Coal-Diesel Demonstration Project at University of Alaska, Fairbanks

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Advanced Cleanup Systems

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

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ABSTRACT

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

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

I. INTRODUCTION

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

This project was the first North American demonstration of the GSA system for flue gas desulfurization (FGD) for a coal-fired utility boiler. This low-cost retrofit project achieved the expected target, which was to remove more than 90% of the sulfur dioxide (SO_2) from the flue gas while achieving a high utilization of reagent lime. TVA furnished its Center for Emissions Research (CER) as the host site and provided operation, maintenance, and technical support during the project. The CER is located at the TVA's Shawnee Fossil Plant near Paducah, Kentucky.

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

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

- Demonstrate SO_2 removal in excess of 90% using high-sulfur U.S. coal.
- Optimize design and operating parameters to maximize the SO_2 removal efficiency and lime utilization.
- Compare the SO_2 removal efficiency of the GSA technology with existing spray dryer/electrostatic precipitator (SD/ESP) technology.

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

- Determine the air toxics removal performance of the GSA technology.
- Compare the SO_2, particulate, and air toxics removal performance between GSA/ESP and GSA/fabric filter systems.

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

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

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

II. HISTORY OF THE GSA TECHNOLOGY

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

In 1985, a GSA pilot plant was built in Denmark to establish design parameters for SO_2 and hydrogen chloride (HCl) absorption for waste incineration applications. The first commercial GSA unit was installed at the KARA Waste-to-Energy Plant at Roskilde, Denmark, in 1988. Currently, there are seventeen GSA installations in Europe; 15 are municipal solid waste incinerator applications, and two are industrial applications (cement and iron ore reduction).

With the increased emphasis on SO_2 emissions reduction by electric utility and industrial plants as required by the Clean Air Act Amendments of 1990, there is a need for a simple and economic FGD process, such as GSA, by the small to mid-size plants where a wet FGD system may not be feasible. The GSA FGD process, with commercial and technical advantages confirmed in this demonstration project, will be a viable alternative to meet the needs of utility and industrial boilers in the U.S.

III. GSA FGD PROCESS DESCRIPTION

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

- A circulating fluidized bed reactor.

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

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

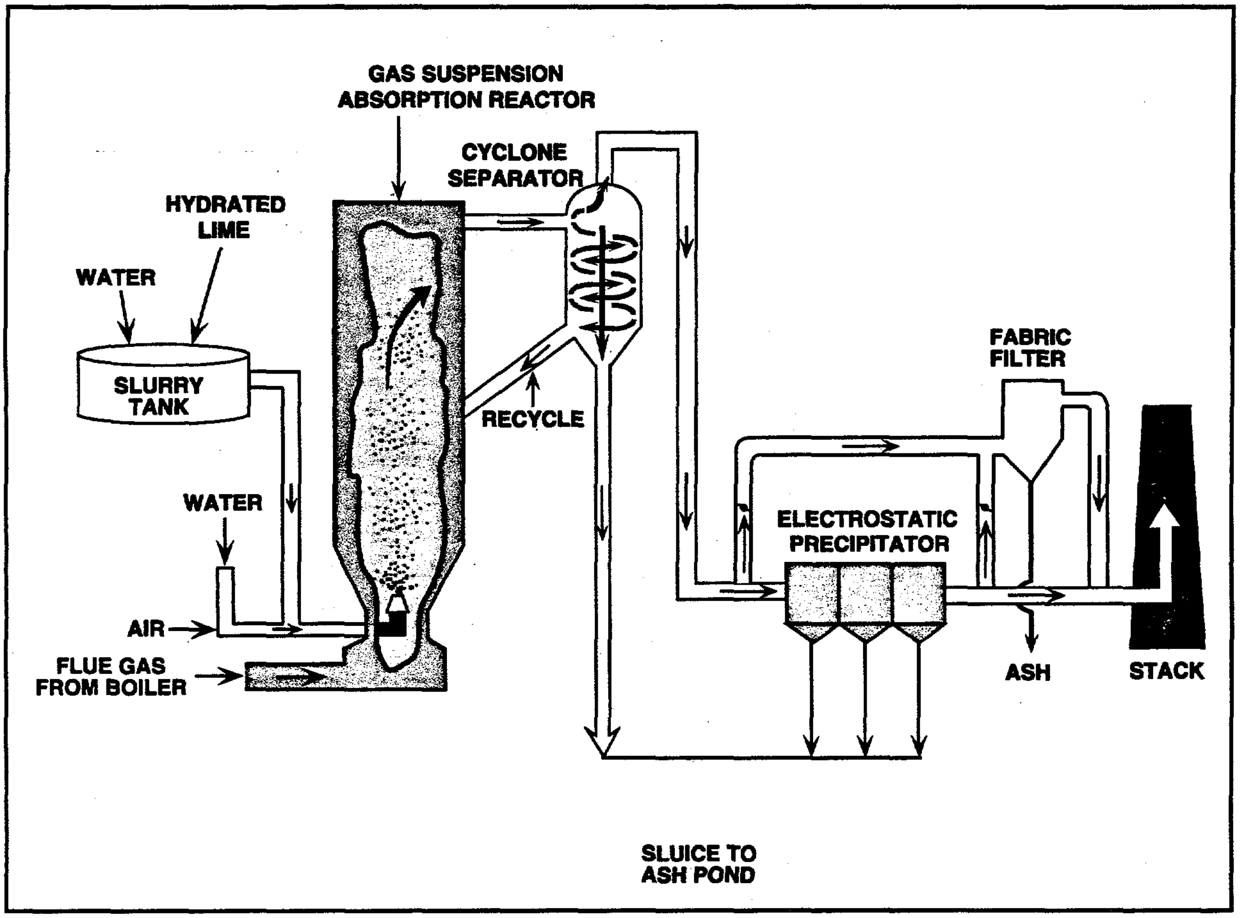


Figure 1. Gas Suspension Absorption Process Flow Diagram

suspended solids, evaporates. Simultaneously, the lime particles in the slurry undergo a chemical reaction with the acid components of the flue gas, SO_2 and HCl, capturing and neutralizing them.

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

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

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

Automatic Process Adjustment

An effective monitoring and control system automatically ensures that the required level of SO_2 removal is attained while keeping lime consumption to a minimum. This GSA control system, which is shown in Figure 2, incorporates three separate control loops:

1. Based on the flue gas flow rate entering the GSA system, the first loop continuously controls the flow rate of the recycled solids back to the reactor. The large surface area for reaction provided by these fluidized solids and the even distribution of the lime slurry in the reactor, provide for the efficient mixing of the lime with the flue gas. At the same time, the large volume of dry material prevents the slurry from adhering to the sides of the reactor.

2. The second control loop ensures that the flue gas is sufficiently cooled to optimize the absorption and reaction of the acid gases. This control of flue gas temperature is achieved by the injection of additional water along with the lime slurry. The amount of water added into the system is governed by the temperature of the flue gas exiting the reactor. This temperature is normally set a few degrees above flue gas saturation temperature to insure that the reactor solids will be dry so as to reduce any risk of acid condensation.

3. The third control loop determines the lime slurry addition rate. This is accomplished by continuously monitoring the SO_2 content in the outlet flue gas and comparing it with the required emission level. This control loop enables direct proportioning of lime slurry feed according to the monitored results and maintains a low level of lime consumption.

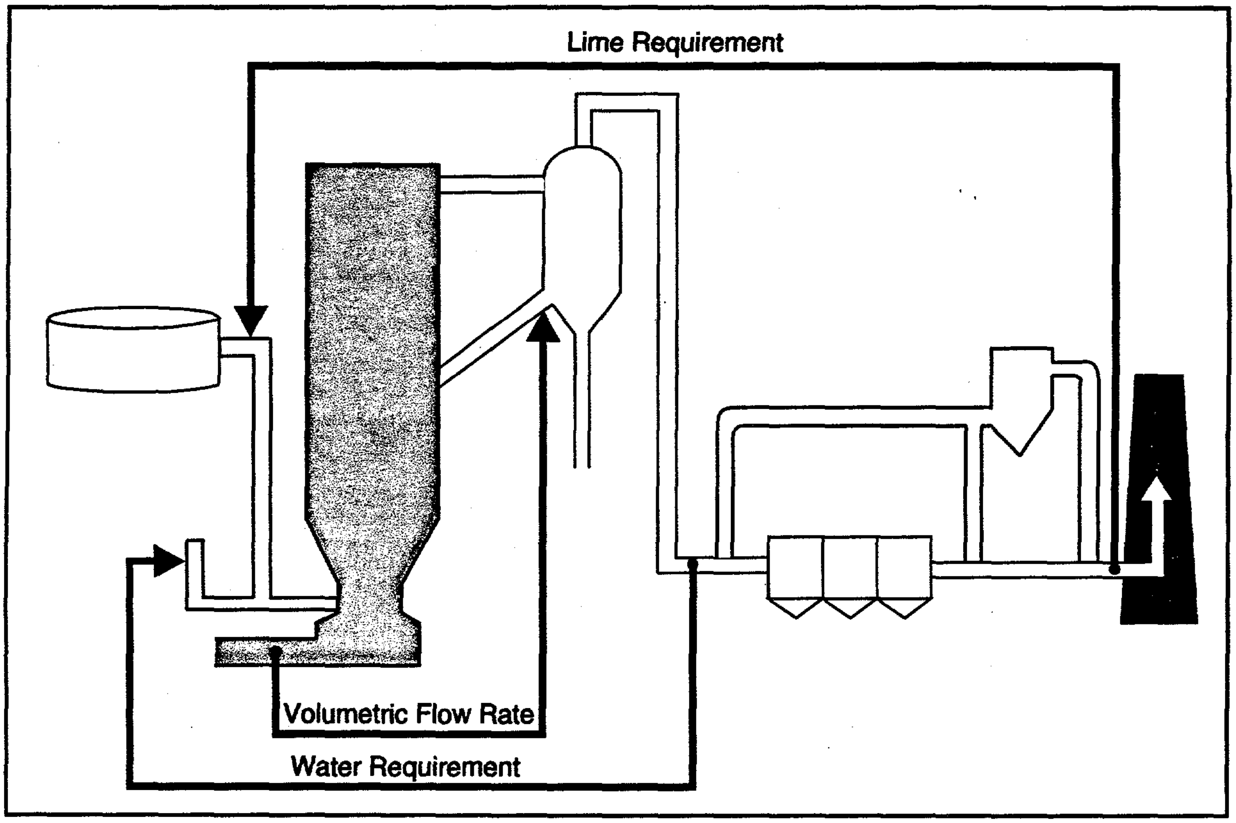


Figure 2. Gas Suspension Absorption Control System

IV. COMPARISON OF GSA PROCESS WITH COMPETING TECHNOLOGY

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

Slurry Atomization

The major difference between GSA and competing technologies lies in the way the reagent is introduced and used for SO_2 absorption. A conventional semi-dry scrubber:

- Requires a costly and sensitive high-speed rotary atomizer or a high-pressure atomizing nozzle for fine atomization,
- Absorbs SO_2 in an "umbrella" of finely atomized slurry with a droplet size of about 50 microns,
- May require multiple nozzle heads or rotary atomizers to ensure fine atomization and full coverage of the reactor cross section, and
- Uses recycle material in the feed slurry necessitating expensive abrasion-resistant materials in the atomizer(s).

The GSA process, on the other hand,

- Uses a low-pressure, dual-fluid nozzle,
- Absorbs SO_2 on the wetted surface of suspended solids with superior mass and heat transfer characteristics,
- Uses only one spray nozzle for the purpose of introducing slurry and water to the reactor, and
- Uses dry injection of recycle material directly into the reactor, thereby avoiding erosion problems in the nozzle or technical limitations on the amount of solids that can be recycled.

Simple and Direct Method of Lime/Solid Recirculation

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

High Acid Gas Absorption

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

Low Lime Consumption / Minimum Waste Byproduct Residue

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

Low Maintenance Operation

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

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

No Internal Buildup

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

Modest Space Requirements

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

Short Construction Period

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

Heavy Metals Removal

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

V. PROJECT STATUS AND KEY MILESTONES

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

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

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

VI. TEST PLAN

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

- Optimization of the operating variables.
- Determination of stoichiometric ratios for various SO_2 removal efficiencies.
- Evaluation of erosion and corrosion at various locations in the system.
- Demonstration of 90% or greater SO_2 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_2 removal efficiency in the GSA system was accomplished through the completion of a statistically-designed factorial test plan. For each test series, the GSA system was set to operate at a certain combination of operating parameters. The results of these test series were analyzed statistically to determine the impact of the parameters, thus arriving at the optimum operating point for the GSA process at the various operating conditions expected in future applications. Operating parameters studied in this phase of the project were the following:

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

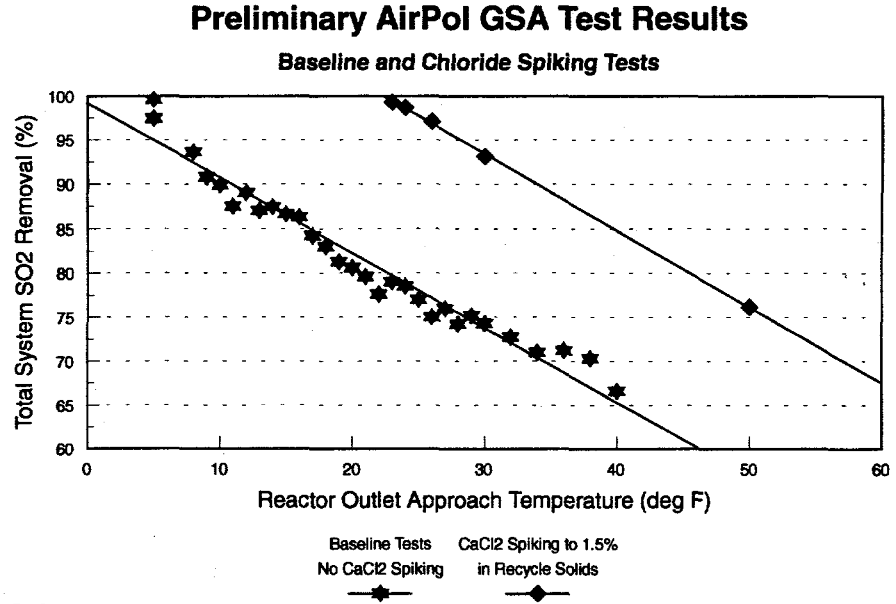
Data Collection

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

- Inlet flue gas flow into the system
- SO_2 loading at the system inlet, SO_2 loading at the ESP inlet and outlet
- Flue gas temperature at the system inlet, the reactor outlet, and the ESP outlet
- Particulate loading at the ESP inlet and outlet
- Fresh lime slurry flow rate and composition (for lime stoichiometry calculation)
- Water flow rate
- Wet-bulb temperature at the reactor inlet (for approach-to-saturation temperature calculation)
- Coal analysis (proximate and ultimate)
- Lime analysis
- Byproduct rate and composition
- Water analysis
- Power consumption

VII. PRELIMINARY TESTING

Immediately after the dedication of the AirPol GSA demonstration plant in late October 1992, a series of preliminary tests was begun. The purpose of these tests was to investigate the operating limits of the GSA system as installed at the CER. The results from several of the preliminary tests completed at the CER in November and December were very interesting, and these results were used as the basis for the design of the factorial test program. During one of the preliminary tests, the approach-to-saturation temperature in the reactor was gradually decreased and the overall system (reactor/cyclone and ESP) SO_2 removal efficiency was monitored over this four-day test. The overall system SO_2 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 $Ca(OH)_2$/mole inlet SO_2 for the lime stoichiometry, and essentially no chloride in the system. The SO_2 removal results from this test are shown in Figure 3.



Test Conditions: Inlet flue gas temperature of 313 to 320F, 1.4 stoich, recycle screw speed of 23 to 38 rpm, 3.0% S/0.03% Cl coal, Miss. lime.

Figure 3. Preliminary AirPol GSA Test Results

The data from this test show that the SO_2 removal efficiency increased dramatically as the flue gas temperature in the reactor more closely approached the saturation temperature of the flue gas, with the incremental increases in SO_2 removal becoming more and more significant as the approach-to-saturation temperature declined. The ability of the GSA system to operate at this close approach-to-saturation temperature without any indication of plugging problems was surprising. Later analysis showed that the moisture level in the solids remained below 1%.

A second extended test was run during December 1992. This test was run at the same conditions as the previous test, except that in this test, calcium chloride was added to the system to simulate the combustion of a high-chloride (about 0.3%) coal. Previous work by TVA at the CER had demonstrated that spiking these semi-dry, lime-based FGD processes with a calcium chloride solution adequately simulated a high chloride coal application. Again, the approach-to-saturation temperature was gradually decreased over a four-day period with all other conditions held constant and the overall system SO_2 removal efficiency was monitored. The results from this second test, which are included in Figure 3 above, show that the presence of chlorides enhances SO_2 removal.

The overall system SO_2 removal efficiency for the chloride-spiked tests increased from about 70% at the high approach-to-saturation condition to essentially 100% at the closer approach-to-saturation temperature (23°F). No attempt was made to operate the system at the close approach-to-saturation temperatures used in the first test because the SO_2 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_2, but also increases the potential for plugging in the system. The easiest method for mitigating this potential for plugging is to increase the approach-to-saturation temperature in the reactor. However, the moisture levels in the solids during this test remained below 1%, even at the closest approach-to-saturation temperature.

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

VIII. FACTORIAL TESTING

The purpose of the statistically-designed factorial test program was to determine the effect of process variables on the SO_2 removal efficiency in the reactor/cyclone and the ESP.

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

Major Variables and Levels for Factorial Testing Table		
Variable		Level
Approach-to-saturation temperature	°F	8[a], 18, and 28
Ca/S	moles $Ca(OH)_2$/mole inlet SO_2	1.00 and 1.30
Fly ash loading	gr/acf	0.5 and 2.0
Coal chloride level	%	0.02 and 0.12
Flue gas flow rate	kscfm	14 and 20
Recycle screw speed	rpm	30 and 45
[a] 8°F level run only at the low-chloride level		

Table 1. Major Variables and Levels for Factorial Testing

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

IX. RESULTS OF FACTORIAL TESTING

SO_2 Removal Efficiency

The overall system SO_2 removal efficiency results from these factorial tests have been analyzed, and several general relationships have become apparent. First, as was expected based on previous testing at the CER, significant positive effects on the SO_2 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_2 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_2 removal. Increasing the flue gas flow rate had a negative effect on the SO_2 removal in the system.

The overall system SO_2 removal efficiency during these tests ranged from slightly more than 60% to nearly 95%, depending on the specific test conditions. The higher SO_2 removal efficiency levels were achieved at the closer approach-to-saturation temperatures (8 and 18°F), the higher lime stoichiometry level (1.30 moles $Ca(OH)_2$/mole inlet SO_2), and the higher coal chloride level (0.12%). The lower SO_2 removal efficiency levels were achieved at the higher approach-to-saturation temperature (28°F), the lower lime stoichiometry level (1.00 mole $Ca(OH_2$/mole inlet SO_2), and the lower coal chloride level (0.02-0.04%). The data from these factorial tests completed at these conditions are shown in Figure 4. The slight scatter in the data in this figure is due to the variations in the other major process variables in these tests (i.e. flue gas flow rate, recycle screw speed, etc.). Most of the SO_2 removal in the GSA system occurs in the reactor/cyclone, with only about 2 to 5 percentage points of the overall system removal occurring in the ESP. There was substantially less SO_2 removal in the ESP than in previous testing at the CER, but the overall system SO_2 removal efficiencies appear to be comparable with the GSA process for most test conditions.

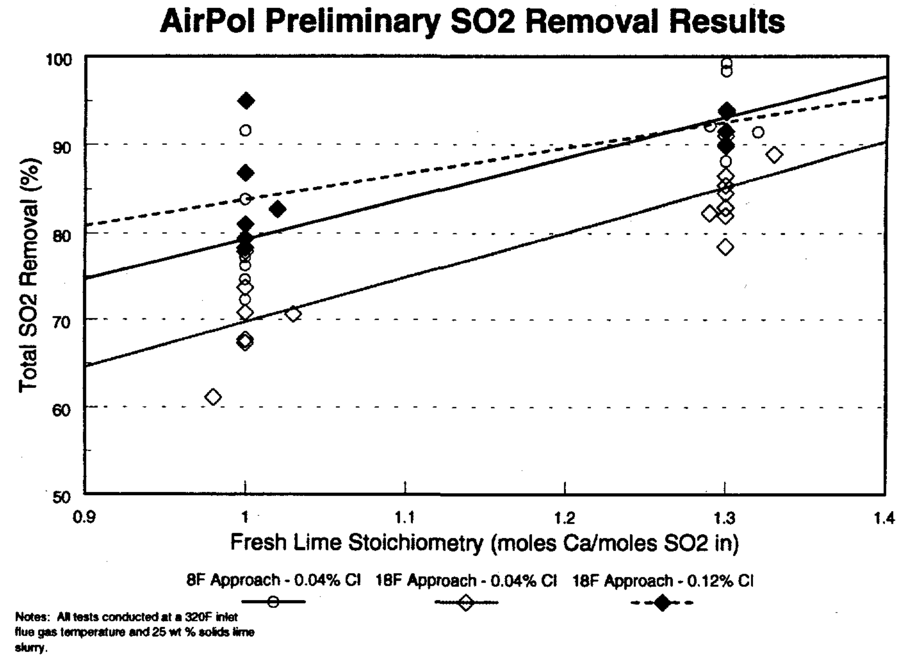


Figure 4. Overall System SO_2 Removal Results from the GSA Factorial Testing

As one would expect, the lime stoichiometry level, which was tested at 1.00 and 1.30 moles $Ca(OH)_2$/mole inlet SO_2, seems to have the most significant effect on the SO_2 removal efficiency in the GSA system.

The approach-to-saturation temperature, which was evaluated at three levels of 8, 18, and 28°F for the low coal chloride conditions and the two levels of 18 and 28°F for the higher coal chloride condition, appears to be the second most important variable in the GSA system in terms of the overall system SO_2 removal efficiency.

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

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

ESP Performance

The ESP installed at the CER is a relatively modern, 4-field unit with 10-inch plate spacing, similar in design to several full-scale ESPs installed on the TVA Power System. This unit has 23-feet-high plates with 8 parallel gas passages. The specific collection area (SCA) of the unit is about 440 ft^2/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^2/kacfm.)

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

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

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

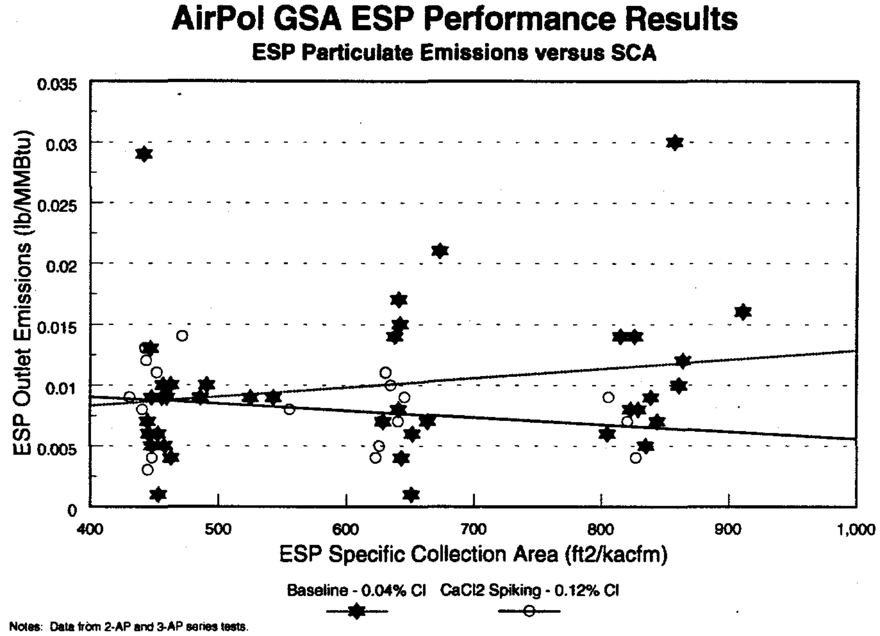


Figure 5. ESP Performance Results from the GSA Factorial Testing

Pulse Jet Baghouse Performance

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

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

SO_2 Removal Efficiency for Reactor/Cyclone/PJBH System

The SO_2 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_2 removal efficiency in the PJBH system was not unexpected given the intimate contact between the SO_2-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_2 removal occurred in the reactor/cyclone; the PJBH SO_2 removal efficiency, based on the inlet SO_2 to the reactor, contributed less than 8 percentage points to the overall system SO_2 removal efficiency during this testing.

Particulate Removal

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

X. AIR TOXICS TESTING

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

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

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

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

Table 2. Sampling Locations For The Air Toxics Tests.

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

XI. RESULTS OF AIR TOXICS TESTING

Tables 3 and 4 present the removal efficiencies and uncertainties of the baseline and demonstration case with varying ESP and baghouse configurations. Removal efficiencies for beryllium and nickel were not determined due to analytical laboratory error. The removal efficiency for most trace metals is generally over 90 percent. Caution is required when reviewing the removal efficiency of antimony, since most of the antimony measurements were below detection limits. Mercury concentration was also low. Only trace levels of mercury, i.e. close to the method detection limits, could be detected in the baseline and parallel tests. The removal efficiency for mercury appears to fall in the 50%-95% range.

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

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

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

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

Table 3. Baseline Tests Removal Efficiencies and Uncertainties

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

Table 4. Demonstration Tests Removal Efficiencies and Uncertainties

XII. DEMONSTRATION RUN

28-day GSA/ESP Demonstration Run

The 28 day demonstration run, with GSA operating in conjunction with ESP only, started on October 25, 1993 and ended on November 24, 1993. This demonstration run began with the boiler burning the high-sulfur (2.7%), low-chloride Andalex coal and test conditions of: 320°F inlet flue gas temperature; 18°F approach-to-saturation temperature; 1.5 gr/acf fly ash injection; 0.12 percent coal chloride level; 20,000 scfm flue gas flow rate; and 30 rpm recycle screw speed. The SO_2 control mode was engaged for this run with an overall system SO_2 removal efficiency set-point of 91 percent. Due to some problems encountered in obtaining the test coal, a switch was made to burning a higher-sulfur (3.5%) coal for a period of time. The Ca/S ratio averaged 1.40 - 1.45 moles of $Ca(OH)_2$/mole inlet SO_2 during this demonstration run.

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

- The overall system SO_2 removal efficiency averaged 90-91 percent, i.e., very close to the set-point. The switch to the higher-sulfur coal demonstrated the flexibility of the GSA system

- The particulate removal efficiency was good at an average of 99.9+ percent, with an emission rate below 0.015 lbs/MBtu.

- The GSA system demonstrated the reliability of this technology by remaining on-line for the entire 28-day period that the boiler was operating.

14-day PJBH Demonstration Run

The purpose of the 14-day demonstration run was to demonstrate that the GSA system (reactor/cyclone/PJBH), as installed at the CER, could operate reliably and continuously, while simultaneously achieving 90+ percent SO_2 removal and maintaining the PJBH outlet emissions below the NSPS for particulates.

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

- The overall system (reactor/cyclone/PJBH) SO_2 removal efficiency averaged more than 96 percent during the entire 14-day demonstration run.

- The average Ca/S level during this run ranged from about 1.34 to 1.43 moles $Ca(OH)_2$/mole inlet SO_2.

- The PJBH particulate removal efficiency averaged 99.99+ percent. The emission rate was 0.001 to 0.003 lbs/MBtu.

XIII. ECONOMIC EVALUATION

Under the scope of this project, Raytheon Engineers & Constructors prepared an economic evaluation of the GSA FGD process using the same design and economic premises that were used to evaluate about 30-35 other FGD processes for the Electric Power Research Institute. The relative process economics for the GSA system were evaluated for a moderately difficult retrofit to a 300-MW boiler burning a 2.6 percent sulfur coal. The design SO_2 removal efficiency was 90 percent.

The resulting capital cost estimate (in 1990 dollars) is shown in Table 5 together with the estimate for the conventional wet limestone, forced-oxidation (WLFO) scrubbing system. The total capital requirement of $149/kW for the GSA process is substantially lower than the $216/kW for the WLFO system. The significant reduction in capital is primarily due to lower costs in the SO_2 absorption area.

Total Capital Investment Comparison
(1990 $, 300-MW, 2.6% S coal)

	$/kW	
Area	GSA	WLFO
Reagent Feed	25	37
SO_2 Removal	38	71
Flue Gas Handling	18	24
Solids Handling	5	7
General Support	1	2
Additional Equipment	4	4
Total Process Capital	91	145
Total Capital Requirement	149	216

Table 5. Total Capital Investment Comparison

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

LEVELIZED COSTS
(300-MW, 2.6% S coal, 15-year levelizing)

	Mills/kWh	
Fixed Costs	GSA	WLFO
Operating Labor	0.52	0.66
Maintenance	1.49	1.74
Administrative and Support Labor	0.34	0.41
	2.35	2.81
Variable Costs		
Raw Material	1.82	0.65
Solids Disposal	0.86	0.57
Water	0.01	-
Steam	-	0.55
Electricity	0.47	1.16
	3.16	2.93
Fixed Charge (Capital)	5.40	7.30
Total	10.91	13.04

Table 6. Levelized Costs

XIV. COMMERCIALIZATION

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

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

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

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

(1) Install the GSA upstream of the ESP and extend the unit to attain sufficient capacity.

(2) Use the ESP as primary dust collector upstream of the GSA with a new final dust collector

(3) Demolish the ESP and replace it with the GSA and new final dust collector

(4) Leave the ESP in place, de-energize it, and connect the GSA with final collector downstream

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

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

The technical data for the Hamilton installation are as follows:

Boiler Capacity	50 MWe
Type of Boiler	Pulverized Coal
Type of Coal	Ohio 3% + sulfur
Gas Volume	224,728 ACFM
Gas Temperature	320 °F
Moisture Content	7.8 % by Volume
Oxygen Content	4.6 % by Volume
Particulate Content	2 gr/SCFD
SO_2 Content	2,612 PPMd
SO_2 Removal	90 % Design

Another GSA installation for a fossil fuel boiler is being installed in Kaohsiung, Taiwan. The installation is in a sugar refinery where two oil fired boilers each have a dedicated GSA. The larger boiler has a steam generating capacity of 100 TPH, while the smaller one generates 35 TPH. Both GSA units are equipped with fabric filters, dual ID fans, and a gas recirculating system. The reason for the dual fans and the recirculation is that both boilers have great load swings, and in order to attain the required SO_2 removal efficiency, the GSA must run with at least 50% of design gas volume. When the boiler capacity is reduced below the design capacity, a portion of the flue gas is recirculated via a separate fan from the outlet of the fabric filter to the inlet of the GSA reactor to maintain minimum design gas flow.

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

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

The technical data for the two plants are as follows:

Boiler Steam Rating	100 TPH	35 TPH
Type of Fuel	Oil #6	Oil #6
Gas Volume	97,554 ACFM	27,748 ACFM
Gas Temperature	298 °F	280 °F
Moisture Content	13.36 % by Volume	3.5% by Volume
Oxygen Content	1.96 % by Volume	1.74 %by Volume
Particulate Content	2.4 gr/SCFD	2.3 gr/SCFD
SO_2 Content	510 PPMd	517 PPMd
SO_2 Removal	80 % Design	80 % Design

In addition to the Hamilton and Taiwan installations, approximately 20 GSA plants for refuse and hazardous waste incineration are in operation, most of them in Europe. Some of these installations have very sophisticated control equipment for NOx, furans, and dioxins with extremely low outlet concentrations.

XV. DISCLAIMER

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

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

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ABSTRACT

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

I. INTRODUCTION

This paper updates the results of work performed on Coal Tech's commercial scale 20 MMBtu/hour air cooled, slagging coal combustor since the last report at the 1995 Clean Coal Conference [1]. During the past year, a second generation, 20 MMBtu/hr combustor has been placed in operation in a coal combustion system. It incorporates all the features of Coal Tech's new low power cost, solid fuel plant. The central feature of this plant is an air cooling combustor whose wall heat transfer loss is recuperated to the combustion air, making this heat available to the thermodynamic cycle. A portion of the SO_2 and NO_x emissions are controlled inside the combustor, which is designed for new and retrofit boiler applications. Coal Tech's development of the air cooled combustor began in the late 1970's in a 1 MMBtu/hr air cooled, cyclone combustor [2], continued in the mid 1980's in a 7 MMBtu/hr water cooled, cyclone combustor [3], and was followed by 2000 hours of operation of a first generation, 20 MMBtu/hr, air cooled combustor between 1987 and 1994 [4-7]. The latter facility was located in an industrial heating plant in Williamsport, PA. Fuels tested include coal, coal water slurry, refuse derived fuel, oil, and gas. Test operations to 1991 were sponsored in part by the United States Department of Energy Clean Coal Technology Program [4].

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

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

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

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

Coal Tech's Advanced Air Cooled, Cyclone Coal Combustor

The cyclone combustor is a high temperature (> 3000^{o}F) device in which a high velocity swirling gas is used to burn crushed or pulverized coal. Figure 1 shows a schematic of Coal Tech's patented, air cooled combustor. Gas and oil burners rapidly preheat the combustor and boiler during startup. Pulverized coal and powdered sorbent for SO_2 control are injected into the combustor in an annular region enclosing the gas/oil burners. Air cooling is accomplished by flowing combustion air through tubes on the outside of a ceramic liner in the combustor. This cooling air provides over 90% of the combustion air in the combustor, and it is introduced tangentially in a swirling manner into an annulus enclosing the fuel injection cylinder in the combustor, (see figure 1). The ash and reacted sorbent melt on the liner and the resultant slag is drained through a tap at the downstream end of the combustor.

Nitrogen oxide emissions are reduced by operating the combustor in a fuel rich mode, with final combustion taking place in the boiler. Operations in the first generation 20 MMBtu/hr combustor in Williamsport yielded, under optimum conditions, about two-thirds NO_x reductions to 0.26 lb of gas/MMBtu, or 200 ppm (at 3 % O_2) at about 70% of stoichiometric air/fuel ratio in the combustor and high combustion efficiencies. The stoichiometric ratio for the combustor/boiler was between 1.25 and 1.5. Sulfur emissions are controlled primarily by sorbent injection into the combustor. Measurement of SO_2 levels at the stack gas outlet from this previous boiler yielded average SO_2 reductions of 50% to 70%, and as high as 85%, with calcium hydrate injected into combustor at Ca/S mol ratios of 3 to 4. Particulate emissions were controlled in part by slag retention in the combustor. It was augmented with a wet particle scrubber which reduced the particle emissions to as low as 0.26 lb of solids /MMBtu.

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

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

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

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

Figure 2 is a side view of the new 20 MMBtu/hr combustor/boiler installation in Philadelphia. PA. Its total size is such that it can be shipped by tractor trailer to any site. Figure 3 shows a plan and side view of the Philadelphia facility. It includes provision for a 25 ton raw coal delivery and storage area, a low cost coal mill, a 4 ton pulverized coal storage bin, sorbent storage bins, pneumatic coal and sorbent delivery, a boiler, the combustor and its auxiliary subsystems, specifically, water cooling, oil, gas, combustion air, cooling air, compressed air, slag removal, and the stack system, including the baghouse, and induced draft fan. The entire system is controlled by programmable logic controllers (PLC) and computer process control. Performance parameters are measured and recorded on a computer. Combustion gases, O_2, CO, NO_x, and SO_2, are measured in the boiler radiant furnace section, boiler stack outlet, and baghouse outlet. Novel Features and Operating Experience of the Second Generation Combustor/Boiler Facility.

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

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

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

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

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

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

A high maintenance item has been the combustor's slag tap assembly, primarily during the initial tests. Subsequent modifications were made which have sharply reduced maintenance to this item. The relay controlled system used in the previous combustor system was replaced with programmable logic controllers (PLC). The PLC assure that the combustor's fuel supply and the boiler's steam supply operate with all safety interlocks functioning. The previous computer process control software was upgraded to account for the changes in the design of the present combustor. As the test effort proceeded, it was found that the combustor could be controlled with a much simpler procedure than was used for the previous combustor, and the software was changed accordingly.

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

20 MMBtu/hr Combustor Operation in the Philadelphia Facility

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

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

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

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

To minimize nitrogen oxide emissions it is necessary to operate the combustor under fuel rich conditions. Final combustion occurs in the furnace section of the boiler where the CO and H_2 rich combustor gas exhaust is mixed with additional air to complete combustion. Optimum No_x reduction occurs at about 70% stoichiometric air/fuel ratio in the combustor. [3,4]. However, operation of the earlier 20 MMBtu/hr combustor at this condition resulted substantially reduced combustion efficiency [3,4].

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

Both these measurements of combustion efficiency do not account for unburned carbon that is carried over to the stack baghouse. Due to the difficulty in obtaining real time sampling of the baghouse fly ash, the carbon content in the fly ash was determined from random grab samples taken from all the ash collected on the day of testing. The carbon content of the ash ranged from 20% to 50% (dry basis). Since on average about one-third of mineral matter injected reported to the baghouse, one can compute the conversion of the solid carbon in the coal to CO2 and CO in the combustor from the amount of unburned carbon in the baghouse fly ash. This yielded a carbon conversion greater than 90% for most of the tests. In several tests small quantities of fly ash in the stack were collected in a filter. Analysis of the carbon content in one of these tests yielded a carbon conversion of 94%.

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

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

The initial test effort this year has been focused on overall combustor performance, with lesser emphasis on SO_2 and NO_x control. The most recent tests have focused on SO2 control and excellent results have been achieved, especially in low sulfur coal. The analysis of these data is incomplete, and the results will be presented at the Conference. Both limestone and hydrated lime were injected into the combustor for slag conditioning and sulfur removal. Previous results in the 20 MMBtu/hr combustor in Williamsport showed that limestone was much less effective than calcium hydrate for sulfur capture. With calcium hydrate injection into the previous combustor, excellent sulfur capture results were achieved, where a maximum reduction in the 85% range was measured [5,7].

The degree of sulfur capture in the present combustor was found to be very sensitive to combustion conditions, the method and quantity of sorbent injection, and the mineral matter injection rate. In recent tests at high slag mass flow rates firing 3+% sulfur coal, SO_2 reductions measured in the end wall of the boiler furnace and in the boiler gas outlet at the stack were in the range of 60% to 75% at Ca/S mol ratios of under 3. Similar reductions were measured with injection of calcium hydrate into the boiler furnace near the combustor gas inlet to the boiler. In this case, the Ca/S mol ratios were in the range of 3.5 to 4.9. However, in the latter case, a substantial amount of the hydrate fell to the floor of the boiler furnace. Therefore, the Ca/S mol ratio is not an accurate measure of calcium utilization in this case.

In very recent tests with 1.5% sulfur coal, the SO_2 reductions were substantially higher. Reductions in the range of 75% to as high as 95% were measured. When expressed in lb/MMBtu, SO_2 emissions as low as 0.22 lb/MMBtu were measured. This is well below the 0.5 lb/MMBtu SO_2 emission standard for Philadelphia, and it near the 0.2 lb/MMBtu that is one of the current test objectives. The reductions were higher at the end wall of the boiler furnace than in the stack at the outlet of the boiler. No conclusive explanation for this behavior has been found. It is suspected the higher SO_2 at the boiler outlet may due to blowby of combustion gases through gaps in between the boiler tubes on the convective tube side of the boiler. This reduces the reaction time of sorbent with combustion gas in the boiler furnace. This matter should be clarified when all the data are analyzed.

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

NO_x emissions are controlled by operating the combustor fuel rich. Maximum reductions to as low as 0.26 lb/MMBtu were measured in the previous combustor at stoichiometric ratio (SR1) in the range of 0.7. In the present combustor, the tests were performed at less fuel rich conditions. At slightly fuel rich conditions, the NO_x, (reported as NO_2) has been in the range of 0.36 to 0.7 lb/MMBtu. The 0.36 lb/MMBtu value was measured at SR1 equal to 0.9.

The above discussion presents a general overview of the performance of the present second generation 20 MMBtu/hr combustor. Its performance has been found to be superior to the previous combustor, especially in the areas of combustion efficiency, slag retention and removal from inside the combustor, durability of the internal combustor wall, and simplicity of control and operation. Considerable data on SO_2 on NO_x control have been measured, but the analysis is not complete. Based on the results obtained to date, the best results achieved in the current combustor (namely SO_2 reductions to 0.22 lb/MMBtu and NO_x reduction to 0.36 lb/MMBtu) match those measured in the prior unit (SO_2 reductions to 0.34 lb/MMBtu and NO_x reduction to 0.26 lb/MMBtu).

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

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

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

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

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

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

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

IV. CONCLUSIONS

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

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

V. ACKNOWLEDGMENTS

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VI. REFERENCES

1. B. Zauderer, R. Frain, K. Peng, and B. Borck, "Demonstration of the Air Cooled Slagging Combustor on a 500 kW Coal Fired Power Plant",, in Proceedings 4th Annual Clean Coal Technology Conference, Denver, CO, Sept., 1995
2. C.S. Cook, Ed., "Evaluation of Closed Cycle MHD Power", DOE Contract Report No. DE-AC01-78-ET-10818, November 1981.
3. B. Zauderer, E. S. Fleming,"Experiments in NO_x and SO_2 Control in a Cyclone Combustor", 6th International Workshop on Coal Liquid and Fuels, Halifax, N.S., October 1986.
4. B. Zauderer, E. S. Fleming, "The Demonstration of an Advanced Cyclone Coal Combustor with Internal Sulfur , Nitrogen, and Ash Control., Vols. 1-3, Coal Tech Corp., August 1991. (NTIS DE92002587 & DE92002588T)

5. B. Zauderer, E. S. Fleming , "Development & Testing of an Industrial Scale, Coal Fired Boiler Slagging Combustor System", in Proceedings 10th Annual Coal Preparation, Utilization, & Environmental Control Conference, (DOE/PETC, Pittsburgh, PA, July, *1994)*
6. B. Zauderer, R. Frain, K. Peng, and B. Borck, "Development & Testing of an Industrial Scale, Coal Fired Boiler Slagging Combustor System", in Proceedings 11th Annual Coal Preparation, Utilization, & Environmental Control Conference, (DOE/ PETC, Pittsburgh, PA, July 12, 1995)
7. B. Zauderer, E. S. Fleming, B. Borck, and C. A. Smith, "Coal Tech's Air Cooled Slagging Combustor-Recent Developments", in Proceedings 3rd Annual Clean Coal Technology Conference, Chicago, IL, Sept.7,1994

Figure 1
Coal Tech's Air Cooled Combustor

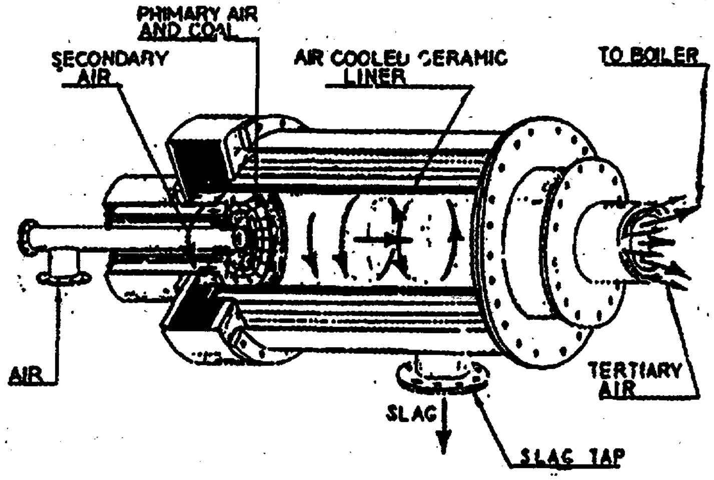
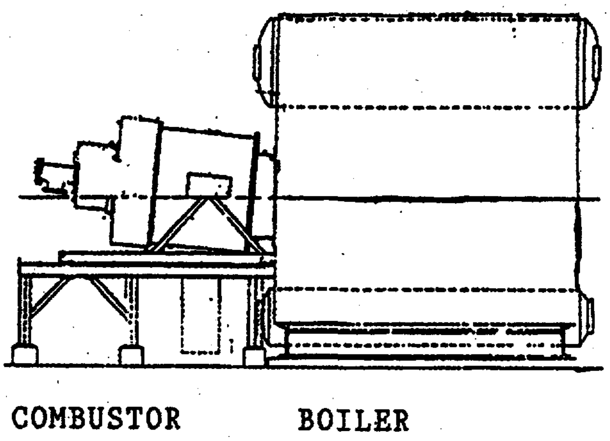


Figure 2:
20 MMBtu/hr Combustor-Boiler Installation at Philadelphia Plant



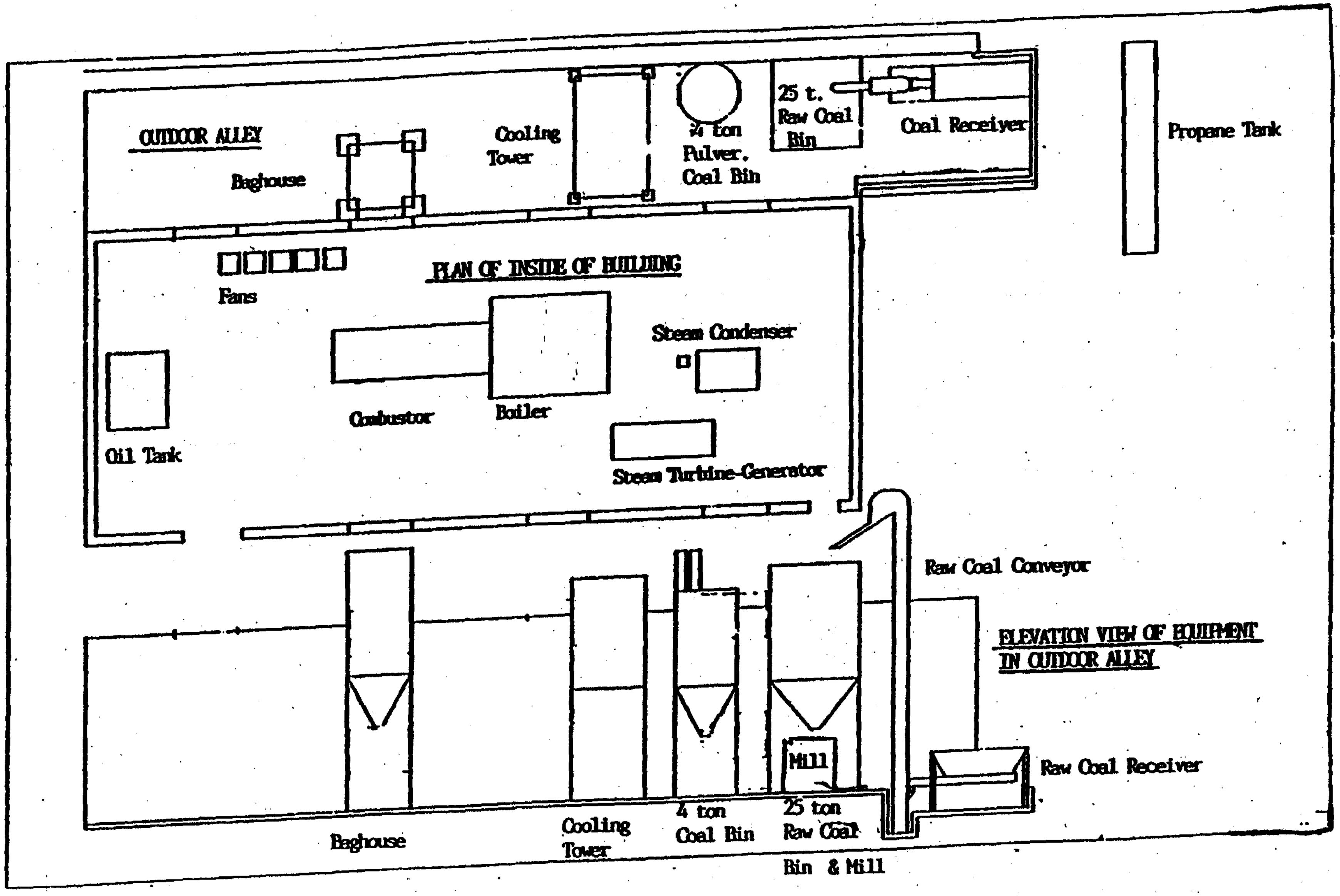


Figure 3: Plot Plan of the 20 MMBtu/hr Combustor-Boiler Test Site in Philadelphia.

CHIYODA THOROUGHBRED CT-121 CLEAN COAL PROJECT AT GEORGIA POWER'S PLANT YATES

PHASE II RESULTS

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ABSTRACT

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

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

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

I. INTRODUCTION

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

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

Parametric testing was also conducted under moderate-ash loading conditions while burning both high- and low-sulfur coals. The data gathered were regressed and multi-variable regression models were developed to provide an accurate prediction of the scrubber's SO_2 removal efficiency under the most likely future operating conditions. As part of the moderate-particulate removal evaluation, limited air toxics measurements were also performed for the second time. The purpose of this additional testing was to evaluate air toxics removal across the CT-121 under elevated ash loading conditions as well as to validate or controvert the findings of an earlier air toxics testing effort that was sponsored by DOE in June of 1993[1].

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

In general, the Yates CT-121 process performed well, exhibiting excellent SO_2 removal efficiency, particulate removal and consistent reliability. In addition to these successes, several possible process improvements were identified during the demonstration that could improve future designs of an already superior process.

II. FACILITY AND OPERATING DESCRIPTION

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

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

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

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

The central feature of the process is Chiyoda's unique absorber design, called a Jet Bubbling Reactor (JBR), which combines concurrent chemical reactions of limestone dissolution, SO_2 absorption/neutralization, sulfite oxidation, gypsum precipitation and gypsum crystal growth together in one vessel. A cut-away view of the JBR is illustrated in Figure 2. Since much of the undesirable crystal attrition and secondary nucleation associated with the large centrifugal pumps in conventional FGD systems is eliminated in the CT-121 design, large easily dewatered gypsum crystals are consistently produced. This design also significantly reduces the potential for gypsum scale growth, a problem that frequently occurs in natural-oxidation FGD systems.

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

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

The JBR slurry reservoir provides about 36 hours of solid-phase residence time, depending on the SO_2 pick-up rate. The slurry from the JBR is pumped intermittently to a gypsum slurry transfer tank (GSTT) for JBR slurry level control and slurry density control. In the GSTT, the slurry is diluted for pumping to a Hypalon®-lined gypsum (or gypsum/ash) stacking area for gravity dewatering and storage. Gypsum stacking is a disposal technique that involves filling a diked area with slurry for gravity sedimentation. Over time, this area fills with settled solids. The filled area is then partially excavated to increase the height of the containment dikes. The repetitive cycle of sedimentation, excavation, and raising of perimeter dikes continues on a regular basis during the active life of the stack. Process water is naturally decanted, stored in a surge pond and then returned to the CT-121 process. There is no blowdown or discharge from the Yates CT-121 process.

During normal operation of the FGD system, the amount of SO_2 removed from the flue gas is controlled by varying the JBR pressure drop (ΔP) or slurry pH. However, changing ΔP is easier and quicker to respond to changing conditions since it is done by adjusting the JBR liquid level. Higher liquid levels result in increased SO_2 removal because of increased contact time between the incoming flue gas and the scrubbing slurry. The pH can also be varied to affect SO_2 removal with higher pH resulting in increased removal efficiency. Boiler load and flue gas SO_2 concentration also affect removal efficiency, but are less controllable.

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

III. PROJECT OBJECTIVES

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

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

Many of these objectives were investigated during this last trimester of the second phase of the demonstration project, also known as the High-Particulate Auxiliary Test block. Two of the test periods in this test block provided data relevant to the focus of this paper:

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

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

IV. RESULTS

The CT-121 scrubber at Plant Yates continued to prove itself a very viable and cost effective technology for use in Clean Air Act, Title IV compliance. It exhibited excellent availability, maintained greater than 97% limestone utilization, and demonstrated the ability to exceed 98% SO_2 removal efficiency with high sulfur coal, while at maximum boiler load. The flexibility of the CT-121 process was also demonstrated through the use of a wide range of coals, varying from 1.2% to 4.3% sulfur content.

Operating Statistics

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

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

1. Reliability = Hours scrubber operated divided by the hours called upon to operate.
2. Availability = Hours scrubber available divided by the total hours in the period.
3. Utilization = Hours scrubber operated divided by the total hours in the period.

Table 1. Summary of Operating Statistics

Effect of Inlet SO_2 Concentration

The SO_2 removal efficiency of the scrubber was measured under five different inlet SO_2 concentration ranges; three during this most recent testing. The coal burned by Unit 1 for a majority of the testing was a blend of Illinois No.5 and No.6 bituminous coal that averaged 2.4% sulfur (as burned), except for a brief, unplanned period when 3.0% sulfur coal was burned. A 4.3% sulfur bituminous coal was burned during the Low-Particulate Alternate Coal Test block, and a 3.8% sulfur coal was burned for the High-Particulate Alternate Coal Test block. The High-Particulate Alternate Limestone Test (last test of the demonstration project) coincided with Plant Yates' compliance-driven transition to a low sulfur coal (approximately 1.2% S). This provided the scrubber project an opportunity to evaluate a fifth coal source.

The effect of inlet SO_2 concentration on SO_2 removal efficiency is quite significant. Figure 4 illustrates the decrease in SO_2 removal as inlet SO_2 concentration increased for the coal sources evaluated. Performance of the scrubber was outstanding during the low-sulfur coal burn. It should be noted that the low-sulfur coal tested limited the JBR pH to a maximum of 3.8 because of Aluminum-Fluoride-inhibited limestone dissolution (Al-F blinding). The Al-F blinding stems from the low-ionic strength of the scrubbing liquor, the elevated ash loading to the JBR and the coal trace metals concentrations. A maximum operating pH of 3.75 was chosen to ensure that near-complete limestone utilization was maintained in the scrubber. The test data from 1000 ppm (inlet SO_2 concentration) operations indicates that SO_2 removal efficiency did not decline at a slightly lower pH.

The evaluation of five different inlet SO_2 concentrations demonstrates the flexibility of the CT-121 process as well as it's exceptional SO_2 removal capability, even when burning fuels with a very high sulfur content. This is even more impressive considering that the maximum designed sulfur content for the demonstration unit was only 3.0%, and that this limit was exceeded by 43% in one test period. Other test data shows that even higher SO_2 removal efficiencies are achievable at higher pH values.

Particulate Removal Efficiency

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

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

Table 2. ESP Configuration during Particulate Testing

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

Although the outlet particulate loading varied from 0.005 to 0.029 lb/MMBtu, analytical results indicate that from 20 to 80 percent of outlet particulate is sulfate (SO_4). Based on the calcium analyses performed on the same material, it is believed that the measured sulfate originated from gypsum carryover and acid mist carryover, so it is scrubber-generated. This finding reduces the estimate of actual ash mass loading at the outlet of the scrubber (actual fugitive emissions) to approximately 70% of the amount captured, measured and recorded during outlet testing.

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

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

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

Table 3. Particulate Removal Testing - Summary of Results

Particulate Removal Efficiency by Particle Size

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

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

As was reported during earlier particulate removal tests, and again observed in Figure 5, the best removal efficiencies were observed for particle sizes greater than 10μm. At all test conditions, there was greater than 99% particulate removal efficiency of particles in this size bin. In some cases, efficiency exceeded 99.99%. As the particle size decreased, there was a drop in observed particulate removal efficiency, but over 90% efficiency was observed at all particle sizes between 1μm and 10μm. Between 0.5μm and 1μm, the particulate removal dropped to sometimes negligible values. In this range, it is believed that acid mist carryover offset the ash particulate removal, resulting in poor particulate removal values. Analyses of the outlet catch indicated that an average of 30% of the outlet particulate can be attributed to gypsum and acid mist carryover. Below about 0.5μm, the particulate removal efficiency increased to above 90%. Also observed in Figure 7 was a higher particulate removal efficiency at the higher JBR ΔP values. This increase in removal efficiency ranged from 1 decade (90%), at the largest particle sized, to less than 1/10th of a decade (10%) at the 0.5μm cut-point. The increased particulate removal at the higher JBR ΔP in this size range results from a deeper sparger tube submergence depth and therefore, a longer gas-phase residence time allowing more opportunity for the particulate to be captured in the slurry.

V. AIR TOXICS TESTING

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

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

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

From the 1993 results, the DOE concluded that:

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

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

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

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

In comparing the results of the two efforts from a macro-perspective, several observations emerge that may effect the use of air toxics data in further rulemaking and health risk determinations:

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

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

VI. GYPSUM QUALITY

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

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

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

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

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

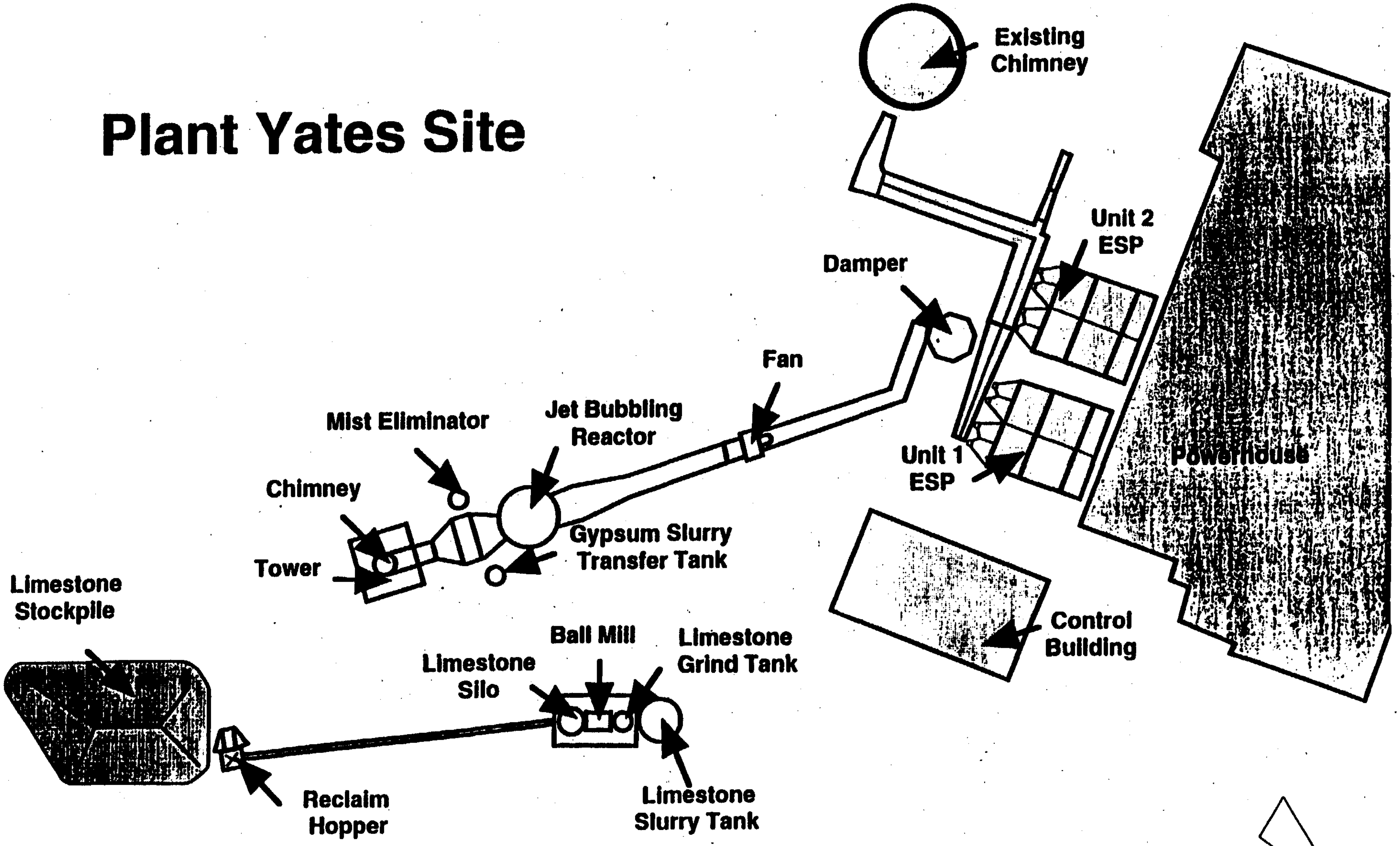
Table 4. Chloride and Moisture Levels in "Clean" Gypsum Stack decline over time

VIII. SUMMARY

Chiyoda's CT-121 FGD process was very successfully tested at conditions far beyond design expectations. From an operating standpoint, the process was reliable, showed consistently high removals (SO_2, particulate, air toxics), was energy efficient and reagent efficient. From a chemical engineering standpoint, the mass transfer interactions are robust and resilient, only limited at conditions far beyond design parameters. This would allow a designer / operator to install a cost effective CT-121 system that would give consistently excellent service, even in periods of difficult operating conditions.

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

Plant Yates Site



307199

FIGURE 1

Chiyoda's Jet Bubbling Reactor

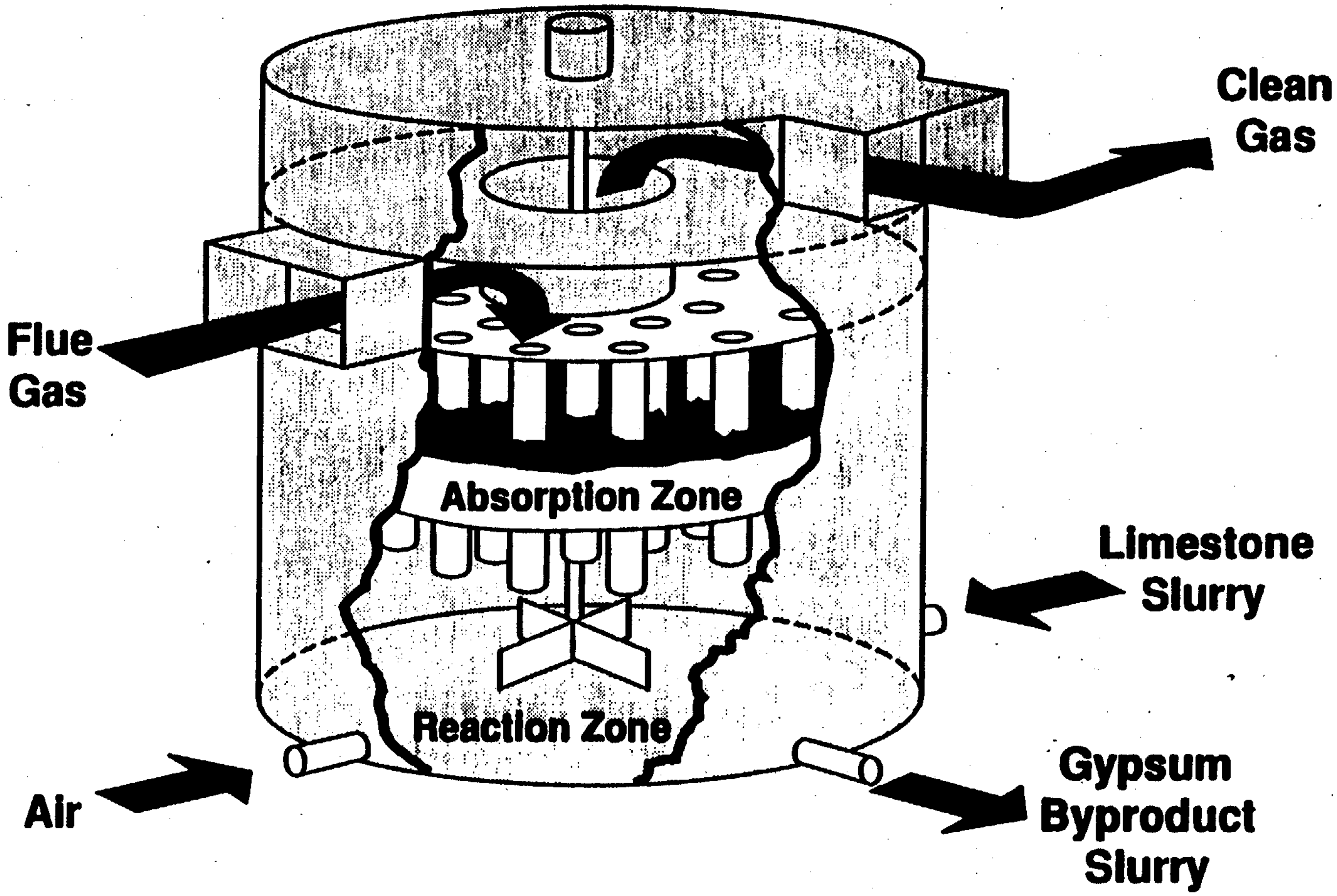


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FIGURE 2

CT-121 Gas Sparger Action

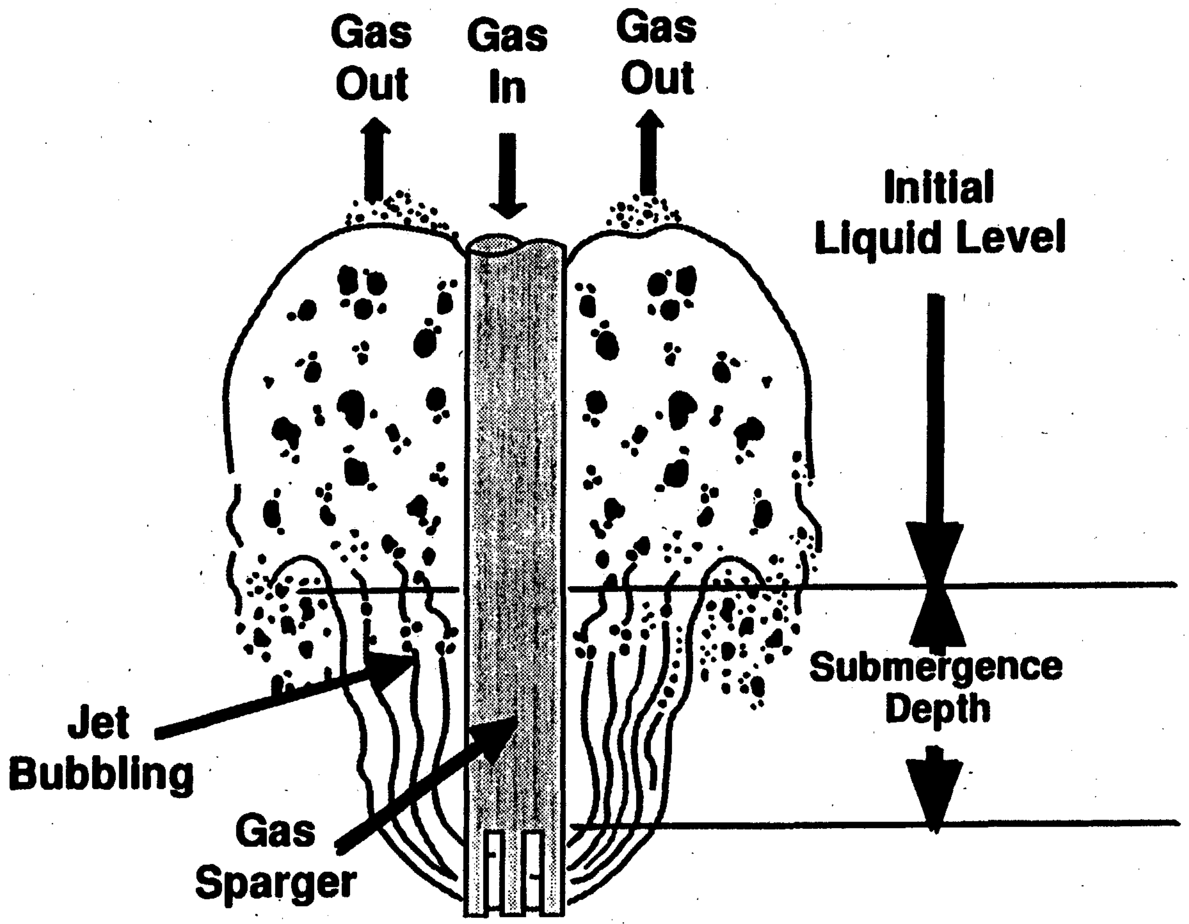
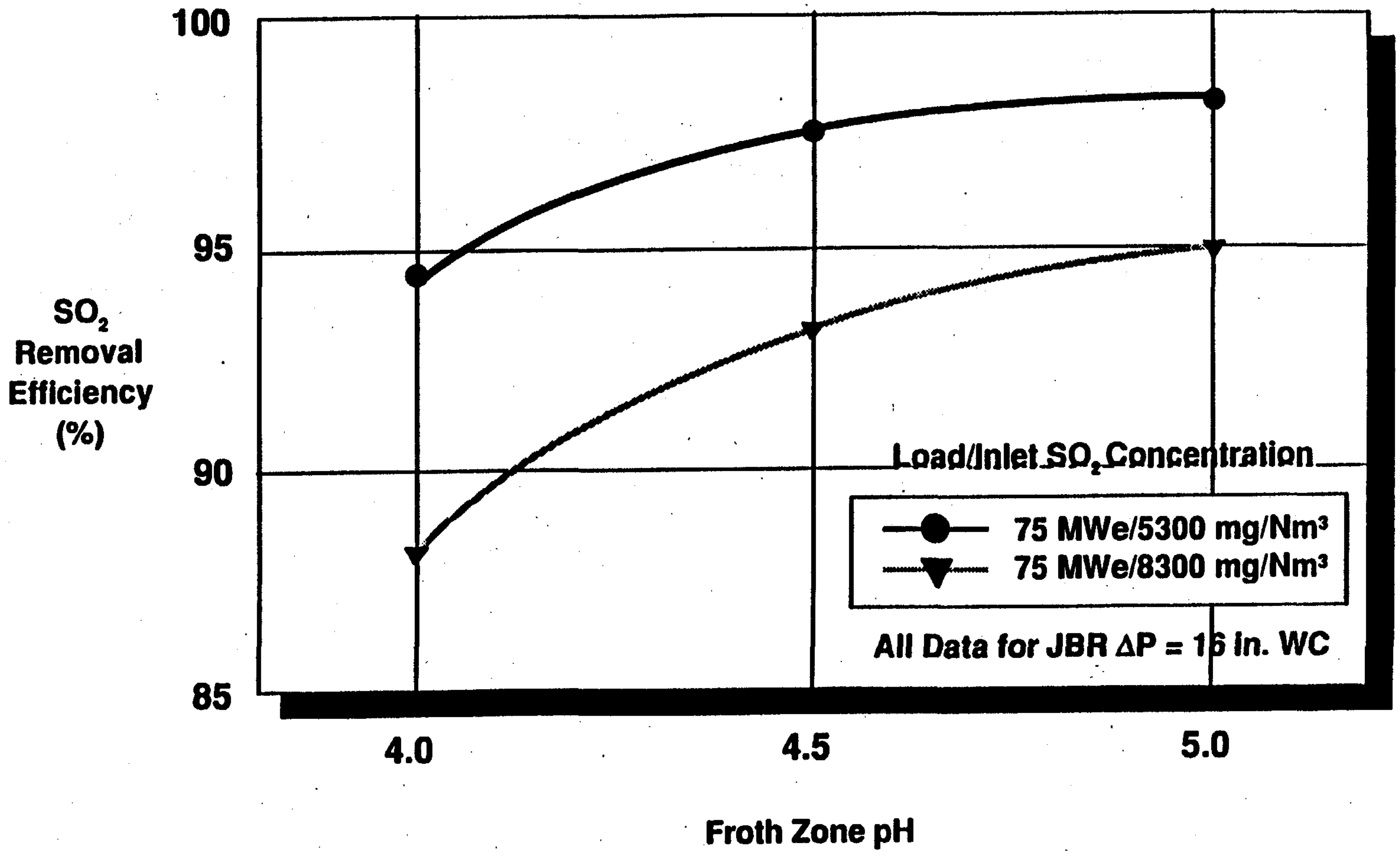


FIGURE 3

High-Sulfur Coal Effects on SO_2 Removal Efficiency



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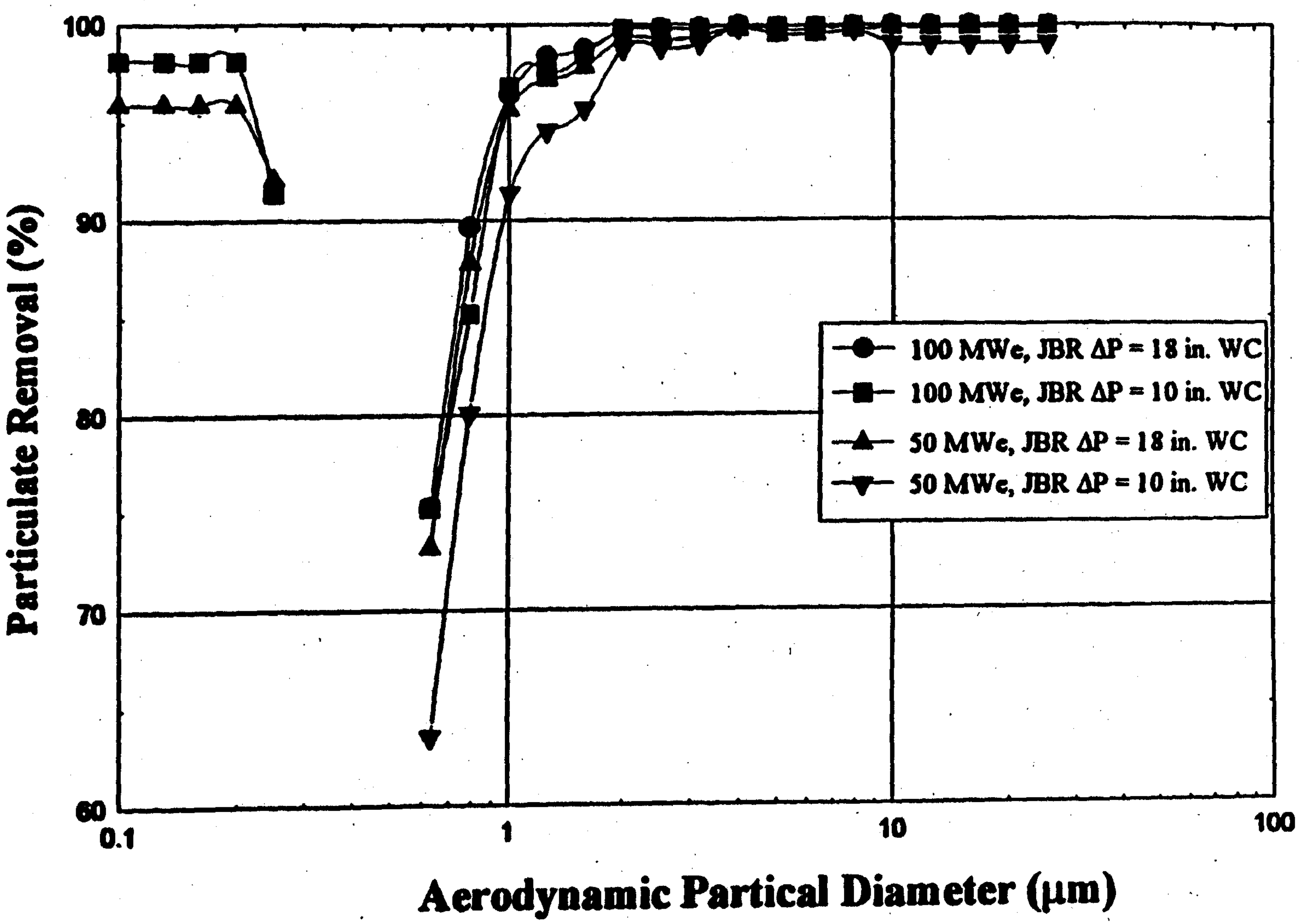
Chiyoda Corporation

Chiyoda CT-121

FIGURE 4



Moderate-Ash Scrubber Particulate Removal Efficiency

FIGURE 5

Air Toxics Removal

Yates CT-121 Project (JBR Components Only)

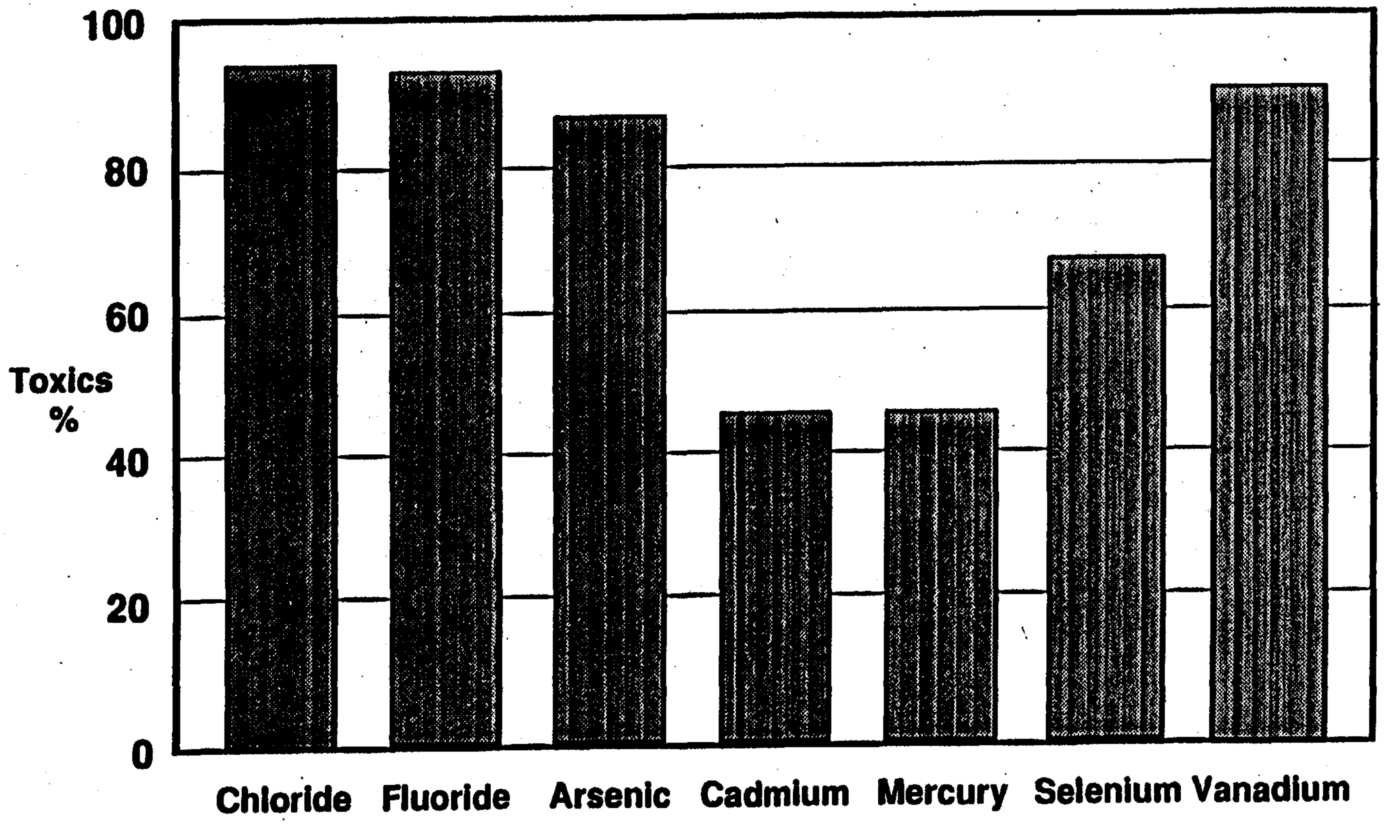


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FIGURE 6

THE HEALY CLEAN COAL PROJECT
AN OVERVIEW

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ABSTRACT

The Healy Clean Coal Project, selected by the U.S. Department of Energy under Round III of the Clean Coal Technology Program is currently in construction. The project is owned and financed by the Alaska Industrial Development and Export Authority (AIDEA), and is cofunded by the U.S. Department of Energy. Construction is scheduled to be completed in August of 1997, with startup activity concluding in December of 1997. Demonstration, testing and reporting of the results will take place in 1998, followed by commercial operation of the facility. The emission levels of NOx, SO_2 and particulates from this 50 megawatt plant are expected to be significantly lower than current standards. The project status, its participants, a description of the technology to be demonstrated, and the operational and performance goals of this project are presented herein.

BACKGROUND

In September 1988, Congress provided $575 million to conduct cost-shared Clean Coal Technology (CCT) projects to demonstrate technologies that are capable of retrofitting or repowering existing facilities. To that end, a Program Opportunity Notice (PON) was issued by the Department of Energy (DOE) in May 1989, soliciting proposals to demonstrate innovative energy efficient technologies that were capable of being commercialized in the 1990's, and were capable of (1) achieving significant reductions in the emissions of sulfur dioxide and/or the oxides of nitrogen from existing facilities to minimize environmental impacts such as transboundary and interstate pollution and/or (2) providing for future energy needs in an environmentally acceptable manner.

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

One of the 13 projects selected for funding is the Healy Clean Coal Project proposed by the Alaska Industrial Development and Export Authority (AIDEA). The project will demonstrate the combined removal of SO2, NOX, and particulates from a new 50 megawatt electric coal-fired power plant using both innovative combustion and flue gas cleanup technologies. AIDEA will own the Project, perform as DOE grant recipient, administer state funds, obtain financing through sale of bonds, and manage the Project. The architect/engineer for the project is Stone & Webster Engineering Corporation.. Fairbanks utility Golden Valley Electric Association (GVEA) will operate the facility and pay for power generated under terms of a power sales agreement.

TECHNOLOGY TO BE DEMONSTRATED

Coal provided by the Usibelli Coal Mine, adjacent to the project site, will be pulverized and burned at the new facility to generate high-pressure steam. The high-pressure steam will be supplied to a steam turbine generator to produce electricity. Emissions of SO2 and NOX from the plant will be controlled using TRW's Entrained Combustor with limestone injection in conjunction with a boiler designed by Foster Wheeler. Further SO2 and particulate removal will be accomplished using the Activated Recycle Spray Dryer Absorber System and Bag Filter developed by Joy Environmental Equipment, Inc.

The TRW Entrained Combustor is designed to operate under fuel-rich conditions, utilizing two staged combustion to minimize NOx formation. These conditions are obtained using a precombustor for heating the fuel-rich main combustor for partial combustion with combustion completion occurring in the boiler. The first and second stages of combustion produce a temperature high enough to generate a slag (liquid ash) while reducing the fuel-bond nitrogen to molecular nitrogen ($N2$). The third and final stage of combustion in the boiler occurs at a combustion temperature maintained below the temperature that will cause thermal NOx formation.

The combustor is also used to reduce SO2 emissions by the injection of pulverized limestone into the hot gases as they leave the combustor and enter the furnace. This technique changes the limestone into lime (flash calcination), which reacts with the sulfur compounds in the exhaust gas to form calcium sulfate. SO_2 is removed with combustor and boiler bottom ash. The flue gas, which contains the remaining sulfur compounds, calcium sulfate, and other solid particles leaves the boiler and passes through a spray dryer absorber and a bag filter for further SO2 and particulate removal prior to exiting through the stack.

The innovative concept to be demonstrated in SO_2 removal is the reuse of the unreacted lime, which contains minimal fly ash, in the second-stage spray dryer SO2 removal. The majority of fuel ash is removed in the combustor in the form of slag. A portion of the ash collected from the spray dry absorber vessel and the bag filter are first slurried with water, chemically and physically activated, and then atomized in the spray dryer absorber vessel for second-stage SO2 removal. Third stage SO2 and particulate removal occurs in the bag filter as the flue gas passes through the reactive filter cake in the bags.

The use of limestone in the combustor, combined with the recycle system, replaces the more expensive lime required by commercial spray dryer absorbers, reduces plant wastes, and increases SO2 removal efficiency when burning high- and low-sulfur coals.

The integrated process is expected to achieve SO2 removal greater than 90%, a reduction in NOx emissions to 0.2 pounds per million Btu. The integrated process is suited for new facilities or for repowering or retrofitting existing facilities. It provides an alternative technology to conventional pulverized coal-fired boiler flue gas desulfurization (FGD) and NOx reduction processes, while lowering overall operating costs and reducing the quantity of solid wastes.

The demonstration project is under construction adjacent to the Golden Valley

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

PROJECT STATUS

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

Coal Reburning for Cost-Effective NO_x Compliance

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ABSTRACT

This paper presents the application of micronized coal reburning to a cyclone-fired boiler in order to meet RACT emissions requirements in New York State. Discussed in the paper are reburning technology, the use of a coal micronizer, and the application of the technology to an Eastman Kodak unit. The program is designed to demonstrate the economical reduction of NO_x emissions without adverse impact to the boiler.

I. INTRODUCTION

The Eastman Kodak Company's Kodak Park Site is one of the largest industrial parks in the nation, spanning an area in excess of 1300 acres. There are over two hundred buildings on the site that produce thousands of different photographic and chemical products. Supporting production are two power plants containing a total of fourteen boilers. Kodak has an agreement with the New York State Department of Environmental Conservation (NYSDEC) in which it states that Kodak will install coal or natural gas reburning systems on all four of its cyclone boilers. Kodak has recently completed installation of a natural gas reburning system on #43 Boiler which is located on the western side of the Kodak Park Site facility. The upgrades of the three remaining boilers (#15, #41, and #42) are planned for the 1996 thru 1998 time frame. #15 Boiler is located apart from the other three cyclone boilers, on the eastern section of the facility, approximately three miles from #43 Boiler. The original schedule for upgrades was #43 by 1996, #41 and #42 by 1997, and #15 by 1998.

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

Eastman Kodak #15 Boiler

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

In February 1996, EER performed a baseline test and measured NO_x emissions at 1.21 lb/10^6 Btu for full load and 0.92 lb/10^6 Btu for low load. Baseline CO emissions were 56 ppm and 34 ppm at full and low loads respectively. The results correlated closely with Kodak's belief that the baseline NO_x emissions are 1.25 lb/10^6 Btu and baseline CO is less than 100 ppm.

Coal Reburning Technology for NO_x Control

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

The technology involves reducing the levels of coal and combustion air in the burner area and injecting reburn fuel (micronized coal) above the burners followed by the injection of overfire air (OFA) above the reburn zone. This three-zone process creates a reducing area in the boiler furnace within which NO_x created in the primary zone is reduced to elemental nitrogen and other less harmful nitrogen species. Each zone has a unique stoichiometric ratio (ratio of total air in the zone to that theoretically required for complete combustion) as determined by the flows of coal, burner air, reburn fuel, and OFA. The descriptions of the zones are as follows:

- *Primary (burner) Zone:* Coal is fired at a rate corresponding to 75 to 90 percent of the total heat input. NO_x created in this zone is slightly lower than normal operation due to the lower heat release and the reduced excess air level.

- *Reburn Zone:* Reburn fuel (micronized coal) is injected above the main burners through wall ports. The reburn fuel consumes the available oxygen and produces hydrocarbon fragments (CH, CH_2, etc.) which react with NO_X from the lower furnace, reducing it to elemental nitrogen, N_2. Optimum NO_X reduction performance is typically achieved when the reburn zone is operated at about 90% of stoichiometric, which is slightly fuel rich (reducing) [2]. NO_X reduction can be adjusted by varying the reburn fuel injection rate, typically over the range of 10-25% of total boiler heat input. To minimize the reburn fuel required to achieve fuel rich conditions in the reburn zone, EER's design utilizes injectors rather than burners, which would have introduced additional air [3].

- *Burnout (exit) Zone:* The oxygen required to burn out the combustibles from the reburn zone is provided by injecting air through overfire air ports positioned above the reburn zone. These ports are similar to conventional overfire air ports except that they are positioned higher in the furnace so as to maximize the residence time for NO_X reduction occurring in the reburn zone. OFA is typically 20 percent of the total air flow. OFA flow rate and injection parameters are optimized to minimize CO emissions and unburned carbon-in-fly ash.

The concept of NO_X reduction via reactions with hydrocarbon fuels has been recognized for some time [4]. The work has progressed from analysis and pilot-scale tests [2] through several full-scale demonstrations including three installations on coal-fired utility boilers as part of the U.S. Department of Energy's Clean Coal Technology Program [5] and a commercial installation at New York State Electric and Gas' Greenidge Plant [6].

Goals of Micronized Coal Reburning Demonstration

The objective of the coal reburning demonstration is to evaluate the applicability of the technology to full-scale cyclone-fired boilers for reduction of NO_x emissions. The project goals are:

- Reduce NO_x emissions at full load from the current established baseline of 1.25 $lb/10^6$ Btu to 0.60 $lb/10^6$ Btu.

- Maintain CO emissions at or below 100 ppm.

- Minimize the impact on boiler efficiency.

- Reduce NO_x without serious impact to cyclone operation, boiler performance or other emissions streams.
- Demonstrate a technically and economically feasible retrofit technology.
- Demonstrate the advantages of micronized coal reburning over conventional coal reburning.

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

II. PROCESS DESIGN

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

Controlling Process Parameters

Since the early 1980's, EER has extensively evaluated the reburning process at bench, pilot and full-scale to identify the parameters that control process performance. The results of these studies have shown that the most critical parameters are: primary NO_x level; reburn zone temperature and residence time; reburn zone stoichiometric ratio; and mixing of the reburn fuel and overfire air with the bulk furnace gases.

Reburn Zone Stoichiometric Ratio: The impact of this parameter on the NO_x emissions achievable with various reburn fuels is shown in Figure 3 [7]. as shown in the figure, overall NO_x reductions are highest when the ratio is approximately 0.9. To minimize the amount of reburn fuel needed to reach the optimum ratio, the primary combustion zone is operated as close to stoichiometric as possible. It should be noted, however, that with cyclone-fired boilers reducing the stoichiometric ratio in the primary zone will disrupt the slagging characteristics of the cyclone. Therefore, the fuel-to-air ratio in this area remains relatively unchanged.

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

- High temperatures in the reburn zone are preferred in order to maximize the rate of NO_x reduction. This suggests that the reburn fuel be injected just downstream of the primary zone.

- The temperature in the burnout zone must be high enough to allow oxidation of carbon monoxide and hydrocarbon fragments from the reburn zone to occur readily.

- The residence time must be of sufficient duration for the reactions to occur. EER has evaluated a number of reburning systems and concluded that a residence time of 0.2 to 0.5 seconds will achieve high efficiency NO_x reduction.

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

Design Approach

The final design was established on the basis of small-scale flow modeling, thermal heat transfer computer analysis, and operation of a pilot-scale micronizer using EER's Boiler Simulator Furnace. The reburn fuel and overfire air injection elevations were selected to provide the maximum amount of residence time possible in the reburn zone in order to maximize the NO_x control performance. This approach involved injection of the reburn fuel at an elevation in the furnace just above the exit of the cyclones and injection of overfire air at a distance downstream of the coal injectors that would provide for a maximum bulk residence time (Figure 4).

The reburn fuel is pneumatically transported to the boiler using recycled flue gas (FGR) as the carrier medium. The fuel is then introduced into the boiler thru injectors that are designed to rapidly mix the small quantity reburn fuel with the furnace gases. FGR is particularly suited as a carrier gas in lieu of air since it consists of a very low level of O_2. Note that any O_2 introduced as carrier gas must be consumed by additional reburn fuel. The use of FGR minimizes this fuel requirement.

III. SYSTEM DESCRIPTION

Coal Micronizer

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

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

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

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

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

The two-mill system for the Kodak project includes:

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

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

Coal Transportation and Injection

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

Two coal micronizers with classifiers are used in the system. Each micronizer is supplied coal from a bunker thru a screw feeder. The FGR system assists in the micronizing process and in operation of the classifiers. The mills are capable of operating singly or as a pair, although, due to capacity limitations, both may be required to produce the targeted NO_x reduction.

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

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

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

Overfire Air System

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

Controls

Kodak installed a new Coen burner management system and replaced the complete boiler control system with a Westinghouse WDPF distributed digital control system. The new controls operate both the existing equipment and the micronized coal reburning system, with all normal start/stop/modulate operator actions occurring in the control room. Critical operations are interlocked to prevent inadvertent operation of equipment when such operation may present an operating hazard or other undesirable condition. The controls are designed to shut down the reburning system while maintaining operation of the boiler. Kodak's insurance carrier, Factory Mutual, has approved this control arrangement. Previous to this project, EER reburning retrofits were approved by Factory Mutual and Hartford Steam.

Operation

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

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

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

IV. SUMMARY

The coal reburning installation at Eastman Kodak Company will permit Kodak to meet RACT emissions requirements in New York State. The project, conducted under the auspices of the U.S. Department of Energy's Clean Coal Technology Program, is designed to demonstrate the economic advantages of using coal micronizer technology versus conventional coal reburning. Testing of the system will verify the target goals of NO_x emissions reduction and determine the full range of operation, including turndown capabilities. The testing will also be used to develop a database of technical information that can be applied to similar boilers.

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

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

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

V. ACKNOWLEDGMENTS

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

VI. REFERENCES

1. J. O. L. Wendt et. al., Reduction of Sulfur Trioxide and Nitrogen Oxides by Secondary Fuel Injection, Fourteenth Symposium (International) on Combustion, pp. 897-904, (1973).

2. S. B. Greene et. al., Bench Scale Process Evaluation of Reburning and Sorbent Injection for In-Furnace NO_x Reduction, Paper No. 84-JPGC-APC-9, presented at the ASME Joint Power Generation Conference (1984).

3. W. R. Seeker, et. al., Controlling Pollutant Emission from Coal and Oil Combustor Through the Supplemental Use of Natural Gas, Final Rpt., GRI 5083-251-0905, 1985.

4. U.K. Patent Office. *Process for Disposal of Oxides of Nitrogen.* Robert D. Reed, John Zink Company. U.K. Patent No. 1,274.637, 1970.

5. B. A. Folsom, et. al., Three Gas Reburning Field Evaluations: Final Results and Long Term Performance, presented at the EPRI/EPA 1995 Joint Symposium on Stationary Combustion NO_X Control, Kansas City, Missouri (May 16-19, 1995).

6. B. A. Folsom, et. al., Advanced Gas Reburning Demonstration and Commercial Gas Reburning System Upgrade, Clean Air '96, Orlando, Florida, November 19-22, 1996.

7. D. K. Moyeda, et. al., Experimental/Modeling Studies of the Use of Coal-Based Reburning Fuels for NO_x Control, Twelfth International Pittsburgh Coal Conference, Pittsburgh, Pennsylvania, September 11-15, 1995.

8. K.T. Wu, et. al., Development and Application of a Gas Reburning Process Model for the Design of Boiler NO_x Reductions, Joint Power Generation Conference, San Diego, California, October 6-10, 1991.

9. B. A. Folsom, et. al., Demonstration of "Gas Reburning-Low NO_x Burner Technology for Cost Effective NO_x Emission Control, Fourth Annual Clean Coal Conference, Denver, Colorado, September 5-8, 1995.

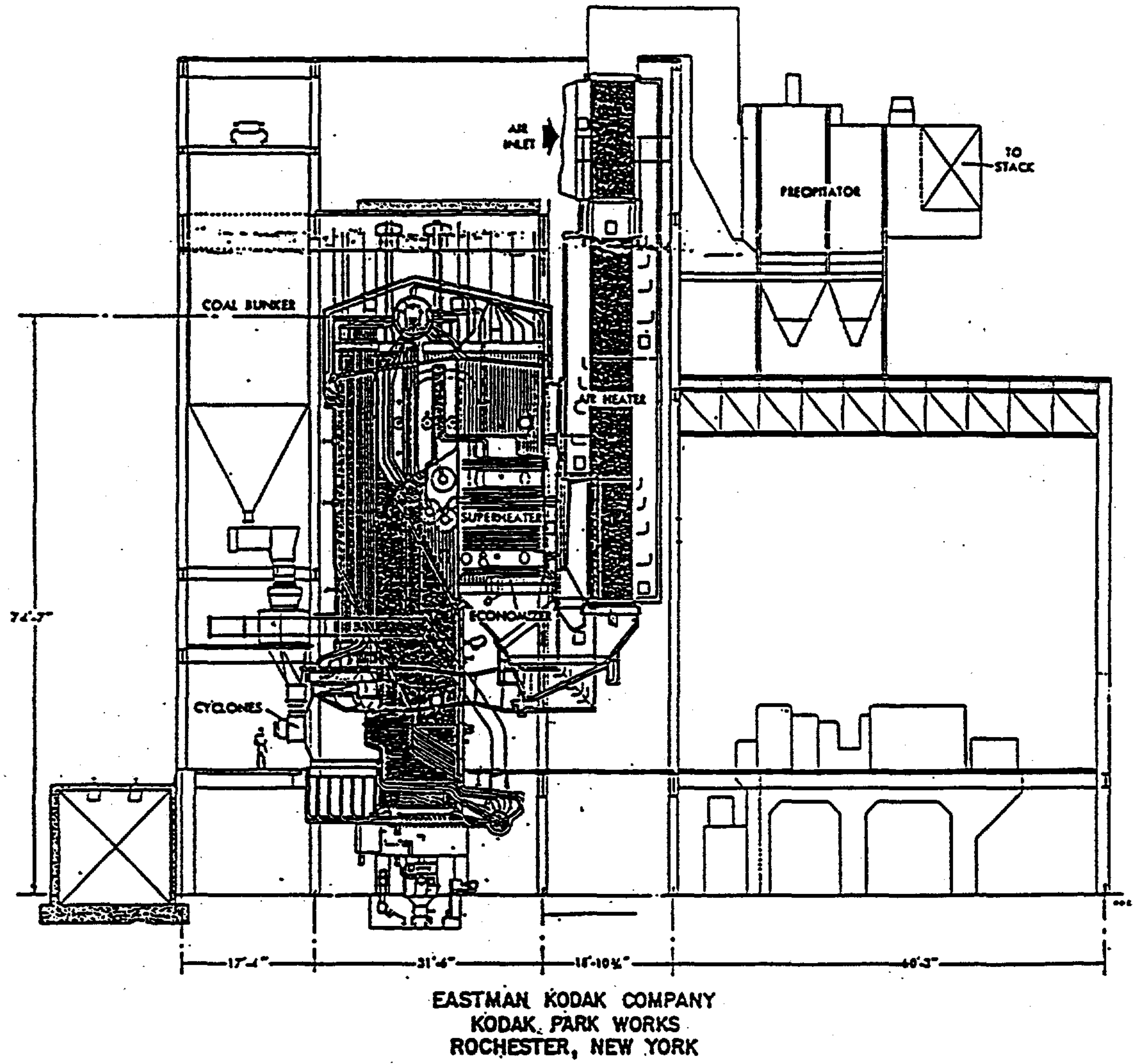

Figure 1. Kodak #15 Boiler.

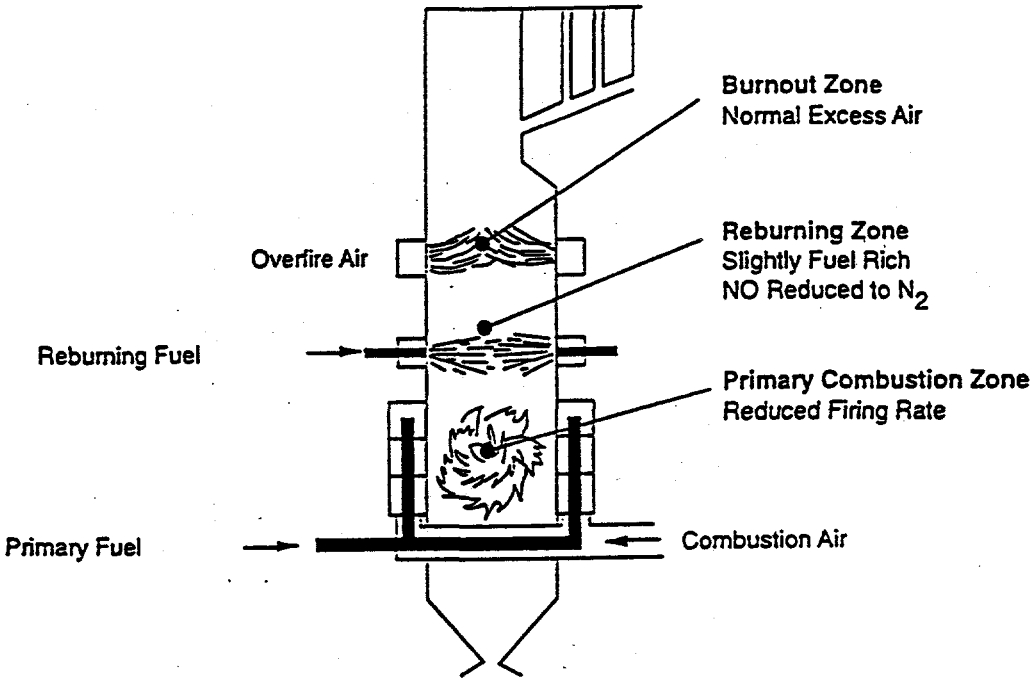

Figure 2. Application of reburning technology to a utility boiler.

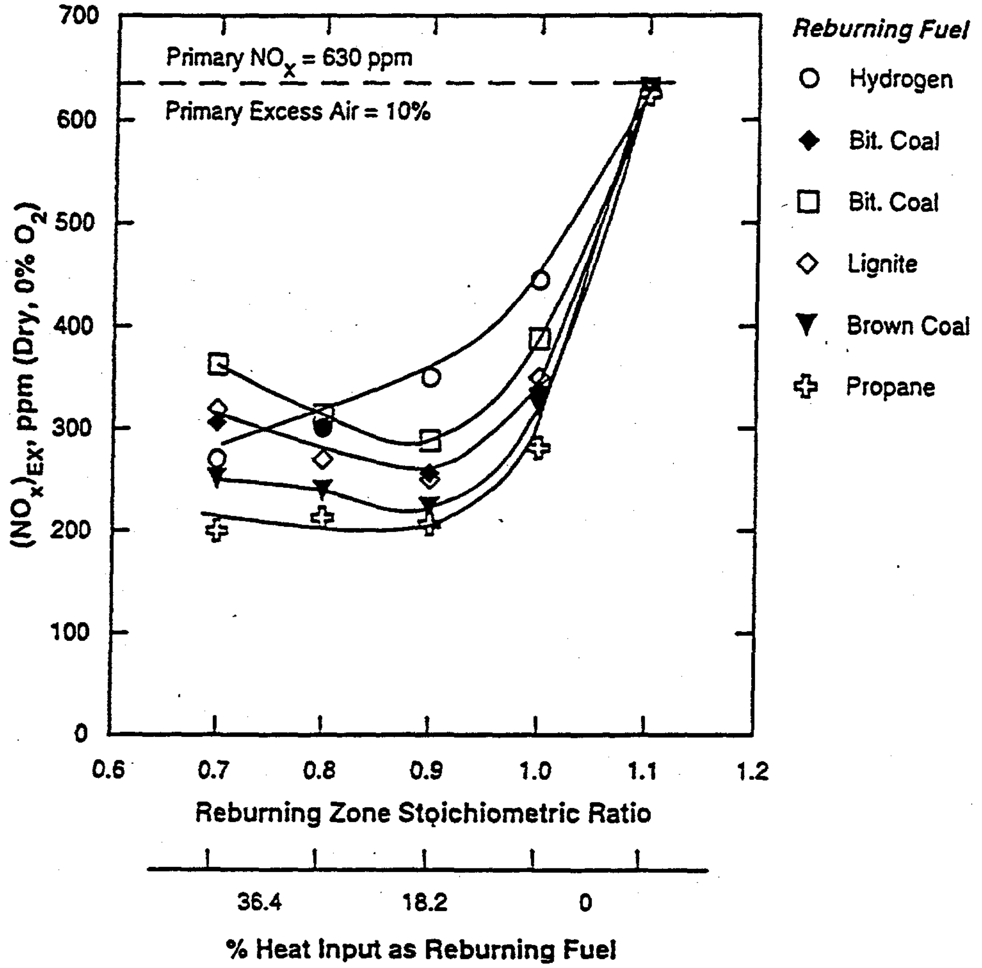

Figure 3. Impact of reburning zone stoichiometric ratio and reburning fuel type on reburning performance.

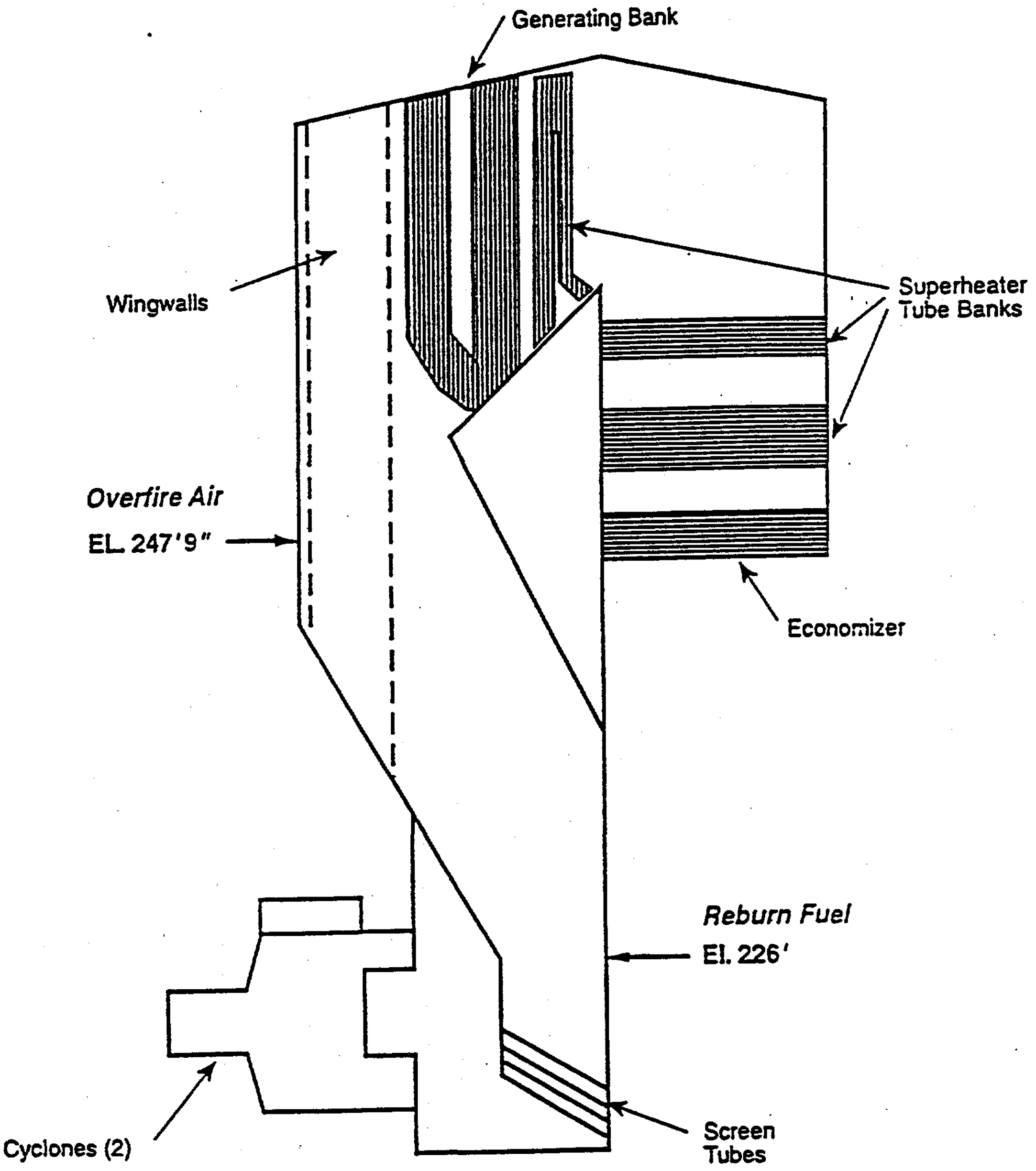

Figure 4. Reburn fuel and overfire air injection elevations for Kodak #15 Boiler.

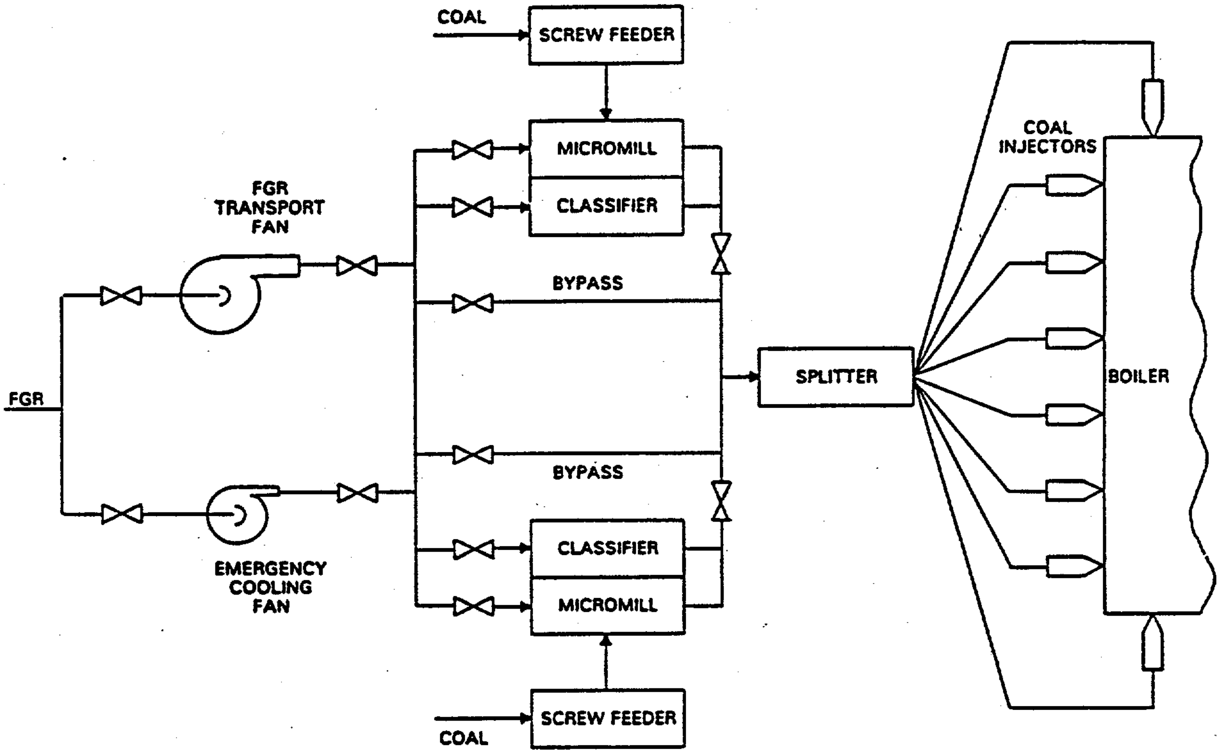

Figure 5. Micronized Coal Feed System

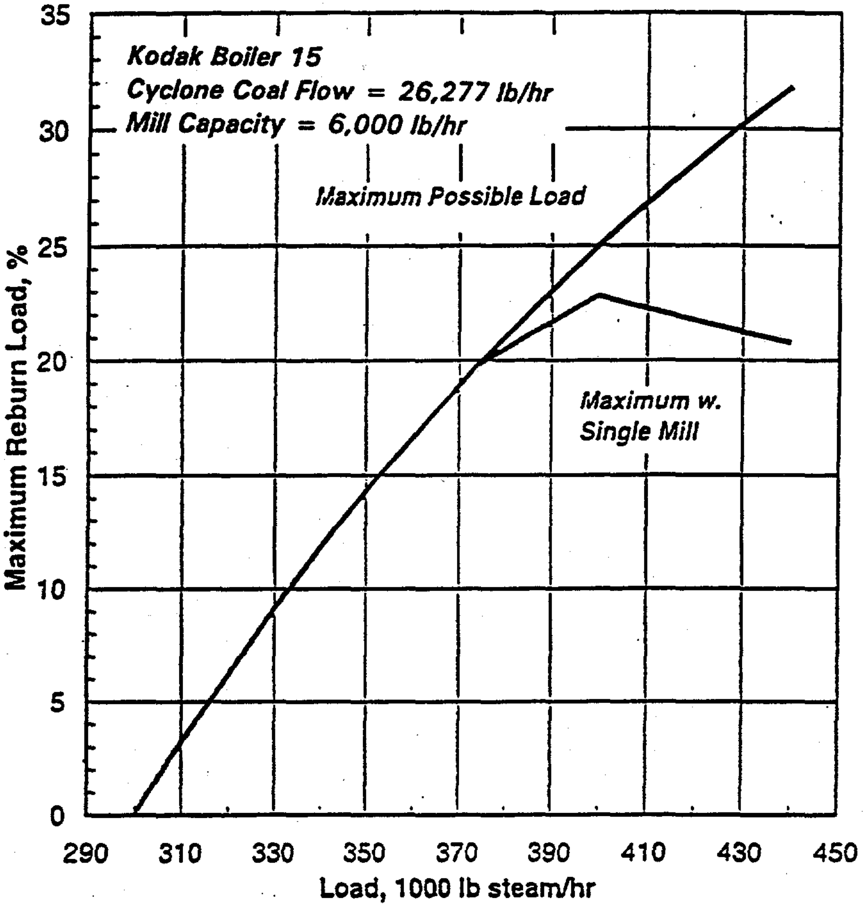

Figure 6. Maximum level of reburning system operation achievable while maintaining minimum coal flow to cyclones.

THE NOXSO CLEAN COAL PROJECT

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ABSTRACT

The NOXSO Clean Coal Project will consist of designing, constructing, and operating a commercial-scale flue-gas cleanup system utilizing the NOXSO Process. The process is a waste-free, dry, post-combustion flue-gas treatment technology which uses a regenerable sorbent to simultaneously adsorb sulfur dioxide (SO_2) and nitrogen oxides (NO_x) from flue gas from coal-fired boilers. The NOXSO plant will be constructed at Alcoa Generating Corporation's (AGC) Warrick Power Plant near Evansville, Indiana and will treat all the flue gas from the 150-MW Unit 2 boiler. The NOXSO plant is being designed to remove 98% of the SO_2 and 75% of the NO_x when the boiler is fired with 3.4 weight percent sulfur, southern-Indiana coal. The NOXSO plant by-product will be elemental sulfur.

The elemental sulfur will be shipped to Olin Corporation's Charleston, Tennessee facility for additional processing. As part of the project, a liquid SO_2 plant has been constructed at this facility to convert the sulfur into liquid SO_2. The project utilizes a unique burn-in-oxygen process in which the elemental sulfur is oxidized to SO_2 in a stream of compressed oxygen. The SO_2 vapor will then be cooled and condensed. The burn-in-oxygen process is simpler and more environmentally friendly than conventional technologies. The liquid SO_2 plant produces 99.99% pure SO_2 for use at Olin's facilities.

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

INTRODUCTION

The NOXSO Process is a waste-free, dry, post-combustion flue-gas cleanup technology which uses a regenerable sorbent to simultaneously adsorb SO_2 and NO_x from flue gas from coal-fired utility and industrial boilers. In the process, the SO_2 is converted to a saleable sulfur by-product (liquid SO_2, elemental sulfur, or sulfuric acid) and the NO_x is converted to nitrogen and oxygen. Since SO_2 and NO_x removal occur at normal flue-gas temperatures (downstream of the combustion air preheater), the NOXSO Process is equally suited for retrofit as well as new installations.

Process development began in 1979 with laboratory-scale tests and progressed to pre-pilot-scale tests (3/4-MW) and a life-cycle test. Each of these test programs [1,2,3] has provided data necessary for the process design. Tests of the NO_x recycle concept, which is inherent to the NOXSO Process, have been conducted on small boilers at PETC and at the Babcock & Wilcox (B&W) Research Center in Alliance, Ohio [4].

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

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

Final processing of the sulfur by-product to make liquid SO_2 will be completed at Olin Corporation's Charleston, Tennessee facility. The SO_2 plant which utilizes a unique burn-in-oxygen process for converting sulfur to liquid SO_2 is complete. The burn-in-oxygen process is simpler and more environmentally friendly than conventional technologies. Midwest Technical, Inc. provided engineering services. Projex, Inc. managed construction of the facility.

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

Meanwhile, mechanical completion, testing and start-up of the liquid SO_2 plant was achieved in December 1997. Feedstock sulfur will be purchased on the market until the start-up of the NOXSO plant, at which time the NOXSO plant will be the sole source of feedstock for the liquid SO_2 plant. Operating and environmental data will be collected during the plant's operation.

Funding for the $82.8 million project will be provided by the DOE, NOXSO, AGC, Warrick County, the Southern Indiana Gas and Electric Company (SIGECO), the Gas Research Institute (GRI), W.R. Grace, and the Electric Power Research Institute (EPRI). NOXSO will raise most of its project funds through the sale of revenue bonds issued and guaranteed by the state of Indiana. The guarantee is made possible by state legislation signed into law on March 28, 1995. NOXSO will repay the bonds from revenue generated by the sale of SO_2 allowances and by the sale of liquid SO_2 to Olin during a thirteen-year time period which includes the two-year demonstration operation period.

II. THE NOXSO COMMERCIAL DEMONSTRATION PLANT

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

Host Site Information

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

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

High sulfur Squaw Creek coal with composition as shown in Table 1 will be burned in Unit 2 after the NOXSO plant is installed. Squaw Creek coal is currently blended with a low sulfur coal for use in Units 1, 2, and 3 to satisfy the Warrick County State Implementation Plan (SIP) limit of 5.11 pounds SO_2 per million Btu of heat input.

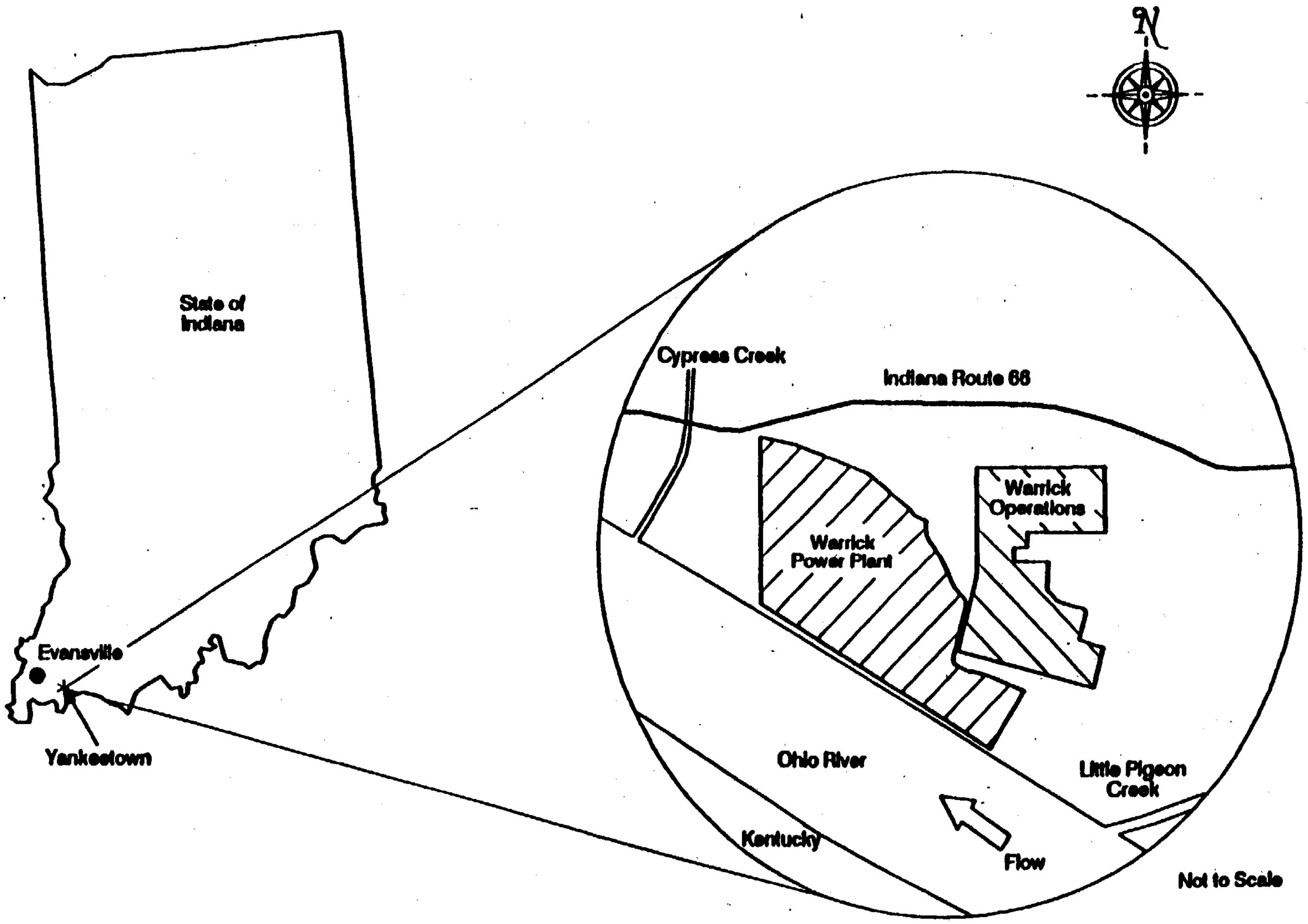

Figure 1. Warrick Power Plant and Warrick Operations Site Location

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

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

AGC has opted-in WPP Units 1, 2, and 3 to the Acid Rain Program of the Clean Air Act (CAA) Amendments of 1990. The Opt-In Program (40 CFR Part 72) allows nonaffected sources, such as AGC's WPP Units 1, 2, and 3, to enter the SO_2 portion of the acid rain program and receive SO_2 emission allowances.

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

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

Table 2. **Unit 2 Design Parameters**

NOXSO Process Description

The NOXSO Process is a dry, post-combustion flue-gas treatment technology which will use a regenerable sorbent to simultaneously adsorb SO_2 and NO_x from the flue gas from Unit 2 of AGC's WPP. In the process, the SO_2 will be converted to liquid SO_2 and the NO_x will be reduced to nitrogen and oxygen. The NOXSO plant is designed to remove 98% of the SO_2 and 75% of the NO_x. Details of the NOXSO Process are described with the aid of Figure 2.

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

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

In heating the sorbent, the NO_x is driven off and carried to the power plant boiler in the NO_x recycle stream. The NO_x recycle stream is cooled from approximately 360°F to 140°F in the feedwater heater. This heat-exchanger heats a slip stream of the power plant's feedwater, thereby reducing the amount of extraction steam taken from the low pressure turbine, enabling the generation of additional electricity. The cooled NO_x recycle stream replaces a portion of the combustion air. The presence of NO_x in the combustion air suppresses the formation of NO_x in the boiler resulting in a net destruction of NO_x.

The heated sorbent is transported through an L-valve to the steam disengaging vessel. Transport steam is separated from the sorbent to reduce the volume of the regenerator off-gas stream. Sorbent gravity flows into the regenerator where it is contacted with natural gas. Through a series of chemical reactions, the sulfur on the sorbent combines with the methane and forms SO_2 and H_2S. Additional regeneration occurs in the steam treater section of the regenerator when the

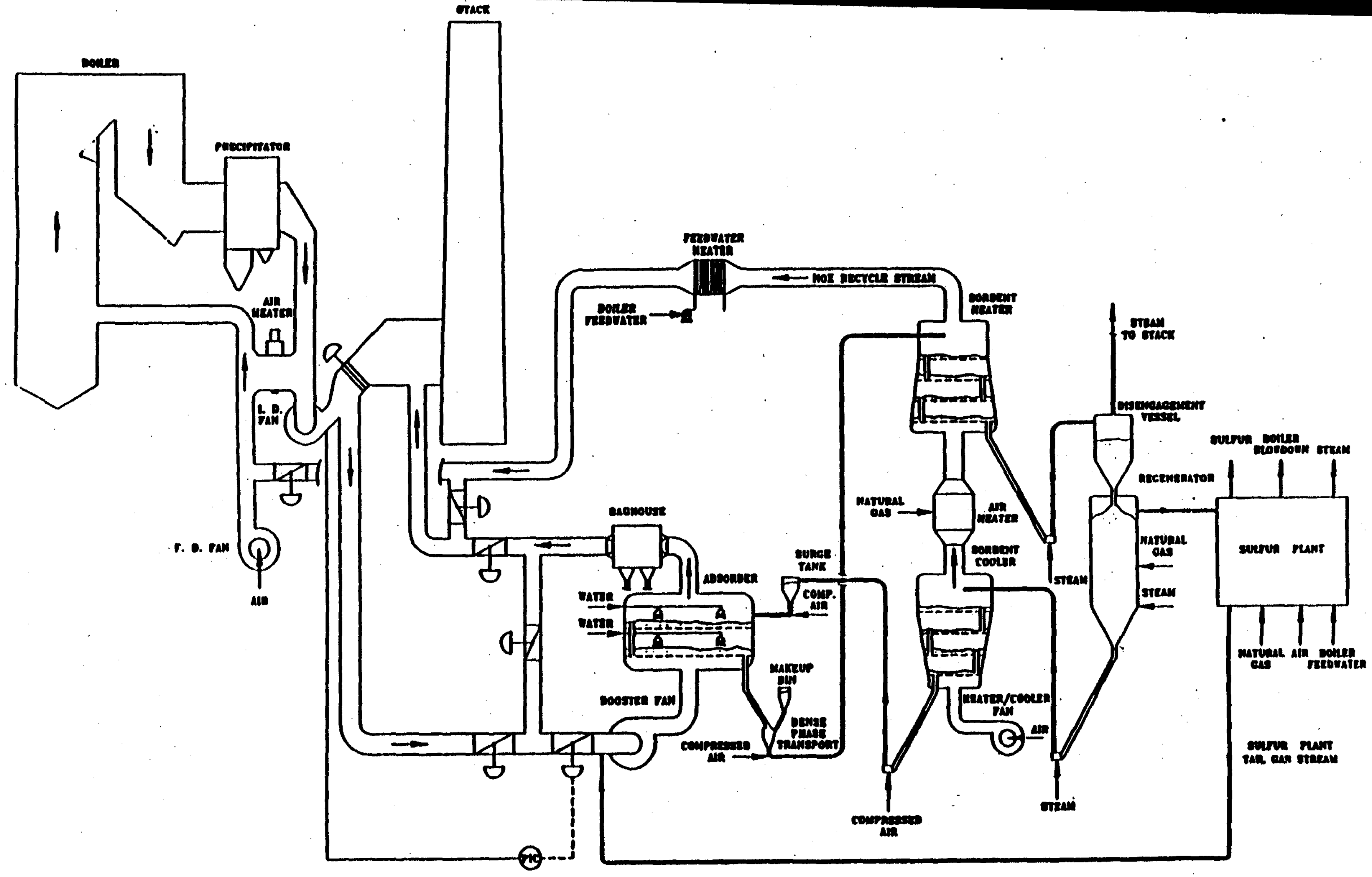

Figure 2. NOXSO Process Diagram - Alcoa Generating Corporation
Warrick Plant Unit 2

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

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

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

NOXSO Plant Description

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

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

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

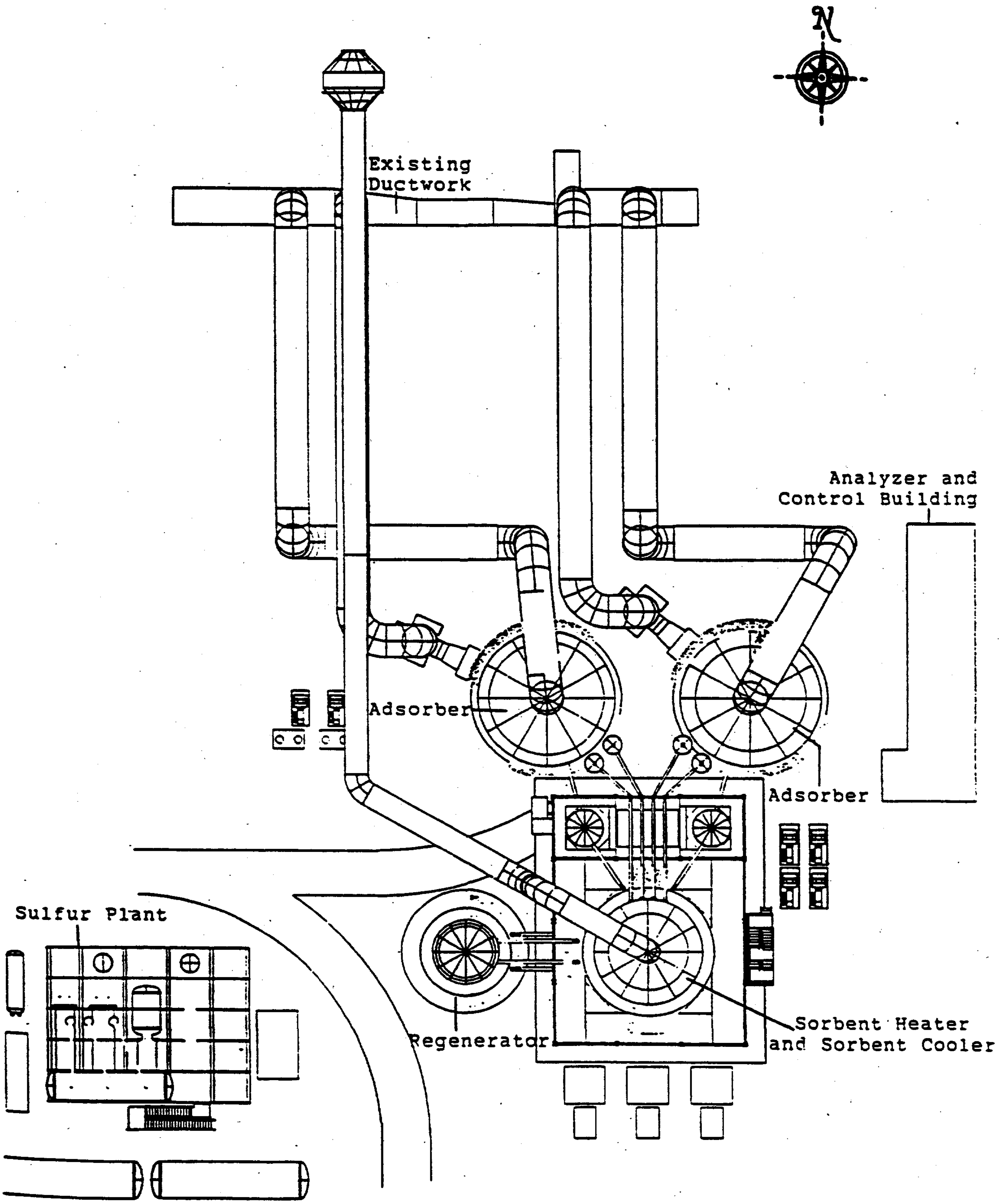

Figure 3. Plant Layout

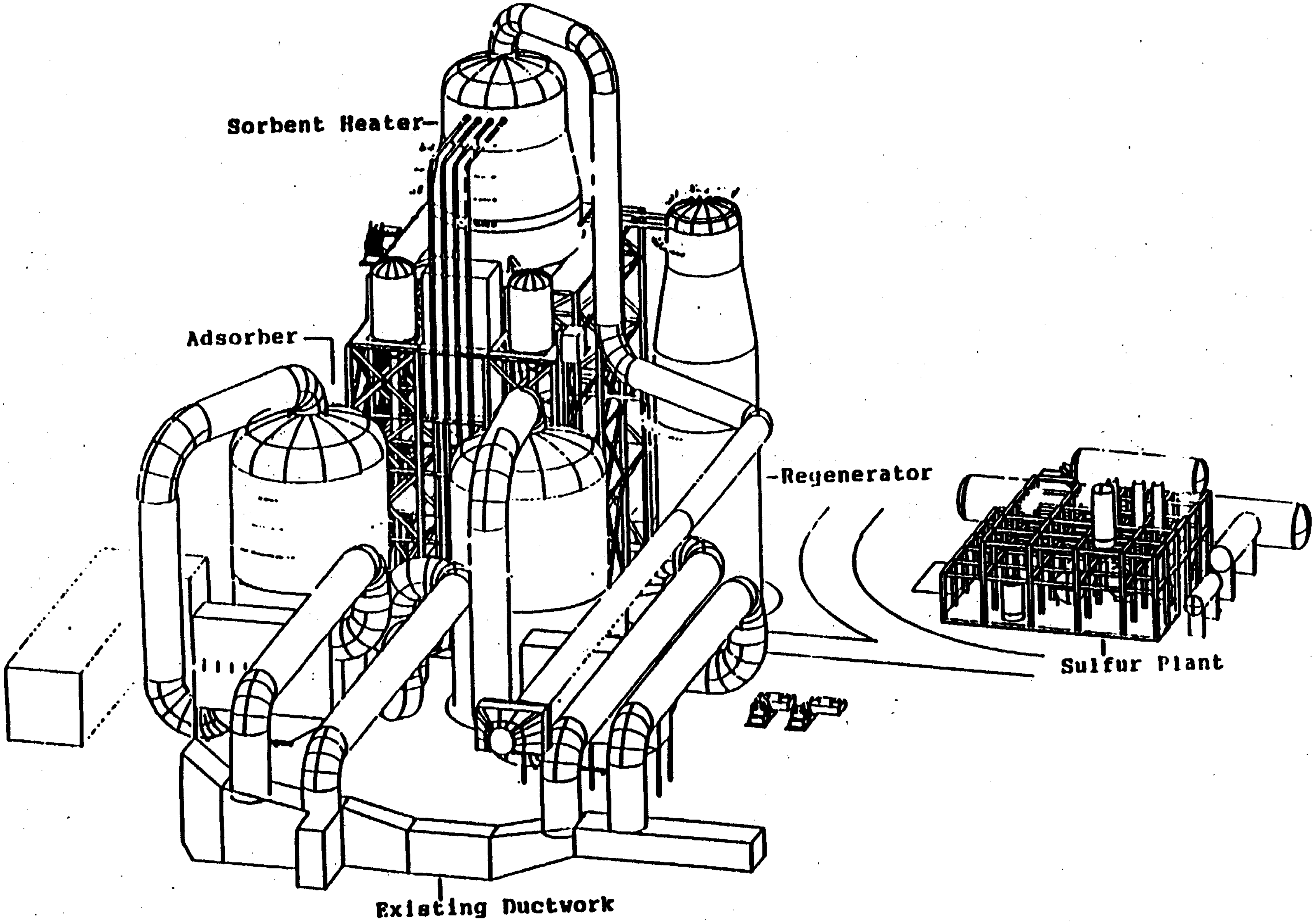

Figure 4. General Arrangement

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

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

III. THE LIQUID SO_2 FACILITY

As discussed previously, the purpose of the NOXSO Clean Coal Project is to demonstrate the NOXSO flue-gas treatment system in a fully integrated commercial scale operation. The NOXSO plant will reduce SO_2 and NO_x emissions from Alcoa Generating Corporation's Warrick Power Plant Unit 2. The removed sulfur will be processed into elemental liquid sulfur. In addition, as part of the project, a liquid SO_2 plant has been constructed at Olin Corporation's Charleston, Tennessee facility to convert the sulfur into liquid SO_2.

Host Site Information

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

As shown in Figure 6, OCP is located in Bradley County, in southeastern Tennessee about 12 miles northeast of Cleveland, Tennessee. Charleston, Tennessee, the closest town to the site, is 1.5 miles southeast of the plant. The OCP consists of roughly 975 acres between Lower River Road and the Hiwassee River (which flows to the northwest). Liquid SO_2 is a primary feedstock at the OCP where it is used to produce sodium hydrosulfite which is sold to the paper industry where it is used as a bleach for paper and clay.

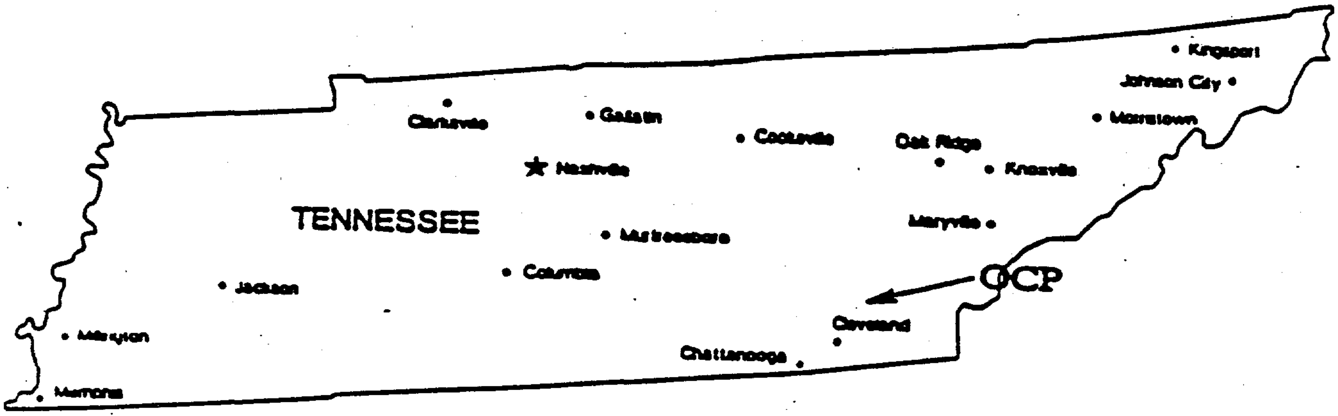

Figure 5. OCP Location

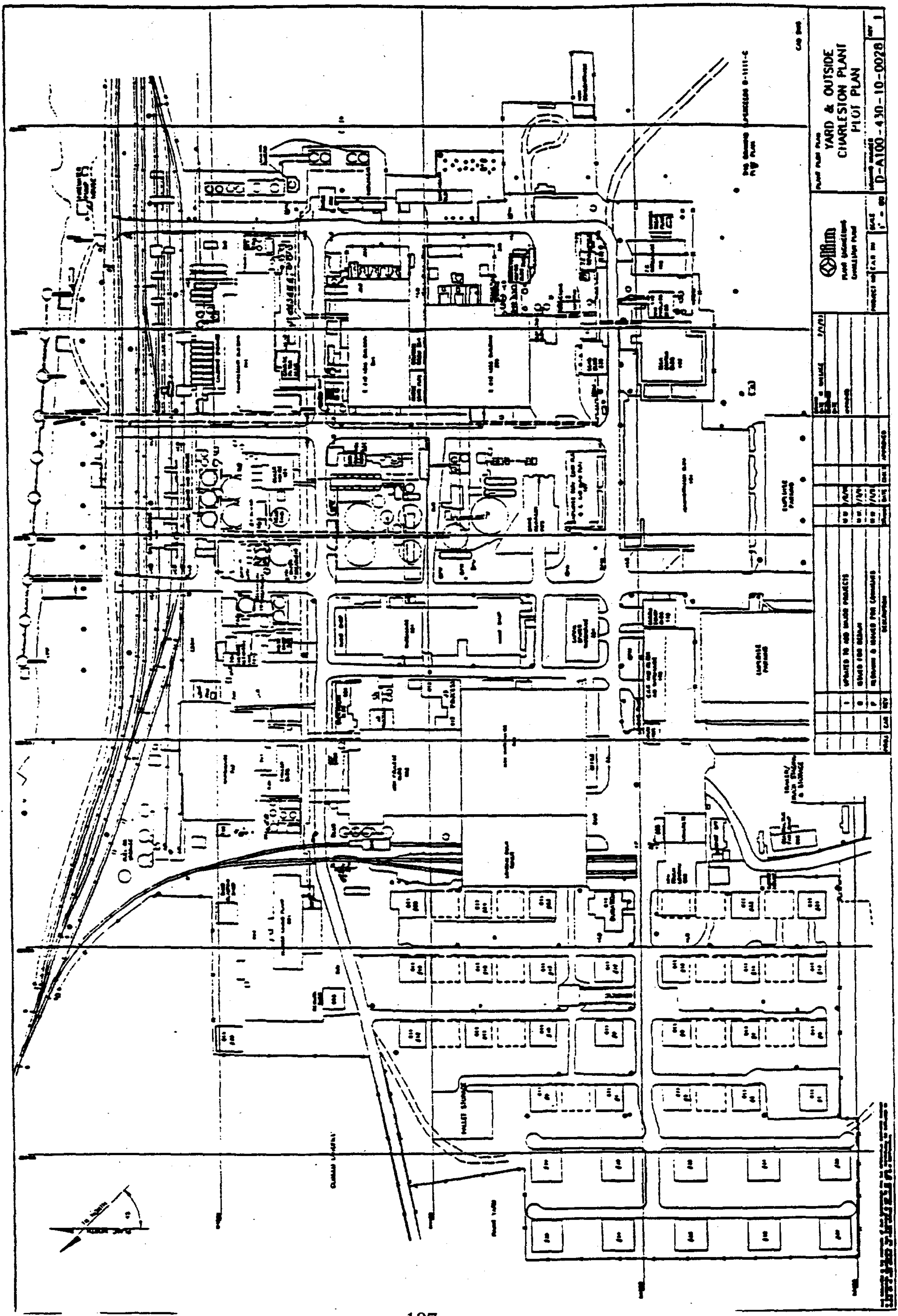

Figure 6. OCP Site Plan

Liquid SO_2 Process Description

The liquid SO_2 facility consists of two components, the liquid SO_2 plant and a cryogenic air separation plant. The facility is located on less than an acre of Olin property east of the existing switchgear building. Figure 7 presents the site plan for the liquid SO_2 facility detailing its relationship within Olin's plant site.

The SO_2 plant, the primary aspect of the liquid SO_2 facility, is an advanced liquid SO_2 production process designed for ease of operation and maintenance and to minimize process waste streams and emissions to the environment. Reliable operation of a 9,000 tpy commercial unit over the last five years has demonstrated and proven the technology. In the basic process, molten sulfur is oxidized to SO_2 vapor in compressed oxygen. The SO_2 vapor is then separated from vaporized sulfur and condensed. Key resources, including molten sulfur, oxygen (O_2), and caustic, are fed to the process. The process in turn produces liquid SO_2, steam, and sodium sulfite.

The cryogenic air separation plant provides 99.5% O_2 to the liquid SO_2 plant. The oxygen is produced by liquefying air and then using fractional distillation to separate it into its components. The air separation plant requires inputs of air, electricity, and cooling water and produces, in addition to the O_2, a small amount of pure nitrogen (N_2).

Liquid SO_2 Plant Description

The facility will have the operating capacity to produce about 125 tpd (45,000 tpy) of liquid SO_2. Figure 8 presents a basic flow diagram of the liquid SO_2 process. Primary unit operations are numerically labelled on this figure and referenced in the following discussion. Liquid sulfur at about 270°F is continuously pumped from two-250 ton capacity sulfur storage tanks (1) to the sulfur day tank (2). Sulfur flows by gravity from the day tank to the SO_2 reactor (3). The sulfur level in the reactor is controlled by equalization with the level in the sulfur day tank.

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

$$S + O_2 \rightarrow SO_2$$

The reaction is spontaneous and exothermic. The reactor temperature rises to about 1100°F, the boiling point of sulfur at 80 psig. The production rate of SO_2 is controlled by the oxygen feed rate to the reactor.

The vapor stream of SO_2 and sulfur is cooled in the sulfur condenser (4) to about 270°F. The condenser is cooled by generating steam at 35 psig. Most of the sulfur vapor condenses and the mixture of condensed sulfur, which flows by gravity, and SO_2 vapor is returned to the molten sulfur day tank. The liquid sulfur drops out in the sulfur day tank and is recycled to the reactor.

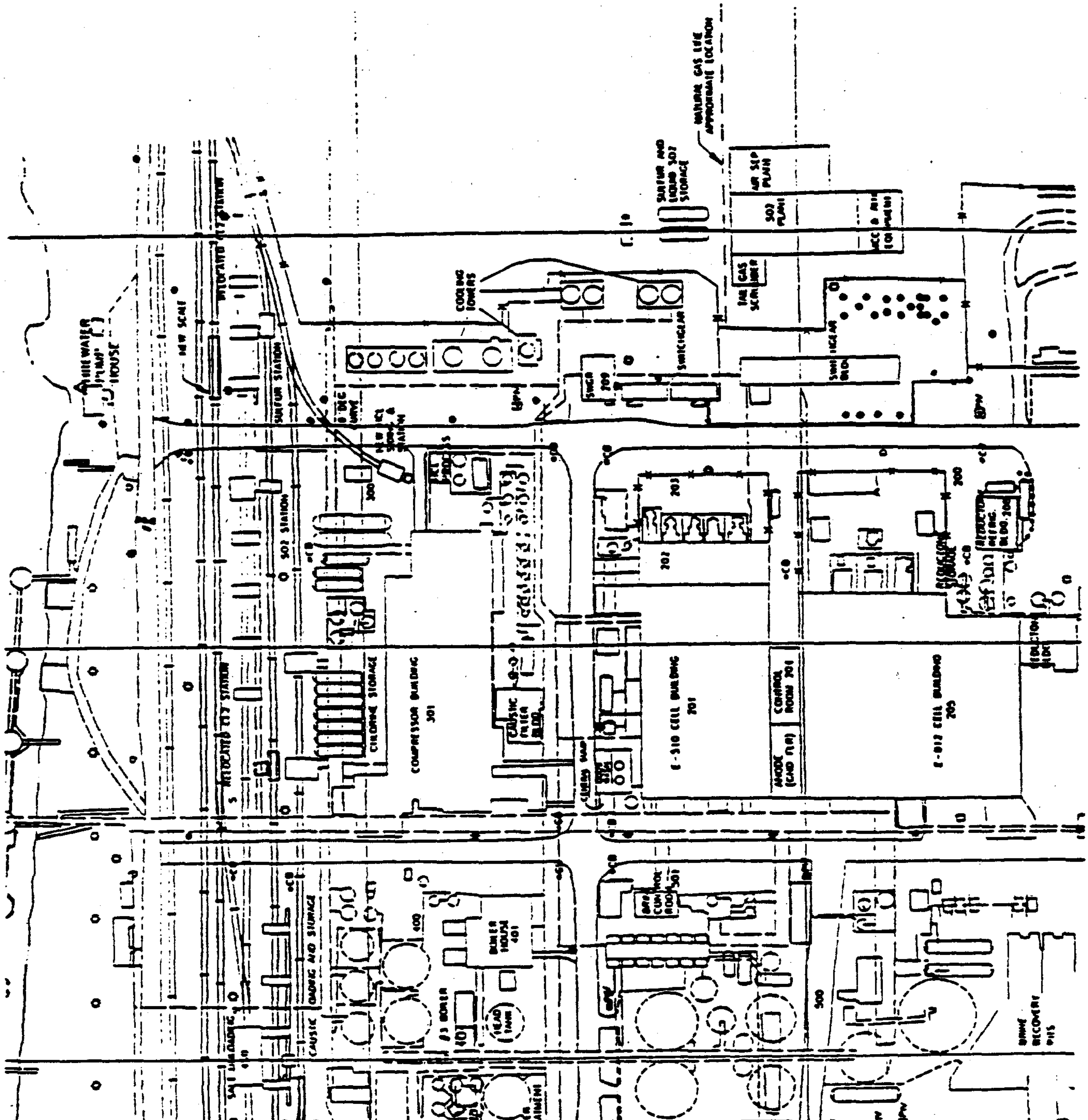

Figure 7. Liquid SO Plant Location

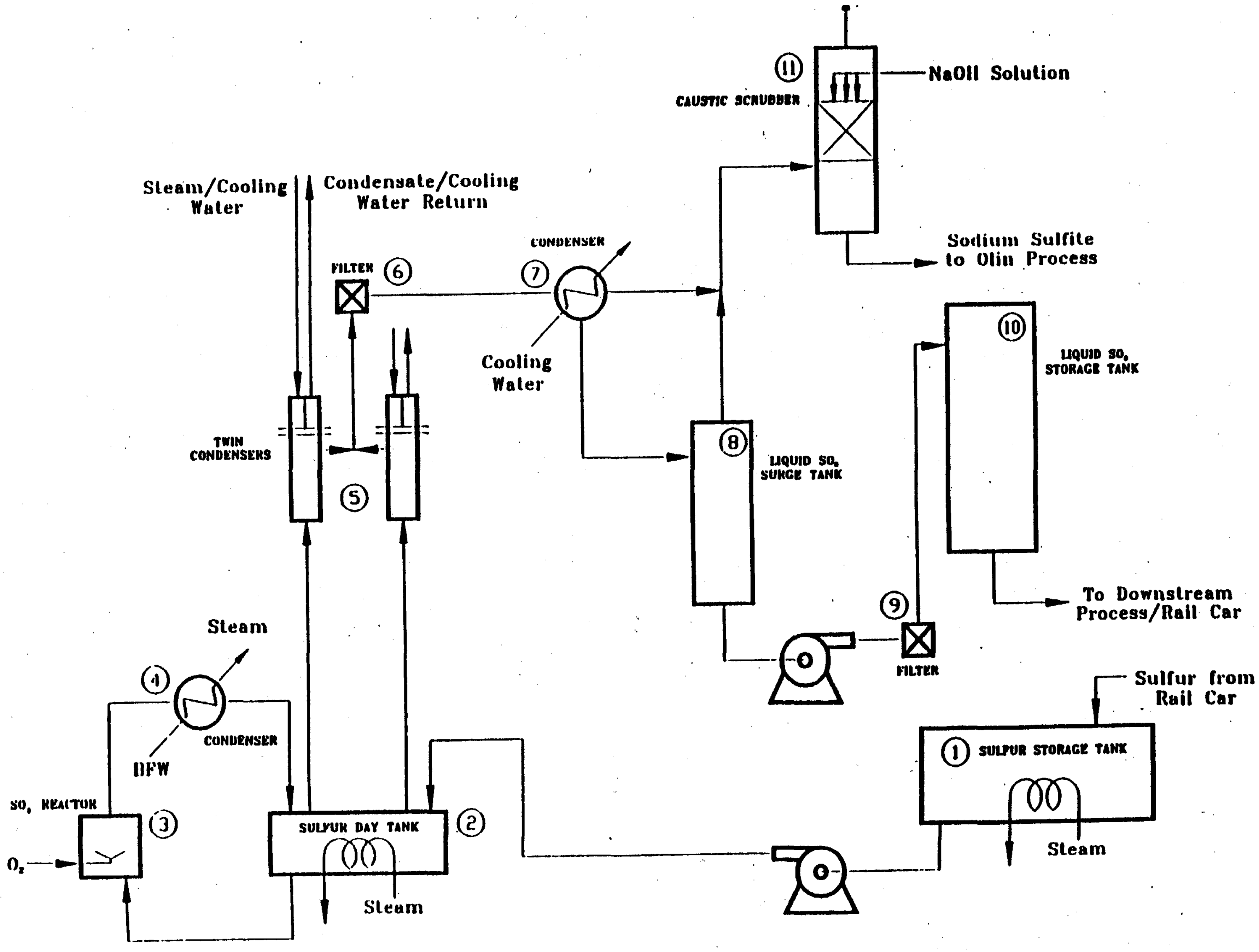

Figure 8. Liquid Sulfur Dioxide Process

The SO_2 vapor does not condense at 270°F and is not significantly soluble in molten sulfur. It is further cooled in the twin condensers (5) to remove additional trace amounts of sulfur. The condensers operate in a two step repeating cycle. In the first step, the condenser cools the SO_2 to 120°F using cooling water. Entrained liquid sulfur and remaining sulfur vapor will collect as a solid on the condenser tube walls. In the second step, the condenser gas outlet is blocked and the sulfur is melted using low pressure steam. The molten sulfur will drain by gravity back to the sulfur day tank. The condensers will alternate between these modes of operation; one condenser will remove sulfur while the second condenser is regenerated using steam.

After filtration (6), the SO_2 vapor is condensed in the SO_2 condenser (7) using cooling water. At the system pressure of 80 psig, the SO_2 condenses at about 104°F. The liquid SO_2 will flow to the liquid SO_2 surge tank (8). From the surge tank it will be pumped through a filter (9) to remove any entrained particulate then to a 200-ton capacity liquid SO_2 storage tank (10). From the storage tank, the liquid SO_2 will be pumped to an existing process liquid SO_2 feed tank or to rail cars for shipment.

A vent stream from the SO_2 condenser and liquid SO_2 surge tank contains non-condensibles, trace amounts of nitrogen and argon introduced to the sulfur reactor with the oxygen, and SO_2 vapor. The SO_2 vapor is removed from the vent stream in a caustic scrubber (11). A sodium hydroxide (NaOH) solution is used to remove the SO_2 vapor from the gas stream. The sodium sulfite formed from the reaction of NaOH and SO_2 will be used by Olin to neutralize a chlorine waste stream from an existing Olin process.

Air Separation Plant Description

Figure 9 presents a basic flow diagram of the air separation plant used to supply O_2 to the liquid SO_2 process. Primary unit operations are numerically labelled on this figure and referenced in the following discussion. As mentioned previously, oxygen is produced by liquefying air and then using fractional distillation to separate the liquefied air into its components. The three fundamental steps in this process are purification, refrigeration, and rectification.

Purification

Atmospheric air contains dirt, water vapor, and carbon dioxide (CO_2) which must be removed from the compressed air stream to prevent plugging of downstream process equipment. The atmospheric air passes through an intake filter (1) to remove entrained particulate and is compressed to 125 psig in a centrifugal compressor (2). After compression the air is cooled in a direct contact after cooler (3) using cooling water. Carbon dioxide, water vapor, and gaseous hydrocarbons are then removed by adsorption on activated alumina and molecular sieve (4). Parallel units are used, like the twin condensers from the liquid SO_2 process, one bed will be regenerated while the other is online. The adsorbents are regenerated using heat and a nitrogen purge gas generated downstream.

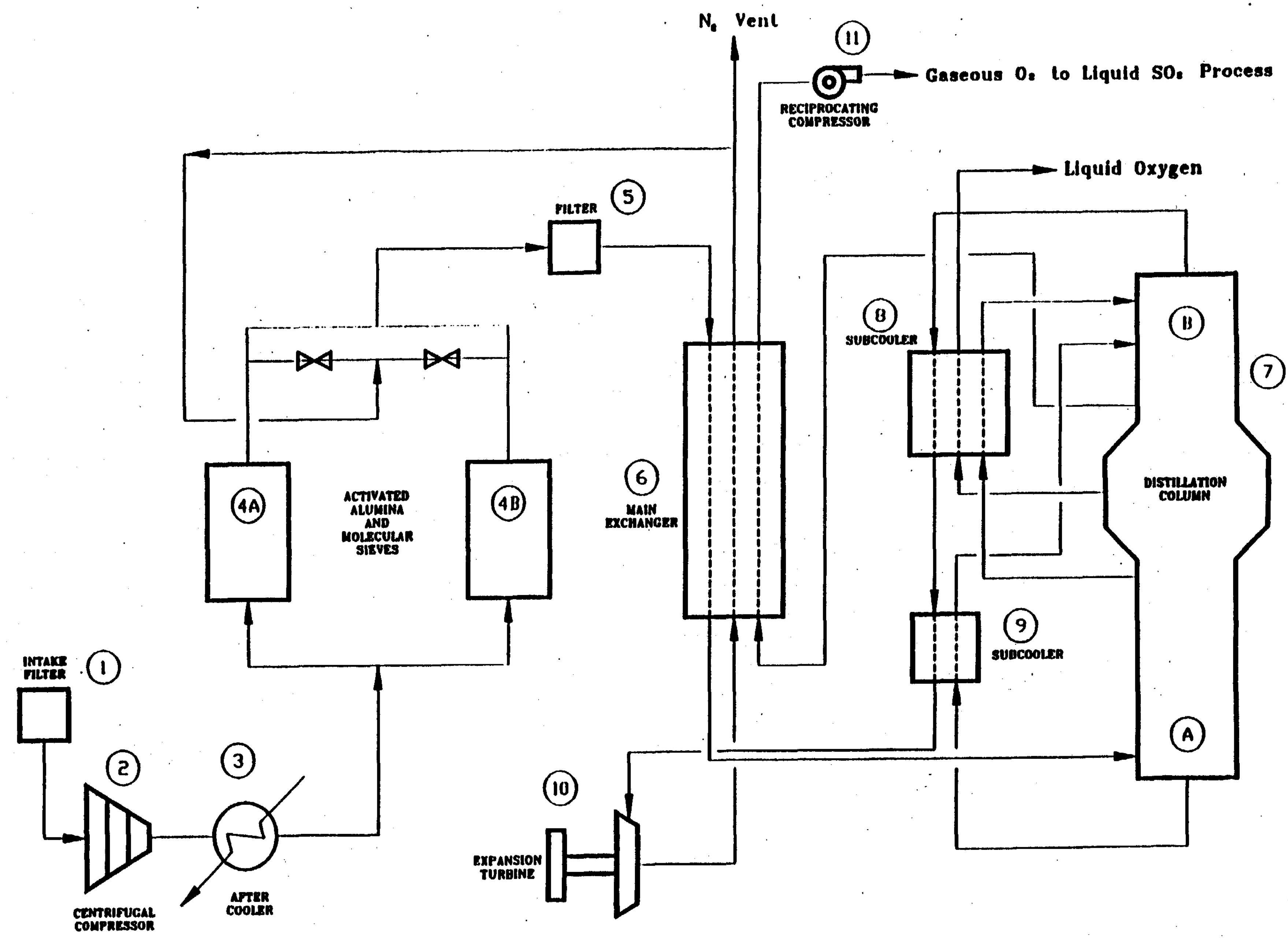

Figure 9. Air Separation Plant

Refrigeration/Rectification

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

Process Alternatives/Advantages

Traditional, older processes used to produce liquid SO_2 from sulfur involve burning the sulfur in air. The resulting gas stream can contain, at best, 16-18 vol.% SO_2 with the balance being mainly nitrogen, oxygen, and water from the combustion air. The SO_2 must then be separated from the other combustion gases. This is done by stripping the SO_2 from the gas stream using either water or an organic solvent like dimethylaniline. Regardless of which stripping liquor is used, these processes are more complex and have greater environmental impacts. To illustrate, the burn in air with water stripping process.

Process advantages of the burn-in-oxygen liquid SO_2 process include the following:

- Process gas at a lower temperature, 1100°F versus about 2500°F.
- Production of lower pressure steam, 35 psig versus 600 psig.
- No acidic wastewater stream which must be neutralized.
- Smaller volume tail gas stream which economically allows for the use of a more efficient scrubber resulting in lower SO_2 emissions.
- No spent acid stream which must be reclaimed or disposed of.
- No solvent emissions or disposal of solvent.

In addition, due to the lower process gas temperature and steam pressure, and simplicity of the process the liquid SO_2 process is inherently more reliable and safe to operate

IV. ACKNOWLEDGEMENT

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V. REFERENCES

1. Haslbeck J.L., C.J. Wang, Neal L.G., Tseng H.P., and J.D. Tucker, "Evaluation of the NOXSO Combined NO_x/SO_2 Flue Gas Treatment Process," U.S. Department of Energy Contract NO. DE-AC22-FE60148, (November 1984).
2. Haslbeck, J.L., W.T. Ma, and L.G. Neal, "A Pilot-Scale Test of the NOXSO Flue Gas Treatment Process," U.S. Department of Energy Contract No. DE-FC22-85PC81503, (June 1988).
3. Ma W.T., J.L. Haslbeck, and L.G. Neal, "Life Cycle Test of the NOXSO Process," U.S. Department of Energy Contract NO. DE-FC22-85PC81503, (May 1990).
4. Zhou Q., J.L. Haslbeck, and L.G. Neal, "An Experimental Study of NO_x Recycle in the NOXSO Flue Gas Cleanup Process," U.S. Department of Energy Contract No. DE-AC22-91C91337, (March 1993).
5. Haslbeck J.L., M.C. Woods, W.T. Ma, S.M. Harkins, and J.B. Black, "NOXSO SO_2/NO_x Flue Gas Treatment Process: Proof-of-Concept Test," presented at the SO_2 Symposium, Boston, MA (August 1993).