

**DEMONSTRATION OF BECHTEL'S CONFINED  
ZONE DISPERSION PROCESS AT PENNSYLVANIA  
ELECTRIC COMPANY'S SEWARD STATION  
PROJECT STATUS**

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**Abstract**

The U.S. Department of Energy and Bechtel Corporation are cooperating in a joint project to demonstrate Bechtel's Confined Zone Dispersion (CZD) Technology. The demonstration is being conducted at Pennsylvania Electric Company's (Penelec's) Seward Station on Boiler Unit #15. This boiler is a 147 MWe pulverized coal-fired unit, which utilizes Pennsylvania bituminous coal (approximately 1.2 to 2.5 percent sulfur). One of the two flue gas ducts leading from the boiler was lengthened and retrofitted with the CZD technology. The new long straight duct replaced the original multi-bend duct to ensure a residence time of about 2 seconds. The goal of this demonstration is to prove the technical and economic feasibility of the CZD technology on a commercial scale. The process can achieve 50 percent sulfur dioxide (SO<sub>2</sub>) removal at lower capital and O&M costs than other systems.

The CZD process involves injecting a finely atomized slurry of reactive lime into the flue gas ductwork of a coal-fired utility boiler. The principle of the confined zone is to form a wet zone of slurry droplets in the middle of the duct confined in an envelope of hot gas between the wet zone and the duct walls. The lime slurry reacts with part of the SO<sub>2</sub> in the gas and the reactive products dry to form solid particles. An electrostatic precipitator (ESP), downstream from the point of injection, captures the reactive products along with the fly ash entrained in the flue gas.

The test program is being conducted in two parts. The first part, parametric testing, started on July 10, 1991, and was completed on August 17, 1992. During this period, Bechtel's objective was to carry out a factorial test program to optimize the performance of the CZD process. The test program was designed to develop operating conditions to achieve a highly reliable and low-cost operation. The second part, from August 17, 1992 to August 30, 1993 will complete the project. Based on the results of the parametric test program, Bechtel performed additional design, procurement, installation, and facility construction as necessary to permit a 12-month continuous demonstration. The CZD system is fully instrumented and integrated with the operation of Penelec's Boiler Unit #15. The purpose of the project is to demonstrate the performance of the CZD process for SO<sub>2</sub> removal without significantly affecting either boiler operations or plant particulate emissions. Penelec is operating the system during the continuous demonstration as a normal part of the operation of Boiler Unit #15. Bechtel is supervising the demonstration and carrying out various performance tests, data acquisition, and chemical analyses.

Results of the demonstration indicate that the CZD process can achieve costs of \$300/ton of SO<sub>2</sub> removed. Based on a 500 MWe plant retrofitted with CZD for a 50 percent SO<sub>2</sub> removal rate, the total capital cost is estimated at less than \$30/kWe. The cost includes lime unloading, lime handling, and fully automated operation. The variable operating cost for this retrofit is estimated at less than 3.0 mills/kWh.

## **Introduction**

The CZD process involves flue gas post-treatment. It is located between the boiler outlet and the particulate collector, an ESP in most of cases.

The features that distinguish the Bechtel CZD process from other similar injection processes are the following:

- Injection of an alkaline slurry directly into the duct. Other processes use injection into a conventional spray-dryer vessel or injection of dry solids into the duct ahead of a fabric filter.
- Use of an ultrafine calcium/magnesium hydroxide, Type S pressure-hydrated dolomitic lime. This commercial available product is made from plentiful, naturally occurring dolomite.
- Low residence time, made possible by the highly effective surface area of the Type S lime.
- Localized dispersion of the reagent. Slurry droplets contact only part of the gas, while the droplets are drying, to remove up to 50 percent of the SO<sub>2</sub>. The process uses dual fluid rather than rotary atomizers.

- Improved ESP performance via gas conditioning from the increased water vapor content and lower temperatures. As a result, supplemental conditioning with SO<sub>2</sub> is not necessary for satisfactory removal of particulate matter.

The waste product is composed of magnesium and calcium sulfites and sulfate, with excess lime and fly ash. The waste fly ash mixture usually has pozzolanic properties. The mixture is self-stabilizing because of the excess lime and tends to retain heavy metals in insoluble forms within the fly ash.

### **CZD-Flue Gas Desulfurization (FGD) Demonstration as Part of DOE's Clean Coal Technology Demonstration Program**

The U.S. Department of Energy and Bechtel Corporation have agreed to a cooperative effort to demonstrate the Bechtel-developed CZD technology at Pennsylvania Electric Company's Seward Station. DOE provided half, or \$5.2 million, of the project's total \$10.4 million cost. Others contributing to the project are Pennsylvania Electric Company (\$3 million), Bechtel (\$3.3 million), the Pennsylvania Energy Development Authority (\$750,000), New York State Electric and Gas Corporation (\$100,000), and Rockwell Lime Company (\$23,000). Pennsylvania Electric Company is providing the project's demonstration test site, Seward Station.

The costs of this demonstration have exceeded the \$10.4 million budget, and Bechtel Corporation has contributed an additional \$1.2 million.

Current CZD activities at Seward Station are directed toward demonstrating the best possible atomization and dispersion of the SO<sub>2</sub> absorbing slurry in flue gas and the performance of the existing precipitator to handle the increased dust load without adverse effects on the stack gas opacity.

The CZD project at Seward Station includes replacement of the original flue gas duct (35-foot-long segments connected with 45° elbows and corresponding turning vanes) with one new 110-foot-long straight duct ahead of the ESP.

The test program consists of two distinct periods:

- In the first period, daily factorial runs were conducted to test different atomizers, limes, and slurry concentrations. First period results were used to set and optimize the second period operations.
- In the second period, the performance of a continuously running CZD system is being demonstrated under actual power plant operating conditions. The CZD demonstration is integrated into one half of the flue gas capacity of the commercial unit (147 MW) operating continuously three shifts a day, seven days a week.

## **Purpose of the Test Program**

The primary objectives of the project are to:

- Achieve an SO<sub>2</sub> removal rate of 50 percent
- Realize SO<sub>2</sub> removal costs below \$300/ton
- Eliminate negative effects on normal boiler operations without increasing particulate emissions and opacity

The CZD process has been automated and integrated with only one of the two existing modules of air preheater, flue gas duct, ESP, and induced draft fan associated with Unit No. 15. All auxiliary subsystems, such as lime slurring, degritting, and lime slurry handling, have also been automated.

The demonstration project is permitting optimization of the system for application at different locations by determining the:

- Degree of atomization (slurry/compressed air ratio) versus length of duct required for evaporation of atomized slurry
- Maximum volume of slurry that can be injected per square foot of duct cross section and the confined zone dimensions of the duct cross section that will prevent deposits on duct surfaces
- Effect of flue gas inlet temperature on the evaporation characteristics, SO<sub>2</sub> removal, and alkali utilization

Other objectives of the demonstration project include:

- Performing comparison tests of hydrated calcitic lime and freshly slaked calcitic lime
- Testing methods for improving ESP performance during lime injection by:
  - Monitoring ESP operating and opacity variations during all injection tests
  - Performing particulate emission measurements on several extended runs
- Testing different slurry atomizers to determine the most energy-efficient and erosion-resistant
- Testing the effect of burning higher sulfur coal on the percentage of SO<sub>2</sub>/NO<sub>x</sub> removal

## **Description of the CZD Process**

The spray of lime slurry is injected close to the center of the flue gas duct parallel to the flow of gas. As a cone of spray moves downstream and expands, the gas within the cone cools and its SO<sub>2</sub> is rapidly absorbed by the liquid droplets.

Spray droplets on the outside of the cone mix with the hot gas and dry very rapidly. With the proper choice of slurry concentration and injection rate, drying will be complete before the droplets contact the walls of the duct. The process does require a sufficient length of straight duct downstream from the sprays, estimated at 100 feet, and the gas flow must be reasonably uniform where the spray is injected. Judicious use of the turning vanes, typically installed to minimize pressure drops, makes the gas flow in the bends more uniform.

By carefully positioning lime slurry atomizers, it is possible to obtain a wet zone in the middle of the duct with an envelope of hot gas between the wet zone and the duct walls. This is the principle of the confined zone.

Gas velocity in large ducts is generally about 60 fps at full load, and the flow is highly turbulent. Thus, spray droplets in the expanding cone are transported outward by eddy diffusion. However, since the outward cone's diffusing droplets continuously contact hot gas at about 300°F, they rapidly achieve surface dryness. Exposed to the highly localized full concentration of SO<sub>2</sub>, the lime reacts extremely rapidly.

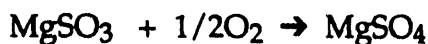
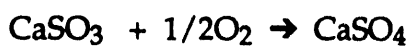
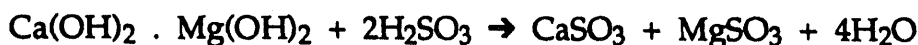
At a certain point downstream, the free moisture in the spray will evaporate completely, and the remaining dry solids can contact surfaces of the duct or the turning vanes without adhering and causing deposits to accumulate.

For removal of 50 percent of the SO<sub>2</sub> from flue gas with a slurry of pressurized hydrated dolomitic lime (PHDL), slurry concentration is a major variable. Enough slurry must be added to achieve the desired results. The demonstration program provides an opportunity to explore and optimize the control of this variable.

The chemical mechanism required for the absorption of SO<sub>2</sub> from the flue gas is simple and very well known. In the presence of water, SO<sub>2</sub> from the flue gas is absorbed as sulfurous acid:



In the presence of water, pressure-hydrated dolomitic lime reacts instantaneously with H<sub>2</sub>SO<sub>3</sub>, producing calcium and magnesium sulfites and sulfates:



### **Past CZD Experience**

Over the last few years, considerable testing of the CZD technology was performed as proof-of-concept on pilot and commercial units. The encouraging results led to the

new phase of demonstrating the process in a commercial unit. References 2, 3, 4, and 5 describe the test programs and the test results of the earlier work.

### Overall System Description

Figure 1 shows a simplified, overall flow diagram of the Seward CZD system and of the Boiler Unit No. 15 flue gas system. The two systems are closely linked. The boiler has twin air and flue gas systems, designated "A" and "B." The CZD system removes SO<sub>2</sub> from the B flue gas stream.

The overall CZD system includes the following process operations and supporting functions:

- Lime slurry preparation
- Lime slurry feed
- Lime slurry injection
- FGD duct (flue gas desulfurization section)
- Atomizing air compression

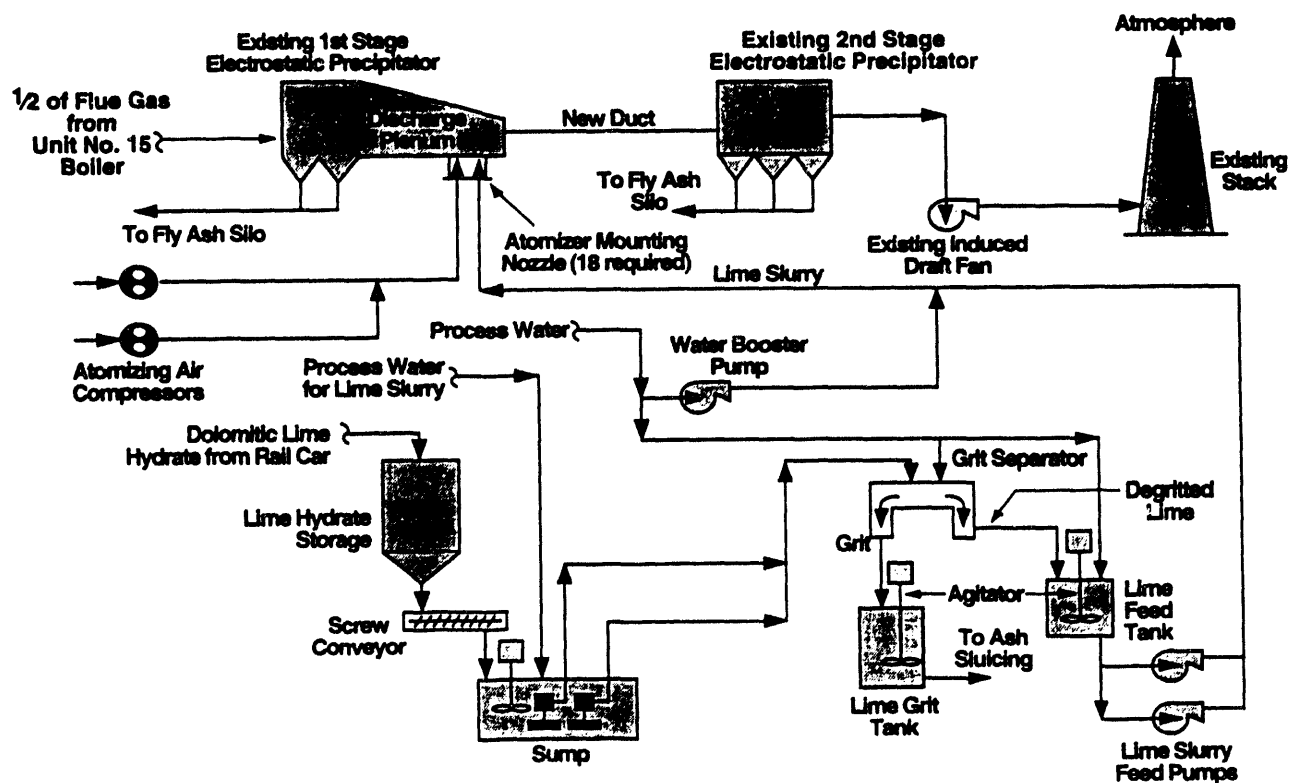


Figure 1 Seward Station Overall Process Flow Diagram

Figure 2 depicts the interrelationship between individual process operations and supporting functions.

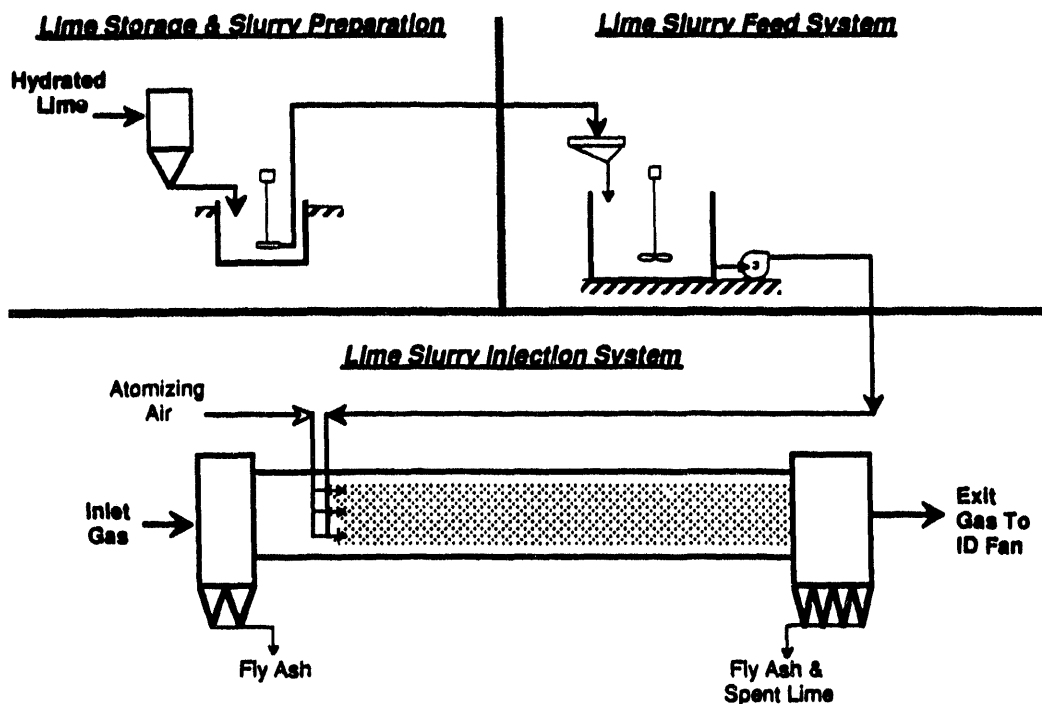


Figure 2 Seward CZD System Overview

### ***Lime Slurry Preparation***

The lime slurry preparation system contains:

- A 50-ton lime silo for receiving and storing lime hydrate, with a vent baghouse filter
- A 5,000-gallon lime hydrate slurring sump with an agitator
- A rotary air lock valve driven by a variable-speed motor and a screw conveyor for transferring the lime hydrate from the lime silo to the sump
- Two sump pumps, one working and one spare, for transferring the lime slurry to the CZD feed system

### ***Receiving and Storing Lime Hydrate***

The lime silo has enough capacity for one day of lime usage. Consequently, daily deliveries of lime are necessary. This silo was recently upgraded for use in the CZD system. Its vent baghouse filter was fitted with new bags and its high- and low-level probes were provided with high- and low-level alarms. The rotary air discharge valve for this silo was equipped with a variable-speed motor for controlling the discharge on the hydrate to the lime slurring sump. The speed of this rotary valve is controlled by the slurry sump density controller.

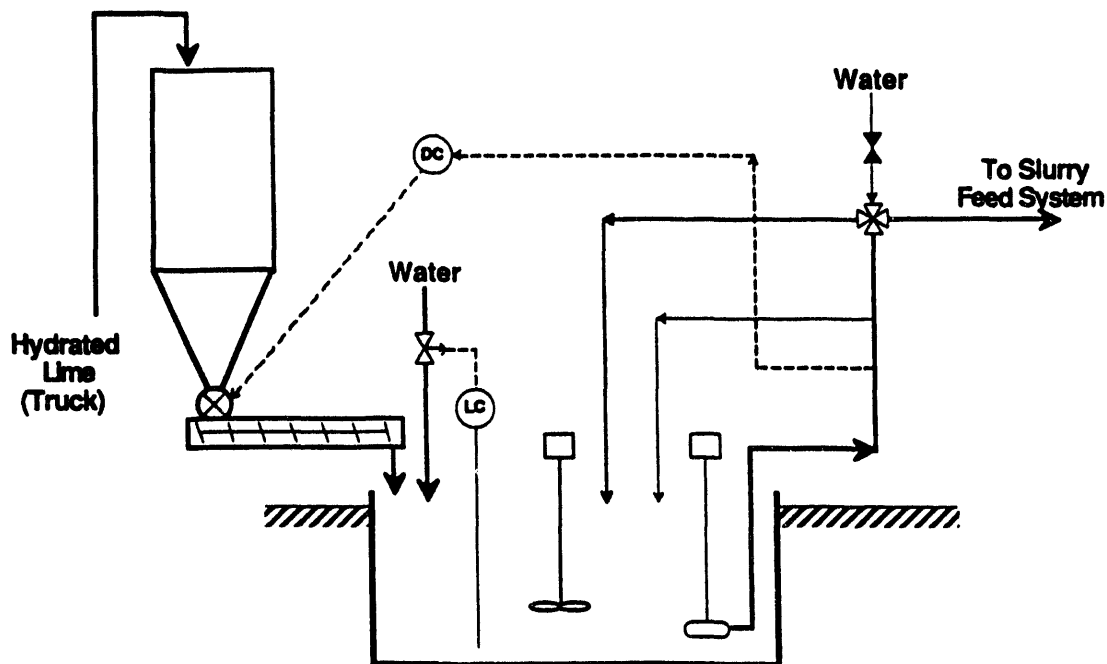
### **Slurrying of Lime Hydrate in the Lime Sump**

The lime slurrying system was designed for fully automatic operation governed by the level controller in the lime feed tank. One of the two sump pumps operate continuously, pumping the lime slurry to the CZD lime feed system. The slurry level in the tank governs the transfer of lime slurry from the sump to the lime feed tank. The tank level controller tends to maintain a constant level in this tank by the operation of a lime flow control valve in the lime transfer line from the sump to the vibrating screen. As the transfer of the lime slurry varies, the lime slurry level in the lime sump also varies.

The lime sump is equipped with a level controller designed to maintain a constant level of slurry in the sump by controlling the sump's water inflow.

The lime sump pump bypass is equipped with a lime slurry density controller which maintains a constant concentration of lime slurry in the sump by controlling the discharge rate of lime hydrate from the silo (speed of rotation of the air lock discharge valve).

Figure 3 shows the lime storage and slurry preparation required for continuous operation.



**Figure 3 Lime Storage and Slurry Preparation – Continuous Operation**



## Lime Slurry Feed

The lime slurry feed system consists of:

- One vibrating screen for the removal of oversized materials from the lime slurry
- Two grits slurry tanks, one working and one spare, both equipped with agitators and level indicators
- Two lime slurry feed tanks, one working and one spare, both equipped with agitators, level controllers, and temperature indicators
- Two lime slurry feed pumps, one working and one spare

The system is designed for intermittent parametric testing as well as continuous plant operation; hence it has double tankage.

The vibrating screen is designed to degrit the lime slurry and is used for the removal of foreign matter from this slurry (sand, trash, etc.). Foreign material drops from the vibrating screen into the collecting gutter from which it is sluiced with water into the grits tank.

The filtered lime slurry is discharged from the vibrating screen into the lime slurry feed tank. The slurry level in the tank is controlled by the tank level controller, which throttles the flow of lime slurry from the lime slurry sump pump to the vibrating screen. The lime slurry feed pump is used to pump the lime from the feed tank to the lime slurry injection header. Figure 4 shows the lime slurry feed system - continuous operation.

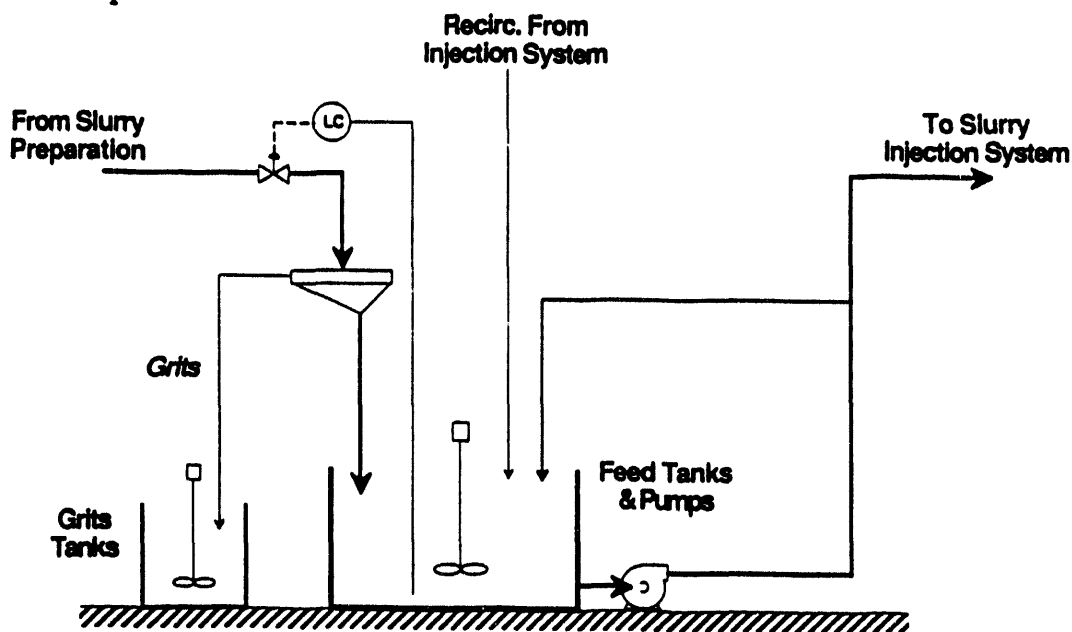


Figure 4 Lime Slurry Feed System - Continuous Operation

### ***Lime Slurry Injection***

The lime slurry injection system consists of:

- The lime slurry and water piping
- The flow controls on the top of the desulfurization duct (other than the lime slurry distribution header and atomizer feeders)
- The water booster pump and associated water piping at ground level

Lime slurry is supplied to the injection lime header from the lime feed system via the loop main, which consists of the feed supply and the excess feed return headers. Operation of the atomizers requires relatively high, constant, lime slurry injection pressure. This pressure is maintained at a constant level at the inlet to the injection header by the back pressure controller in the lime slurry return header. The flow of the lime slurry to the atomizers' distribution header is controlled by a flow controller which is reset by the Section C temperature controller. In the duct, Section C is in the turning vanes area.

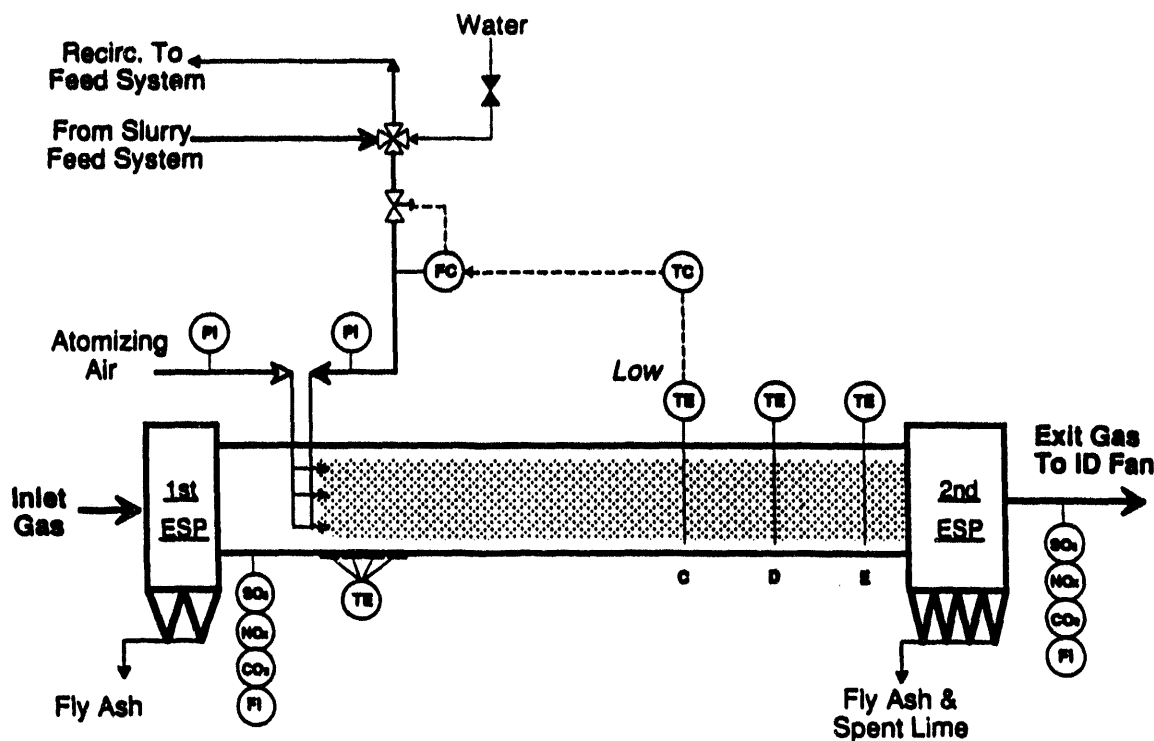
The lime slurry injection header is connected to the lime slurry feed loop via a four-way valve. This valve connects the lime injection header to the water supply piping from the water booster pump. The use of the four-way valve permits the lime injection header to be flushed with water whenever the lime injection is interrupted. The water supply header is furnished with a water flow controller that can be reset by the Section C temperature controller. Both the lime and water flow controller valves are connected to the low-pressure switch on the atomizing air supply header so that the operating flow control valve(s) will close in case of low atomizing air pressure. The arrangement protects the flue gas handling system from being flooded with unatomized lime slurry or water.

The power plant domestic water distribution system provides water for flushing the atomizers and their lime slurry supply piping and for injection into the flue gas stream. Because the operating pressure of this system is inadequate for the operation of atomizers, the CZD injection system is equipped with a water booster pump to maintain an adequate water supply pressure.

Figure 5 shows the lime slurry injection, continuous operation.

### ***Flue Gas Desulfurization Section***

Seward Boiler Unit No. 15 is a balanced draft boiler provided with two F.D. fans, two Ljungstrom air heaters, two twin-chamber ESPs, and two I.D. fans. The two ESPs are joined by twin flue gas ducts that form twin flue gas treating trains, referred to as A and B trains.



**Figure 5 Lime Slurry Injection – Continuous Operation**

The desulfurization duct has a 110-foot-long straight run for injection of the atomized lime slurry. This duct length is necessary for the boiler unit for the absorption of  $\text{SO}_2$  from flue gas and for drying out the absorption products. The atomizing nozzles are located at the duct inlet.

The operating instrumentation includes a low-pressure switch which will stop lime or water from being injected if the air pressure is too low to ensure adequate atomization. This instrumentation is essential for the protection of the flue gas system from the formation of wet deposits, plugging, and flooding.

A ready/standby switching system allows the lime slurry feed to the atomizers to be diverted back to the feed tanks, while water is supplied to flush the atomizing nozzles and lime supply header. The ready/standby system can be used to temporarily suspend lime injection without shutting down the CZD system and can be activated from the plant control room.

### **Atomizing Air Compression System**

This system contains two screw-type air compressors (which can be operated singly or in parallel) and an air receiver. Each of the two compressors can supply up to 2,250 scfm of air at 120 psig and is driven by a 500 bhp motor. Each compressor is equipped with air intercoolers and after-coolers using 100 gpm of cooling water. The

compressors are of the oilless type and provide oil-free compressed air. The operation of the CZD system requires continuous operation of at least one of the two compressors.

### ***Instrumentation and Control (I&C) System***

Instrumentation and control (I&C) is broken down according to the plant locations at which CZD equipment and systems are found. These five operational areas/systems are:

- Lime slurry preparation system
- Lime slurry feed system
- Lime injection system
- Duct flue gas monitoring and controls
- Atomizing air compression system

Startup, operation, and monitoring of the equipment and systems within these areas are accomplished by a combination of actions performed locally or in the control room (remote operation). In general, initial startup of all pumps, mixers, and systems must occur locally. In this way the operator can visually verify the condition of the equipment in the area and determine whether it is safe to put the equipment or system into operation. Once a system or equipment is in operation, monitoring the condition of equipment and changing the system setpoints can be done remotely in the control room, or locally through panel-mounted switches and controllers.

The ready/standby system is also part of the CZD I&C and operates through the Bailey Distributed Control System (DCS). The ready/standby switch gives the operator a means of controlling whether or not lime slurry is injected into the duct without unnecessarily upsetting CZD controls, and safeguards the operation of the Buell ESP. Low atomizing air pressure also activates the standby mode of operation.

Remote monitoring and control of the CZD process from the control room are provided by the existing combustion management control system (MCS) and are supplemented by the process control view station (PCV). Additional plant and process operating information is available from Leeds & Northrup (L&N) recorders located in the ESP control room and in the duct B instrumentation room.

### **Parametric Test Results**

The parametric tests included duct injecting atomized lime slurry made of dry hydrated calcitic lime, fresh slaked calcitic lime, and pressure-hydrated dolomitic lime. All three reagents removed SO<sub>2</sub> from the flue gas, requiring different concentrations in the lime slurry for the same percentage of SO<sub>2</sub> removal. The most efficient and easiest to operate is the pressure-hydrated dolomitic lime. The lime

slurry duct injection does not adversely impact the stack opacity. On the contrary, it substantially reduces stack opacity during the lime injection.

### **Continuous Operation Results**

Table 1 shows typical results when using pressure-hydrated dolomitic lime.

**Table 1 Typical Results  
Continuous Demonstration Tests with Pressure-Hydrated Dolomitic Lime (PHDL)**

Date	8/21/92	8/21/92	8/24/92	8/26/92	8/26/92	8/27/92	11/12/92
Item	21:15	22:15	18:15	14:45	21:15	5:45	15:30
Boiler Load, MW	139.5	144.5	143.5	144.4	142.8	142.6	120.4
Stack Opacity, percent	7.8	7.8	11.5	6.8	10.2	5.9	14.2
Flue Gas Temp. in, °F	305.2	304.4	309.8	310.2	304.2	305.4	284.7
Flue Gas Temp. out, °F	192.1	190.9	191.1	197.8	190.7	194.1	197.3
SO <sub>2</sub> in, ppm	910	910	869	905	944	901	703
SO <sub>2</sub> out, ppm	333	351	343	401	368	399	377
Lime Slurry Flow Rate, gpm	55.8	54.9	58.9	52.4	55.3	54	37.7
Lime Slurry Conc., percent	9.0	9.0	12.0	6.4	12	6.5	10
SO <sub>2</sub> Removed, percent	55.3	53.1	50	48	51.2	45.3	39.3
Lime Utilization, percent	41.0	40.1	23.7	49.8	27.8	46.4	25.3

It is clear that a high lime slurry injection rate is required to achieve a high percentage of SO<sub>2</sub> removal. Following are the conditions that limit the injection rate:

- The flue gas flow rate and temperature
- The residence time in the straight duct

The design of the CZD system at Seward Station is based on the following assumptions:

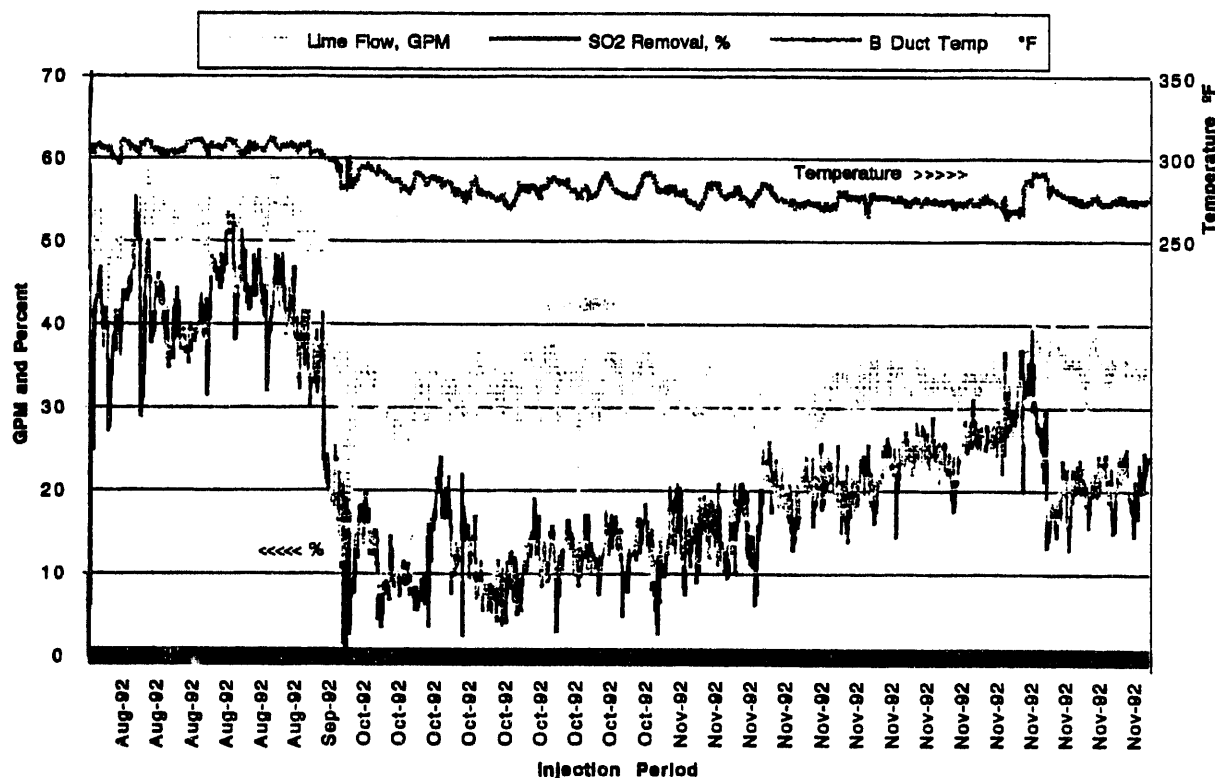
- An inlet flue gas temperature in the B-duct of 300°F
- A boiler load of 147 MW, equivalent to a flue gas velocity in the duct of 55-66 ft/sec
- A 110-foot-long straight duct, equivalent to 2 seconds' residence time

During the second half of August 1992, the flue gas temperature was 300-310°F, the boiler load was 145-147 MW, the residence time in the duct was 2 seconds, and the injection of lime slurry was 52-57 gpm with SO<sub>2</sub> removal above 50 percent.

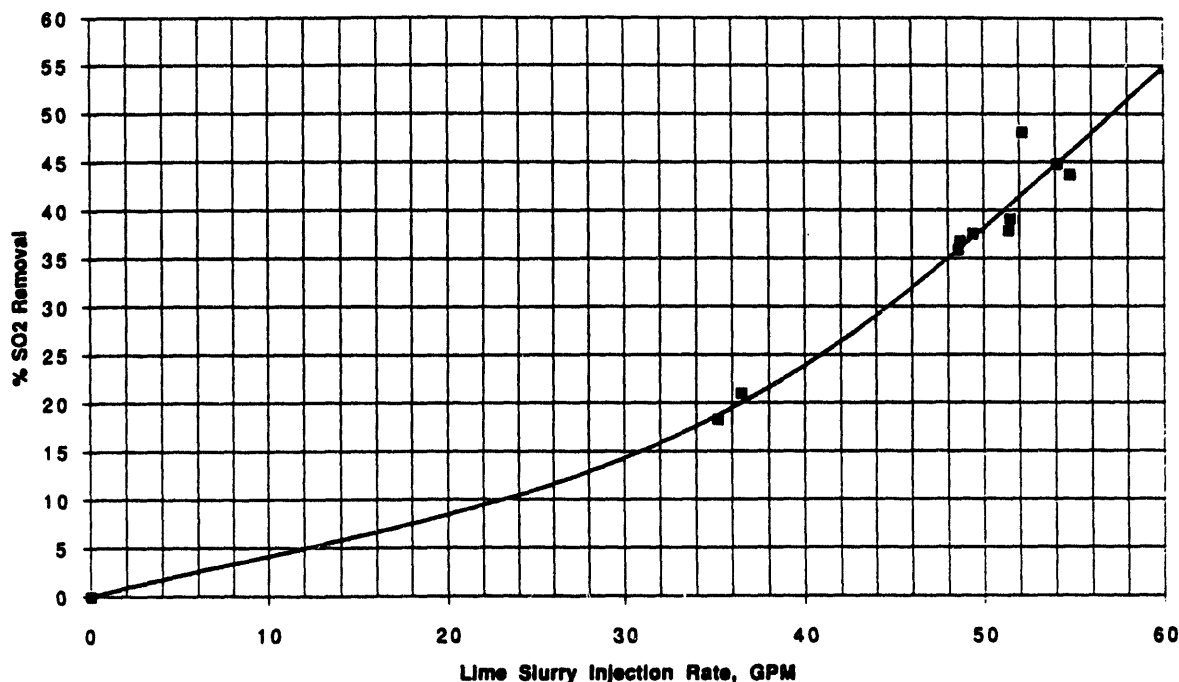
Figure 6 shows the CZD duct temperature versus the lime slurry injection rate and percentage of SO<sub>2</sub> removal during the period August through November 1992. In order to include the duct inlet temperature in this graph, it has been divided by 5 – i.e., 300°F is represented by 60°F. The graph shows that during the months of October and November, the inlet flue gas temperature continued to descend from above 300°F to 260-280°F due to air leakage in the boiler air heater. This low temperature limited the lime slurry injection rate to 30-40 gpm and resulted in a 15-30 percent variation in the percentage of SO<sub>2</sub> removal.

The percentage of SO<sub>2</sub> removal is dependent on the lime slurry injection rate and that the percentage of lime slurry concentration above a specific level does not affect the percentage of SO<sub>2</sub> removal.

Figure 7 shows SO<sub>2</sub> removal versus the lime slurry injection rate, expressed in gpm and based on daily averages during the period August 17 to September 16, 1992. This graph shows that for a low lime slurry injection rate — for example, 36 gpm — the percent of SO<sub>2</sub> removal was only about 20 percent. By increasing the injection rate to 50 gpm, SO<sub>2</sub> removal increased to 38 percent. At a 54 gpm injection rate, the SO<sub>2</sub> removal rate increased to 45 percent. By extrapolation, with a 60 gpm injection rate, the percentage of SO<sub>2</sub> removed will reach 55 percent. As mentioned above, the percentage of SO<sub>2</sub> removed represents daily averages and not peak values.



**Figure 6 Temperature vs Lime Flow and SO<sub>2</sub> Removal August 17 to November 16, 1992**



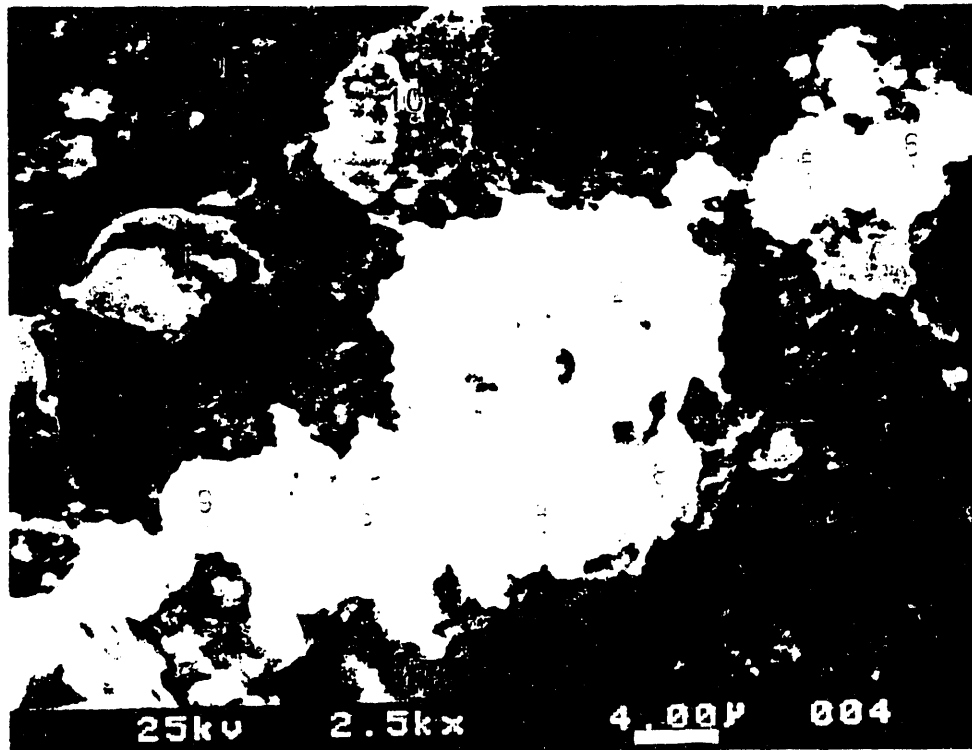
**Figure 7 Removal vs Injection Rate, Based on August 17 to September 16, 1992 Daily Rates**

### ***Chemical Analyses of Fly Ash and CZD Reacted Products***

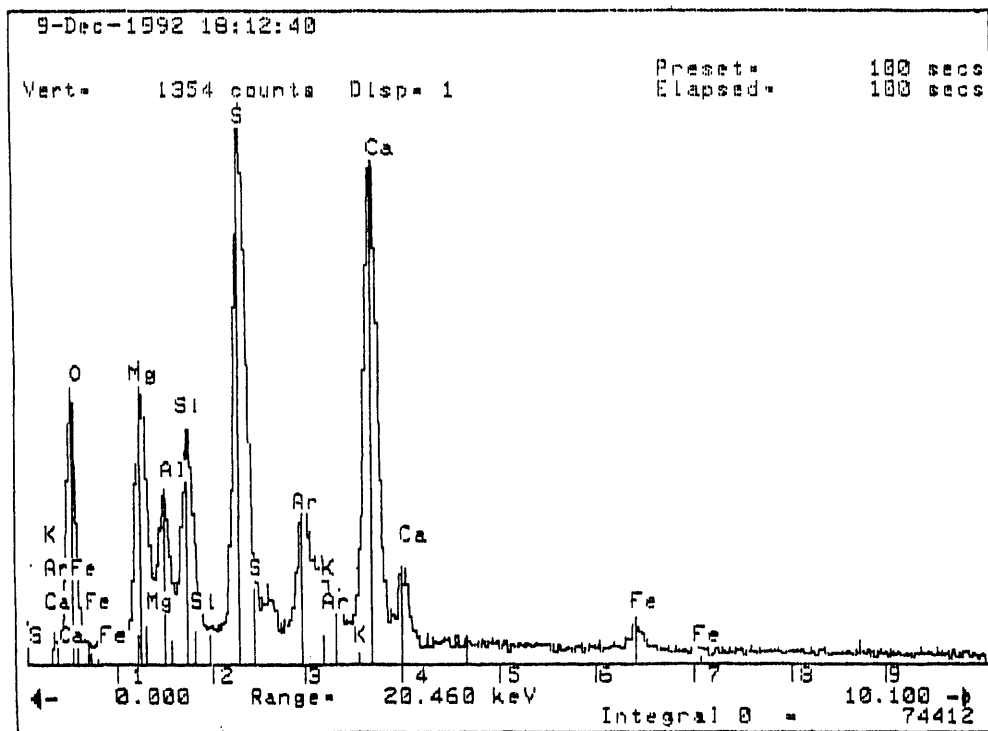
Samples of solids containing fly ash, CZD reacted products, and unreacted lime were collected and analyzed. The samples were collected during the CZD continuous operation with pressure-hydrated dolomitic lime. The analyses were made by scanning electron micrographs (SEM) and energy-dispersive X-ray analyses (EDAX) for elements in fly ash and CZD reacted products. The analyses were performed at Pennsylvania State University, Department of Material Science and Engineering, Fuel Science Program.

Figure 8 is an SEM and Figure 9 is an EDAX of a sample of fly ash with CZD reaction products. The fly ash particles on the SEM appear as gray spheres 1-20 microns in diameter. The reaction products (calcium sulfite, calcium sulfate, and magnesium sulfate) appear as separate white crystals from submicron size to about 2 microns in diameter. The very fine crystals of CZD reaction products agglomerated on the larger fly ash spheres, creating larger particles that are more easily removed by the ESP from the flue gas.

The EDAX analyses show the main elements and their concentration. Position 5, shown in Figure 9, shows calcium, magnesium, sulfur and oxygen as principal elements of the reaction products. Other positions show silica, aluminum, iron, and titanium as main elements (the fly ash constituents), and calcium, magnesium, and sulfur (reaction-product constituents) at lower concentrations.



**Figure 8 Scanning Electron Micrograph of Samples of Fly Ash, CZD Reaction Products, and Unreacted Lime**



**Figure 9 Energy-Dispersive X-ray Analysis of Position 5 In Figure 8**



### ***Percentage of Lime Utilization and the Factors that Could Affect It***

The percentage of lime utilization in the CZD system significantly affects the total cost per ton of SO<sub>2</sub> removed. For this reason, we have examined methods that can improve the percentage of lime utilization. An analysis of the continuous operational data compiled to date indicates that the percentage of lime utilization is directly dependent on two factors:

- The percentage of SO<sub>2</sub> removal
- The lime slurry concentration

For operating conditions at Seward Station, data show that a 40-50 percent SO<sub>2</sub> removal and a 6-8 percent lime or dolomitic lime slurry concentration will assure a 40-50 percent lime utilization rate. That is, 2 to 2.5 moles of CaO or CaO.MgO are required for every mole of SO<sub>2</sub> removed; or assuming 92 percent lime purity, 1.9-2.4 tons of lime are required for every ton of SO<sub>2</sub> removed.

### **Technology Applicability and Limitations**

#### ***Commercial Application***

CZD technology is particularly well suited for retrofitting existing boilers, regardless of type, age, and size, type of coal burned, or the percentage of sulfur in the coal. Compared to currently available flue gas desulfurization systems, CZD technology can be more easily and economically integrated into existing power plants.

The inherent advantages of the CZD process relative to currently available commercial technologies are:

- Substantially lower capital cost and total cost per ton of SO<sub>2</sub> removed
- Ease of retrofit because CZD eliminates the need for chimney alterations, boiler reinforcements, and modifications to boiler draft controls
- No increase in flue gas pressure drop; therefore, no extra fans are needed
- Minimal space requirements in the stack area
- No dewatering or liquid waste treatment required
- No flue gas reheating requirement
- No congestion close to the boiler or stack
- Easily disposable reaction products that are dry, free flowing, and are removed with the fly ash
- Reduced labor and maintenance requirements
- Applicable to wide range of: geographic locations, load profiles, and particulate collectors (either ESP or baghouse)

- Raw materials (dolomitic rock is well distributed in many areas of the United States)

A disadvantage of the CZD process is the limitation in SO<sub>2</sub> removal to about 50 percent.

The CZD process requires that drying and SO<sub>2</sub> absorption take place within 2 seconds. A long straight flue gas duct of about 100 feet is required to ensure a residence time of 2 seconds. If this length of flue gas duct is not available and there are space limitations for a new horizontal long duct, then a vertical duct of the required length can be built to enable utilization of the CZD technology.

Reference 6 describes flue gas desulfurization by the CZD process on a comparative basis with the economies of other clean coal technologies.

### ***Commercial Demonstration***

The CZD project is designed to demonstrate:

- Reliable operation of the CZD process when integrated with a power station
- Absence of any detrimental effect on normal boiler operations
- Capability to operate with high- and low-sulfur coal

Bechtel intends to commercialize the CZD process when the present demonstration is successfully concluded. During the CZD demonstration, papers giving technical and economic data, results, and conclusions, will be presented at various conferences. These papers will be made available for publication in appropriate journals of technical societies, the electric generating industry, and in other publications. Representatives of utilities will be invited to visit the demonstration site and learn how SO<sub>2</sub> can be removed cost-effectively using the CZD process.

### **Conclusions**

- Parametric tests with key variables were clearly determined and used as a basis for the continuous operation system
- The fully automated and integrated system with the power plant operation proved that the CZD process responds very well to automated control operations
- CZD system availability is very good. For example, during the period October 17 to November 16, 1992, the automated systems were on line 100% of time - 744 hours

- During normal CZD operations, no deposits of fly ash and reaction products took place in the flue gas duct. On a few occasions, deposits accumulated in the duct and had to be manually removed. The failures that caused these deposits were identified as follows:
  - A power failure on the air compressor supplying the compressed air to the lime slurry atomizers
  - Breakage of the ceramic tips, causing coarse sprays
  - Leakage of the lime slurry atomizers around their flanges and spray tips, causing the formation of deposits on the atomizer tips

These problems were corrected, and we expect that they will not be repeated. A new control system will monitor the operation of each atomizer and will stop operations if one of them is not operating correctly.

- At Seward Station, stack opacity is not detrimentally affected by the CZD system
- Results of the demonstration indicate that the CZD process can achieve costs of \$300/ton of SO<sub>2</sub> removed when operating a 500 MW unit burning 4% sulfur coal. Based on a 500 MWe plant retrofitted with CZD for a 50 percent rate of SO<sub>2</sub> removal, the total capital cost is estimated at less than \$30/KWe

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**PROJECT UPDATE: ADVANCED FGD DESIGN FOR NORTHERN INDIANA  
PUBLIC SERVICE COMPANY'S BAILLY GENERATING STATION**

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Atlanta, Georgia**

**Abstract**

**In 1988, the U.S. Department of Energy selected Pure Air to demonstrate a state-of-the-art sulfur dioxide scrubber, under DOE's Clean Coal Technology demonstration program. The scrubber is located at the Northern Indiana Public Service Company's (NIPSCO's) Bailly Generating Station, approximately 40 miles southeast of the city of Chicago, Illinois in the U.S.A. It boasts a number of advanced technical features, including a single absorber vessel which is sized nominally at 600 MWe - the largest in the United States, and the largest in the world to operate with high-sulfur coal. Other project features are:**

- high-velocity co-current scrubber design, which contributes to reduced space requirements;**
- direct injection of dry, pulverized limestone into the scrubber;**
- production of a high quality gypsum by-product, instead of disposable scrubber sludge;**
- wastewater treatment and evaporation system, and;**

- the first-of-a-kind agreement (Own-and-Operate) to design, engineer, construct, fabricate, own, operate, maintain and finance a power plant scrubber, by a non-utility.

These and other features allow the scrubber to have improved environmental performance, reduced space requirements, better energy efficiency, and lower costs than conventional first (or second) generation scrubbers. With specific regard to environmental management, this project seeks to demonstrate that air pollution control need not have deleterious solid waste and/or wastewater consequences.

Construction of the scrubber is complete; operations began in June 1992, ahead of schedule and within budget. The Clean Coal demonstration project calls for three years of operations. After the three-year demonstration period, Pure Air will continue to Own-and-Operate the scrubber for the next 17 years.

This paper reviews the advanced wet flue gas desulfurization (FGD) design features, and the environmental and business features of the project. This paper also includes operations data, project costs and schedule.

## **Background**

Pure Air began development efforts in early 1988 for an Own-and-Operate Advanced FGD facility serving NIPSCO. With the cooperation of NIPSCO, the project was submitted to the United States Department of Energy for consideration under the Innovative Clean Coal Technology Program (Solicitation II).

Pure Air's "Advanced Flue Gas Desulfurization (AFGD) Demonstration Project" was one of 16 projects selected under DOE's second Clean Coal solicitation. Following selection in September 1988, DOE entered into negotiations with Pure Air, and a Cooperative Agreement was awarded on 20 December 1990.

In September 1989, a flue gas processing agreement was signed with NIPSCO, whereby an AFGD facility would be constructed at its Bailly Generating Station located in Porter, Indiana, forty miles east of Chicago on the southern shore of Lake Michigan. The project has since progressed through design and construction, and into operation. Indiana law required that NIPSCO obtain from the Indiana Utility Regulatory Commission (IURC) a "Certificate of Public Convenience and Necessity". This law was originally written to create incentives for utilities to install clean coal technology. The Certificate process was initiated on 15 August 1989 and a certificate was granted on 11 April 1990. In doing so, the IURC concurred that by installing the AFGD facility at Bailly, NIPSCO was using the most cost-effective solution for SO<sub>2</sub> reduction.

Bailly Station consists of two coal-fired boilers -- Unit #7, rated at 183 MW net capacity, went into operation in 1962; and Unit #8, rated at 345 MW net capacity, started in 1968. Both units burn bituminous coal, typically ranging from 2 to 4 percent sulfur.

First, a brief description of the plant and the installed FGD system (i.e., absorber design, limestone feed system, gypsum dewatering and wastewater treatment system) are outlined. Next, a summary of the results of the performance testing completed on the AFGD system is presented. Also included is the DOE demonstration test matrix and results of the first DOE test conducted in September 1992, along with relevant operations information.

## **Plant Description**

A simplified process flow diagram for the Bailly Advanced FGD plant is illustrated in Figure 1.

This facility includes a single absorber designed to process the maximum quantity of flue gas generated from both NIPSCO's Units #7 and 8. The absorber is equipped with a co-current open grid tower with two levels of slurry distribution, an integrated reaction tank and a two-stage vertical mist eliminator in the horizontal duct. The absorber is designed to accomplish several process steps, including gas quenching, absorption of SO<sub>2</sub>, reaction with limestone and oxidation of sulfite to gypsum in a single vessel. The co-current absorber is designed at higher than conventional counter-current flue gas flow rates which makes the co-current vessel very compact. The flue gas from both boilers (approximately 2,200,000 scfm) through the existing I.D. fans is combined in a common duct feeding a single absorber tower where it contacts the recirculation slurry. Quenching and humidification of the gas and absorption of the SO<sub>2</sub> occur simultaneously. The zone where the flue gas initially contacts the recirculation slurry is called the "wet-dry" interface and is washed intermittently with fresh water to prevent the formation and growth of deposits.

Recirculating slurry is split between the two levels of distribution provided. The slurry and flue gas pass co-currently over the open grid packing located in the absorber tower. The grid packing, made of polypropylene, provides a large surface area for gas-to-liquid contact to enhance the SO<sub>2</sub> removal efficiency. The SO<sub>2</sub> of the flue gas is absorbed into the slurry, and the amount of gas phase SO<sub>2</sub> reaction is reduced as the gas flows through the tower. Partial oxidation of the absorbed SO<sub>2</sub> occurs in the tower. Oxidation of the SO<sub>2</sub> is completed in the reaction tank. After flowing downward through the absorber tower, the scrubbed flue gas makes a 90° turn and passes over the liquid in the reaction tank where gas-liquid disengagement occurs. The gas exits the absorber by passing through a highly efficient, two-stage mist eliminator located vertically in the horizontal outlet duct, where the cleaned flue gas exits through a newly built stack. The reaction tank is located beneath the absorber tower, so the recirculating slurry with absorbed SO<sub>2</sub>

falls directly into the tank. The reaction tank is designed to hold an adequate liquid volume to ensure efficient usage of limestone and to provide adequate residence time for complete oxidation of calcium sulfite to calcium sulfate.

Three air rotary spargers (ARSs) are provided to maximize the efficiency of the oxidation of calcium sulfite. The patented ARS is an innovative and unique design which combines the process functions of agitation and oxidization. By simultaneous rotation of the ARS and the introduction of air through the sparging nozzles, very fine bubbles are produced.

In addition to the three ARSs, a small section of the absorber reaction tank is also fitted with two rows of fixed air spargers (FAS). Incursion of the FAS helps ensure complete oxidation at maximum levels of coal sulfur.

To neutralize the absorber slurry, dry pulverized limestone is pneumatically conveyed to the absorber reaction tank. The SO<sub>2</sub> content in the flue gas at the stack is monitored and controlled by regulating the quantity of limestone injection into the reaction tank.

During humidification of the flue gas, water is consumed from the reaction tank by evaporation. To ensure that the solids concentration in the reaction tank is maintained at 25 percent, slurry is transferred from the reaction tank to the gypsum dewatering equipment.

## **Summary of Project Features**

### **Large, Single-Module Scrubber**

The AFGD facility is sized to scrub all of the flue gases from both of the Bailly Station's two coal-fired units. Unit 7 has a nameplate capacity of 194 MWe, and is permitted at 183 MWe. Unit 8 has a nameplate capacity of 422 MWe and is permitted at 345 MWe. Thus, the AFGD facility is currently operating at 528 MWe (or 1,420,000 SCFM); but it is sized to handle Bailly's nameplate capacity of 616 MWe, if the power plant is re-permitted. This point is important to keep in mind, when performing economic assessments.

With a nominal capacity of 600 MWe, the Bailly scrubber is the largest single SO<sub>2</sub> absorber in the United States. Under the Clean Air Amendments of 1970, as amended by the Clean Air Act Amendments of 1977, newly constructed power plants were required to scrub SO<sub>2</sub> emissions; otherwise, they would not be allowed to operate. As a result, the conventional U.S. practice during the 1970s and early 1980s was to install several small absorber modules, including one or two spare/backup modules. (For example, four 175 MWe modules might be installed at a 600 MWe power plant.)



The AFGD facility at Bailly Station is the first large (i.e., >500 MWe) single-module absorber in the United States, and the largest in the world to operate with high-sulfur coal. It has no spare or backup modules, and treats all of the combined flue gases from Units 7 and 8 at the Bailly generating station.

### **High Velocity, Co-Current SO<sub>2</sub> Absorber**

As the elimination of spare/backup modules serves to reduce the scrubber's cost and relative size, so does its advanced design. Figure 1 is a simplified flow diagram of the AFGD system at Bailly Station. Pure Air utilizes a co-current absorber, in which the limestone slurry moves in the same direction as the power plant flue gases. This enables the Bailly scrubber to operate at a flue gas velocity of approximately 18 to 20 feet/second, versus 8 to 10 feet/second for a more conventional counter-current scrubber. As a result of the higher flue gas velocity, the absorber vessel is smaller, with commensurate cost savings.

### **Direct Limestone Injection**

At the Bailly scrubber, pulverized limestone is injected directly into the SO<sub>2</sub> absorber. This direct injection of pulverized limestone eliminates the need for on-site wet grinding systems, thereby reducing both space requirements and capital costs. Direct limestone injection is particularly attractive at power plants with limited space availability. The Bailly scrubber is one of only a few scrubbers in the world to utilize direct limestone injection, and the first in the United States. The direct limestone injection system has operated without problems to date.

### **High Quality Gypsum By-Product**

Conventional first and second generation scrubbers produced a mixture of calcium sulfite (CaSO<sub>3</sub>) and calcium sulfate (gypsum, CaSO<sub>4</sub>). This mixture is commonly called "scrubber sludge". It caused scaling problems in many early scrubbers, and must be stabilized prior to its disposal as a solid waste.

Most of today's scrubbers either inhibit the oxidation of CaSO<sub>3</sub> or force the oxidation to gypsum, because scaling problems are generally not encountered with either pure CaSO<sub>3</sub> or pure gypsum. Forced oxidation to gypsum has a potential advantage over inhibited oxidation in that the gypsum by-product can often be utilized commercially, depending upon market conditions.

For the AFGD project, NIPSCO has entered into a long-term contract with U.S. Gypsum (USG), whereby USG is purchasing all of the by-product gypsum for use as a feedstock at its East Chicago wallboard manufacturing plant. Wallboard-grade gypsum specifications are shown in Table 1, along with the Bailly gypsum characteristics measured during a 100-hour performance test in August 1992. From start-up through the end of June 1993, the AFGD project produced 216,344 tons of wallboard-grade gypsum.

Note that scrubber gypsum can be used in other applications such as cement, and the gypsum purity requirement is generally lower for cement than for wallboard.

### **Wastewater Treatment and Evaporation Systems**

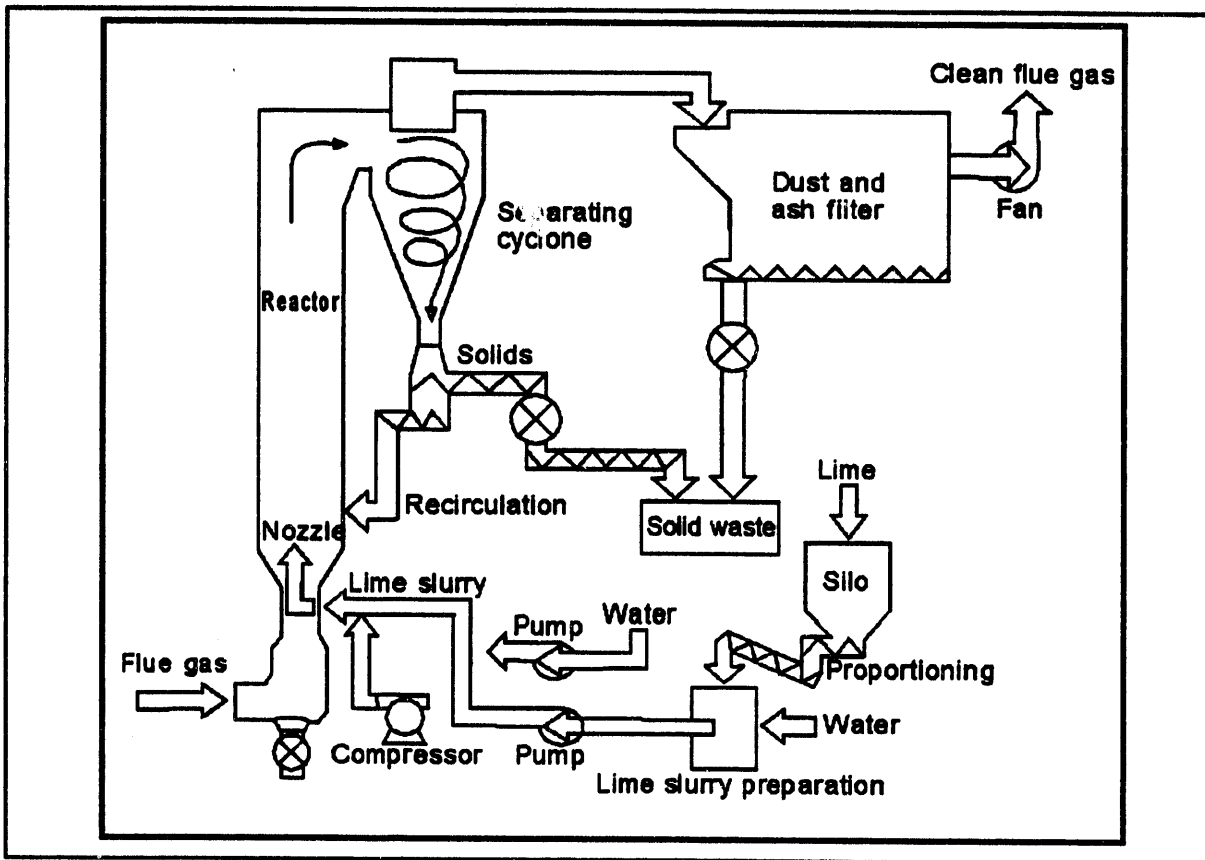
The Bailly generating station is situated on the southern shore of Lake Michigan, and the AFGD facility utilizes process water taken from the lake. Much, though not all, of the process water is recycled within the AFGD system. Treated wastewater is discharged into Lake Michigan. Wastewater requirements are shown in Table 2, along with measured wastewater characteristics at the Bailly Station.

Chloride content is a critical parameter for wallboard-grade gypsum. Removal of chlorides from the gypsum can be accomplished easily by washing the gypsum, but the resultant wastewater can often exceed permit requirements. To avoid this potential problem, Pure Air will demonstrate a novel Wastewater Evaporation System (WES). Part of the process water stream is bled off to maintain an acceptable chloride level within the absorber vessel, and then injected into the power plant ductwork where hot flue gases evaporate the water. Upon evaporation of the water, any dissolved salts will crystallize, so that they can be collected along with fly ash by the power plant's particulate control devices. The salts are then easily disposed of with the power plant fly ash.

At the Bailly Station, the WES will be demonstrated only on Unit 8 (422 MWe nameplate; 345 MWe permit capacity). Taken together, gypsum utilization and wastewater evaporation will demonstrate that SO<sub>2</sub> control need not have deleterious solid waste and/or wastewater consequences.

### **PowerChip™ Gypsum Demonstration**

The AFGD by-product gypsum is in a finely powdered form. However, Pure Air will demonstrate a process to agglomerate and flake part of the by-product gypsum stream, in an attempt to improve the marketability of scrubber gypsum to end-users who are accustomed to using natural gypsum rock. This PowerChip gypsum can be transported more easily and handled with existing equipment at most wallboard and/or cement plants. As an add-on to the AFGD project, Pure Air will attempt to blend fly ash and wastewater treatment solids into the PowerChip gypsum by-product. Although these impurities would make the gypsum unacceptable for wallboard applications, it could still be used in cement. Pilot tests have indicated that maximum fly ash loadings of 20% to 30% may be achieved. In combination with wastewater evaporation and the coproduction of wallboard-grade gypsum, this process may bring coal-fired power generation technology one step closer to the goal of a zero-discharge power plant. Demonstration of the PowerChip gypsum process is scheduled to begin in mid-1993.



**Figure 1. Gas Suspension Absorption Process Flow Diagram**

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 flyash. During the drying process in the reactor, the moisture in the fresh lime slurry, which coats the outer surface of the 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 then 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.

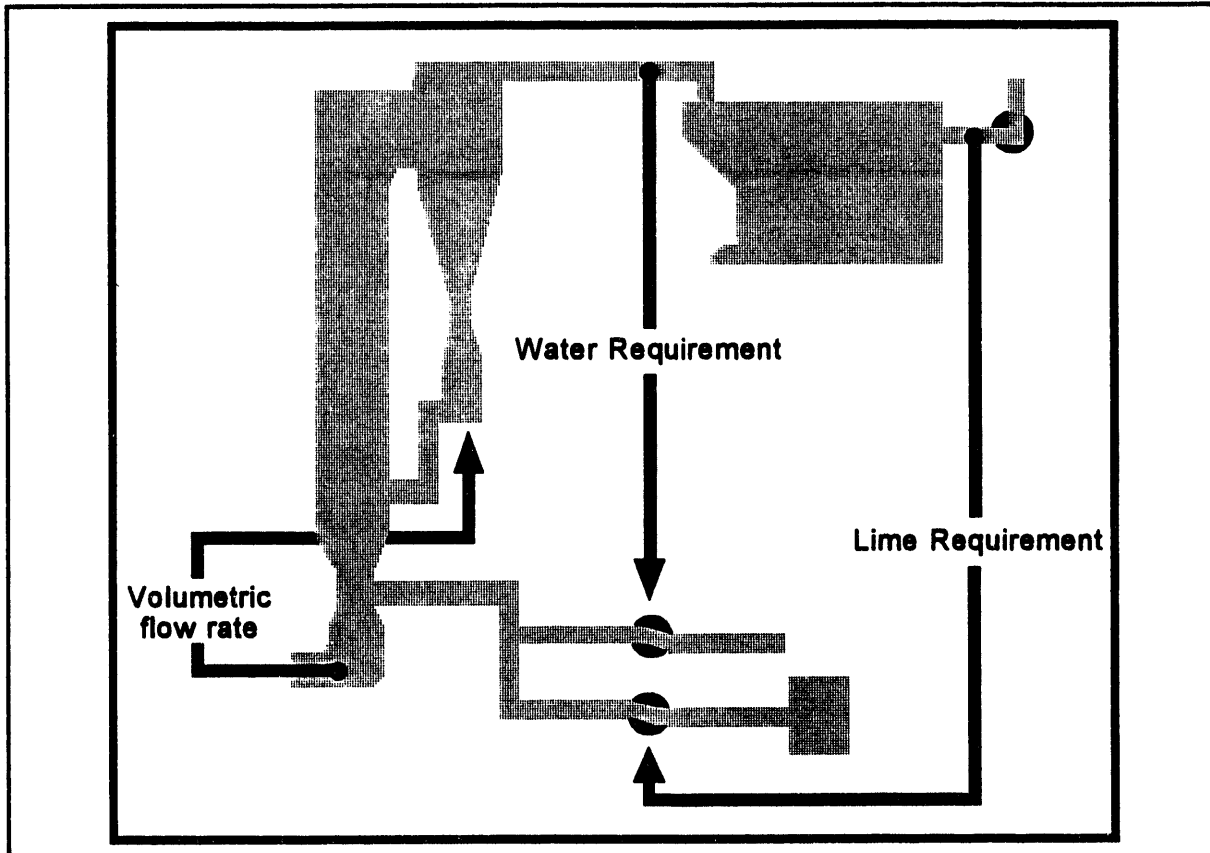
The pebble lime is slaked in a conventional, off-the-shelf system. The resulting fresh slaked lime slurry is pumped to an interim storage tank and then to the dual-fluid nozzle. The slurry is diluted with trim water prior to being injected into the reactor.

### Automatic Process Adjustment

An effective monitoring and control system automatically ensures that the required level of SO<sub>2</sub> 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, provides 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 SO<sub>2</sub> 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**

## **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 the GSA and the competing technologies lies in how the reagent is introduced and used for SO<sub>2</sub> 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 SO<sub>2</sub> in an "umbrella" of finely atomized slurry with a droplet size of about 50 microns,
- May require multiple nozzle heads or rotary atomizer 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 SO<sub>2</sub> 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 limitation 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 problem 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 that 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 the acid gas removal in the GSA system and helps it achieve SO<sub>2</sub> removal levels which are comparable to a wet scrubber.

### 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 the 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, more than adequate 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 and 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 NCER 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 NCER will be evaluated later in the test program.

### **PROJECT STATUS AND KEY MILESTONES**

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

<b>Phase I - Engineering and Design</b>	<b>Start - End</b>
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
<b>Phase II - Procurement and Construction</b>	
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
<b>Phase III - Operating and Testing</b>	
3.1 Project Management	10/01/92-02/28/94
3.2 Start-up and Training	10/01/92-10/14/92
3.3 Testing and Reporting	10/15/92-02/28/94

The progress of this project has been on or ahead of schedule. The parametric optimizing tests are scheduled for completion in August 1993. Following the air toxics testing, which is scheduled to be conducted in September, there will be a one month around-the-clock demonstration run.

## **TEST PLAN**

A test plan was prepared to depict in detail the procedures, locations, and analytical methods to be used in the tests. The following objectives are expected to be 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 will be accomplished through the completion of a statistically-designed factorial test plan. For each test series, the GSA system is set to operate at a certain combination of operating parameters. The results of these test series are analyzed statistically to determine the impact of the operating parameters, thus arriving at the optimum operating point for the GSA process at the various operating conditions expected in future applications. Operating parameters that may be varied in different test series for process optimization purposes are the following:

- Inlet flue gas flow rate
- Inlet SO<sub>2</sub> concentration (dependent on availability of different coal)
- Inlet flue gas temperature
- Inlet dust loading
- Solids recirculation rate
- Stoichiometric ratio
- Approach-to-saturation temperature
- Coal chloride level

Preliminary tests were conducted prior to the factorial testing to determine the ranges that these parameters could be varied. The results from the preliminary testing were used as the basis for the design of the factorial test program.

### Data Collection

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

- Inlet flue gas flow into the system
- SO<sub>2</sub> and HCl 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

### 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 NCER. The results from several of the preliminary tests completed at the NCER in November and December were very interesting. During one of these 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.

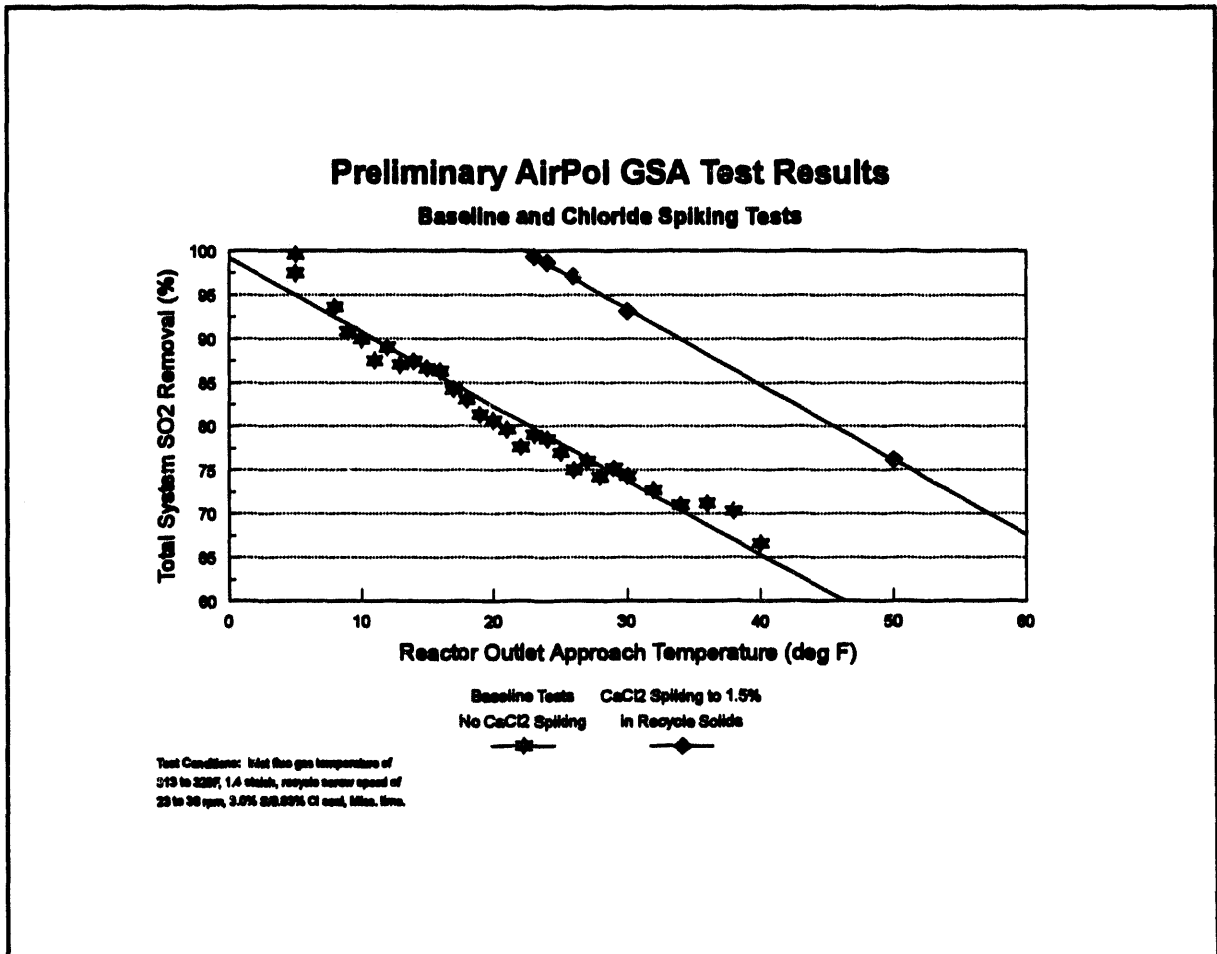


Figure 3. Preliminary AirPol GSA Test Results

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 the 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. 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 NCER 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 preliminary results from this second test are also shown in Figure 3 above.

The overall system SO<sub>2</sub> removal efficiency 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.

## **Factorial Testing**

The primary focus of the recent GSA testing was the completion of the statistically-designed factorial test program. The purpose of this factorial testing was to determine the effect of the 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, flyash loading, coal chloride level, flue gas flow rate, and recycle screw speed. Two levels were determined for nearly all of the variables and these variables and levels are shown in the table below. The one exception was the approach-to-saturation temperature where three levels were defined, but the third level was only run for those tests at the lower coal chloride level.

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

**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 is decreased at the higher chloride levels. However, based on previous test work at the NCER, 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.

## **RESULTS OF FACTORIAL TESTING**

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

The overall system SO<sub>2</sub> removal efficiency results from these factorial tests are currently being analyzed, but several general relationships have become apparent. First, as was expected based on the previous testing at the NCER, 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), 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%). 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 is substantially less SO<sub>2</sub> removal in the ESP than in the previous testing at the NCER, but the overall system SO<sub>2</sub> removal efficiencies appear to be higher 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 Ca(OH)<sub>2</sub>/mole inlet SO<sub>2</sub>, seems to have the most significant effect on the SO<sub>2</sub> 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 SO<sub>2</sub> 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 NCER is a relatively modern, 4-field unit with 10 in. plate spacing, similar in design to several full-scale ESPs installed on the TVA Power System. This unit has 23-ft.-high plates with 8 parallel gas passages. The specific collection area (SCA) of this unit is about 440 ft<sup>2</sup>/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 ft<sup>2</sup>/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 particulate (0.03 lb/MBtu) at all of the test conditions evaluated. 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 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+% 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.

### Pulse Jet Baghouse Performance

Although not part of the original GSA project, TVA and EPRI had cofunded the installation of a 1-MWe PJBH pilot plant at the NCER 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 in. 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

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 and 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+% 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.

### Demonstration Run

Based on the findings during the factorial testing, the GSA system will be operated at optimum settings for a four-week consecutive period of around-the-clock operation to demonstrate the reliability of the system operation as well as its SO<sub>2</sub> removal capability. During the demonstration run, all controls will be switched to automatic mode with set points determined from the optimizing tests.

## **COMMERCIALIZATION**

One of the objectives of this demonstration project is 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 ready to be commercialized for the industry in order to meet the Phase II Clean Air Act Amendments 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. An effort was also made during the design phase to achieve simplicity in the equipment design, which later proved to contribute to reduced material and construction costs. Another major effort being made at AirPol now is to scale up the GSA design to accommodate a utility plant up to 200 MWe with a single GSA reactor. Having gained the confidence that the GSA system is capable of achieving the required levels of performance, the current effort being made at AirPol is to develop standard design of scale-up units. Meanwhile, field operating experience and findings continue to help perfect the design process.

## **DISCUSSION**

As of September 1993, the design, fabrication, installation, and performance optimization of the GSA system for the Clean Coal Technology demonstration project will have been completed. AirPol will have successfully demonstrated the technical performance of the GSA FGD process.

It is expected that the results of the air toxics test will confirm that GSA is also capable of removing heavy metals such as mercury, cadmium, and lead. As this demonstration program is coming close to its completion, it can be concluded now that the GSA process is a viable solution to the SO<sub>2</sub> removal problem of coal-fired boiler plants, and AirPol is ready to offer the technology for commercial application.

## **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.

# **LIFAC SORBENT INJECTION FOR FLUE GAS DESULFURIZATION**

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## **Abstract**

This paper discusses the demonstration of LIFAC sorbent injection technology at Richmond Power and Light's (RP&L) Whitewater Valley Unit No. 2 under the auspices of the U.S. Department of Energy's (DOE) Clean Coal program. LIFAC is a sorbent injection technology capable of removing 75 to 85 percent of a powerplant's SO<sub>2</sub> emissions using limestone at calcium to sulfur molar ratios of between 2 and 2.5 to 1. The site of the demonstration is a coal-fired electric utility powerplant located in Richmond, Indiana, which is between Indianapolis, Indiana and Dayton, Ohio. The project is being conducted by LIFAC North America, a joint venture partnership of Tampella Power Corporation and ICF Kaiser Engineers, in cooperation with DOE, RP&L, and several other organizations including the Electric Power Research Institute (EPRI), the State of Indiana, and Black Beauty Coal Company.

## **Introduction**

The Clean Coal Technology Program (CCT) has been recognized in the National Energy Strategy as a major initiative whereby coal will be able to reach its full potential as a source of energy for the nation and the international marketplace. Attainment of this goal depends upon the development of highly efficient, environmentally sound, competitive coal utilization technologies responsive to diverse energy markets and varied consumer needs. The CCT Program is an effort jointly funded by government and industry whereby the most promising of the advanced coal-based technologies are being moved into the marketplace through demonstration. The CCT Program is being implemented through a total of five competitive solicitations. This paper discusses the LIFAC sorbent injection technology which was selected in the third round of CCT solicitations.

LIFAC North America, a joint venture partnership of ICF Kaiser Engineers, Inc. and Tampella Power Corporation of Finland, will demonstrate the LIFAC flue gas desulfurization technology developed by Tampella. This technology provides sulfur dioxide emission control for powerplants, especially existing facilities with tight space limitations. Sulfur dioxide emissions are expected to be reduced by up to 85% by using limestone as a sorbent. The limestone is injected into the upper regions of a furnace, where calcining to lime and partial absorption of SO<sub>2</sub> occur. Subsequently, the combustion gas is passed through a unique piece of equipment known as the activation reactor. This is a vertical elongation of the ductwork between the air preheater and ESP where the combustion gas is humidified and SO<sub>2</sub> absorption is completed. The LIFAC technology will be demonstrated at Whitewater Valley Unit No. 2, a 60-MWe coal-fired powerplant owned and operated by Richmond Power and Light (RP&L) and located in Richmond, Indiana. The Whitewater plant consumes high-sulfur coals with sulfur contents ranging from 2.0 - 2.9 percent.

The project, co-funded by LIFAC North America and DOE, is being conducted with the participation of Richmond Power and Light, the State of Indiana, the Electric Power Research Institute, and the Black Beauty Coal Company. The project has a total cost of 21.4 million dollars and a duration of 48 months from the preliminary design phase through the testing program.

The sponsors of this project believe that LIFAC has the potential to be a new and important SO<sub>2</sub> control option for U.S. utilities subject to the Clean Air Act's acid rain regulations. To be considered as a commercially feasible option in this particular emissions control market, LIFAC must demonstrate a high SO<sub>2</sub> removal rate while remaining competitive with other options on a cost per ton of SO<sub>2</sub> removed basis. To this end, the sponsors of this project have designed the demonstration with the following goals in mind:

- **Sustained High SO<sub>2</sub> Removal Rate** - Incorporated into the test plan are several periods of long term testing which are intended to demonstrate LIFAC's SO<sub>2</sub> removal and reliability characteristics under normal operating conditions.
- **Cost** - LIFAC must compete with both low capital cost, low SO<sub>2</sub> removal rate options such as sorbent injection and high capital cost, high SO<sub>2</sub> removal rate options such as wet

scrubbing. This project will demonstrate LIFAC's competitiveness on a cost per ton of SO<sub>2</sub> removed basis with these currently available alternatives.

- **Retrofit Adaptability** - The host site chosen required a retrofit with tight construction conditions that will prove LIFAC's ability to be installed where other technologies might not be possible. Construction was also to demonstrate LIFAC's ability to be built and brought on-line with zero plant down time other than scheduled outages.
- **System Compatibility** - A major concern of utilities is the degree of compatibility of SO<sub>2</sub> removal systems with their existing operations. This demonstration will show LIFAC's minimal impact on the host site's boiler and associated subsystems.

### **LIFAC Process History and Description**

In 1983, Finland enacted acid rain legislation which applied limits on SO<sub>2</sub> emissions sufficient to require that flue gas desulfurization systems have the capability to remove about 80 percent of the sulfur dioxide in the flue gas. This level could be met by conventional wet limestone scrubbers but not by then available sorbent injection technology. Tampella, therefore, began developing an alternative sorbent injection system which resulted in the LIFAC process.

Initially, development first involved laboratory and pilot plant tests, then full-scale tests of sorbent injection of limestone. Using high-ash, low-sulfur coal and a Ca/S molar ratio of three to one, Tampella was unable to achieve a 50 percent SO<sub>2</sub> removal rate at its 160 megawatt Inkeroinen facility. Substitution of lime for limestone was rejected due to its high cost.

Subsequent research and development by Tampella led to the addition of a humidification section after the furnace which became known as the LIFAC process. The sorbent injection process was installed full scale on a 220 megawatt boiler located at Kristiinankaupunki, Finland and a side-stream representing 2.5 megawatts was used to test a small scale humidification reactor. SO<sub>2</sub> removal rates of up to 84 percent were achieved at this plant. Additional tests at the Neste Kulloo combustion laboratory were conducted at 8 megawatts and also achieved 84 percent removal rates.

In 1986, the first large full scale test was performed at Imatran Voima's Inkoo powerplant using a 70 megawatt side-stream from a 250 megawatt boiler. A 76 percent SO<sub>2</sub> removal rate with 1.5% sulfur coal was reached. A second LIFAC activation reactor was constructed to handle an additional 125 megawatt side-stream. This newer reactor is achieving removal rates of 75 to 80 percent while using Ca/S molar ratios of between 2 and 2.5 to 1. Also, in 1988 the first tests with high-sulfur U.S. coals were run at the Neste Kulloo Laboratory. A Pittsburgh No. 8 Seam coal containing 3 percent sulfur was tested and an SO<sub>2</sub> removal rate of 77 percent was achieved at a Ca/S molar ratio of 2 to 1.

## **LIFAC Process Description**

The LIFAC system combines conventional limestone injection into the upper furnace region with a post-furnace humidification reactor located between the air preheater and the ESP. The process produces a dry, stable waste product that is removed from both the bottom of the humidification reactor and the ESP.

Finely pulverized limestone is pneumatically conveyed and injected into the upper region of the boiler where temperatures are approximately 1800 to 2200 degrees Fahrenheit. At these temperatures the limestone ( $\text{CaCO}_3$ ) calcines to form lime ( $\text{CaO}$ ) which readily reacts with the  $\text{SO}_2$  to form calcium sulfate ( $\text{CaSO}_4$ ). All of the sulfur trioxide ( $\text{SO}_3$ ) reacts with the  $\text{CaO}$  to form  $\text{CaSO}_4$ .

Approximately 25 percent of the sulfur dioxide removal occurs in the boiler with the remaining 75 percent and the unreacted lime passing through the air preheater to the humidification reactor. There the flue gas is sprayed with water that allows the unreacted lime to hydrate to  $\text{Ca}(\text{OH})_2$  which more readily reacts with the sulfur dioxide and forms  $\text{CaSO}_3$ . A combination of the proper water droplet size and residence time allows for effective hydration of the lime and complete water evaporation to create a dry reactor bottom product.

After exiting the humidification reactor, the flue gas is reheated before entering the ESP. The humidification and lower gas temperature enhance the efficiency of the ESP. Seventy-five percent of the LIFAC-produced spent sorbent and fly ash is collected by the ESP with the other 25 percent collected by the humidification reactor. Both the reactor and ESP ash may be recycled to a point ahead of the reactor to improve sorbent utilization and to improve the  $\text{SO}_2$  removal efficiency of the system to the range of 75 to 85 percent. A schematic of the LIFAC process is shown in Figure 1 along with the typical sampling locations used during the demonstration.

## **Process Advantages**

LIFAC is similar to other current sorbent injection technologies but has unique advantages with its use of a patented vertical humidification reactor. And while LIFAC's sulfur dioxide removal efficiency is not as high as traditional wet flue gas desulfurization systems, its cost and simplicity of design, construction and operation offer other advantages over these alternative systems. In particular the advantages of the LIFAC system are:

- High  $\text{SO}_2$  removal rates - Currently available sorbent injection systems have been unable to sustain high  $\text{SO}_2$  removal rates with any consistency. LIFAC has proven in the past and intends to demonstrate during this project the ability to achieve and sustain high  $\text{SO}_2$  removal rates of 75 to 85 percent over long operating periods.
- By-products - Wet lime and limestone scrubbing systems create a wet byproduct ash that must be further treated before disposal. LIFAC produces a dry solid waste ash



containing calcium sulfide, calcium sulfate and fly ash. This waste is easily disposed of under U.S. regulatory requirements, may be recycled to increase LIFAC's efficiency and may have commercial applications in the cement industry.

- **Compatibility and Adaptability** - LIFAC has minimal impact on the host's site and systems, primarily the boiler, ESP and ID fan. In addition, LIFAC requires little space and few utilities and therefore is easily installed even in small or cramped powerplant sites.

### **Construction and Systems Integration**

Construction of the LIFAC system has occurred in two phases over a period of one and a half years. The first phase of construction was completed during a routine plant outage in March, 1991. The period was utilized to install tie-ins to the host site's existing systems.

Ductwork and three dampers were installed between the air preheater and ESP to allow flue gas flow to the LIFAC activation reactor. Tie-ins were also made to the powerplant's high-pressure steam, condensate and river-water supplies. The high-pressure steam is required to reheat the flue gas exiting the LIFAC reactor and the water is needed for flue gas humidification inside the reactor. Injection ports were also installed in the boiler walls about 10 feet above the nose elevation.

The second phase of construction began in the Fall of 1991 with the driving of reactor piling and the installation of underground conduit runs. Work continued through to the Summer of 1992 with no need for plant downtime other than normally scheduled outages. During this time the limestone storage area was completed and the injection system was installed on Unit #2. The activation reactor was constructed and then tested with both cold air during a scheduled Unit #2 outage and hot flue gas during a low electricity demand period. Other powerplant tie-ins such as the steam and condensate system were also tested during low demand periods in the evening or on weekends.

All of the construction work associated with the LIFAC system was performed in close proximity to the exterior of the powerplant or in cramped areas inside the plant. The ductwork tie-ins and new steelwork required inside the plant are located in small, difficult to access work areas. The reactor structure is approximately ten feet from the powerplant with the outside ductwork and piping crossing overtop of offices and the plant maintenance area. All of these new structures and equipment were constructed with no interference to daily plant operations.

### **Schedule**

The current schedule for the LIFAC demonstration program extends over a four year period from the beginning of preliminary design in August 1990 through the testing program to be completed in early August 1994 (see Figure 2). The LIFAC system was originally scheduled

to come on-line in June of 1992 but due to delays in receiving construction permits and some minor startup problems, this date was moved to March 1993. Although testing is scheduled to continue through the Summer of 1994, preliminary test results are now available.

Currently the demonstration project is on track with the revised schedule shown in Figure 2. All construction work was completed at the beginning of August 1992. Equipment check-out was performed in July and August and the first limestone delivery was received in early September. Initial tests with limestone injection into the boiler along with post-furnace humidification were conducted in October to December 1992. Having overcome all the normal operational problems that accompany retrofit installations, the project team was prepared to conduct the test plan beginning in early 1993.

### **Test Plan**

The test plan for the LIFAC demonstration is composed of five distinct phases, each with its own objective. The first of these phases, which has already been completed, consisted of the initial baseline testing portion of the project. Measurements were taken to characterize the operation of the host's boiler and associated subsystems prior to the use of the LIFAC system. The results will be used for comparison purposes with the LIFAC system in operation and with data collected at the end of the project to determine any changes in the host's systems.

The second, or parametric, phase of testing is currently underway and will be performed to determine the best combination of LIFAC process variables for SO<sub>2</sub> removal. The variables being studied include the limestone injection nozzles' angle and location, the Ca/S molar ratio, the need for supplemental injection air at the boiler, the water droplet size and injection nozzle arrangement in the reactor, the ash recycling ratio and the approach to saturation temperature of the flue gas exiting the activation reactor. The best combination of these variables will be chosen at the conclusion of this phase and used for the remainder of the test program.

Optimization tests will be conducted to examine the effects of different coal and limestone feeds on the SO<sub>2</sub> capture rate. Coals with sulfur contents as high as 3.3 percent will be tested to determine LIFAC's compatibility with high sulfur U.S. coals. Limestones with different compositions will also be tested to determine the LIFAC system's adaptability to local sorbent sources.

Long term testing will also be performed to demonstrate LIFAC's performance under commercial conditions. The LIFAC system will be in operation 24 hours per day for several weeks using the powerplant's baseline coal, high calcium limestone and the optimum combination of process variables. In addition to process performance measurements, during this phase the operation and maintenance requirements of the system will be examined. Long term (two to three weeks) tests will also be conducted with two other coals; one lower sulfur coal (1.5%) and one higher sulfur content coal (3.3%).

The final phase of testing is composed of the post-LIFAC tests. The baseline tests will be repeated to gather information on the condition of the boiler and its associated subsystems. Comparisons will be made to the original baseline data to identify any changes either caused by the LIFAC system or independent of its operation.

### **Preliminary Results**

Once startup and checkout were complete and the operational problems overcome, the project team initiated parametric testing. During the early tests with limestone addition and humidification, increases in opacity levels prevented the system from being operated as intended. Test work conducted by EPRI and Southern Research Institute identified the cause as lower ash resistivity resulting from reduced operating temperatures in the ESP. The activation reactor is designed to humidify the flue gas and drop the gas temperature to slightly above saturation temperature in order to maximize SO<sub>2</sub> capture. The flue gas is then reheated to above 175° Fahrenheit as it exits the activation reactor. Due to the relatively small size of the ESP (only 200 SCA) and because of lower ash resistivities, it was determined the ESP needed to be operated at about 200° Fahrenheit to avoid any problems with increased opacity. Having determined this, the operating procedures were revised to insure an ESP operating temperature above 200° Fahrenheit.

Parametric testing was initiated at 60 MW to assess the broad impacts of limestone injection, flue gas humidification, and sorbent recycle. Figure 3 shows average reductions achieved throughout the LIFAC process. About 22 percent SO<sub>2</sub> reduction is achieved in the boiler. This is increased to about 52 percent with humidification, and further raised to 75 percent with the use of sorbent recycle from the ESP ash hoppers. These tests were conducted with a fine grind limestone (80% minus 325 mesh) with a Ca content above 90 percent. A Ca/S molar ratio of 2.0 was held near constant and a 4 to 5° Fahrenheit approach to saturation was maintained in the activation reactor.

Figure 4 shows the impacts of varying the Ca/S molar ratio. The majority of the tests have been conducted at 2.0, but the trends are as expected. The higher the Ca/S ratio, the higher the SO<sub>2</sub> reduction. Results show, however, that SO<sub>2</sub> reductions of 75 to 85 percent are possible when spent sorbent is recycled and a 3 to 5° Fahrenheit approach to saturation temperature is maintained.

Figure 5 shows the impact of recycling spent sorbent under various boiler loads. The Ca/S molar ratio was maintained at about 2.0 and the level of humidification is high (4 to 5° Fahrenheit above saturation). Generally, there is an 18 to 25 percentage point increase in SO<sub>2</sub> reduction as a result of sorbent recycle. With recycle, total SO<sub>2</sub> reductions ranged from 75 to 85 percent depending on boiler load.

Although only preliminary testing has been completed, the results are encouraging. Additional work will be conducted to optimize these process parameters in hopes of maintaining a minimum of 80 percent SO<sub>2</sub> reduction at all boiler loads.

At this point it has also been shown at RP&L and other LIFAC installations that the system can be installed and operated without affecting normal powerplant operations. It will also be shown that the system can economically reduce SO<sub>2</sub> emissions when compared with other flue gas desulfurization technologies.

### References

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2. DOE Contract No. DE-FC22-90PC90548, "LIFAC Demonstration at Richmond Power and Light White Water Valley Unit #2", March 1992.
3. Rose, J., Hervol, J., Viiala, J., Koskinen, J., Patel, J. and Huffman, I., "LIFAC Flue Gas Desulfurization", March 1992.

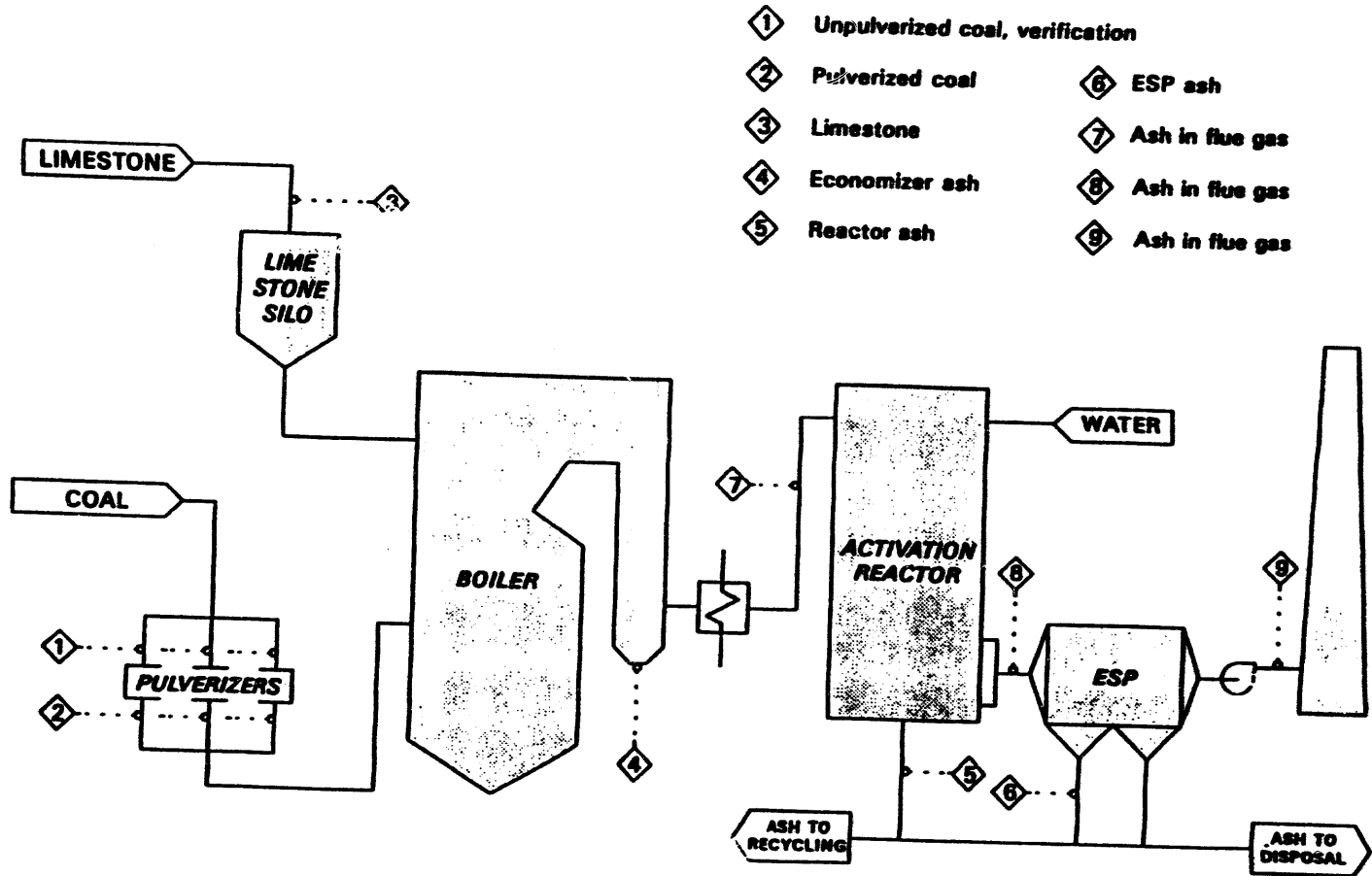


Figure 1. LIFAC RP&L DEMONSTRATION -- SOLIDS SAMPLING LOCATIONS

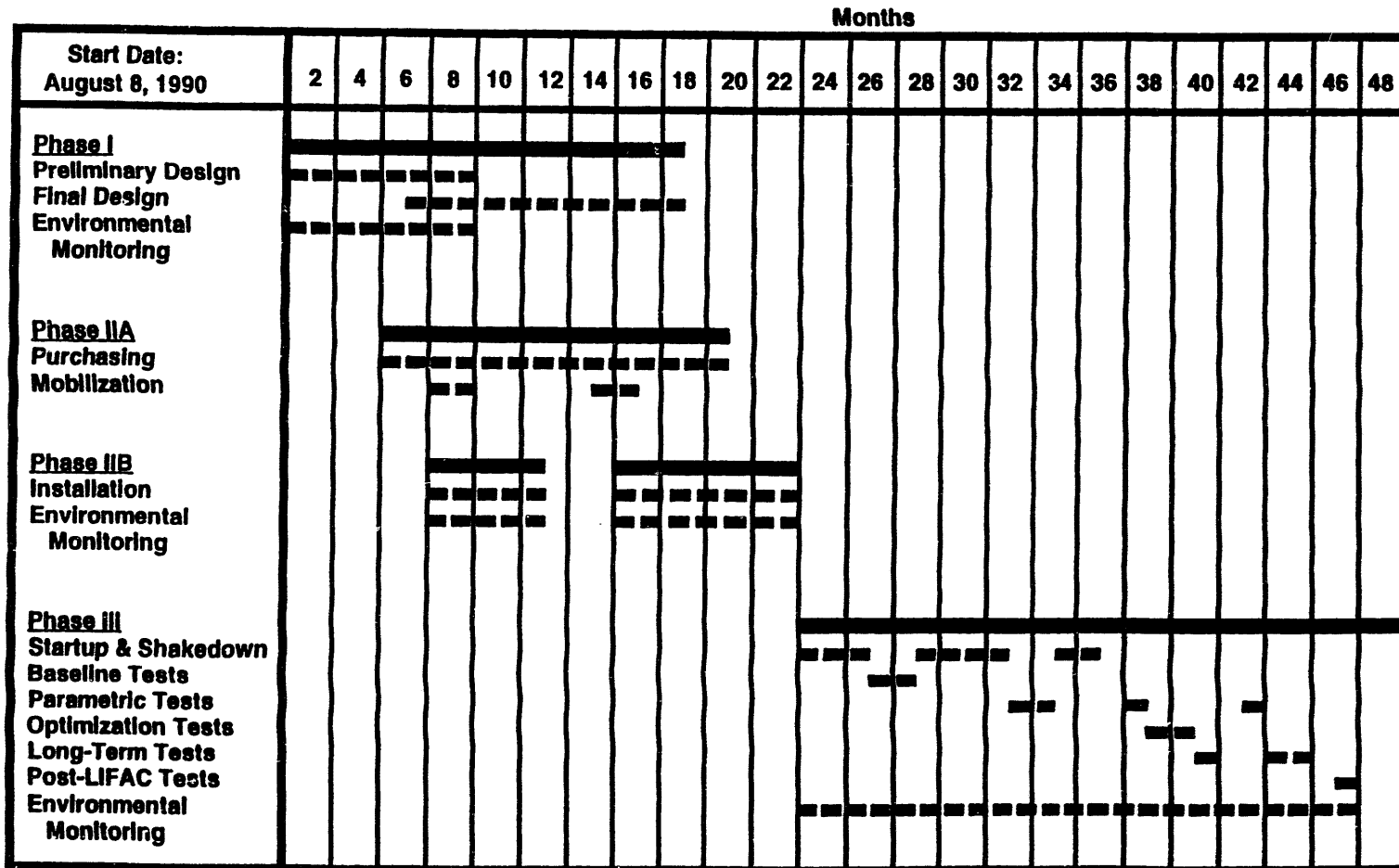
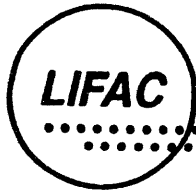
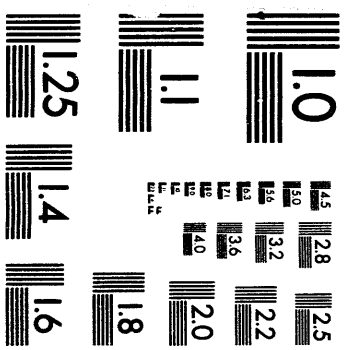


Figure 2. LIFAC DEMONSTRATION -- CURRENT PROJECT SCHEDULE



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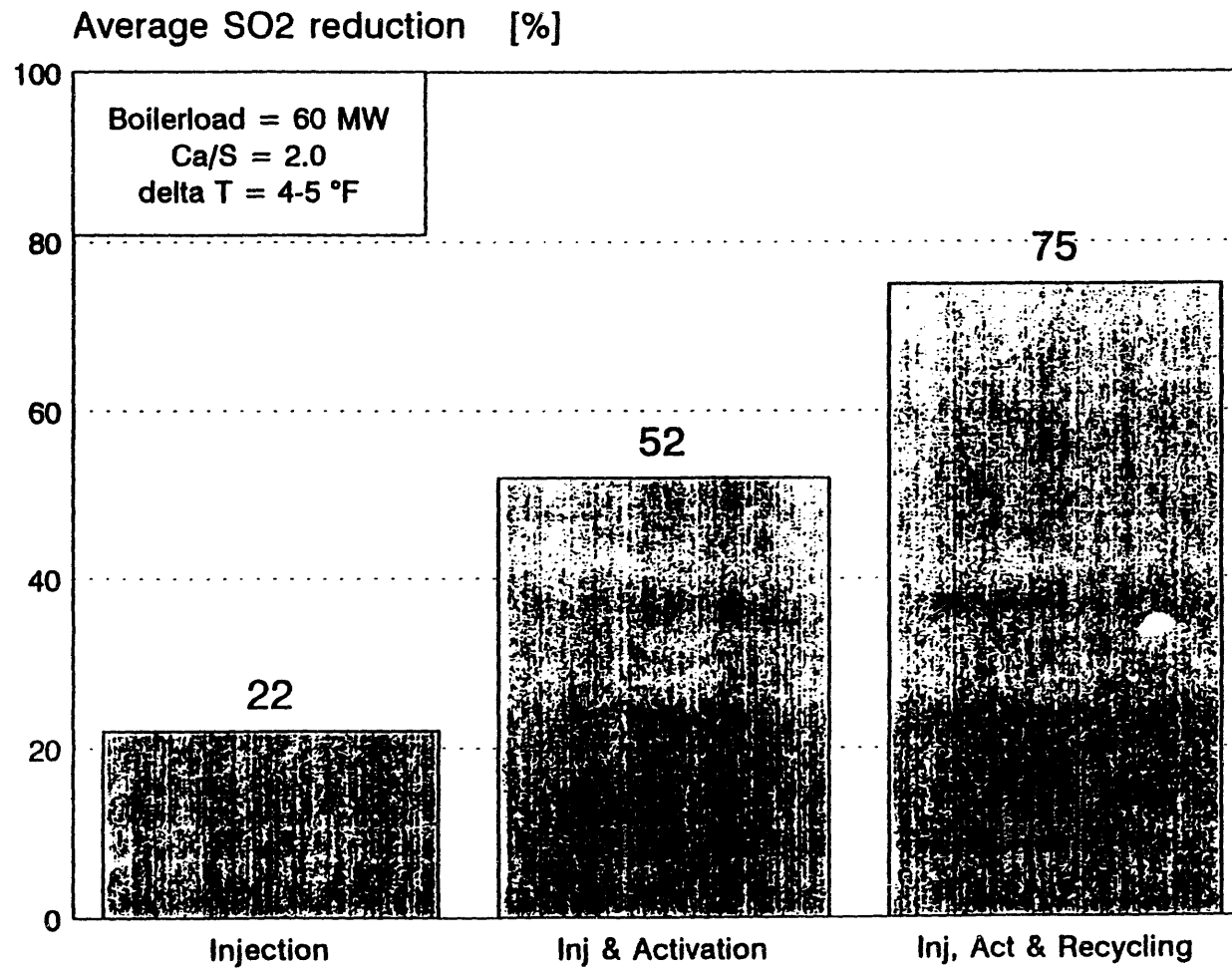


Figure 3. PARAMETRIC EFFECTS

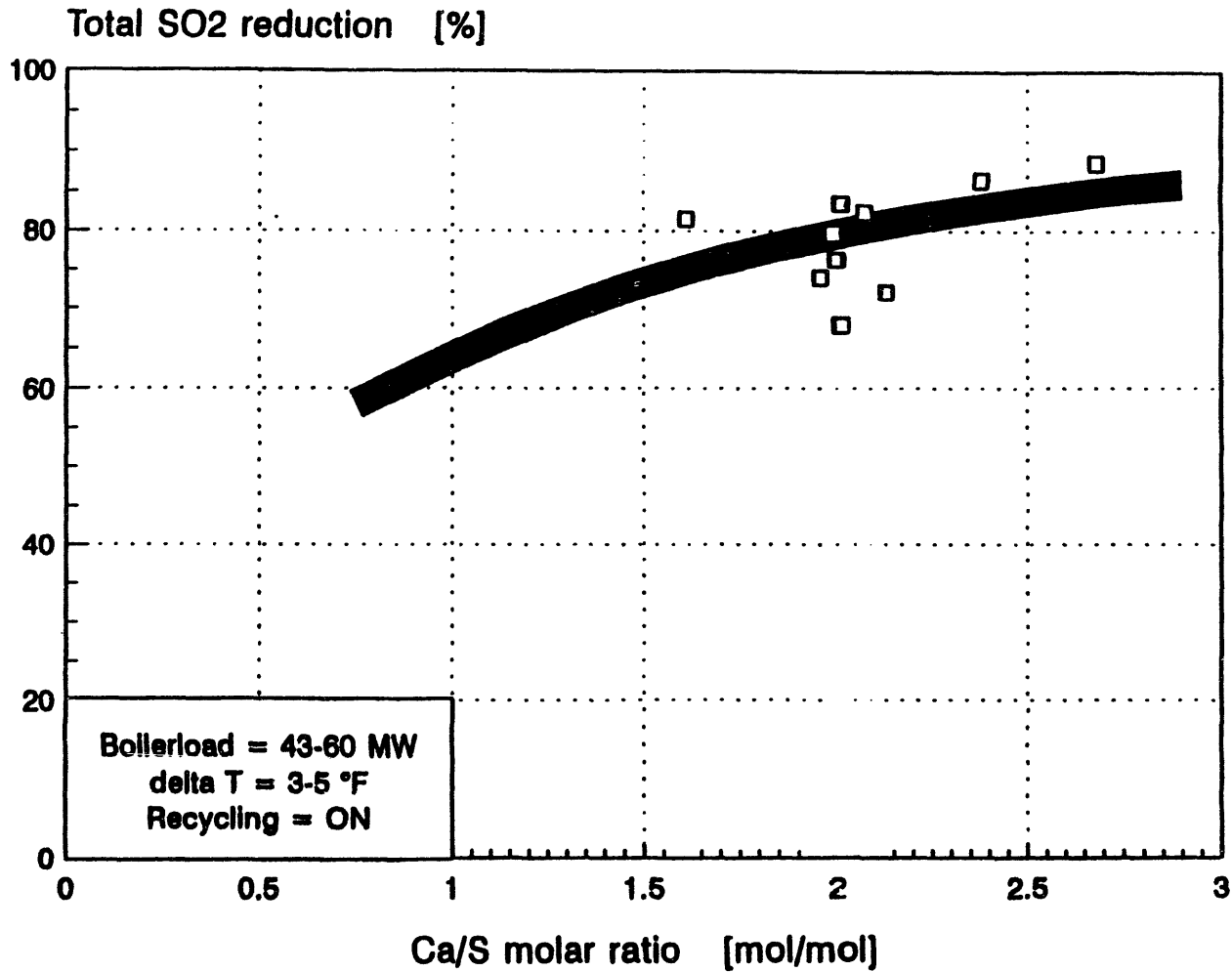


Figure 4. SO<sub>2</sub> REDUCTION vs. Ca/S MOLAR RATIO

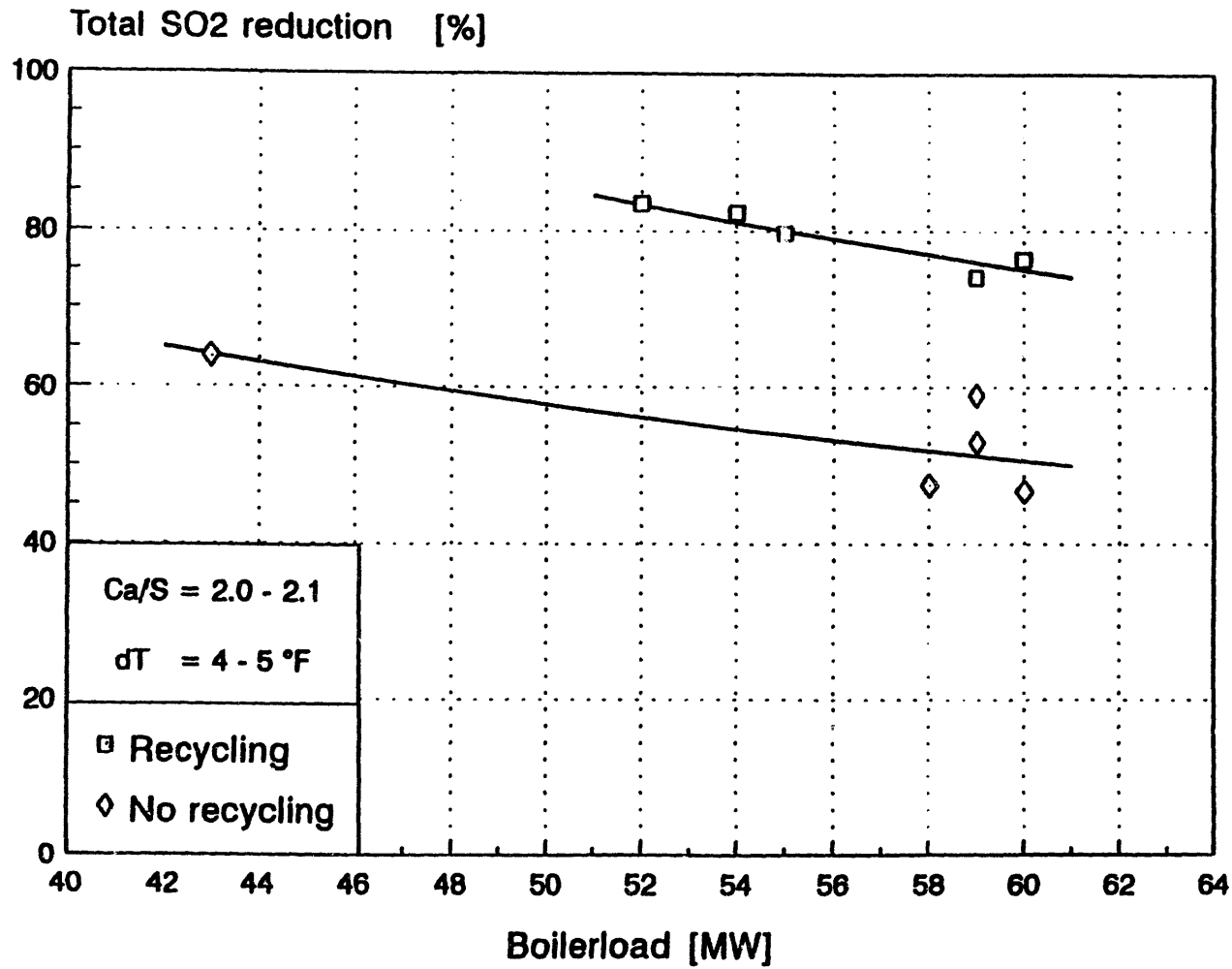


Figure 5. SO2 REDUCTION vs. BOILERLOAD



**CHIYODA THOROUGHbred 121  
INNOVATIVE CLEAN COAL TECHNOLOGY  
DEMONSTRATION PROJECT INITIAL TESTING RESULTS**

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**ABSTRACT**

The Chiyoda CT-121 Project at Georgia Power's Plant Yates Unit #1 is a Round II, \$36 million, Innovative Clean Coal Technology project co-funded by the Department of Energy, the Electric Power Research Institute and The Southern Company. The CT-121 scrubbing process features a single SO<sub>2</sub> absorption module, called the Jet Bubbling Reactor (JBR), made entirely of fiberglass reinforced plastics (FRP) where several chemical reactions (absorption/neutralization/oxidation/crystal growth) take place concurrently. The 100 MWe flue gas scrubber uses ground limestone to remove up to 95% of the SO<sub>2</sub> in the flue gas from a pulverized coal-fired boiler, producing a high quality gypsum by-product. Gypsum will be

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deposited in a gypsum "stack", a disposal technique commonly used in the phosphate fertilizer industry. Operational testing, continuing through early 1995, will include sustained high performance testing, simultaneous particulate removal in the JBR, alternate limestone and alternate higher sulfur coal. Initial results from parametric testing have demonstrated the excellent SO<sub>2</sub> removal and particulate removal performance of this unique flue gas desulfurization technology.

## **INTRODUCTION**

As part of the Innovative Clean Coal Technology (ICCT) program, the Department of Energy (DOE), the Southern electric system, and the Electric Power Research Institute (EPRI) are sponsoring a 100 MWe demonstration of the Chiyoda Thoroughbred (CT-121) process. The \$36 million project is located at Georgia Power Company's Plant Yates Unit 1, near Newnan, Georgia.

This demonstration project began with the retrofit construction of a CT-121 wet-limestone scrubber to a 100 MWe pulverized coal-fired boiler. The CT-121 process involves the use of a unique process vessel called the Jet Bubbling Reactor (JBR) in place of the traditional spray tower/reaction tank arrangement of most conventional FGD processes. Start-up occurred in October, 1992 and the first phase of the demonstration began in January, 1993. The demonstration project is scheduled to continue through early 1995.

The demonstration is divided into two major periods with the first utilizing the pre-existing electrostatic precipitator (ESP), operating at full capacity. The second period, scheduled to begin in early 1994, will involve the field-by-field deenergization of the ESP and repetition of the tests executed in the first period. The second period will be used to evaluate the process' ability to remove uncontrolled particulate from a coal-fired boiler and the effect of high fly-ash concentrations in the slurry on scrubber performance. The two periods are further divided into parametric testing and long-term load-following testing. It is the results of the low-fly-ash parametric testing which are addressed in this paper (Performance, operability, and reliability evaluation are the focus of this demonstration, with performance characterization the specific focus of the parametric testing).

Thus far, the performance of the CT-121 process has been excellent, with SO<sub>2</sub> removal greater than 90% easily achievable and 98% SO<sub>2</sub> removal achieved under some operating conditions. Additionally, 90% of the particulate matter in the flue gas following the ESP was removed by the JBR. Long-term testing, currently in-progress, should help establish CT-121 as a highly reliable FGD process at a U.S. electric utility plant.

## **HISTORICAL OVERVIEW**

The Yates Chiyoda project was one of four successful proposals from The Southern Company in Round II of the DOE's Clean Coal solicitation in 1988. Design and engineering began in 1989 at SCS and the Cooperative Agreement was executed in April of 1990. Construction began with ground breaking in August of 1990 and was completed in October of 1992 following a significant delay in obtaining permits from the State of Georgia's Environmental Protection Division. Operations began in late October of 1992 and continue today.

The interest in the Chiyoda process stems from The Southern Company's previous experience with five different FGD systems at Gulf Power's Plant Scholz in the late 1970's. The CT-121 process was selected because of its reliability and potential to offer significant cost reductions over other FGD processes. Lessons learned at Scholz and from other Chiyoda CT-121 FGD systems have been incorporated and expanded into the aggressive evaluation program now underway at Plant Yates.

## **FACILITY DESCRIPTION**

The equipment comprising the demonstration facility can be divided into five major systems: boiler/ESP; CT-121 scrubber/wet chimney; limestone preparation circuit; byproduct gypsum stacking area; and process control system.

Plant Yates' Unit 1, with a rated capacity of 100 MWe, is the source of flue gas for the CT-121 process. All of the flue gas from this unit is treated by the CT-121 wet FGD process with no provision for flue gas bypass (The CT-121 process must remain in service whenever

the boiler is operating). During the low fly-ash phase of parametric testing, the existing ESP for Unit 1 is being used for particulate control. The design efficiency for this ESP is 98%.

A simplified process flow diagram for the CT-121 process is presented in Figure 1. The CT-121 employs a unique absorber design, called a Jet Bubbling Reactor (JBR), to combine conventional SO<sub>2</sub> absorption, sulfite oxidation, and gypsum crystallization in one reaction vessel. This significantly reduces the potential for gypsum scaling, a problem that frequently occurs in natural-oxidation FGD systems. Since much of the crystal attrition and secondary nucleation associated with the large centrifugal pumps in conventional FGD systems is also eliminated in the CT-121 design, large, easily dewatered gypsum crystals are produced.

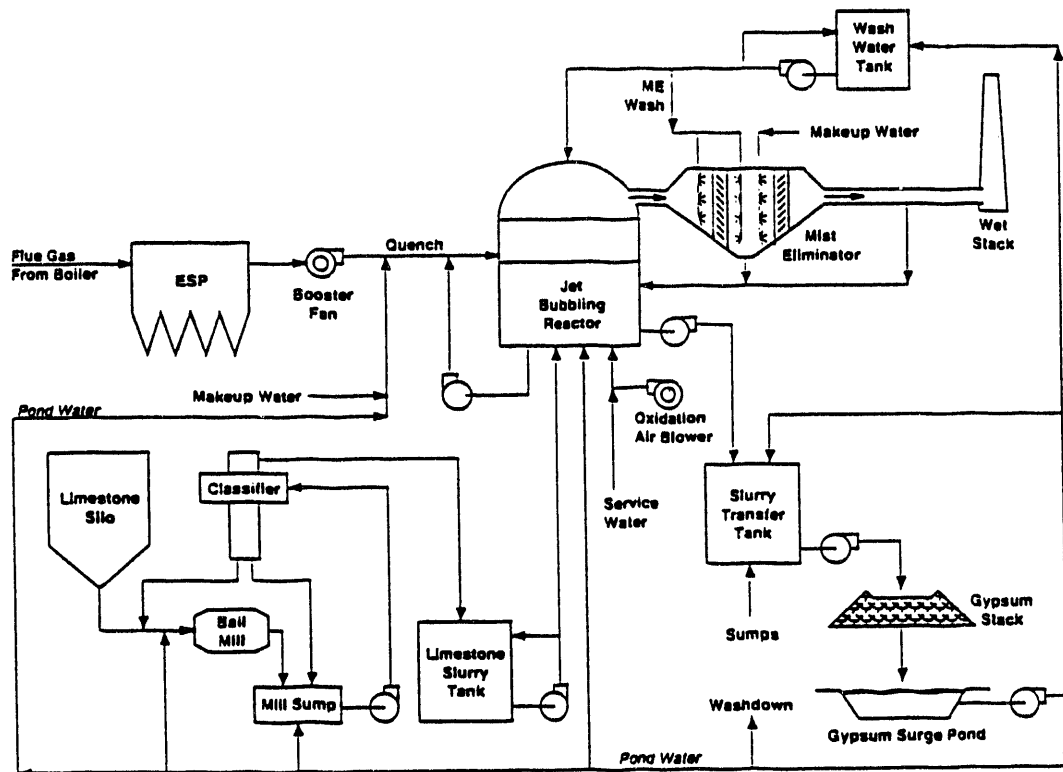


Figure 1. Simplified CT-121 Process Flow Diagram

In the Yates application, the flue gas enters the inlet gas cooling section after the I.D. 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 the JBR, the central feature of the CT-121 process. The gas enters the JBR through an enclosed plenum chamber formed by an upper



deck plate and a lower deck plate. Sparger tube openings in the lower deck plate force inlet flue gas beneath the slurry contained in the jet bubbling (froth) zone of the JBR vessel. After bubbling through the slurry, the gas flows upward through gas risers which pass through both the lower and upper deck plates. Entrained droplets in the gas disengage in a second plenum above the upper deck plate, and the cleaned gas passes to the mist eliminator.

After leaving the mist eliminator, the clean gas exits the system through a wet FRP chimney. Since the gas enters the chimney saturated with water, any heat loss in the chimney will result in gas cooling and condensation. Condensate in the chimney is collected by a system of aerodynamically designed internal "gutters" and is returned to the JBR.

A closed-circuit wet ball mill limestone preparation system is used to grind the limestone to a small enough particle size so that the amount of unreacted limestone needed in the JBR can be kept to a minimum. The baseline particle size criterion is 90% less than 200 mesh.

Slurry from the gypsum slurry transfer tank is diluted and pumped to a lined gypsum stacking area for dewatering and storage. The stacking technique involves filling a diked area with slurry for gravity sedimentation. The filled area is then partially excavated to increase the height of the containment dikes. The process of sedimentation, excavation, and perimeter dike formation continues on a regular basis during the active life of the stack. Process water is decanted, stored in a surge pond and returned to the process.

During normal operation of the FGD system, the amount of SO<sub>2</sub> removed from the flue gas is controlled by varying the JBR pressure drop ( $\Delta P$ ). The  $\Delta P$  is adjusted by varying the JBR liquid level. Higher liquid levels result in increased SO<sub>2</sub> removal. The pH can also be varied to affect removal with higher pH's resulting in increased removal. Boiler unit load and flue gas SO<sub>2</sub> concentration also affect removal efficiency.

## **TECHNOLOGICAL ADVANTAGES**

The CT-121 process, as constructed at Plant Yates, offers several technological and economic advantages over both conventional spray tower scrubbers as well as previously

constructed CT-121 systems. The innovations which provide these advantages include the use of fiberglass reinforced plastics, the unique JBR absorber, a novel wet FRP chimney design, and the use of a gypsum stack for byproduct dewatering and storage.

### Fiberglass Reinforced Plastic

Many of the vessels comprising the Yates CT-121 installation are made of fiberglass reinforced plastics (FRP) to withstand the corrosiveness of the FGD process streams. Two of the vessels (the JBR and the limestone slurry storage tank) were constructed on-site in 1990 since their large size precluded roadway shipment. In a temporary facility, the contractor built the two vessels by layering materials on a slow moving "mandrel", alternating and mixing resin applications, fiberglass mat, fiberglass weave, fiberglass strand and chopped fiberglass in a carefully designed sequence over several week's time. FRP internals and wall penetrations (nozzles, man-ways, sample ports, etc.) were then installed by hand with additional FRP components to complete these homogeneous plastic vessels. Other FRP components that were small enough for shipment (two tanks, inlet duct section, chimney sections, etc.) were assembled and joined on-site at about the same time. Baseline data on the FRP vessels' acoustic emissions and a finite element analysis were taken for comparison to their future condition.

A distinct advantage of the FRP construction was that it eliminated the need for a flue gas prescrubber to remove chlorides, because the corrosion resistance properties of fiberglass are superior to those of alloys. This represented a large capital cost savings to the project.

### JBR

The inherently high reliability of the CT-121 process eliminated the need for a spare absorber. This results in significantly reduced capital costs compared to spray tower systems which typically are built with 'spare' absorbers. Additionally, the JBR offers the distinct advantage of providing simultaneous SO<sub>2</sub> removal and particulate control. The JBR's high particulate removal efficiency may allow elimination of the ESP in new plant designs or make it a good choice for a retrofit to a plant with a marginally performing ESP.

### Wet Chimney

The lower than normal mist loading inherent in the JBR design, combined with the unique design of the wet chimney, eliminated the need for flue gas reheat, another cost saving feature. Fluid dynamic modeling was performed to assist in the wet duct and chimney design. The resulting FRP wet chimney has a specially designed system of "gutters" which collect the liquid as it becomes un-entrained or disengaged from the flue gas and returns it to the JBR. This design effectively eliminates rain-out from the chimney.

### Gypsum Stack

The FGD byproduct gypsum solids are disposed of by stacking. Stacking combines the advantages of ponding and landfills. Ponding has low operating costs and low capital equipment requirements, while landfills require less space and have less environmental impact. During the low fly ash parametric test period, handling, stackability, and trafficability of the gypsum stack were carefully monitored.

### **TEST OBJECTIVES**

The parametric testing portion of the low fly ash test period has recently been completed.

The major objectives of the parametric test program were to:

- Correlate the effects of pH and  $\Delta P$ , and evaluate the effects of boiler load on system performance;
- Correlate the effect of limestone grind on system performance;
- Monitor solids properties and gypsum stack operation;
- Evaluate particulate removal efficiency; and
- Demonstrate reliable operation of the CT-121 FGD system.

The test schedule was a full factorial matrix of the three primary test parameters ( $\Delta P$ , pH, and load) which affect  $SO_2$  removal efficiency. A full factorial matrix was designed to eliminate the need for a complex statistical analysis to evaluate the collected data. In addition to the full factorial matrix, the test plan also includes selected tests to evaluate limestone grind.

## **SCHEDULE/MILESTONES**

After project selection in 1988 and execution of the Cooperative Agreement in 1990, construction of the Yates CT-121 took approximately 26 months. Operations began in October of 1992 with system shakedown, and parametric testing was conducted from January through March of 1993.

In March, long-term testing began (with the ESP fully energized) and will continue through the end of 1993. In early 1994, high-particulate operations (ESP deenergized) will begin and continue through early 1995. The high fly ash test phase will also contain both parametric and long-term testing periods. Extended monitoring of the groundwater and the gypsum stacking area will carry through 1996.

## **RESULTS**

The parametric test period proceeded well, beginning on January 17, 1993 and ending on March 31, 1993. In general the process responded quickly and smoothly to load and process parameter changes, usually stabilizing within one hour of the change. The process' availability and reliability were both 98% due to a low equipment failure rate.

SO<sub>2</sub> removal efficiency in the JBR is a function of JBR  $\Delta$ P, pH, load, and inlet SO<sub>2</sub>. The parametric test results indicate that SO<sub>2</sub> removal increases with increasing pH and JBR deck  $\Delta$ P, and that removal decreases with increasing load and inlet SO<sub>2</sub> concentration. Figure 2 illustrates the increase in SO<sub>2</sub> removal with increasing JBR  $\Delta$ P as well as with pH. SO<sub>2</sub> removal however, does not increase with pH above a pH of 4.5 at the conditions tested. The figure shows that performance in excess of 90% SO<sub>2</sub> removal is easily achievable for the CT-121 process without the use of additives. Additionally, limestone utilization was consistently greater than 97% in the selected pH operating range (4.0 - 5.0) and exceeds that historically achieved in spray tower scrubbers. Oxidation, one of the keys to the excellent performance of this scrubber, was consistently 100%.

Although slurry pH and deck  $\Delta$ P are the primary operational parameters for controlling SO<sub>2</sub> removal efficiency, inlet SO<sub>2</sub> concentration and boiler load (flue gas flow rate) have an effect

on scrubber performance. Figure 3 shows that SO<sub>2</sub> removal decreases with increasing boiler load at constant pH and ΔP. Because of decreased dynamic head (i.e., decreased pressure drop in the duct, plenum, etc.) at lower loads (due to lower gas flows), the JBR slurry level must increase to maintain a constant JBR ΔP at lower loads. Therefore, higher loads, which require a lower JBR level, result in decreased removal.

The SO<sub>2</sub> removal efficiency was seen to decrease with increasing inlet SO<sub>2</sub> at pH of 4.5. This trend is readily apparent in the regression analysis, but can also be seen in Figure 4 which compares tests at similar conditions with the exception of differing inlet SO<sub>2</sub> concentrations.

#### Particulate Testing Results

JBR particulate removal (ESP on) was evaluated in January, 1993, concurrent with parametric testing. The results indicate that the CT-121 process is an excellent particulate control device with removal efficiencies of 90% measured for those particles not collected by the existing ESP. Additionally, 99.9% of the particles greater than 10μ and 90% of those in the 1 -10 μ range were collected. Based on this performance, it is anticipated that the high fly-ash operating period will also demonstrate excellent particulate removal characteristics as well.

#### Results Regression and Empirical Model

The parametric test series developed data on system performance over the entire range of expected operating conditions. These data were then used to construct an empirical model of system performance. Besides being a simple and efficient way of presenting the results of this test program, such a model is useful for choosing the most efficient operating conditions for a desired level of performance. It may also be possible to use this model to estimate emissions in situations where the CEM system has failed.

Linear models were investigated since linear is much simpler than non-linear regression analysis. The goal of this effort was to have a regression model in which all of the terms

were statistically significant and sensible from a technical perspective. The following equation was found to meet these criteria (Note that it is linear with respect to Number of Transfer Units (NTU) and becomes exponential when expressed in terms of efficiency):

$$\text{SO}_2 \text{ removal (\%)} = 100 * (1 - \exp(-A * pH + B * \Delta P + C * \text{ISO}_2 + D * \text{LOAD} + E * pH^2 + F * pH * \Delta P + G))$$

Where:      A,B,C,... = numeric coefficients  
               $\Delta P$         = JBR differential pressure, in WC  
               $\text{ISO}_2$       = Inlet  $\text{SO}_2$  (@ 3%  $\text{O}_2$ ), ppm  
              Load        = Unit Gross Load, MWe

With an  $R^2$  of 0.99 and no evidence of autocorrelated residuals, this model can easily be used to make accurate predictions of  $\text{SO}_2$  removal performance over the range of operating conditions tested.

## ECONOMIC EVALUATION

An important result from the parametric test series is the relationship between operating conditions,  $\text{SO}_2$  removal performance, and operating costs. This information can be used to choose lowest cost operating conditions for a desired  $\text{SO}_2$  removal, and for comparison with other system designs such as open spray towers. The analysis presented here is limited to limestone and power cost (e.g., I.D. fan, motors, pumps, ball mill) data. It does not consider O&M costs which are assumed to be constant.

Figure 5 shows that the cost per ton of  $\text{SO}_2$  removed was relatively constant over the range of conditions tested with a standard deviation of 0.6. Most of the variability is the result of operation at different loads. At a given load, the fan power costs (normalized to \$/ton  $\text{SO}_2$  removed) are seen to be relatively constant. Fan power does increase with both load and  $\Delta P$  as expected; however, the increased  $\text{SO}_2$  removal at higher  $\Delta P$ 's results in a relatively constant normalized power cost. A fan power credit, equivalent to the documented I.D. fan power consumption prior to the CT-121 scrubber construction, was taken when calculating power costs.

## **CONCLUSION**

The Chiyoda CT-121 scrubber has exhibited excellent performance throughout the demonstration project, including the parametric test period. It easily exceeds 90% SO<sub>2</sub> removal, while maintaining consistently low variable operating costs and power consumption less than 1.5% unit Maximum Rated Capacity (MRC). Limestone utilization in excess of 97% is achievable at any pH within the established operating range of 4.0 to 5.0. The scrubber's ease of operation allows even the most inexperienced operator to quickly become familiar with system operations. Follow-on testing in a long-term load-following mode should firmly establish the viability of this unit as both an SO<sub>2</sub> reduction process, as well as a particulate control device. Initial indications are that the process' load following capabilities are excellent.

## **ACKNOWLEDGEMENTS**

The authors wish to thank all those involved in the success of the Chiyoda CT-121 project at Plant Yates. Specifically, we are grateful for the tireless efforts and continued support of Messrs. Rob Moser and Jeff Stallings of EPRI, Mr. Hiroshi Yanagioka of Chiyoda-Japan, Messrs. Mike Knowles, Ross Kist, Steve Lee, Bill Mashburn, and Flip Kee (retired) of Georgia Power Company, Messrs. Randall Rush, Pete Honeycutt, Zach Looney, Lamar Larrimore, Kamyar Vahkshoorzadeh, and Steve Kimble of Southern Company Services, Inc., Messrs. Buddy Hargrove, Jack Burke, and Greg Stevens of Radian Corporation, and certainly the men and women of Georgia Power Company's Plant Yates who, on a daily basis, meet and overcome unfamiliar obstacles in support of scrubber operations.

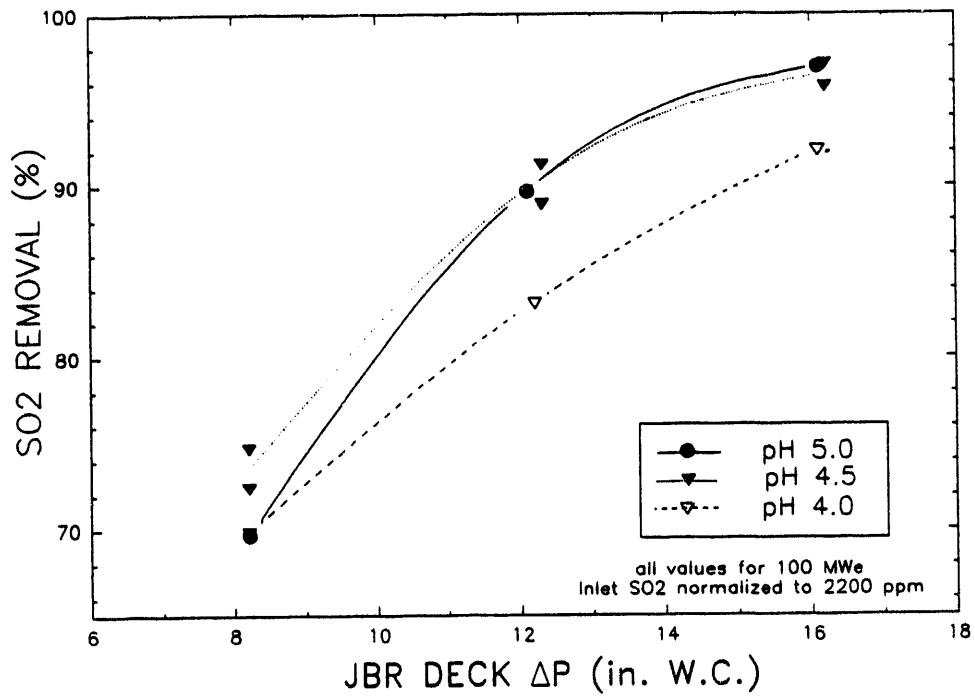


Figure 2. SO2 Removal vs JBR Deck ΔP

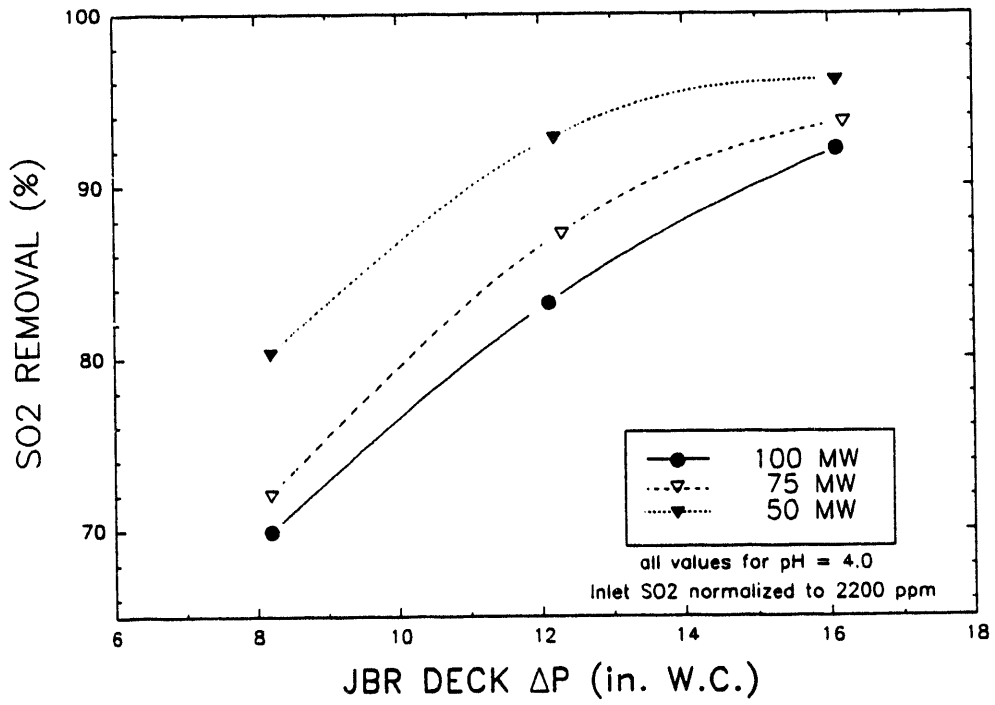


Figure 3. Effect of Load on SO2 Removal



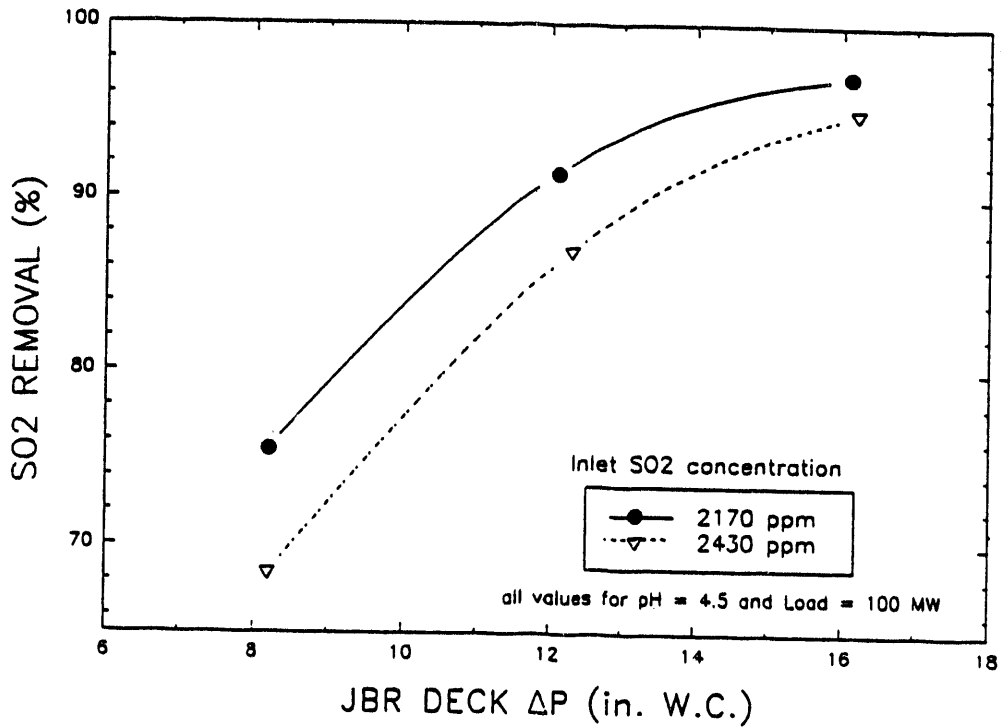


Figure 4. Effect of Inlet SO<sub>2</sub> on SO<sub>2</sub> Removal

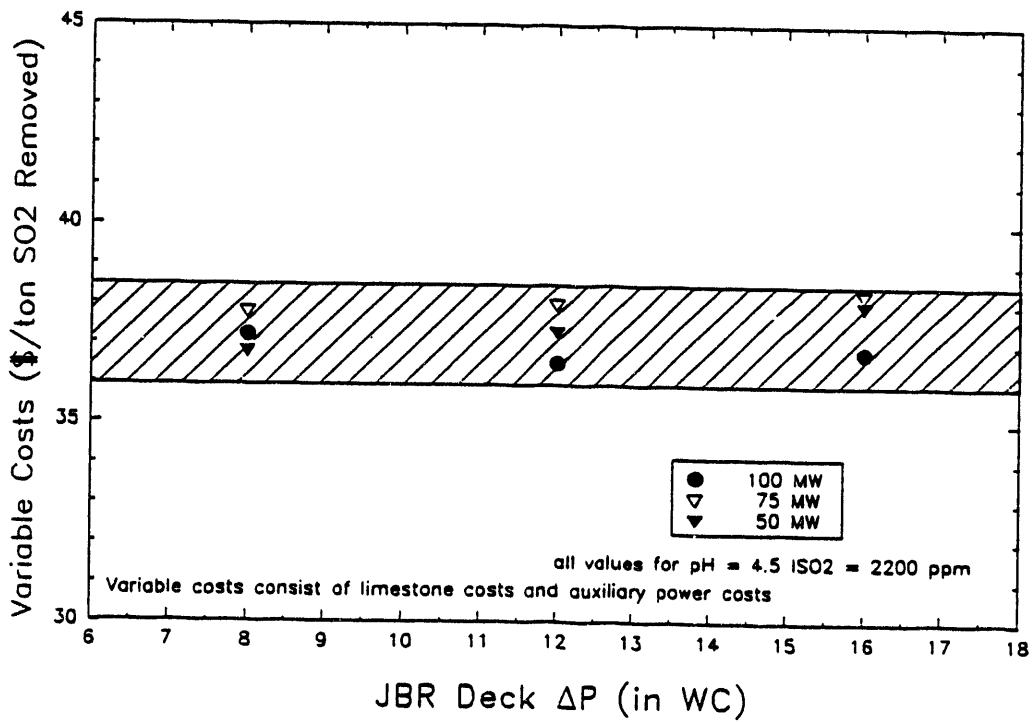


Figure 5. Variable Cost Deviation



## **Session 4**

# **Industrial Applications**

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