc oxide based sorbent which also contains nickel oxide reduces the sulfur content in the gas to less than 20 ppmv. Fuel gas enters the mixing zone at the bottom of the transport absorber riser where it mixes with the sorbent recirculated from the absorber cyclone. Absorption of the gaseous sulfur compounds occurs in the riser section as the fuel gas and sorbent flow upward into the absorber cyclone. A slip stream of sulfurized sorbent is withdrawn from the absorber standpipe and enters the bottom of the transport regenerator along with preheated air. The sulfur rich gas exits the transport regenerator at about 1400°F and flows to the sulfator. Regenerated sorbent is returned to the absorber by a controlled gravity flow.

Particulates are removed from the hot desulfurized gas in a Westinghouse ceramic candle filter system before the product gas is burned in the combustion turbine. Filter fines are burned in the fines combustor in the sulfator system.

With the exception of a very small quantity of sulfur in the fuel gas to the gas turbine (about 20 ppmv) all the sulfur in the coal is ultimately disposed of in the sulfator system which includes an air fluidized bubbling bed reactor. The sulfator is operated at about 1600°F to maximize capture of the sulfur dioxide released from combustion of residual char in the LASH and that contained in the transport regenerator effluent gas. This high temperature also maximizes oxidation of the calcium sulfide to the sulfate form. Sulfator off gas is used to quench the fines combustor effluent stream which is vented through a baghouse after cooling the gas by raising additional steam. Sulfator solids are cooled, combined with fines from the baghouse, and stored in a silo for intermittent transfer by truck to landfill or for other uses.

The GE model 6FA gas turbine forms the heart of the power island. The technologically advanced firing temperature (2350°F) and cooling system of the F-Class machines provide such units operating in combined cycle configuration with the highest total cycle efficiencies of any proven type of fossil fueled electric power generation system. Mechanical power will be converted to electrical power in a once through air cooled synchronous generator which will provide 61 MW of power.

Thermal energy in the exhaust gases from the combustion turbine is captured in a heat recovery steam generator (HRSG), the steam from which will drive a condensing steam turbine generator to produce an additional 46 MW of power. In-plant power use is expected to be only 7 MW which is less than oxygen blown IGCC processes due to the lack of a requirement for an air separation unit.

## Technology Development

The Piñon Pine IGCC project integrates a number of technologies fostered by the DOE over a period of many years. These include the KRW gasifier with in-bed desulfurization using limestone, external hot gas desulfurization and hot gas particulate removal.

The DOE and its predecessors supported the KRW gasifier development from 1972 to 1988. Westinghouse Electric Corporation originally proposed the technology and in 1975 completed the construction at Waltz Mill, Pennsylvania of a 25 ton/day Process Development Unit (PDU). Testing proceeded for more than a decade thereafter. In 1986, The M W Kellogg Company acquired the process. The PDU was operated in the air and oxygen blown modes and tested many different types of coal from many parts of the world ranging from lignite to anthracite. Tests also included in-bed desulfurization using limestone and dolomite sorbents, the use of ceramic (and sintered metal) filters for particulate removal and external-bed desulfurization using zinc ferrite in a fixed bed reactor.

Based on experience at Waltz Mill the initial process design for the Piñon Pine Power Project included a fixed bed hot gas desulfurization system using zinc ferrite sorbent. In support of this project following its selection under Round 4 of the Clean Coal Program, DOE at their Morgantown Energy Technology Center tested zinc ferrite, zinc titanate, and Z-Sorb®, developed by Phillips Petroleum Company. From these screening tests Z-Sorb® proved to have the best properties. These tests also strongly indicated economic and technical reasons for not using a fixed bed system. A transport type system was then evaluated by MWK.

After a series of tests and a detailed technical review of both fixed bed and transport systems, the latter was chosen for demonstration in the SPPCo project. The transport system represents both a technical and cost improvement over the fixed bed arrangement. The inventory of sorbent is greatly reduced and the overall system is much simpler with no cycling valves in hot product gas streams, and no opportunity for process upsets during bed switching which could occur with the multi-vessel fixed bed system.

## **Project Schedule**

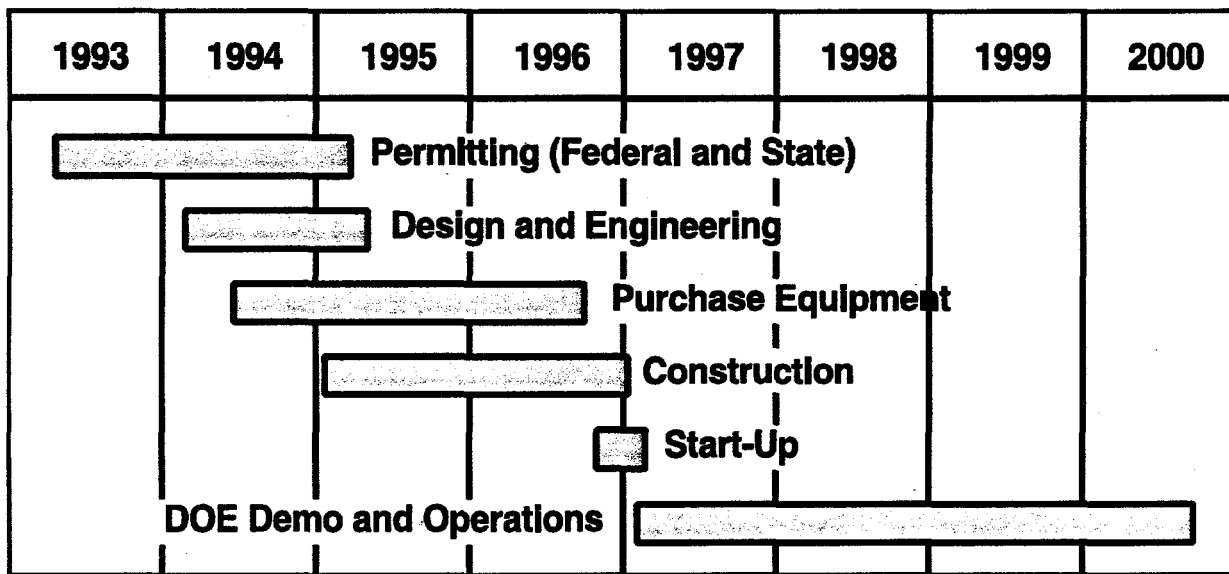
The project schedule is shown in **Figure 4**. The Cooperative Agreement was signed in August 1992, which signaled the commencement of phased execution of the project. With the design phase completed, the construction and startup phase started in January 1995. The first firing of the combustion turbine was successfully carried out as scheduled on August 15, 1996 using natural gas. Mechanical completion of the combined cycle power plant section of the project was achieved on August 29, 1996. Work is proceeding to complete commissioning of the gasifier island and supporting balance of plant in January 1997 with the commencement of Phase 3, operation and testing in February 1997.

## Future Plans

The 42 month Phase 3 of the Piñon Pine IGCC Project will include extensive tests of process, equipment, and controls using the design coal, (Southern Utah low sulfur bituminous), as well as a specific campaign on midwestern or eastern high sulfur coal.

Key aspects of the facility that will be demonstrated include the air blown KRW gasifier with in-bed desulfurization, the hot gas transport desulfurization system and the high temperature ceramic candle filter system, as well as the GE Model MS6001FA gas turbine operating with low Btu coal gas and natural gas (and the ability to switch from one to the other), and the capability of the facility as a whole to operate in base load mode and follow load demand variations in accordance with utility standard requirements. In addition the facility will demonstrate that it complies with acceptable emissions and with the required efficiency, reliability, availability, and maintainability.

Successful demonstration of this technology will provide the power generation industry with design, construction and operating information for assessing new power generation options. The performance of the KRW-based IGCC technology together with its modular design concept will offer an attractive way to satisfy future demand for greenfield electricity generating facilities. In addition, with the large number of existing boilers reaching the end of their useful lives, the air



**Figure 4: Schedule of Piñon Pine Project**

blown KRW based IGCC process with its relatively simple configuration and limited space requirement offers an excellent method to repower these stations. The resulting facility could produce up to three times the power currently generated by the existing steam turbine.

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# **McIntosh Unit 4 PCFB Demonstration Project**

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## **Introduction**

The City of Lakeland, Foster Wheeler Corporation and Westinghouse Electric Corporation have embarked on a utility scale demonstration of Pressurized Circulating Fluidized Bed (PCFB) technology at Lakeland's McIntosh Power Station in Lakeland, Florida. The U.S. Department of Energy will be providing approximately \$195 million of funding for the project through two Cooperative Agreements under the auspices of the Clean Coal Technology Program. The project will involve the commercial demonstration of FOSTER WHEELER PYROFLOW PCFB technology integrated with Westinghouse's Hot Gas Filter (HGF) and power generation ® technologies.

The total project duration will be approximately eight years and will be structured into three separate phases; two years of design and permitting, followed by an initial period of two years of fabrication and construction and concluding with a four year demonstration (commercial operation) period. It is expected that the project will show that Foster Wheeler's Pyroflow PCFB technology coupled with Westinghouse's HGF and power generation technologies represents a cost effective, high efficiency, low emissions means of adding greenfield generation capacity and that this same technology is also well suited for repowering applications.



## Background

The City of Lakeland, Department of Electric & Water Utilities (Lakeland) is a municipally owned and operated electric and water utility in Central Florida. Lakeland is conveniently located between Tampa and Orlando which has allowed Lakeland to grow and prosper over its 92 year history. Lakeland is the third largest municipal utility in the State of Florida serving more than 104,000 electric customers and also has residential rates that are currently the second lowest of all Florida utilities. Despite enjoying low electric rates and steady load growth, Lakeland is not immune to competition. Competition is driving all utilities to find ways not only to prevent cost growth but to also lower costs. A heightened awareness of the environment by the general public and Lakeland's customers is also maintaining the pressure for "clean" electric generation. Traditionally these two goals have not been complimentary in that environmental compliance normally has meant an increase in generation costs to achieve that compliance. This raises the question each utility must soon face: how to provide new generating capacity, needed for growth and replacement of retired capacity, at a competitive cost while meeting stringent environmental requirements.

Lakeland has experienced and is forecasting steady load growth within its municipal system of approximately 15 MW per year which will result in a capacity shortfall in the year 2000 of approximately 45 MW. In addition to the pending capacity shortfall, Lakeland wishes to retire 50 MW of very old and inefficient existing generating capacity. Considering both of these issues and future needs, Lakeland needs to bring on line at least 150 MW of additional generating capacity by the year 2000.

In today's competitive environment, the prospects of adding additional capacity in itself can bring many uncertainties. With the majority of Lakeland's capacity already tied to one fuel that has greater uncertainties in such areas as price and availability, the need to add more capacity led Lakeland to look closely once again at America's most abundant fuel source, coal. Lakeland's current mix of resources include approximately 200 MW of base load pulverized coal and 450 MW of intermediate/peaking gas capacity. This capacity is divided between two power stations that Lakeland owns which are located within the city limits on the shores of Lake Parker. The larger of the two power stations is the McIntosh station on the north side of Lake Parker with approximately 590 MW of generating capacity while the smaller Larsen station on the south side of the lake has about 230 MW of<sup>1</sup> generating capacity.

Lakeland was a pioneer of sorts when the 334 MW McIntosh 3 unit went on-line in 1982. The unit was one of the first "scrubbed", zero-discharge coal units in the nation. Today, Lakeland is looking to be a pioneer again by partnering with Foster Wheeler Corporation and Westinghouse Electric Corporation to build and operate a utility scale demonstration of PCFB technology (unit 4) at Lakeland's McIntosh Plant site. The addition of McIntosh unit 4 will provide Lakeland with new, cost competitive and environmentally clean coal based capacity for the 21st Century. The added capacity that this unit will provide will not only add to Lakeland's fuel diversity, but will provide energy at some of the lowest costs per megawatt hour of any generating source in the Southeast. These factors combined with the state of the art pollution controls provided by the Foster Wheeler PCFB process and the Westinghouse HGF technology will ensure that McIntosh unit 4 will keep Lakeland very competitive and environmentally acceptable well into the future.

The successful construction and operation of this technology will provide utilities with a means of adding needed generating capacity in a manner that is consistent with the competitive and environmental challenges that all are facing.

<sup>1</sup>McIntosh Unit 3 is a 334 MW pulverized coal unit that is jointly owned by Lakeland and the Orlando <sup>1</sup>Utilities Commission.

## **Project Structure**

The proposed McIntosh Unit 4 PCFB Demonstration Project would be constructed as two sequential demonstrations that would demonstrate both PCFB and Topped PCFB technology. There are two primary reasons for this proposed project structure:

- (I) The DOE funding being provided for the project results from a combination of two previous Clean Coal awards: the DMEC-1 PCFB Repowering Project (DMEC-1) selected under Round III and the Four Rivers Energy Modernization Project (FREMP) selected under Round V. The DMEC-1 project was intended to demonstrate PCFB technology while the FREMP project was planning to demonstrate Topped PCFB technology. By utilizing a sequential approach with the McIntosh Unit 4 PCFB project, it will be possible to demonstrate both PCFB (1st Demonstration) and Topped PCFB (2nd Demonstration) technology in the same project, thereby satisfying the objectives of both the DMEC and FREMP projects.
- (II) Additional development work is required on certain components of the Topped PCFB cycle prior to the construction of the same components at a commercial scale. Specifically, additional development is required for the Westinghouse topping combustor (multi-annular-swirl-burner or MASB) including the demonstration of MASB operation at low outlet oxygen levels. Important aspects of Westinghouse's MASB development work have been and will be conducted at the University of Tennessee Space Institute. Some additional development work may also be performed for other components of the carbonizer system. Development on the carbonizer system has been performed at Foster Wheeler's John Blizzard Research Center in Livingston, New Jersey. Both of these systems are incorporated in the Wilsonville Power Systems Development Facility (PSDF) facility at a Southern Company operated site in Wilsonville, Alabama that will shortly be starting operation. The combination of the above programs is expected to provide Westinghouse and Foster Wheeler with the necessary information required to finalize the design of the carbonizer and MASB's in time to support the demonstration of Topped PCFB technology.

The project schedule (discussed in more detail below) anticipates the start of commercial operation of the 1st Demonstration in the winter of the year 2000. In parallel with the first two years of operation of the 1st Demonstration will be the design, fabrication and construction of the 2nd Demonstration culminating in a planned start of operation of late 2002 for the combined facility.

## **Project Objectives**

Through the sequential demonstration of both PCFB and Topped PCFB technology it has been possible to preserve the objectives of both the original Cooperative Agreements described in the preceding section. The objectives governing the agreement relating to PCFB technology include the demonstration of PCFB technology to provide for the potential commercialization of the technology in the 21st century and to provide the capability of achieving significant reductions in the emissions of sulfur oxide and nitrogen oxides from existing facilities when they are repowered with PCFB technology.

The objectives for the agreement relating to Topped PCFB technology call for the demonstration of the technology in a "fully commercial power generation setting" which is certainly the case at the McIntosh site as is further explained below. All the key components of the Topped PCFB technology will be demonstrated thereby paving the way for future plants that will operate at higher gas turbine inlet temperatures and that are expected to provide cycle efficiencies in excess of 45%. Additional objectives relating to the Topped PCFB technology that will be proven through a successful demonstration include reductions in sulfur oxide emissions of as much as 95% and nitrogen oxide emissions as low as 0.17 lb/MMBTU of heat input.

## **Process Description**

PCFB technology is a combined cycle power generation system that is based on the pressurized combustion of solid fuel to generate steam in a conventional Rankine cycle combined with the expansion of hot pressurized flue gas through a gas turbine in a Brayton cycle. The technology can be subdivided into the basic PCFB cycle ("First Generation") and Topped PCFB cycle ("2nd Generation" or "Advanced PCFB"). In the PCFB cycle, hot pressurized flue gas is expanded through the gas turbine at a temperature of less than 1650°F. Topped PCFB cycles include a coal carbonizer (mild gasifier) to generate a low BTU fuel gas which is used to fire the inlet of the gas turbine (in a topping combustor or MASB) and increase the gas turbine inlet temperature from a less than 1650°F up to 1900° - 2300°F or higher. Both versions of PCFB technology offer high cycle efficiencies and ultra low emissions. More detailed descriptions of the PCFB and Topped PCFB cycles are provided below.

Figure 1 presents a simplified schematic of the 1st Demonstration of the McIntosh Unit 4 PCFB Demonstration Project incorporating a PCFB cycle. Combustion air is supplied from the compressor section of the gas turbine to the PCFB combustor located inside a pressure vessel.

Coal and limestone are mixed with water into a paste which is pumped into the combustion chamber using piston pumps commonly used in the concrete industry. The same type of pumps have been successfully proven in a number of pressurized fluidized bed combustion (PFBC) coal projects around the world.

Combustion takes place at a temperature of approximately 1560° - 1600°F and at a pressure of about 200 psig. The resulting flue gas and fly ash leaving the cyclone enter the hot gas filters where dust removal takes place. The hot gas filters are a Westinghouse design based closely on the filter supplied to the Sierra Pacific Piñon Pine project in Tracy, Nevada. In addition to the Piñon Pine project, a Westinghouse filter has undergone approximately 6000 hours of testing at Ohio Power's Tidd PFBC Demonstration facility in Brilliant, Ohio (Round I project). A full scale commercial module of this type of ceramic candle filter has also undergone more than 6000 hours of extensive testing at Foster Wheeler's PCFB test facility in Karhula, Finland.

The hot clean gas leaving the filter is expanded through the gas turbine before passing through a heat recovery unit and entering the stack. Heat recovered from the cycle from both the combustor and the heat recovery unit is used to generate steam to power a reheat steam turbine. Approximately 15% of the gross power output is derived from the gas turbine with the steam turbine contributing the remaining 85%.

The gas turbine technology is based on a standard Westinghouse 251B12, single shaft, cold end drive industrial machine that has had the center section of the turbine modified. A scroll section has been added to allow for the removal of compressor discharge air from the casing for external firing in the PCFB combustor and to allow for the introduction of hot clean gas back through the casing into the expander section. This air outlet/gas inlet configuration has been previously applied in recuperative gas turbine cycles. The gas inlet temperature of less than 1650 F allows for a simplified turbine shaft and blade cooling system. This combined with low excess air operation in the PCFB combustor provides a maximum amount of steam generation per unit mass of air from the gas turbine and therefore maximizes power output from the cycle.

Figure 2 shows the process flow arrangement of the 2nd Demonstration of the McIntosh Unit 4 PCFB Demonstration Project. This involves the addition of a carbonizer island which includes a topping combustor (MASB) to convert the PCFB cycle to a Topped PCFB cycle. Through the addition of this equipment, the inlet temperature to the gas turbine is increased via the combustion of coal derived "syngas". This has the effect of increasing the cycle power output while simultaneously improving the net plant heat rate. Natural gas can also be used as the topping fuel thereby providing a backup to the operation of the carbonizer island.

In the top right hand corner of Figure 2, the carbonizer island is shown. Dried coal and limestone are fed via a lock hopper system to the carbonizer together with part of the gas turbine compressor discharge air. The coal is partially gasified or carbonized at about 1700°F to produce a syngas and char solids stream. The limestone is used to absorb sulfur compounds generated during the mild gasification process and to catalyze the gasification process. After cooling the

syngas to about 1200°F, the char and limestone entrained with the syngas are removed by a Westinghouse hot gas filter. The char and limestone are transferred to the PCFB combustor for complete carbon combustion and limestone utilization. The hot clean filtered syngas is then fired in a topping combustor (MASB) to raise the turbine inlet temperature to almost 2000°F. The gas is expanded through the turbine, cooled in a heat recovery unit and exhausted to the stack. As in the case of the previous cycle, combustion air is supplied to the PCFB combustor from the compressor section of the gas turbine. Coal and limestone are again fed to the PCFB combustor in paste form but are supplemented by the char transferred from the carbonizer as discussed above.

## **Performance**

The First Demonstration would involve a basic PCFB cycle that would come on line in the year 2000 and would provide approximately 157 MW of coal-fired generating capacity. The cycle would have a gas turbine inlet temperature of approximately 1550°F. Following the completion of some additional development work, the Second Demonstration of the project would be constructed and brought on line approximately two years later. This would entail the conversion of the 1st Demonstration PCFB system to a Topped PCFB system through the addition of a carbonizer island and a topping combustor. The addition of the carbonizer system would generate a coal derived, low BTU synthesis gas that would be fired at the inlet of the gas turbine to raise the turbine inlet temperature to approximately 1975°F. The net impact of this equipment addition would be an additional 12 MW of power output with an associated improvement in heat rate of about 600 BTU/kWhr for the entire plant.

The project would be constructed as McIntosh Unit 4 within the boundaries of existing station on land owned by the city. The new unit will be designed to burn a range of coals including both the current Eastern Kentucky coal burned in unit 3 and high ash, high sulfur coals that are expected to be available in the future at substantially lower prices than mid to low sulfur bituminous coals. Limestone would be sourced from a number of nearby Florida limestone quarries while ash would be disposed of in a landfill or marketed to others.

The majority of the project's water makeup requirements will be met using secondary treated sewage effluent for cooling tower makeup while the use of sewage "sludge" (3 - 4% solids) is being considered for preparation of the coal-water paste mixture that is pumped into the PCFB. Service water will be used only for boiler water makeup feed to the demineralizer system. Wastewater from the unit will be treated on site for neutralization and removal of heavy metals before being returned to the Glendale waste water treatment facility (owned by Lakeland) for discharge. Gaseous emissions from the plant will be controlled using state of the art technology and will be representative of recent best available control technology (BACT) determinations in Florida.

## **Project Schedule**

The City of Lakeland wishes to have the 1st Demonstration plant enter commercial operation during the winter of the year 2000. Prior to commencing fabrication and construction (Phase 2) of the new facility, the permitting and licensing process required by the state of Florida must be completed. In addition, DOE requires that the National Environmental Policy Act (NEPA) process be completed prior to DOE providing any funds for the purpose of fabricating and constructing the facility.

The NEPA and permitting/licensing processes are each expected to take 20 months to complete and are parallel critical path activities dictating the duration of Phase 1 of the project. At the time of writing, Phase 1 was expected to begin around December 1, 1996 following the formal execution of the Cooperative Agreements by Lakeland and DOE. Phase 2 begins with the general release for fabrication and construction for the 1st Demonstration and lasts for a total of 53 months. Phase 3 has an overall duration of 48 months. The first 29 months of Phase 2 cover the period from the end of Phase 1 through to the start of Phase 3 during which the 1st Demonstration facility is fabricated and constructed. The second 24 months of Phase 2 overlap with Phase 3 and cover the time required to design, engineer, fabricate and construct the 2nd Demonstration equipment.

Phase 3 will be structured in two segments: an initial two year period while the PCFB technology of the 1st Demonstration is demonstrated, and a subsequent two year period during which the Topped PCFB technology of the 2nd Demonstration will be operated. The additional equipment required for the 2nd Demonstration will be engineered, procured and constructed in parallel with the operation of the 1st Demonstration during the first two years of Phase 3. All efforts will be made to minimize the amount of downtime of the facility required to connect the 2nd Demonstration equipment to the 1st Demonstration plant.

Figure 3 presents an overview of the anticipated project schedule.

## **Project Cost and Funding Summary**

The total cost and funding summaries for McIntosh Unit 4 PCFB Demonstration Project in "as spent" dollars are shown below. The total project costs include the total cost to construct the facility, certain project related offsite costs, 4 years of operation and maintenance (O&M) costs, owner's costs and permitting costs.

(\$1000)

**COSTS** Total Project Costs 387,970

Lakeland In-Kinds 2,030

**TOTAL COSTS** 390,000

**FUNDS** Lakeland In-Kinds 2,030

Lakeland 192,970

DOE 195,000

**TOTAL FUNDS** 390,000

The total McIntosh Unit 4 PCFB Demonstration project costs have been divided between the two Cooperative Agreements.

### **Participant Project Financing**

The City of Lakeland has a number of financing alternatives to use for the project. Lakeland has accumulated reserves for future expansion and system general purpose uses. These funds are available for use by the City's Department of Electric & Water Utilities and part of them have been earmarked for the McIntosh Unit 4 PCFB Demonstration Project.

Lakeland also enjoys very favorable bond ratings due to its long-standing financial health. Recently, the drop in interest rates was found to be financially favorable for Lakeland's financing team to issue tax exempt revenue bonds in order to provide funding for several projects listed in Lakeland's current capital forecast. As with any bond issue, this issue has been rated by the bond rating agencies. Lakeland had the bonds rated by Standard and Poor's Group (AA-) and Moody's Investors Service, Inc. (Aa). Lakeland has maintained these ratings since 1989 when the Moody's rating was upgraded to the current level.

The payments for operating costs of Lakeland's Department of Electric & Water Utilities are funded through revenue generated by the sale of electricity and water. The amount of revenue is in part determined by the rates charged for these products. The Department of Electric & Water Utilities, through its long range forecasts, identifies when rate increases are expected. These are identified years in advance of the actual need and are then implemented when, and at the level necessary to continue the financially sound operations of Lakeland. The City Commission for the City of Lakeland has the rate making authority for the Department of Electric & Water Utilities.

Detail revenue and expense budgets are prepared and reviewed each year. The approved budgets are then used to update the long range forecast to determine their impact on future years. This process has been very successful for Lakeland in avoiding unplanned rate increases. In fact, since 1989, Lakeland has been able to implement lower rate increases than originally forecast. Lakeland also believes that the Pressurized Circulating Fluidized Bed generator that this project will involve will operate more efficiently than any of its current generators, further strengthening Lakeland's financial position, and aiding it in providing cost effective power to its customers. The revenue anticipated from operating the new generator is based on the expected demand from existing customers and is not contingent on any future negotiations or sales to another utility.

### **Project Organization**

The City of Lakeland is anticipating entering into an engineer, procure, construct (EPC) contract with a Foster Wheeler/Westinghouse consortium for the entire McIntosh Unit 4 PCFB project with the exception of certain specific items such as a 90 car unit train that would be handled by Lakeland's staff. Through the execution of a single EPC contract, Lakeland would have a single point of contact and single point of responsibility for all issues associated with the project. In order to assist Lakeland in reviewing and monitoring the performance of the EPC contractor,

Lakeland is in the process of entering into an additional contract with a company who will act as the "Owner's Engineer". This company will safeguard Lakeland's interest on the project and conduct an ongoing prudency review.

In order to obtain the required permits and licenses for the construction and operation of McIntosh Unit 4, the City of Lakeland has retained the services of a qualified environmental consulting firm with particular expertise in the state of Florida. This same firm will be empowered to prepare the necessary information required by DOE to complete the NEPA process and is expected to liaise closely with DOE's chosen NEPA consultant or subcontractor.

### **Project Status**

At the time this paper was written, DOE had recently announced approval of the project and efforts were underway to have all the Cooperative Agreements and related project agreements formally executed by the parties. Completion of this activity will trigger the formal start of Phase 1 of the McIntosh Unit 4 project. In parallel with this activity, the scope of work of each of the project participants, and their role within the project structure, is currently being fine tuned and finalized. The agreements necessary for each project participant to fulfil their project obligations are in the process of being negotiated. Two important project activities that will be initiated shortly are the permitting and NEPA activities.



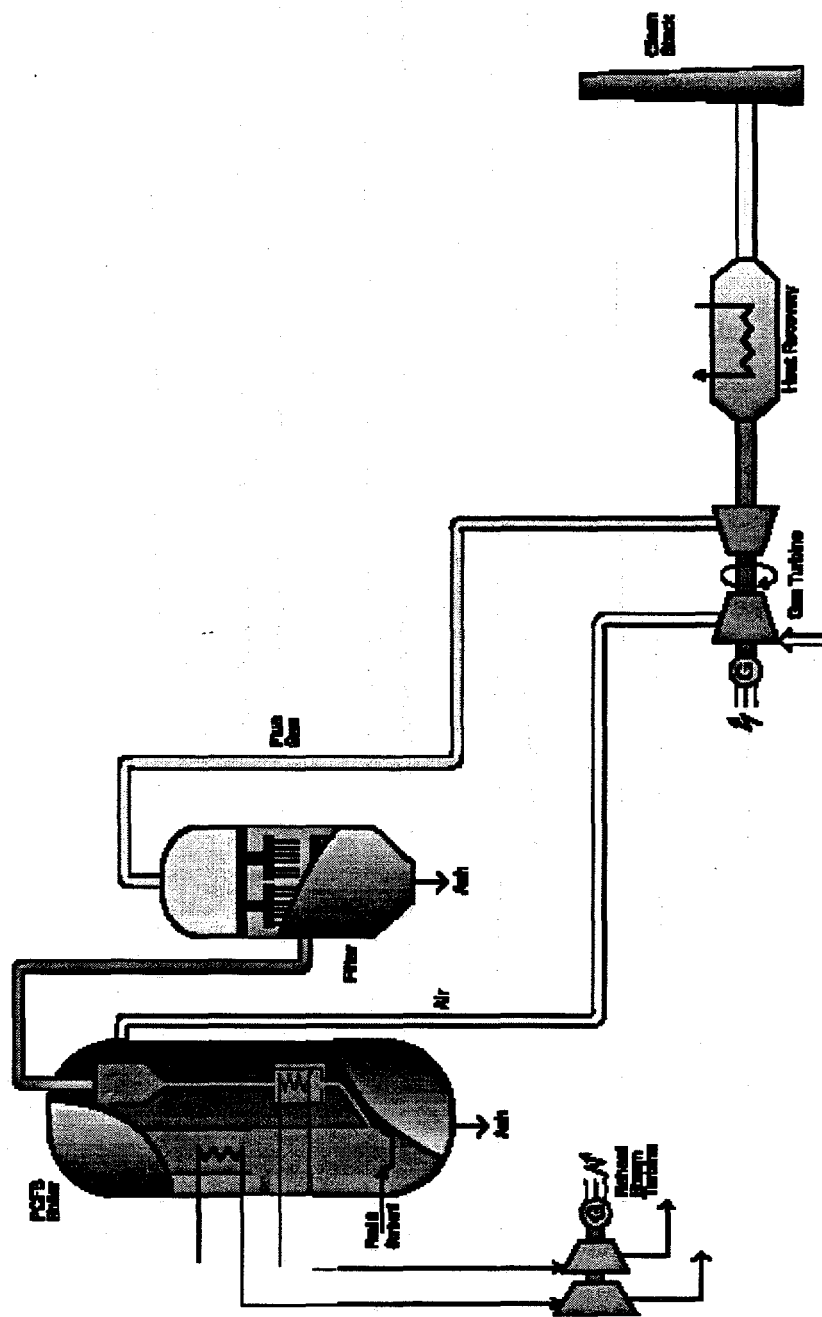


Figure 1.  
PCFB Cycle - 1st Demonstration

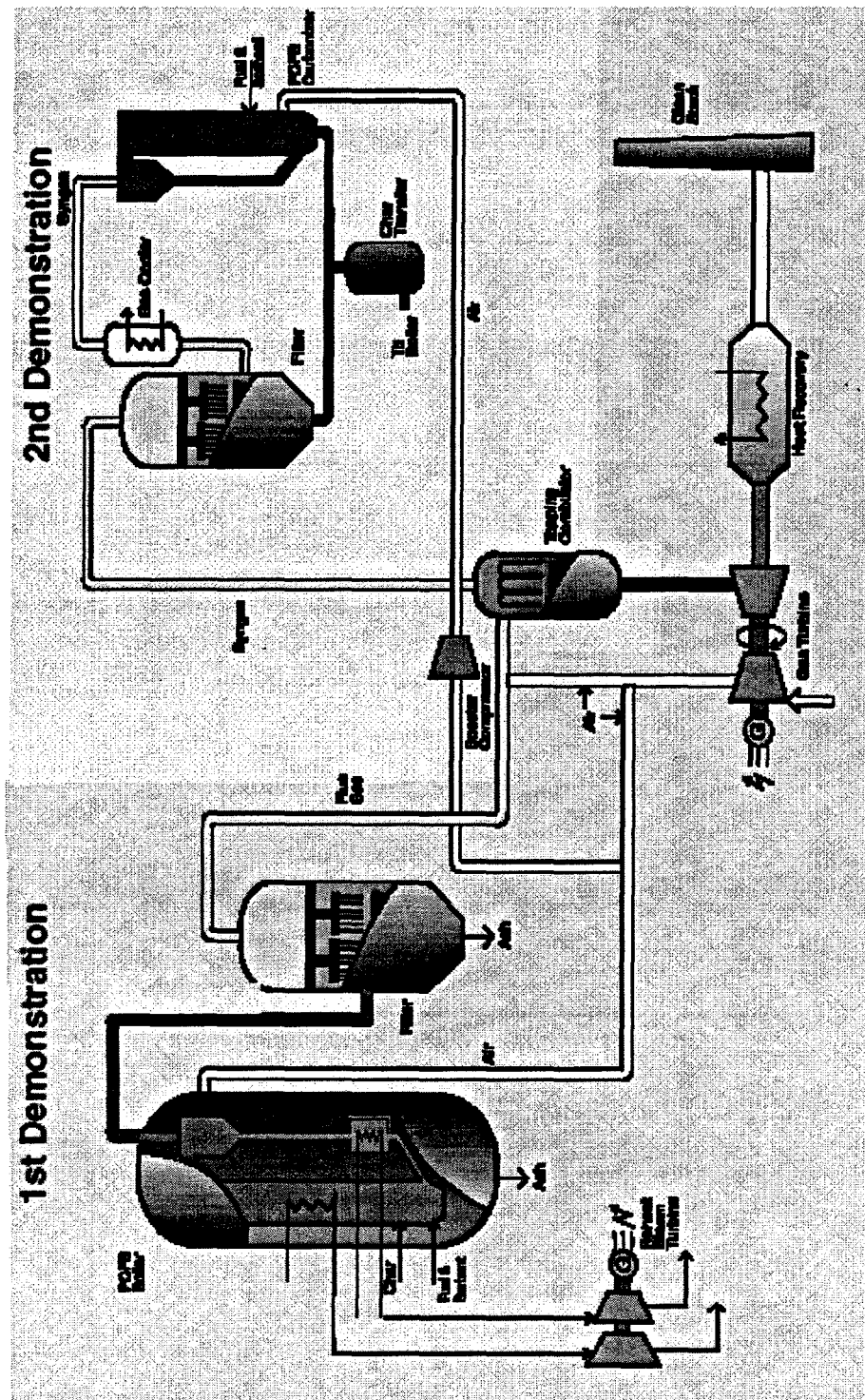
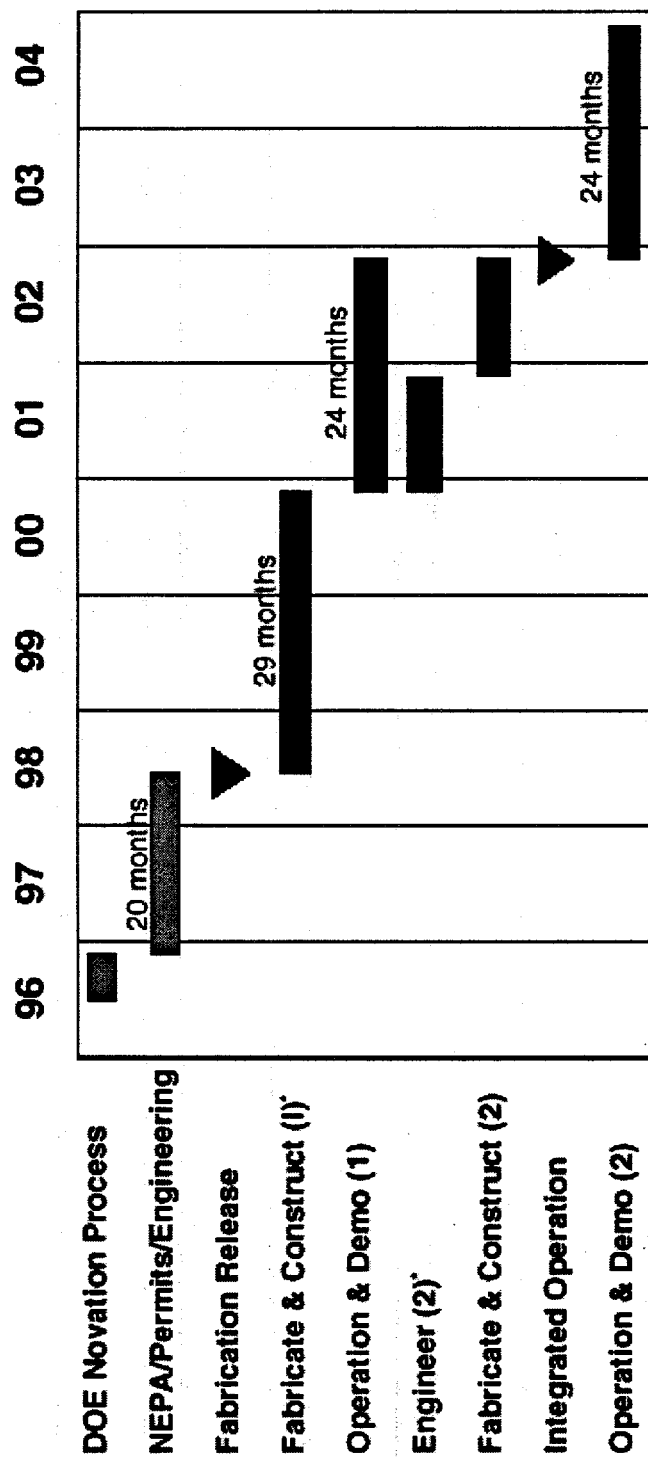


Figure 2  
Topping PCFB Cycle - 2nd Demonstration

**Figure 3**  
**McIntosh Unit 4 Summary Schedule**



\* (1) refers to 1st Demonstration and (2) to 2nd Demonstration

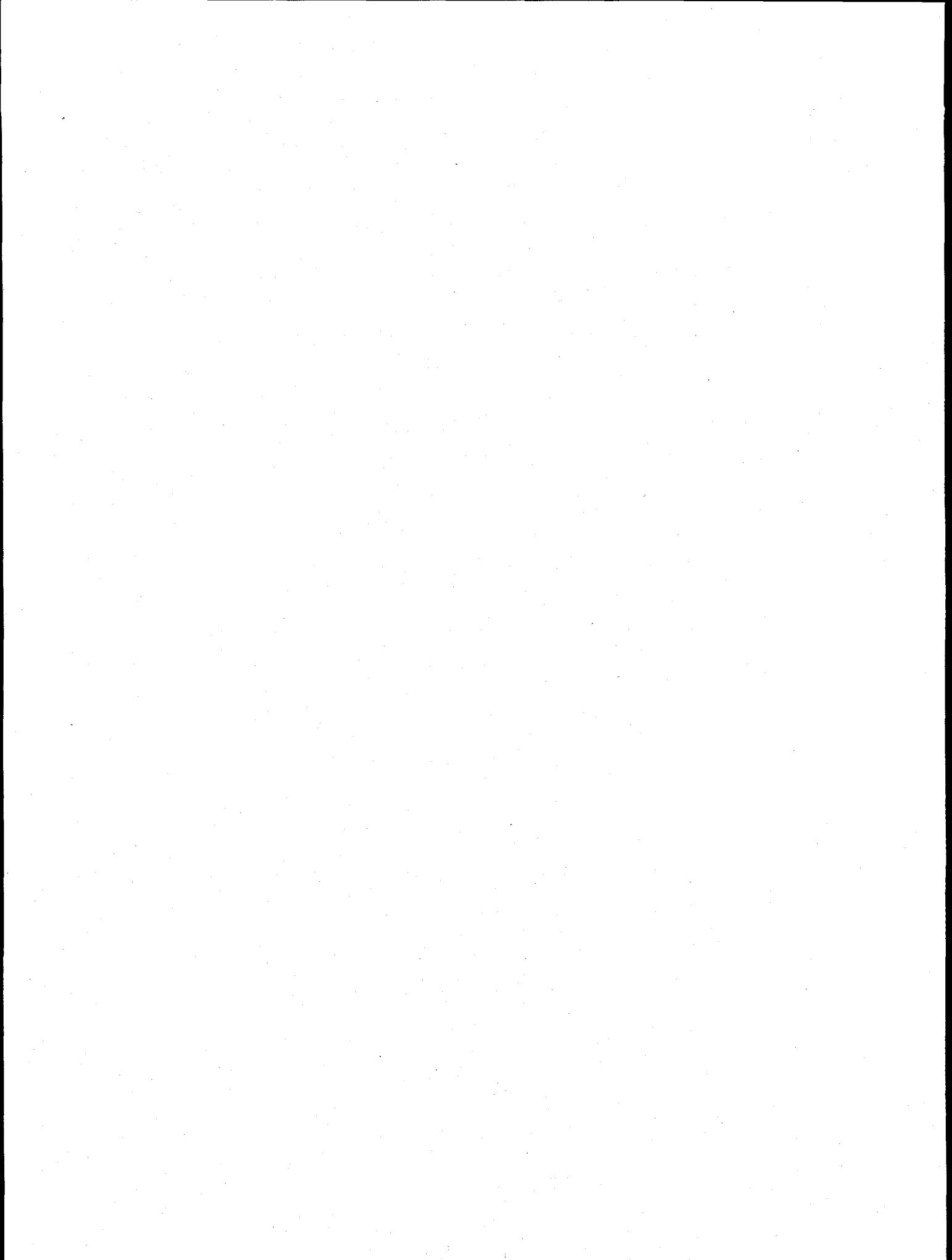
# **Poster Session Abstracts**

# **THE HEALY CLEAN COAL PROJECT AN OVERVIEW**

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## **ABSTRACT**

The Healy Clean Coal Project, selected by the U.S. Department of Energy under Round III of the Clean Coal Technology Program is currently in construction. The project is owned and financed by the Alaska Industrial Development and Export Authority (AIDEA), and is cofunded by the U.S. Department of Energy. Construction is scheduled to be completed in August of 1997, with startup activity concluding in December of 1997. Demonstration, testing and reporting of the results will take place in 1998, followed by commercial operation of the facility. The emission levels of NO<sub>x</sub>, SO<sub>2</sub> and particulates from this 50 megawatt plant are expected to be significantly lower than current standards. The project status, its participants, a description of the technology to be demonstrated, and the operational and performance goals of this project are presented herein.



# Tidd PFBC Demonstration Project

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## ABSTRACT

On March 30, 1995, one of the nation's pioneering Clean Coal Technology Projects -- the Tidd Demonstration Plant<sup>1</sup> in Brilliant, Ohio -- completed its 4-year test run, producing more than 11,500 hours of data for the power industry and establishing the technical foundation for cleaner, more efficient power plants in the 21st century.

The Tidd project was one of the first joint government-industry ventures to be approved by the U.S. Department of Energy (DOE) in its Clean Coal Technology Program. In March 1987, DOE signed an agreement with the Ohio Power Company, a subsidiary of American Electric Power, to refurbish the then-idle Tidd plant on the banks of the Ohio River with advanced "pressurized fluidized bed technology."

Testing ended after 49 months of operation, 100 individual tests, and the generation of more than 500,000 megawatt-hours of electricity. The demonstration plant has met its objectives. The project showed that more than 95 percent of sulfur dioxide pollutants could be removed inside the advanced boiler using the advanced combustion technology, giving future power plants an attractive alternative to expensive, add-on scrubber technology. During its test program, the Tidd Plant earned national honors for its innovative approach for power generation. In 1991, the plant was named Power Magazine's Power Plant of the Year. In 1992, the National Energy Resource Organization presented American Electric Power with a national award for its efforts in promoting energy efficient power technology.

In addition to its sulfur removal effectiveness, the plant's sustained periods of steady-state operation boosted its availability significantly above design projections, heightening confidence that pressurized fluidized bed technology will be a reliable, baseload technology for future power plants. The technology also controlled the release of nitrogen oxides to levels well below the allowable limits set by Federal air quality standards. It also produced a dry waste product that is much easier to handle than wastes from conventional power plants and will likely have commercial value when produced by future power plants.

At the time the 70-megawatt Tidd Plant was built, it represented a 13:1 scaleup from the earlier pilot plant facility. Future commercial PFBC plants will likely be in the 100 to 300 mega-

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<sup>1</sup>Research sponsored by the U.S. Department of Energy's Morgantown Energy Technology Center under Cooperative Agreement No. DE-FC21-87MC24132 with the American Electric Power Service Corporation as agent for Ohio Power Company, 1 Riverside Plaza, Columbus, OH 43215

watt size range and feature efficiencies over 40 percent. More than 50 percent of new capacity added between 2000 and 2010 in the U. S. will be coal-based. High coal market capture rates are also anticipated in the international market. Compared to conventional technology, PFBC will have superior environmental and economic performance and is clearly a technology which will be used to meet the growing electricity demand worldwide.

The Tidd Project also served as the testing station for future devices that can clean unburned particles from the hot combustion gases with minimal losses in efficiency. The DOE used a "slip stream" of hot gases from the boiler to test advanced, ceramic barrier filters. Data acquired during 6,000 hours of operation will help in the design of the hot gas cleanup devices that will be needed as the technology further evolves.

The Tidd Project also gave the U.S. company, The Babcock & Wilcox Company, headquartered in Barberton, Ohio, the opportunity to strengthen its leadership role in developing high-technology boiler systems, demand for which is growing throughout the world.

Total project cost, including design, construction, and operation of the demonstration plant, was nearly \$190 million, with DOE supplying \$67 million, or 35%, and the project's co-sponsors providing nearly \$123 million.

The materials shown at the Poster Session highlight the quantitative results of the testing and the commercial version of this technology at a utility scale. Information about obtaining the Final Report for the project will be available at the Poster Session.

### **Acknowledgments**

The success of this project would not have been possible without the support and expertise of the project co-sponsors -- the State of Ohio and two major technology vendors, ASEA Brown Boveri Carbon and Babcock & Wilcox. The author also wishes to acknowledge the contribution of the Project Mangers Larry Carpenter and Donald Geiling at DOE's Morgantown Energy Technology Center.



# **McIntosh Unit 4 PCFB Demonstration Project**

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Lakeland, Florida**

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Project Manager  
Foster Wheeler Development Corporation  
San Diego, California**

**Harry T. Morehead  
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Westinghouse Electric Corporation  
Orlando, Florida**

## **Introduction**

The City of Lakeland, Foster Wheeler Corporation and Westinghouse Electric Corporation have embarked on a utility scale demonstration of Pressurized Circulating Fluidized Bed (PCFB) technology at Lakeland's McIntosh Power Station in Lakeland, Florida. The U.S. Department of Energy will be providing approximately \$195 million of funding for the project through two Cooperative Agreements under the auspices of the Clean Coal Technology Program. The project will involve the commercial demonstration of FOSTER WHEELER PYROFLOW PCFB technology integrated with Westinghouse's Hot Gas Filter (HGF) and power generation ® technologies.

The total project duration will be approximately eight years and will be structured into three separate phases; two years of design and permitting, followed by an initial period of two years of fabrication and construction and concluding with a four year demonstration (commercial operation) period. It is expected that the project will show that Foster Wheeler's Pyroflow PCFB technology coupled with Westinghouse's HGF and power generation technologies represents a cost effective, high efficiency, low emissions means of adding greenfield generation capacity and that this same technology is also well suited for repowering applications.

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The project is being partially funded under the Clean Coal Technology Program by the US Department of Energy through its Morgantown Energy Technology Center under contracts DE-FC21-91MC27364 and DE-FC21-94MC21261 between DOE and the City of Lakeland.

The paper will provide a general description of the project including its objectives, structure and the roles of the various participants. The technology to be demonstrated will be described together with the project design basis and predicted performance. Current project activities will be discussed and planned future activities will be summarized.

## **Acknowledgment**

The assistance of Mr. Don Geiling, METC's Project Manager for the McIntosh Unit 4 PCFB Demonstration Project, during the preparation and review of this paper is hereby gratefully acknowledged.

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The project is being partially funded under the Clean Coal Technology Program by the US Department of Energy through its Morgantown Energy Technology Center under contracts DE-FC21-91MC27364 and DE-FC21-94MC21261 between DOE and the City of Lakeland.

# Anatomy of an Upgraded Pulverized Coal Facility: Combustion Modification Through Flue Gas Scrubbing

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## ABSTRACT

POWER PLANT ANATOMY 101 Regeneration is a biological term for formation or creating anew. In the case of Milliken station, a species of steam generation (*Tangentus coali*) regeneration refers to refitting critical systems with the latest technological advances to reduce emissions while maintaining or improving performance. The plant has undergone a series of operations which provided an anatomical changes as well as a face lift. Each of the two units were placed in suspended animation (outage) to allow these changes to be made.

The digestive system (combustion) was renewed from the molars to the sphincter the system which grind the food (coal) prior to digestion (combustion) were replaced with efficient, larger and stronger molars. All four molars (coal pulverizers) for each unit replaced with D.B. Riley MPS 150 mills with dynamic classifiers. In order to improve

*The paper describes the project and could be on file*

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<sup>1</sup>Research sponsored by the U.S. Department of Energy's Pittsburgh Energy Technology Center, under contract DE-FCC 92PC92642 with New York State Electric & Gas Corporation, P.O. Box 5224, Binghamton, NY 13902-5224; telefax 607-762-8457.

delivery to the stomach (boiler), a new esophagus (coal piping) has been installed. The stomach lining (boiler wall) has been fitted with ABB LNCFS III firing system which will increase energy and vitality while reducing indigestion and the formation of noxious gas ( $\text{NO}_x$ ).

As with any well operating digestive system, gas and solids are by products of the process. The gas will be handled in a sensitive manner. Before expulsion to the atmosphere it will be conditioned through the intestines (back pass of the boiler, precipitator and scrubber). The small intestine (back pass of the boiler and precipitator) continue the digestive process by recovering additional calories and removing solids. A portion of the small intestine (precipitator) was enlarged to allow for its regeneration from a conventional weights and wires system to a Belco wide spaced ridged frame unit. ABB Air Preheater International is demonstrating a Q-Pipe Airheater for even greater heat recovery on Unit 2.

The digestion products are then passed through the large intestine (scrubber). Through this formic acid enhanced wet limestone process, process gas emissions ( $\text{SO}_2$ ) are reduced and solids are processed into a useable cake (gypsum). Since Milliken was born without a large intestine, the scrubber was the most visible change which allowed for several cosmetic improvements.

The brain (control room) was considered to be A.B. Normal. Major surgery was performed to improve the units logic and memory capabilities. Additional nerve centers (DCS) and nerve sensors were added to improve efficiency, coordination and response time. After enduring all these changes Milliken has been given a TOPAZ system by DHR Technologies for on-line optimization and strategies for least cost plant operation.

Twenty-one reports will be issued prior to completion of testing in 1998. A listing of report titles and sample reports will be available at the poster display.

### **Acknowledgement**

Mr. James U. Watts, DOE Project Manager; Demonstration Team Members: CONSOL Inc., Saarberg-Holter-Umwelttechnik. (SHU), NALCO Fuel Tech., DHR Technologies Inc., Stebbins Engineering, ABB Air Preheater. Project cofunders include NYSEG, CONSOL, Electric Power Research Institute, New York State Energy Research and Development Authority and Empire State Electric Energy Research Corporation. Parsons Power Group is the Architect/Engineer and Construction Manager for the flue gas desulfurization (FGD) retrofit portion of the project.

# PROJECT facts

DEPARTMENT OF ENERGY  
OFFICE OF FOSSIL ENERGY

ADVANCED power  
SYSTEMS

## DB RILEY—LOW EMISSION BOILER SYSTEM (LEBS): SUPERIOR POWER FOR THE 21ST CENTURY

### PRIMARY PROJECT PARTNER

DB Riley, Inc.  
Worcester, MA

### MAIN SITE

Worcester, MA

### TOTAL ESTIMATED COST

\$116,000,000

### COST SHARING

DOE \$42,500,000

Non-DOE \$73,500,000

### Project Description

In conjunction with the U.S. Department of Energy, DB Riley, Inc., is developing a highly advanced coal-fired power-generation plant called the Low Emission Boiler System (LEBS). By the year 2000, LEBS will provide the U.S. electric power industry with a reliable, efficient, cost-effective, environmentally superior alternative to current technologies.

LEBS incorporates significant advances in coal combustion, supercritical steam boiler design, environmental control, and materials development. It employs the combustion expertise of Deutsche-Babcock, the University of Utah, and Reaction Engineering International; the pollution-control experience of Thermo Power Corporation; and the plant design practices of Sargent & Lundy Engineers.

The system will include a state-of-the-art steam cycle operating at supercritical steam conditions; a slagging combustor that produces vitrified ash by-products; low nitrogen oxide (NOx) burners; a new, dry, regenerable flue gas cleanup system (copper oxide process) for simultaneously capturing sulfur dioxide (SO<sub>2</sub>) and nitrogen oxides (NOx); a pulse-jet fabric filter for particulate capture; and a low-temperature heat-recovery system.

The copper oxide flue gas cleanup system, which has been under development at DOE's Pittsburgh field center, removes over 98% of SO<sub>2</sub> and 95% of NOx from flue gas. A new moving-bed design provides efficient sorbent utilization that lowers the cleanup process cost. The captured SO<sub>2</sub> can be converted to valuable by-products such as sulfuric acid and/or elemental sulfur, and the process generates no waste.

### Program Goal

DOE's strategic plan aims not only to ensure a reliable and affordable energy supply for the U.S., but also to minimize adverse environmental impact. The highly advanced coal-fired LEBS will achieve significantly lower emissions and higher plant efficiencies than conventional units. Performance objectives of LEBS include plant thermal efficiencies of 42%; lower emission levels of SO<sub>2</sub>, NOx, and particulates; and a cost of electricity equal to or less than that of conventional coal-fired power plants.

### Project Partners

SARGENT & LUNDY ENGINEERS  
Chicago, IL  
(plant design)

THERMO POWER CORPORATION  
Waltham, MA  
(pollution control)

UNIVERSITY OF UTAH  
Salt Lake City, UT  
(combustion)

REACTION ENGINEERING INTERNATIONAL  
Salt Lake City, UT  
(combustion)

## DB RILEY—LOW EMISSION BOILER SYSTEM (LEBS): SUPERIOR POWER FOR THE 21ST CENTURY

### CONTACT POINTS

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(508) 792-4817 fax  
rileyrb@aol.com

Lawrence A. Ruth  
U.S. Department of Energy  
Pittsburgh, PA  
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(412) 892-5971 fax  
ruth@perc.doe.gov

### Project Benefits

In the near future, the United States will have to build a new generation of coal-based power plants to replace its aging units. Coal supplies more than 56% of the nation's electricity, and, because of our abundant reserves, it will remain the dominant source of fuel for power generation well into the next century. A national cap on sulfur and nitrogen oxide (NOx) emissions, however, will require future coal technologies to be much cleaner than current technology.

DOE is sponsoring the Low Emission Boiler System Program to meet these power and environmental needs. Without significantly departing from the traditional design features of pulverized coal-firing systems, this technology will:

- Reduce sulfur dioxide and NOx emissions to a sixth of the levels allowed by today's Federal air quality standards (New Source Performance Standards).
- Lower emissions of flyash and other particulates to a third of those allowed by today's standards.
- Significantly improve power-plant efficiency—from today's level of 35% up to 48%.
- Produce electricity at costs equal to or less than those of a modern-day coal plant.

LEBS is one of several advanced power generation systems that are being developed with support from DOE. Of these systems, LEBS offers the nearest-term commercial option for utilities to meet these performance goals for new installations. In addition, many of the technologies that are being developed in the LEBS Program will be available for retrofit or repowering applications at existing facilities.

DB Riley, along with ABB-Combustion Engineering and Babcock & Wilcox, are leading independent teams to develop low emission boiler systems that incorporate each team's unique, preferred technologies. In mid-1997, one of the teams will be selected to construct and operate a proof-of-concept (POC) test facility to provide the engineering data for commercializing its system by the year 2000.

### Cost Profile (Dollars in Millions)

	Prior Investment	FY93	FY96	FY97	Future Funds**
Department of Energy*		\$2.1	\$2.0	\$8.5	\$20.0
Private Sector Partners		\$0.8	\$0.7	\$6.9	\$20.0

\* Appropriated funding

\*\* If DB Riley is selected to design, construct, and operate a proof-of-concept power plant, a total of about \$100 million (\$20 million DOE) will be required with \$6 million received in FY97.

### Key Milestones

FY91	FY92	FY93	FY94	FY95	FY96	FY97	FY98	FY99	FY00
Planning and Development				Design and Testing		Design	Construction and Operation		
Concept development Preliminary R&D Component testing Commercial generating unit preliminary design				Subsystem testing Proof-of-concept facility: preliminary design Selection of host site for POC facility		POC facility: revised design Commercial generating unit: revised design	Construction and operation of POC facility		