Integrated Dry NO_x/SO_2 Emissions Control System Performance Summary

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ABSTRACT

The Integrated Dry NO_x/SO_2 Emissions Control System was installed at Public Service Company of Colorado's Arapahoe 4 generating station in 1992 in cooperation with the U.S. Department of Energy (DOE) and the Electric Power Research Institute (EPRI). This full-scale 100 MWe demonstration combines low-NO_x burners, overfire air, and selective non-catalytic reduction (SNCR) for NO_x control and dry sorbent injection (DSI) with or without humidification for SO_2 control. Operation and testing of the Integrated Dry NO_x/SO_2 Emissions Control System began in August 1992 and will continue through 1996. Results of the NO_x control technologies show that the original system goal of 70% NO_x removal has been easily met and the combustion and SNCR systems can achieve NO_x removals of up to 80% at full load. Duct injection of commercial calcium hydroxide has achieved a maximum SO_2 removal of nearly 40% while humidifying the flue gas to a 20°F approach to saturation. Sodium-based dry sorbent injection has provided SO_2 removal of over 70% without the occurrence of a visible NO_2 plume. Recent test work has improved SNCR performance at low loads and has demonstrated that combined dry sodium injection and SNCR yields both lower NO_2 levels and NH_3 slip than either technology alone.

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INTRODUCTION

Public Service Company of Colorado (PSCC) was selected by DOE for a CTT-III project in December 1989 to demonstrate an Integrated Dry NO_x/SO_2 Emissions Control System. The demonstration project is taking place at PSCC's Arapahoe Unit 4, a 100 MWe top-fired unit which fires a low sulfur (0.4%) Colorado bituminous coal as its main fuel, but also has 100% natural gas capability. Figure 1 shows a boiler elevation drawing.

The Integrated Dry NO_x/SO_2 Emissions Control System combines five major control technologies to form an integrated system to control both NO_x and SO_2 emissions. The system uses low-NO_x burners, overfire air, and urea injection to reduce NO_x emissions, and dry sorbent injection using either sodium- or calcium-based reagents with (or without) humidification to control SO_2 emissions. The goal of the project was to reduce NO_x and SO_2 emissions by up to 70%. The combustion modifications were expected to reduce NO_x by 50%, and the SNCR system was expected to increase the total NO_x reduction to 70%. Dry Sorbent Injection was expected to provide 50% removal of the SO_2 emissions while using calcium-based reagents. Because sodium is much more reactive than calcium, it was expected to provide SO_2 removals of up to 70%. Figure 2 shows a simplified schematic of the Integrated Dry NO_x/SO_2 Emissions Control System at Arapahoe Unit 4.

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

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

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

TECHNOLOGY DESCRIPTION

This section will provide a brief description of the technologies used in the integrated NO_x/SO_2 Emissions Control System with emphasis on the SNCR and sodium dry sorbent injection system. The reader is referred to prior publications for more complete descriptions of the low NO_x combustion system and calcium dry sorbent system with humidification.(1-11)

Low NO_x Combustion System

B&W's DRB-XCL® (Dual Register Burner-aXially Controlled Low-NO_x) burner had been successfully used to reduce NO_x emissions on wall-fired boilers but had never been used in a vertically-fired furnace. The burner utilizes dual registers to control near burner mixing and a sliding air damper to control air flow to each individual burner independent of swirl. Twelve of these burners were installed on the roof of Arapahoe Unit 4. The low NO_x combustion system also incorporated three B&W dual zone NO_x ports which were added to each side of the furnace approximately 20 feet below the boiler roof. These ports can inject up to 28% of the total combustion air through the furnace sidewalls.

Arapahoe Unit 4 was originally designed with the ability to fire 100% natural gas. Natural gas firing capabilities were maintained with the DRB-XCL® burners by installing a gas ring header at the tip of the burner. However, the burner is not specifically designed to be a low-NO_x burner with natural gas firing.

Selective Non-Catalytic Reduction

The purpose of the SNCR system at Arapahoe was two-fold. First, to further reduce the final NO_x emissions obtained with the combustion modification so that the goal of 70% NO_x removal could be achieved. Second, the SNCR system is an important part of the integrated system interacting synergistically with the dry sodium injection system. During this program, it was shown that when both systems are used simultaneously, both NO_2 emissions from the sodium system and NH_3 slip from the SNCR system are reduced.

When the SNCR system was originally designed and installed, it incorporated two levels of wall injectors with 10 injectors at each level. These two separate levels were intended to provide load following capability. The locations of these two levels were based on flue gas temperature measurements made with the original combustion system. However, the retrofit low-NO_x combustion system resulted in a decrease in the furnace exit gas temperature of nominally 200°F. This decrease in temperature moved the cooler injection level out of the SNCR temperature window. With only one operational injection level, the load-following performance of the system was compromised.

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

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

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

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

Figure 4 shows the location of the new ARIL lances relative to the two original SNCR injection locations. Level 2 is the location that became unusable as a result of the flue gas temperature decrease after the low-NO_x combustion system retrofit. The ARIL system consists of two retractable lances and two retractable lance drive mechanisms. Each lance is nominally 4 inches in diameter and approximately 20 feet in length. Each lance has a single row of nine injection nozzles spaced on two-foot centers. A single division wall separates the Arapahoe Unit 4 furnace into east and west halves, each with a width of approximately 20 feet. When each lance is inserted, the first and last nozzles are nominally one foot away from the division and outside walls, respectively.

Each injection nozzle is composed of a fixed air orifice (nominally one-inch in diameter), and a replaceable liquid orifice. The liquid orifices are designed for easy removal and cleaning, because they can become plugged. This ability to change nozzles also allows adjustments in the chemical injection pattern along the length of the lance in order to compensate for any significant maldistributions of flue gas velocity, temperature, or baseline NO_x concentration.

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

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

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

Dry Sorbent Injection

PSCC designed and installed a dry sorbent injection system that can inject either calcium- or sodium-based reagents into the flue gas upstream of the fabric filter. Figure 5 shows a simplified flow diagram of the equipment. The reagent is fed through a volumetric feeder into a pneumatic conveying system. The air and material then pass through a pulverizer where the material can be pulverized to approximately 90% passing 400 U.S. Standard mesh. The material is then conveyed to the duct and evenly injected into the flue gas. After the original results suggested that the duct flue gas temperature was too low for effective SO_2 removal with sodium bicarbonate, the dry injection system was modified to allow injection of sodium-based compounds at the entrance to the air heater where the flue gas temperature is approximately 600°F. The pulverizer can be bypassed allowing calcium hydroxide to be fed from the silos and injected either ahead of the fabric filter or into the boiler economizer region where the flue gas temperature is approximately 1000°F.

To improve SO_2 removal with calcium hydroxide, a humidification system capable of achieving a 20°F approach to saturation temperature has been installed. The system was designed by B&W and includes 84 I-Jet humidification nozzles which can inject up to 80 gpm of water into the flue gas ductwork. The humidifier is located approximately 100 feet ahead of the fabric filter and there is no bypass duct.

Balance of Plant

Besides the major environmental equipment, the project also included required upgrades to the existing plant. A new distributed control system was installed to control the boiler and other pollution control equipment added as part of the integrated system. The fly ash collection system was also converted from a wet to a dry collection system to allow dry collection of the ash and injection waste products. A Continuous Emissions Monitor (CEM) system was installed at Arapahoe Unit 4 to collect data for the extensive test program. This monitor allows measurements of N_2O, NH_3, NO_2, and H_2O in addition to the more common pollutants.

RESULTS

Fossil Energy Research Corporation (FERCo) of Laguna Hills, California, has been performing all testing of the Integrated Dry NO_x/SO_2 Emissions Control System. The test program is nearing completion and the individual testing of the low-NO_x burners, overfire air, urea injection, calcium duct injection, calcium economizer injection, and sodium injection has been completed. Testing of the SNCR lances and the complete integrated system while firing low-sulfur coal is in progress. In addition to efficiency and emissions measurements, four tests were conducted to determine baseline and removal capabilities of the system for many common air toxic emissions. Prior papers (1-11) also discussed the performance of the individual technologies. This paper will overview these prior results and focus on the performance of the new SNCR injection lances and the performance of the integrated system. In addition, the paper will also present some recent information on NH_3 absorption on fly ash and the impacts on fly ash handling.

Review of NO_x and SO_2 Reduction Performance

This section will provide a brief overview of the NO_x and SO_2 reductions from the individual technologies used in the integrated system. The reader is referred to prior publications for more detailed discussions (1-11).

Low-NO_x Combustion System Performance

The following section describes the performance of the low-NO_x combustion system (low-NO_x burners and OFA ports).

Low-NO_x Burners

Figure 6 compares the Arapahoe Unit 4's NO_x emissions before and after the retrofitting of the low-NO_x combustion system. Note, NO_x ($NO + NO_2$) and NO are used interchangeably when discussing the performance of the combustion system, since NO_2 levels are very low for this unit. The original combustion system produced nearly uniform NO_x emissions of 800 ppmc (corrected to 3% O_2, dry) or 1.1 lb/MMBtu across the boiler's load range. The low-NO_x combustion system reduced NO_x emissions by over 63 percent, to less than 300 ppmc, across the boiler's load range. Note that all testing was conducted under steady-state conditions and with the careful supervision of test technicians to achieve the maximum possible reduction in NO_x emissions. Under load following conditions, NO_x emissions were about 10 to 25% higher. Additional modifications to the control system and additional operator training may decrease the difference.

Overfire Air

At full load, opening the OFA control dampers to full open (maximum OFA) diverts about 24% of the total combustion air to the OFA ports and at low load (60 MWe) it diverts about 28% of the combustion air. At maximum OFA, the low-NO_x combustion system reduces NO_x emissions by 62 to 69% across the load range. Since the OFA ports are located in a very hot part of the boiler, a significant amount of cooling air is required and the minimum amount of OFA is limited to about 15% of the total combustion air at full load and about 8% at low load. At minimum OFA, the retrofitted combustion system reduces NO_x emissions by 60 to 63%. Arapahoe Unit 4 cannot be tested at 0% OFA, but the small difference in NO_x reduction between maximum and minimum OFA indicates that the low-NO_x burners are responsible for most of the NO_x reduction.

SNCR System Performance

As mentioned previously, in addition to reducing NO_x emissions significantly, the low-NO_x combustion system also reduced the temperature of flue gas at the furnace exit by about 200°F. Since SNCR systems are very sensitive to changes in flue-gas temperatures, this reduction made the flue-gas temperature too cold for one row of injection nozzles, so all testing was performed using the row of injection nozzles originally designed for loads below 80 MWe. Figure 7 shows the SNCR performance achievable over the load range for a 10 ppm NH_3 slip limit with this single row of injectors. At full load, NO_x reductions of 45% are achieved. However, the performance decreased dramatically as the load decreased; at 60 MW, NO_x removals were limited to 11% for a 10 ppm NH_3 slip.

Calcium-Based Economizer Injection Performance

SO_2 removal has been less than expected with calcium hydroxide injection at the economizer. Initial testing at a Ca/S molar ratio of 2.0 without humidification resulted in SO_2 removals in the range of 5 to 8%. Note that the stoichiometric ratio for the Ca/SO_2 reaction is 1.0, since one mole of Ca reacts with one mole of SO_2 to form calcium sulfate, $CaSO_4$ It was found that the sorbent distribution was very poor, and only approximately one-third of the flue gas was being treated. New nozzles that increased reagent distribution only increased the SO_2 removal to 15% at a Ca/S molar ratio of 2.0. Although distribution of the calcium reagent is far from perfect, it appears that high levels of SO_2 removal are not possible at Arapahoe Unit 4 using the current $Ca(OH)_2$ material, even in areas with high Ca/S molar ratios.

Calcium-Based Duct Injection Performance

Higher SO_2 removal was achieved with duct injection of calcium hydroxide and humidification with SO_2 removals approaching 40% at Ca/S ratios of 2.0 and approach to saturation temperatures of 20 to 30°F. These levels of SO_2 removal are consistent with the prior DOE study at Ohio Edison's Edgewater Station.[13] Immediately after this test, problems developed with the dry fly ash transport system, and it is suspected that the low approach temperature contributed to this problem. Then, after a short period of 24 hour/day testing during load following operation, fabric filter pressure drop significantly increased due to the buildup of a hard ash cake on the fabric filter bags which could not be cleaned during normal reverse air cleaning. The heavy ash cakes were caused by the humidification system, but it was not possible to determine if the problem was caused by operation at a 30°F approach temperature or a short-excursion to a lower approach temperature caused by a rapid decrease in boiler load.

Sodium-Based Injection

Sodium-based reagents are much more reactive than calcium-based sorbents and can achieve significantly higher SO_2 removals during dry injection.[14,15] Figure 8 shows the SO_2 removal for dry sorbent injection for sodium bicarbonate and sodium sesquicarbonate. In Figure 8, SO_2 removals are plotted as a function of Normalized Stoichiometric Ratio (NSR). This corresponds to the amount of sodium compound injected relative to the amount of sodium required to form sodium sulfate, Na_2SO_4 (i.e., two moles of Na per mole SO_2). Sodium bicarbonate provided the highest SO_2 removal and was also the most efficient reagent in terms of sodium utilization. Flue gas temperature at the fabric filter inlet duct at Arapahoe Unit 4 varies from 250 to 280°F. The dry sorbent injection (DSI) system was originally designed for duct injection before the fabric filter only. However, initial testing with sodium bicarbonate showed that SO_2 removal was erratic, which was attributed to the low flue gas temperatures. The DSI system was modified to inject sodium sorbents at the air heater inlet where the flue gas temperature is approximately 600°F. It should be noted that sodium sesquicarbonate does not exhibit this slow reaction rate when injected ahead of the fabric filter.

A major disadvantage of sodium-based injection is that it converts some existing NO in the flue gas to NO_2. In addition, during the conversion process a small amount of the total NO_x, 5 to 15%, is removed. However, the net NO_2 exiting the stack is increased. While NO is a colorless gas, small quantities of the brown/orange NO_2 can cause a visible plume to develop. The chemistry of the conversion is not well understood but it is generally accepted that NO_2 increases as SO_2 removal increases. Figure 8 shows that NO_2 emissions are generally higher with sodium bicarbonate, although a significant amount of data scatter exists. The threshold NO_2 level that forms a visible plume is site specific; at Arapahoe Unit 4, a visible plume appears when NO_2 concentrations reach 30 to 35 ppm. Also, the NO_2 levels were found to depend on conditions in the fabric filter with NO_2 levels increasing dramatically after each cleaning cycle.(11)

SNCR Lance Performance Results

The recent test work has focused on the performance of the SNCR lances, both the NOELL ARIL lances and a comparison of the performance of the alternate DPSC lance to the ARIL lance.

ARIL Lances

Prior to incorporating the ARIL lances into the SNCR control system, a series of parametric tests was conducted to define the optimum injection angle at each load. As shown in Figure 4, each lance can rotate to inject urea into a different region of the furnace in order to follow the SNCR temperature window as the boiler load changes. The minimum injection angle is 22° (0° corresponds to injection vertically downward), at which point the chemical is injected parallel to the tube wall located below the lances. Smaller injection angles are not used to avoid direct liquid impingement on these tubes. An injection angle of 90° corresponds to injection straight across the furnace toward the front wall, and an angle greater than 90° results in injection of the solution in a direction up toward the roof-mounted burners.

While the primary focus of the parametric tests was to define the injection angle versus load, the tests also investigated the effects of:

- coal mill out-of-service patterns
- coal mill biasing
- biasing the urea flow along the length of the lances
- independent adjustment of the injection angles for each lance

The results of these tests are described below.

Effect of Lance Angle

One of the primary attributes of the ARIL lance system is the inherent flexibility of accessing the optimum flue gas temperature location by simply rotating the lance. Figure 9 shows the effect of varying the lance injection angle at loads of 43 and 50 MWe. All of the tests shown in these figures were performed at a N/NO_x ratio of 1.0, with two mills in service. At 43 MWe, varying the injection angle had little effect on NO_x removal, and the maximum removal occurred at an angle of 35 degrees (Figure 9a). However, Figure 9a shows that the lance angle had a large effect on NH_3 slip; decreasing from 46 ppm at an angle of 22° to under 5 ppm at an angle of 135°. This overall behavior at 43 MWe suggests that, on average, injection is occurring just on the high side of the SNCR temperature window. In fact, the optimum temperature, in terms of NO_x removal, appears to correspond to an angle of 35°. However, since it is desirable to maintain the NH_3 slip less than 10 ppm, an injection angle of 90° is a more appropriate operating angle at this load.

At a slightly higher load of 50 MWe (Figure 9b), the effect of lance injection angle was markedly different. At this load, where the average flue gas temperatures were higher, injection angle had little effect on NH_3 slip. However, at the higher temperature, lance angle had a large effect on NO_x removal. The relative insensitivity of the NH_3 slip and large sensitivity of the NO_x removal to lance angle suggests that at 50 MWe, chemical injection is occurring far on the high side of the SNCR temperature window for injection angles ranging from 22° to 135°.

The results at 43 and 50 MWe shown in Figure 9 illustrate how varying lance angle can be used to optimize the SNCR performance over the load range. As the load increases, the preferred injection angle will decrease. Again, the minimum angle is 22°, where the chemical is injected parallel to the tube sheet located below the lances.

Performance over the Load Range

The SNCR performance using the ARIL lances over the load range from 43 to 80 MWe is shown in Figure 10. Note that for this particular lance location, the flue gas temperatures are too high for the lances to be effective at loads greater than 80 MWe. As the load increases, the preferred lance angle decreases in order to inject the urea into a lower temperature region.

As discussed above, at 43 MWe with an angle of 90°, injection occurred on average just on the high temperature side of the window. At $N/NO_x = 1$, NO_x removals were 35% with less than 10 ppm NH_3 slip. At 50 MWe, a 45° injection angle was on average at a better location in the SNCR window, with NO_x removals of 40% and NH_3 slip less of 5 ppm at $N/NO_x = 1$. As the load increased to 60 MWe, a decrease in lance angle to 34° resulted in SNCR performance similar to a load of 43 MWe. At higher loads of 70 and 80 MWe, injection was clearly occurring on the high side of the temperature window. Note that the NH_3 slip at 80 MWe was higher than the slip at 70 MWe even though the chemical was injected into a region of higher overall temperature (i.e., compare the NO_x removals at 70 and 80 MWe in Figure 10). This effect was a result of temperature stratification in the furnace, and the way in which the stratification varies with different coal mill patterns. This effect is discussed in more detail below. However, comparing Figures 9 and 10 to the low load performance of the wall injectors in Figure 7 clearly shows that the lances have markedly improved the low-load performance of the SNCR system.

Effect of Boiler Operation on SNCR Performance

As mentioned above, local changes in temperature due to variations in boiler operating parameters (excess O_2, mill pattern, mill biases, etc.) can have a major impact on SNCR performance. This is particularly true at Arapahoe Unit 4 where the 12 burners are located on the roof of the furnace. Each of the four coal mills feeds three burners, two burners on one side of the furnace and a single burner on the other side of the furnace. Since the furnace has a division wall, there is an imbalance in heat release across the furnace, and a corresponding variation in flue gas temperature, when only three mills are in service. These temperature variations impact the performance of both the wall injectors and the ARIL lances. In this paper, the effect will be illustrated by looking at the performance of the ARIL lances with varying mill out-of-service patterns. During normal operation, Arapahoe Unit 4 operates with four mills in service over the load range from 80 to 110 MWe (although the unit can operate up to 100 MWe with only 3 mills). From 60 to 80 MWe, the unit typically operates with three mills in service. Below 60 MWe, the unit is usually operated with only two mills in service.

Figure 11 shows the effect of various mill out-of-service (OOS) patterns on east/west imbalances across the furnace. The bottom of Figure 11 shows a plan view of the in-service burners (numbered) and out-of-service burners (filled circles) for a given mill pattern. Note that the left side of these figures corresponds to the west wall of the furnace (adjacent to burners 1, 2 and 3), and the right side corresponds to the east wall (adjacent to burners 10, 11 and 12). With either A mill or C mill out-of-service, more heat release occurs on the east side of the furnace, while the west side has more heat release with either B mill or D mill out-of-service.

The change in lance metal temperatures provides a general indication of changes in flue gas temperatures on the east and west sides of the furnace. As seen in Figure 11, the changes in lance metal temperatures reflect the variations in heat release in the furnace with differing mill out-of-service patterns. Correspondingly, the NO_x removals and NH_3 slip levels also reflect these variations in temperature. For instance, NH_3 slip decreased on the west side when D mill was out-of-service, since more coal was fired (and the flue gas temperatures were higher) on the west side. The lance metal temperatures also indicated that, in general, the east side of the furnace was hotter than the west side. Figure 12 shows the overall impact of various mill out-of-service patterns on SNCR performance at 60 MWe. As can be seen, NO_x removals varied from 30% to 52% (@ $N/NO_x = 1.5$) depending on which particular mill was out-of-service. Comparably, the NH_3 slip varied from under 5 ppm to over 30 ppm with different mill-in-service patterns. This behavior made overall optimization of the SNCR system quite challenging.

In addition to the temperature variations that occur with the various mill out-of-service patterns, day-to-day variations can occur as a result of changes in the performance of the individual mills, or changes in any other variables which affect the flue gas temperature distribution. Three operational changes were investigated to deal with these types of temperature variations.

- varying the urea flow along the length of each lance
- independently varying each lance angle
- biasing the in-service coal mills

Varying the urea flow between the two liquid zones in each lance provided minor improvements in the performance of the SNCR system. Independently varying the lance angles as a function of the mill-in-service pattern also provided minor improvements. Unfortunately the implementation of either of these strategies would significantly complicate the automatic control system. On the other hand, biasing the in-service coal mills, which is relatively easy to implement, resulted in major improvements in the performance of the SNCR system. Arapahoe Unit 4 is equipped with four O_2 monitors at the economizer exit. Biasing the coal mills to provide a balanced O_2 distribution at this location is a fairly simple exercise for the boiler control operator. Figure 13 shows the improvements in SNCR performance that can be achieved by biasing the coal mills. These tests were performed at a load of 60 MWe with both lances at an injection angle of 22° and A mill out-of-service. The "biased" condition in Figure 13 corresponds to a negative 10% bias on B mill and D mill, and a positive 10% bias on C mill. This has the net effect of moving coal from the east side to the west side of the furnace to compensate for A mill being out-of-service (see bottom of Figure 11). Biasing the mills increased NO_x removals from nominally 27% to 42% at an NH_3 slip limit of 10 ppm.

Overall System Performance

The parametric tests were conducted to determine at which loads the ARIL lances should be used, as well as the optimum injection angle for each of these loads. Based on the parametric tests, the control system has been set up to operate with the Level 1 wall injectors at loads above 80 MWe. Below 80 MWe, the ARIL lances are used. Figure 14 compares the NO_x removal over the load range for injection at the two locations with an NH_3 slip limit of 10 ppm. It is evident that the installation of the ARIL lances has improved low-load performance of the SNCR system. Currently, NO_x removals of more than 30% are achievable over the load range with less than 10 ppm NH_3 slip. The minimum NO_x removal of 30% occurs at 80 MWe, which corresponds to the point where the temperature becomes too high for the ARIL lances and too low for the Level 1 injectors. With continuing operation of the system, it is anticipated that further optimization will take place as the operators gain more experience balancing the furnace.

Alternate Lance Design

While the NO_x removal performance of the ARIL lances has been good, their location in the furnace has resulted in some operational problems. At this particular location in the furnace, the lances are exposed to a large differential heating between the top and bottom surfaces. The top surface receives a high radiant load from the burners, while the bottom of the lance radiantly communicates with the relatively cold tube wall immediately below. This uneven heating pattern causes a great amount of thermal expansion along the upper surface, and the lances bend downward toward the tubes. Within 30 minutes of insertion, the tip of each lance would drop by approximately 12 to 18 inches. Within less than six weeks of operation, the lances became permanently bent, making insertion and retraction difficult. This was partially addressed by adding additional cooling slots at the end of the lance.

An alternate lance design supplied by Diamond Power Specialty Company (DPSC) was evaluated during this test period. As mentioned previously, this design sprays the urea solution through a single atomizer at the entrance to the lance. This provides evaporative cooling to supplement the air cooling. The evaporative cooling was expected to help minimize the lance bending discussed above. This alternate lance design was evaluated by installing a single lance on the west side of the boiler in place of one of NOELL's ARIL lances. The two different lance designs were compared during a nominal three week test program.

Overall, the DPSC lance performed mechanically well. The lance exhibited less bending than the ARIL lance, and eliminated air injection on the outside of the boiler.

Figure 15 compares the performance of the ARIL and DPSC lances on the west side of the furnace. In terms of NO_x reduction and NH_3 slip performance, the DPSC lance was not quite as good as the ARIL lance. With the B mill OOS, the ARIL lance yielded 42 percent NO_x removal with less than 5 ppm slip on the west side (60 MWe, $N/NO_x = 1$). Under comparable conditions, the DPSC lance yielded 36 percent NO_x removal and less than 5 ppm slip. This slight difference in performance is primarily attributable to the urea distribution along the lance. The ARIL lance uses a separate liquid circuit with individual liquid orifices at each air nozzle. This results in a fairly uniform liquid distribution along the length of the lance. The DPSC lance, on the other hand, sprays the urea solution into the cooling air stream at the inlet to the lance. Impingement on the walls and incomplete evaporation results in the liquid tending to be carried toward the far end of the lance, with part of the urea exiting as a stream of liquid rather than a finely atomized spray. In fact, this explains why the optimum angle for the DPSC lance is 34° compared to 22° for the ARIL lance at 60 MWe. The higher temperature associated with the 34° angle is needed to evaporate the liquid stream. In addition, the feed tube geometry of the DPSC lance created an additional pressure drop, restricting the amount of cooling air flow. This resulted in less penetration of the air jets, although this was partially compensated for by the unatomized portion of the urea solution, which carried the urea farther into the furnace before decomposing and releasing the reactive nitrogen components.

Overall, the results of the short test program of the DPSC lance were sufficiently positive that a second DPSC lance has been ordered. An additional three weeks of testing is planned.

Integrated System Performance

An important part of the test program was demonstrating the integrated performance of the various NO_x reduction and SO_2 removal technologies. In particular, a key element of the program was documenting the synergistic benefits of simultaneous operation of the SNCR and sodium-based dry sorbent injection system. When operated together, it was expected that the SNCR system would reduce NO_2 emissions from the sodium DSI system, while the sodium DSI system would in turn reduce NH_3 slip from the SNCR system.

Ideally, it would have been desirable to parametrically evaluate the merits of the integrated system over a range of operating conditions. This was not entirely possible for a number of reasons. With sodium-based dry sorbent injection, NO_2 levels are not only dependent on the amount of sodium injected, but also depend on the particulate in the fabric filter and the cleaning intervals.(11) Likewise, the time required for NH_3 levels to stabilize at the exit of the fabric filter, both before and after sodium injection, was greater than the 10-hour a day period during which the load from Arapahoe 4 could be blocked. Thus, characterizing the integrated performance relied on a limited number of parametric tests followed by a series of "long term" tests under normal load following conditions. During these "long-term" tests, the NO_x reduction and SO_2 removal systems were operated in automatic while the unit was operated according to system dispatch requirements. Data were collected at regular intervals using a data logger. No effort was made to set up specific test conditions, as these tests were designed to simulate operation of these systems once they are turned over to the plant at the completion of this program.

The results of a parametric test with sodium sesquicarbonate injection and the SNCR system are shown in Figure 16. During these tests, the DSI system was started first, followed by the SNCR system. For this test, the DSI system was set at an NSR of 2.0 (i.e., 4 moles of sodium per mole of SO_2) and the SNCR system at $N/NO_x = 0.6$. Following the start of the DSI system, the SO_2 removal stabilizes at nominally 70% removal and the NO_x removal at 12%. This level of NO_x removal is consistent with previous tests of the DSI system. The NO_2 levels increased to only about three ppm at the point that the SNCR system was started. With the SNCR system started, the NO_x removal increased to 35 to 40% and the NO_2 levels remained constant at three ppm. Even following a cleaning cycle, the NO_2 levels did not increase with the SNCR system in operation. Just before 1800 hours, the SNCR system was turned off and an immediate increase in NO_2 was noted.

Figure 17 shows the results of a parametric test with sodium bicarbonate injection ahead of the air preheater. With sodium bicarbonate injection alone at an NSR of 1.1, NO_2 levels on the order of 50 ppm are expected (see Figure 8). For the test results shown in Figure 16, the SNCR system was started at $N/NO = 1.1$ nominally two hours before the DSI system. As can be seen, the NO_2 levels remained near zero for the entire test. Further, it can be seen that following the start of the DSI system, the NH_3 slip levels continued to decrease.

The results shown in Figures 16 and 17 clearly show that there is a synergistic benefit of operating the SNCR and sodium-based DSI systems simultaneously.

Because of the difficulties encountered running these short term integrated tests, the balance of the integrated tests were run under normal load following conditions. During these tests the integrated system was operated 24 hours per day. Figure 18 shows the data collected during one 24-hour period (February 25, 1996). During these tests, the integrated system was utilizing sodium sesquicarbonate injection ahead of the fabric filter, and the SNCR system was load following with both the wall injectors and ARIL lances.

On this day, the boiler load was nearly constant for the first 17 hours of the day. The N/NO_x ratio and NH_3 emissions were also relatively steady during this time. At 1600 hours, the DSI system was started with a 75 percent SO_2 removal setpoint with the hope that the load would remain steady and it would be possible to assess the beneficial effects of running the integrated system. Although, the load increased significantly about two hours after the DSI system was started, it eventually settled back down to a level similar to the level before the increase. Figure 18 shows that the average NH_3 emissions with and without sodium injection were similar, which was expected since the NH_3 trim control was functioning during both of these tests. However, the results also show that there was a substantial increase in the N/NO_x ratio. Since the SNCR control system was set to maintain the NH_3 emissions within the range of 7 to 8 ppm, it should have increased the urea injection rate if the DSI system reduced NH_3 emissions. A temporary increase was expected as a result of the load swing, but the N/NO_x ratio should have returned to the pre-swing level within two to three hours (as was seen after the "morning demand peak" between 0800 and 0900 hours). When the DSI system was started at 1600 hours, there was an immediate 10 percent increase in the NO_x removal, which is consistent with the increases seen during sodium-based DSI-only tests. After this initial NO_x removal increase, there was another slower increase (amounting to nominally 10 to 15 percent removal) which occurred as the N/NO_x ratio increased. Although the scaling of the data makes it difficult to see, Figure 18 indicates that the N/NO_x ratio basically doubled after the DSI system was started. The increase in N_2O emissions (from nominally 8 to 16 ppm), confirms that the N/NO_x ratio was increased by roughly a factor of two. These results clearly indicate that there was a substantial reduction in the stack NH_3 slip, when the SNCR and DSI systems were run concurrently.

Figure 19 shows data collected during the 24-hour period on March 4, 1996. The DSI system was operated for the entire period and the SNCR system was started at 1420 hours. The boiler load was fairly steady at this time, and was low enough for the control system to insert the ARIL lances. Although the DSI feedrate was not very consistent, Figure 18 shows that there was nominally a 50 percent reduction in the NO_2 emissions when urea injection began. The load remained steady for nearly four hours; then it increased for the usual "evening demand peak" at 1800 hours. When the lances retracted, the N/NO_x ratio dropped as demanded by the control system, and the NO_2 emissions were also seen to decrease. By 1900 hours, the NO_2 emissions had been reduced to near-zero levels. This effect is due to the difference in the NH_3 emissions between injection at the Level 1 and ARIL locations. Although effort was made to set up the SNCR control system such that the NH_3 slip was limited to 10 ppm throughout the load range, the Level 1 location is "cooler" overall than the ARIL location; thus injection at Level 1 is more sensitive to variations in the flue gas temperature profile. Therefore, in general, urea injection at the Level 1 location results in higher NH_3 slip levels at the fabric filter inlet. Since the NH_3 emissions are generally higher with urea injection at the Level 1 location, it would be expected that the reduction in stack NO_2 emissions would also be higher (relative to injection at the ARIL location). The hypothesis is further supported by the decrease in NO_2 emissions seen when the urea injection switched from the lances to Level 1 at 1800 hours in Figure 19, but also by the increase in NO_2 seen when the lances were reinserted at 2000 hours. When the lances went in at this time, the NO_2 emissions were essentially zero. After an hour, however, the NO_2 emissions slowly began to increase, finally leveling out at approximately 8 ppm.

The above (Figures 16 through 19) demonstrate the synergistic benefits of the integrated process. The NH_3 slip from the SNCR process suppresses the NO_2 emissions associated with NO to NO_2 oxidation by dry sodium injection. Concurrently, the sodium reduces the NH_3 slip from the SNCR process. (Note: In the present case, the control system adjusts the urea injection rate to maintain a set NH_3 slip level, and the tendency to reduce NH_3 slip is manifested in a higher N/NO_x ratio for a given NH_3 slip.)

Ammonia Absorption on the Fly Ash

An issue that needs to be addressed with any post-combustion NO_x reduction technology with NH_3 slip is the absorption of ammonia on the fly ash. This can have a number of impacts ranging from personnel safety while handling the ash, odor problems, or impacting the salability of the ash for future use as a cement aggregate. In the latter, a salable product becomes a disposal problem with an attached economic penalty. At the Arapahoe Station, the ash is not sold for use in cement. Thus, the only problems that have been encountered have been an occasional NH_3 odor around the ash handling area and potential concern with worker safety should the concentrations become too high.

At Arapahoe Unit 4, ash is removed from the fabric filter hoppers with a vacuum system and transported dry to an ash silo. When loaded onto trucks for transport to the disposal site, the ash is wetted with about 20% water (by weight) in order to minimize fugitive dust emissions. Depending upon the specific ash characteristics, this wetting process can result in the release of NH_3 vapors from the ash. Whether or not NH_3 is released from the ash depends primarily on the pH of the aqueous phase on the surface of the ash particles. As the pH increases above a level of 9 to 9.5, there is an increased release of vapor-phase ammonia.

During the test program with urea injection alone, the ammonia concentration in the ash varied over the range of 100 to 200 ppm (measured on a weight basis). The ash ammonia content appeared to be primarily related to the NH_3 slip levels from the SNCR system and, to some extent, the fabric filter cleaning cycles. During long-term testing with the SNCR system alone, and a 10 ppm NH_3 slip limit at the stack, there were no incidents of excessive NH_3 odors during the ash handling process.

Testing has shown that when the SNCR system is operated in conjunction with the dry sodium injection system, the urea injection rate could be increased substantially while maintaining a 10 ppm NH_3 slip level at the stack. This is one of the synergistic benefits of the patented Integrated

Dry NO_x/SO_2 Emissions Control System discussed above. However, during these tests, the ammonia concentration in the ash increased to the range of 400 to 700 ppm (weight basis), and there were frequent occurrences of NH_3 odors at the ash silo. Reducing the NH_3 slip set point to the range of 4 to 5 ppm reduced the ammonia concentration of the fly ash down to the 100 to 200 ppm range (weight basis), but the odor problem persisted.

At first, it was thought that the odor problem was a result of the sodium changing the pH of the ash. The pH resulting from placing 0.5 gram of ash in 200 ml of distilled water was 9.3 for an ash sample without sodium injection. The same test run with an ash sample from a test with sodium injection resulted in a pH of 10.3. While the sodium did indeed increase the pH, which in turn would tend to release more NH_3 from the aqueous to the vapor phases, the pH difference did not appear significant enough to account for the ash handling problems encountered.

An interesting observation was made during the pH measurements. While the presence of sodium was found to slightly increase the final pH, it was also found to have a large effect on the rate at which the pH changed as the ash was wetted. Figure 20 shows the change in pH versus time after 0.5 gram of ash is placed into 200 ml of distilled water and stirred. With the coal ash alone, almost 30 minutes are required for the soluble components of the ash to dissolve and change the pH to a final value of 9.3. However, with sodium present in the ash sample, the pH develops almost instantaneously, presumably because of the higher solubility of the sodium compounds in the ash. This more rapid development of the high pH level can result in more rapid and localized release of the ammonia vapor, and may explain the odor problem encountered when concurrently operating the SNCR and sodium systems. Other than decreasing the level of NH_3 slip from the SNCR system, additional approaches to dealing with this issue have not been explored.

CONCLUSIONS

Public Service Company of Colorado, in cooperation with the U.S. Department of Energy and the Electric Power Research Institute, has installed the Integrated Dry NO_x/SO_2 Emissions Control System. The system has been in operation for over three years and preliminary conclusions are as follows:

- NO_x reduction during baseload operation of the unit with low-NO_x burners and overfire air ranges from 63 to 69% with no increase in unburned fly ash carbon or CO emissions.
- With the addition of retractable lances to the SNCR system, improved low load performance of the system urea-based SNCR injection allows an additional 30 to 52% NO_x removal with an ammonia slip limit of 10 ppm at the fabric filter inlet. This increases total system NO_x reduction to greater than 80% at full load, significantly exceeding the project goal of 70%.
- The ability to follow the temperature window by rotating the ARIL lances has been demonstrated and also proved to be an important feature in optimizing the performance of the SNCR system.

- SO_2 removal with calcium-based dry sorbent injection into the boiler at approximately 1000°F flue gas temperature was disappointing with less than 10% removal achieved.

- SO_2 removal with calcium-based dry sorbent injection into the fabric filter duct has been less than expected with a maximum short term removal rate approaching 40%.

- Sodium bicarbonate injection before the air heater has been very effective with short term SO_2 removals of over 80% possible. Longer term testing has demonstrated removal near 70% at an approximate NSR of 1.0.

- Sodium sesquicarbonate injection ahead of the fabric filter can achieve 70% removal on a long term basis, at an approximate NSR of 2.0.

- NO_2 emissions are generally higher when using sodium bicarbonate than when using sodium sesquicarbonate. The NO_2 generated during sodium-based injection is related to SO_2 removal and the cleaning cycle of the fabric filter, but all factors important to NO_2 generation are not fully understood.(11)

- Long term testing of the integrated system demonstrated the synergistic benefit of operation with SNCR and sodium-based dry sorbent injection (i.e., reduce NO_2 and NH_3 emissions).

- When the SNCR and dry sodium systems were operated concurrently, an NH_3 odor problem was encountered in the area around the unit 4 ash silo. This problem appears to be related to the rapid change in pH due to the presence of sodium in the ash.

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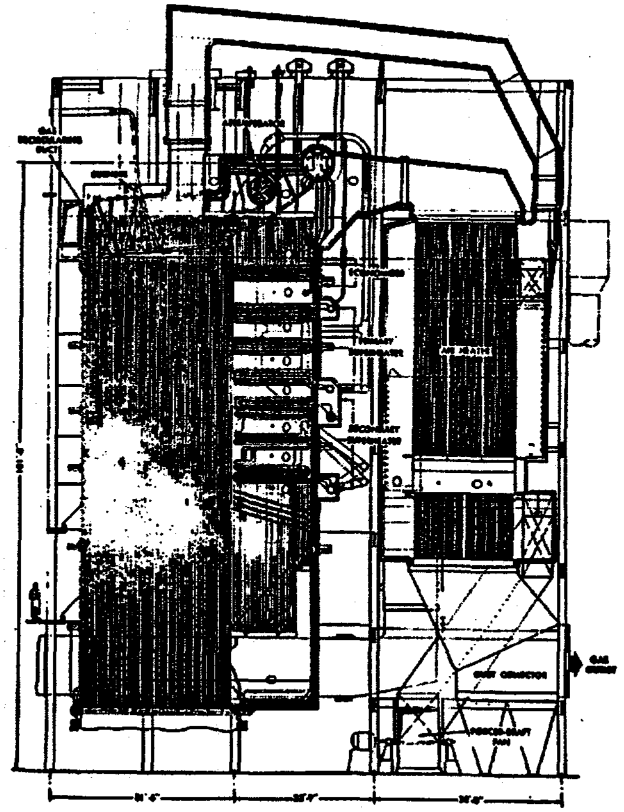


Figure 1. Boiler Elevation

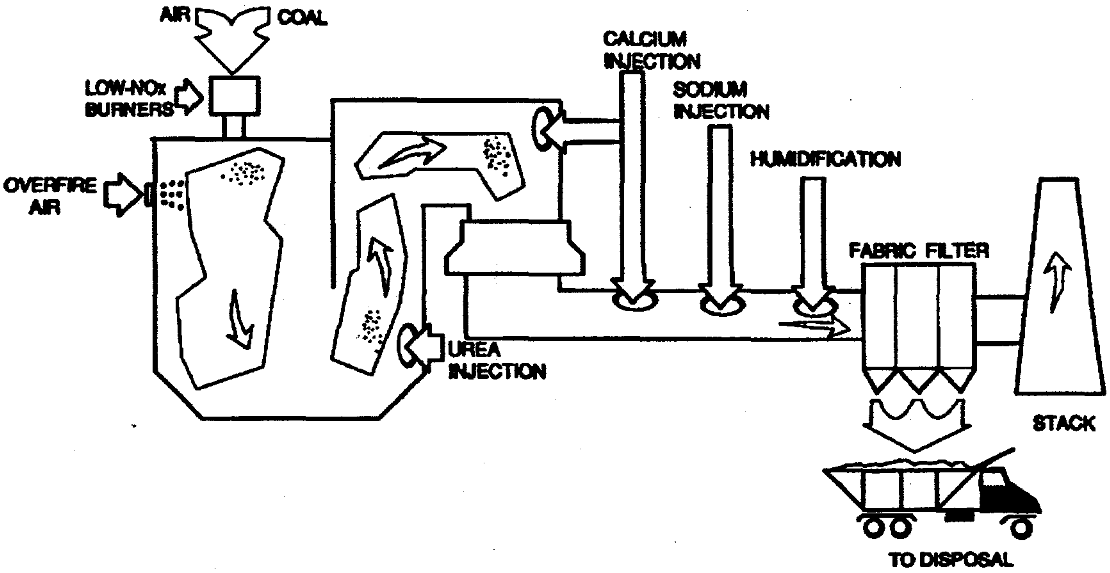


Figure 2. Process Flow Diagram of the Integrated NO_x/SO_2 Emission Control System

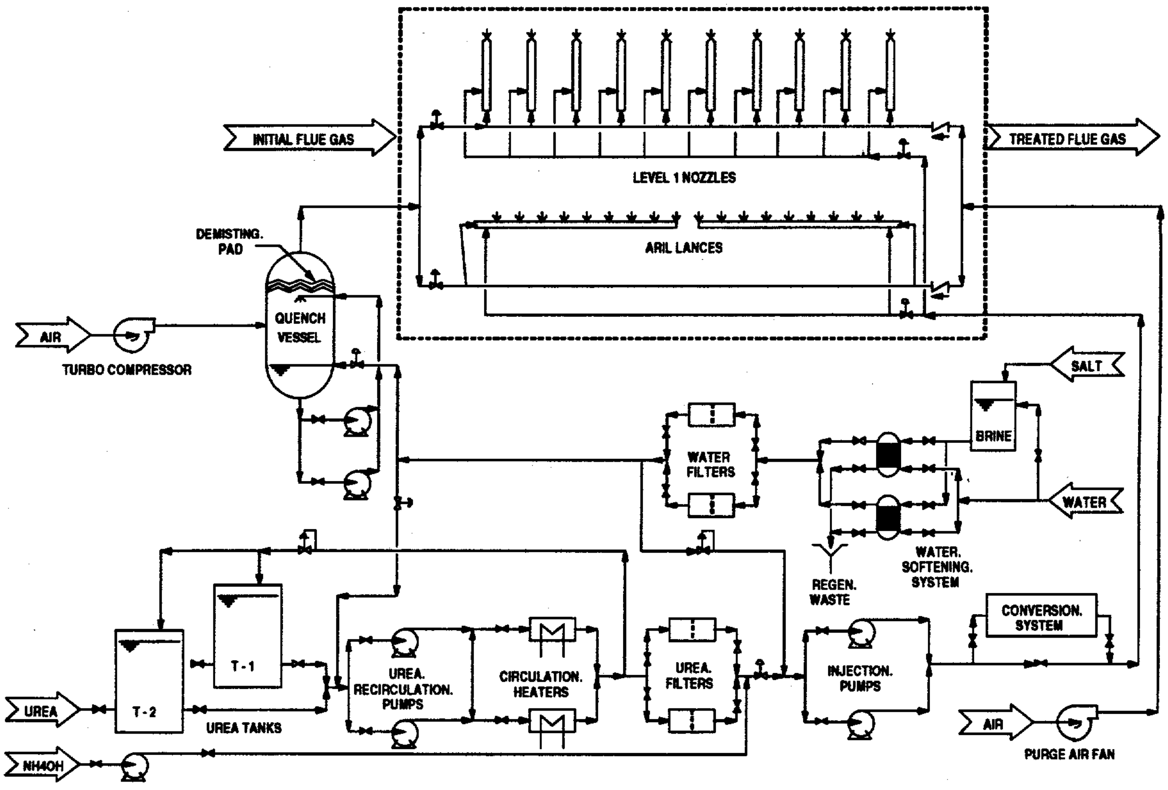


Figure 3. SNCR System Flow Diagram

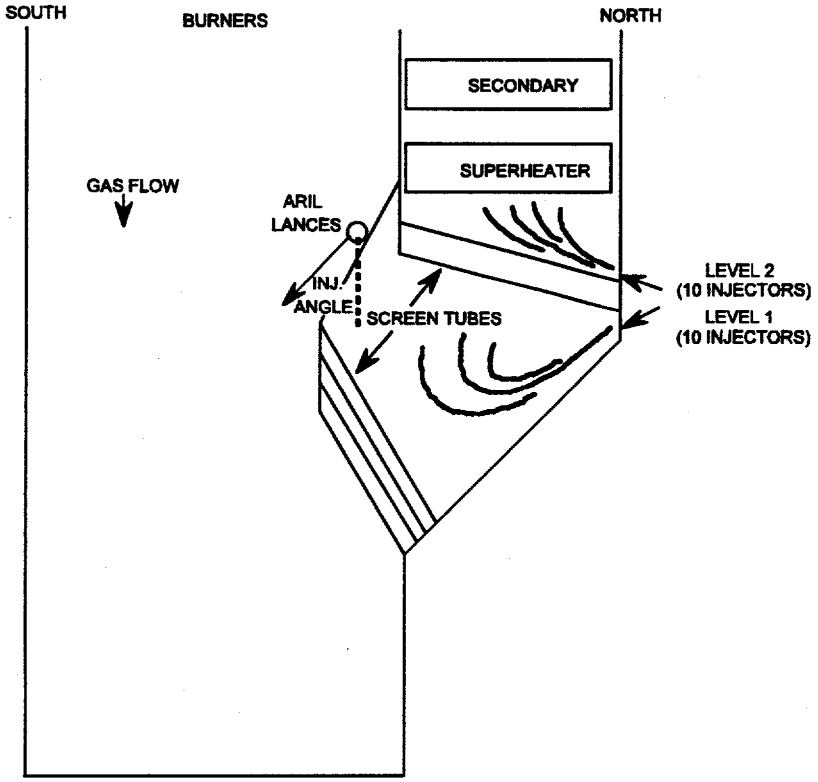


Figure 4. SNCR Injection Locations

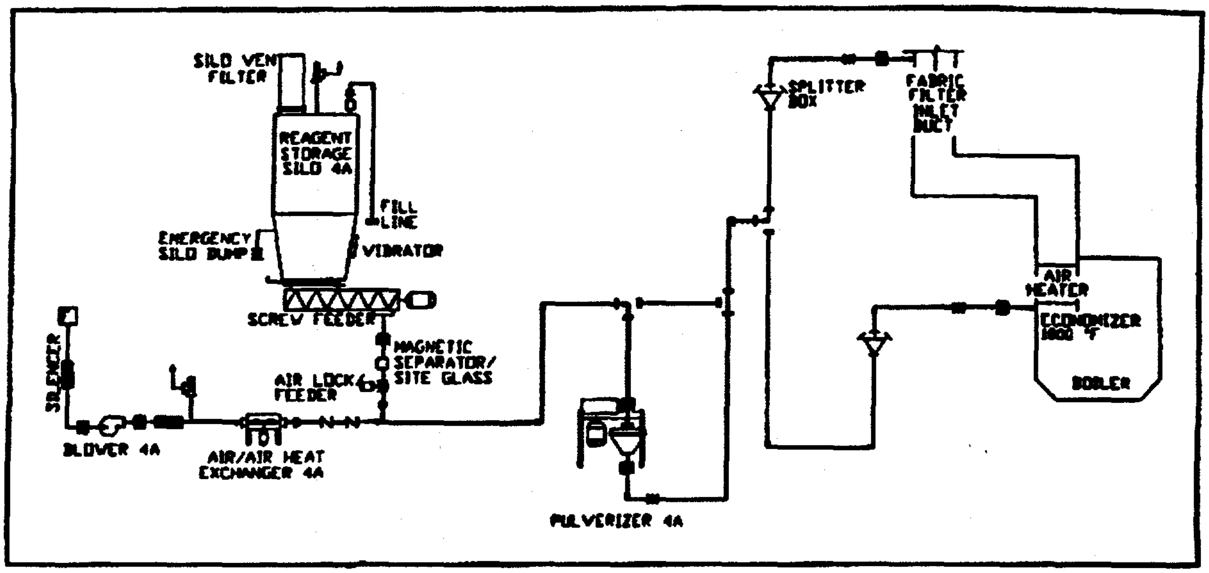


Figure 5. Dry Sorbent Injection Flow Diagram

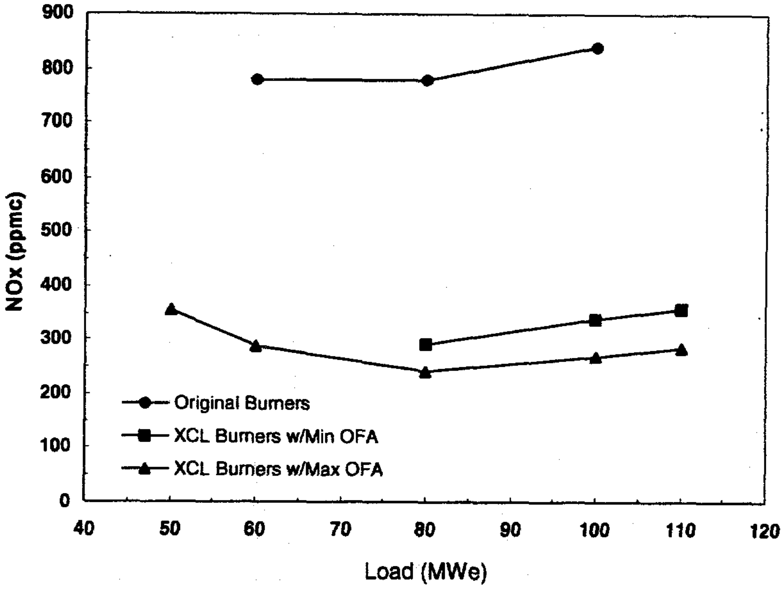


Figure 6. NO_x Emissions Before and After Low-NO_x Combustion System Retrofit

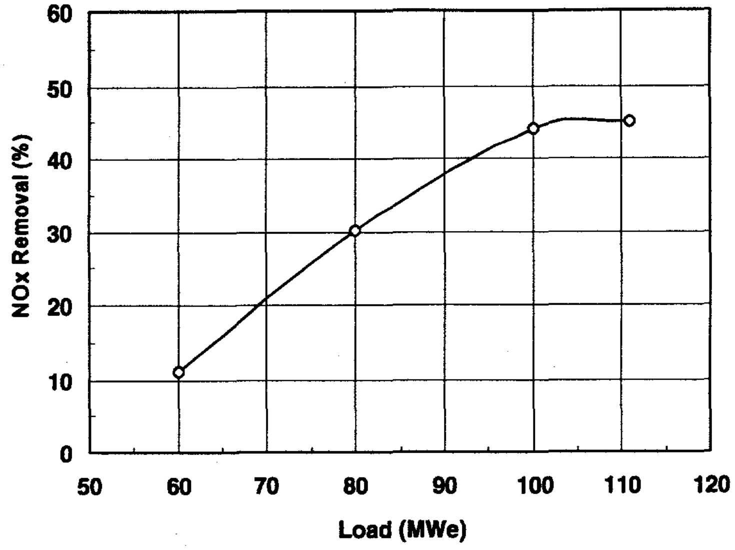


Figure 7. Performance of the Original SNCR System as a Function of Load (10 ppm NH_3 slip limit)

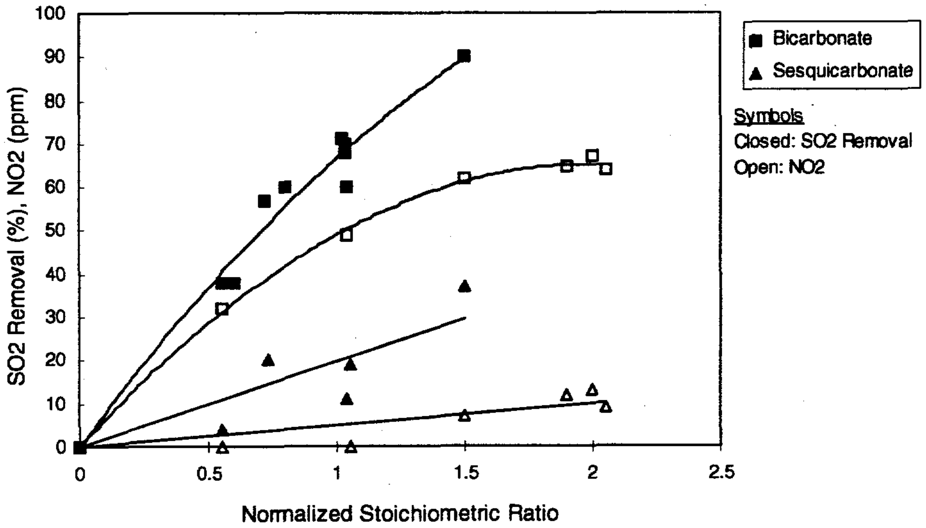
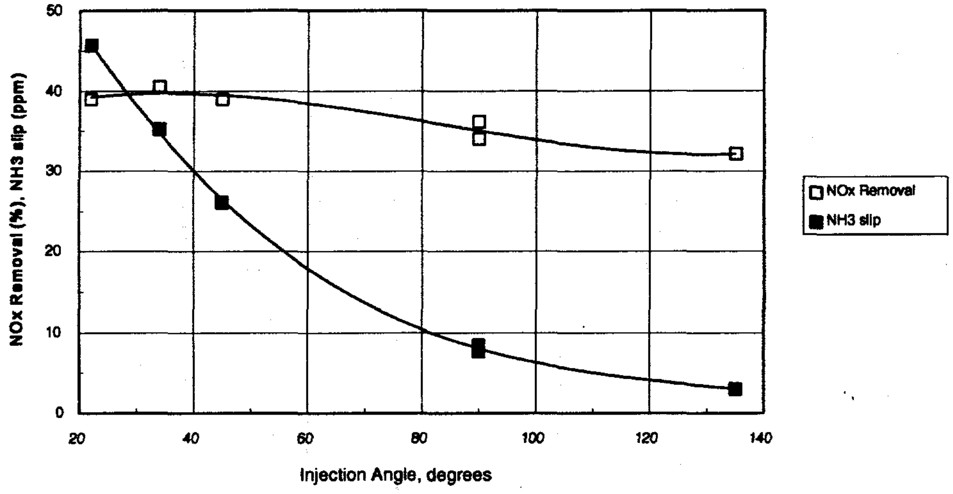
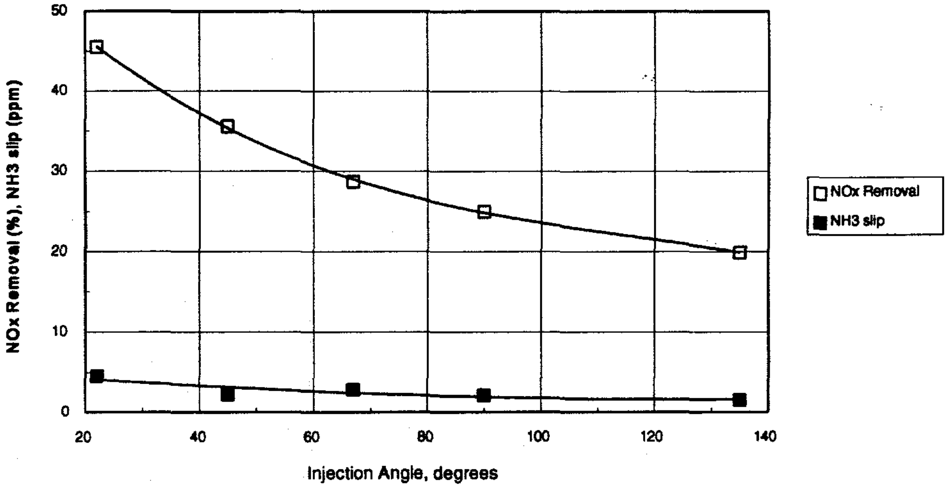


Figure 8. Performance of Sodium Dry Sorbent Injection (sodium sesquicarbonate injected ahead of the fabric filter; sodium bicarbonate injected ahead of the air preheater)

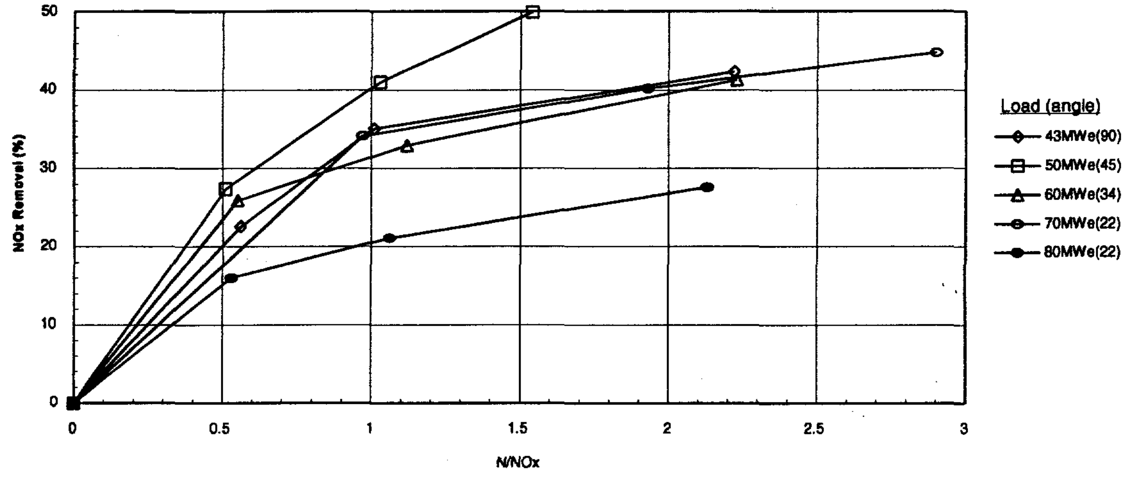


(a) 43 MWe

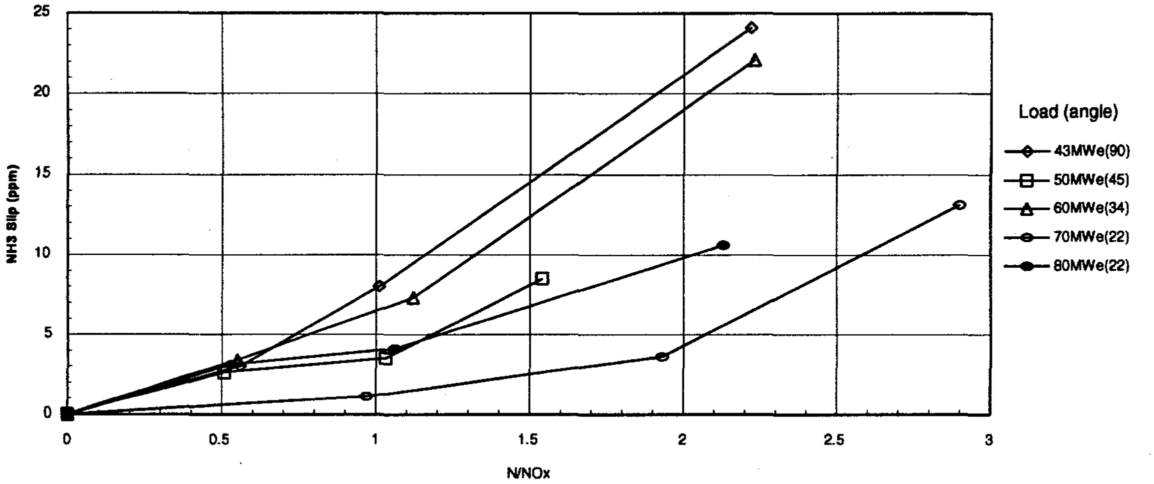


(b) 50 MWe

Figure 9. Effect of Injection Angle on NO_x Removal and NH_3 Slip
(Loads: 43 and 50 MWe, $N/NO_x = 1.0$)



(a) NOx Removal



(b) NH3 slip

Figure 10. ARIL Lance Performance Over the Load Range: 43 to 80 MWe

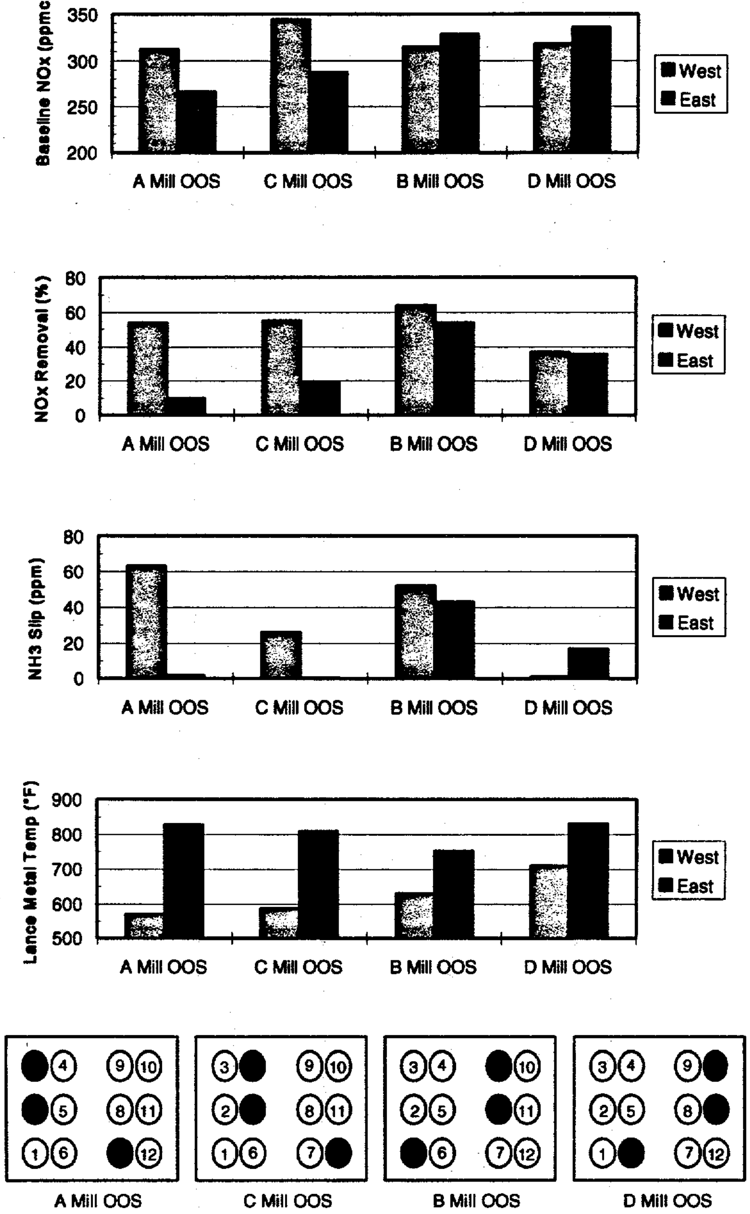


Figure 11. East-West Baseline NO_x Emissions, NO_x Removals, NH_3 Slip Levels and Lance Metal Temperature Distributions as a Function of Mill-in-Service Pattern at 60 MWe ($N/NO_x = 2.0$, 22° Injection Angle)

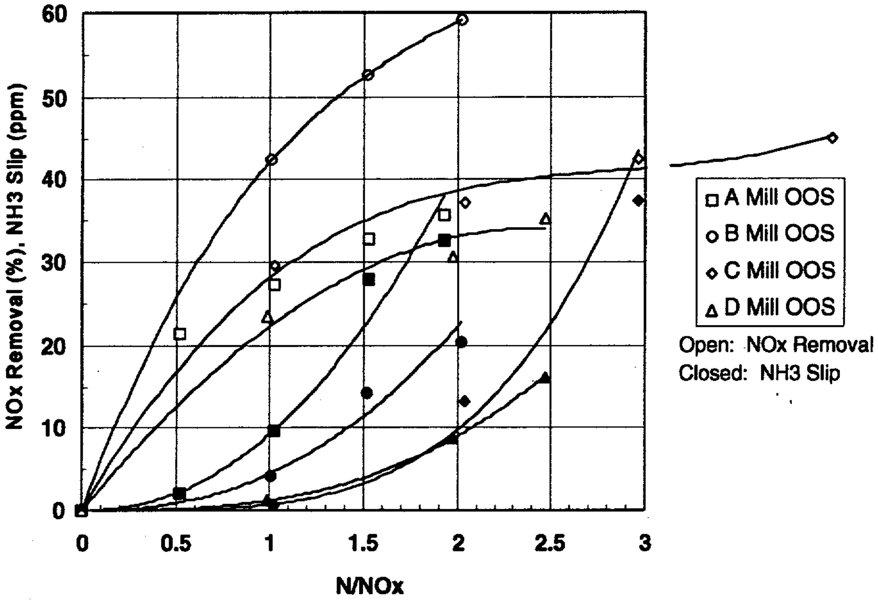


Figure 12. Effect of Mill-in-Service Pattern on ARIL Lance Performance at 60 MWe (22° Injection Angle)

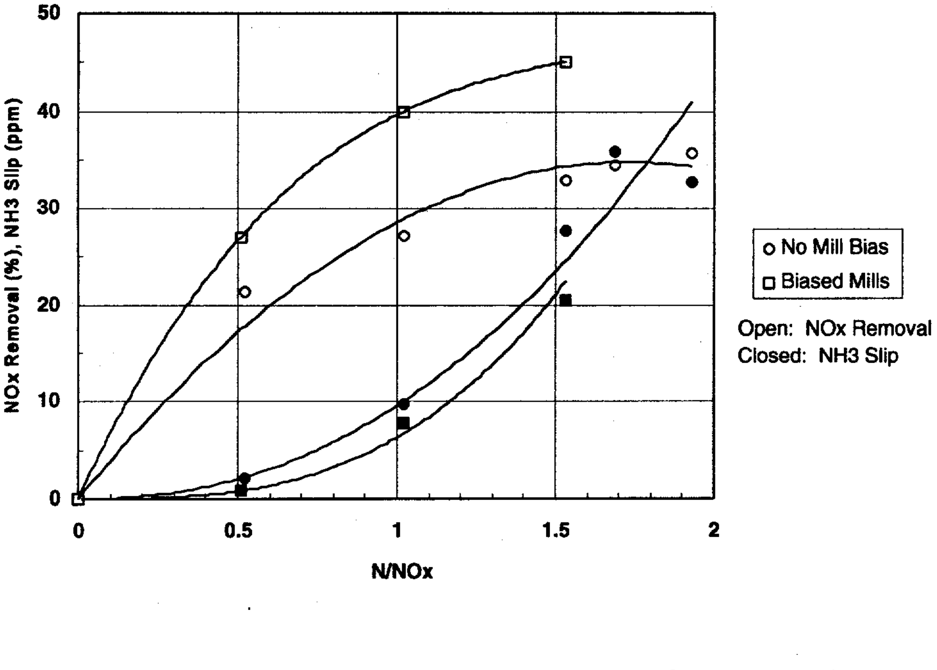


Figure 13. Effect of Coal Mill Bias on ARIL Lance Performance at 60 MWe (A Mill OOS, 22° Injection Angle)

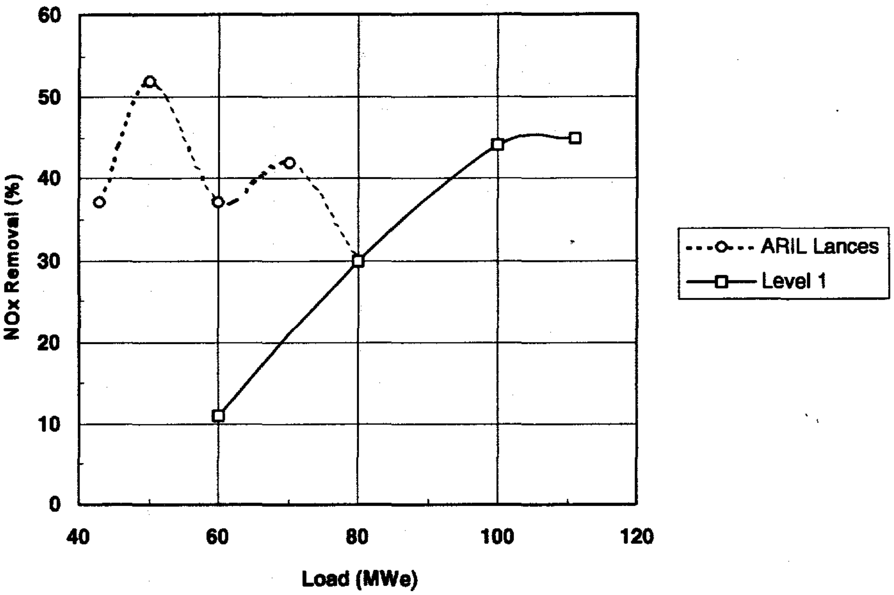


Figure 14. NO_x Removal as a Function of Load for an NH_3 Slip Limit of 10 ppm

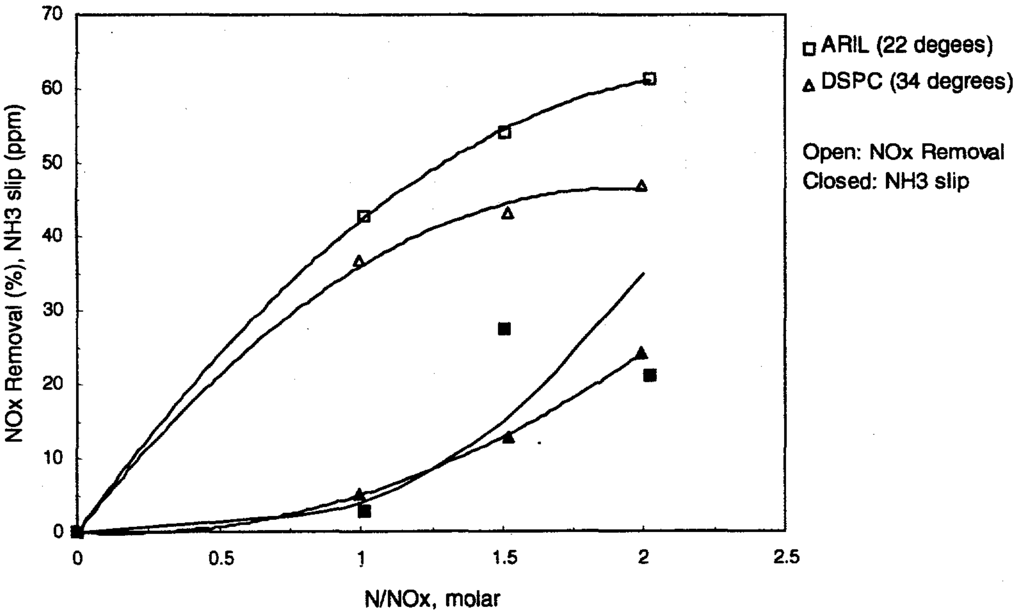


Figure 15. Comparison of the ARIL and DPSC Lance Performance

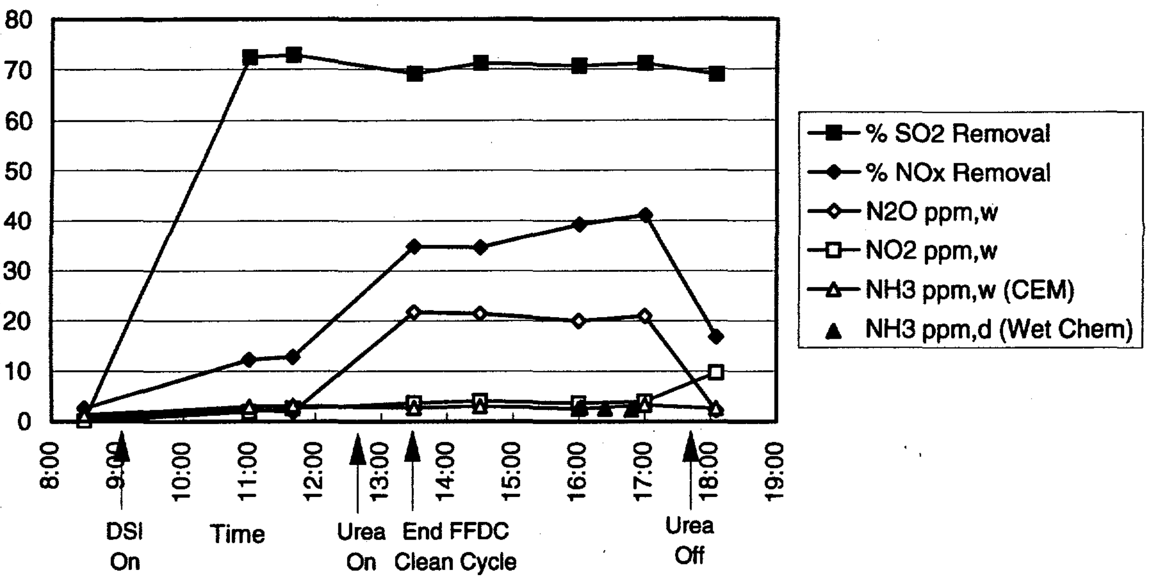


Figure 16. Time History of 100 MWe Integrated Test with Sodium Sesquicarbonate Injection (2Na/S = 2.0, N/NO$_x$ = 0.6, A Mill OOS)

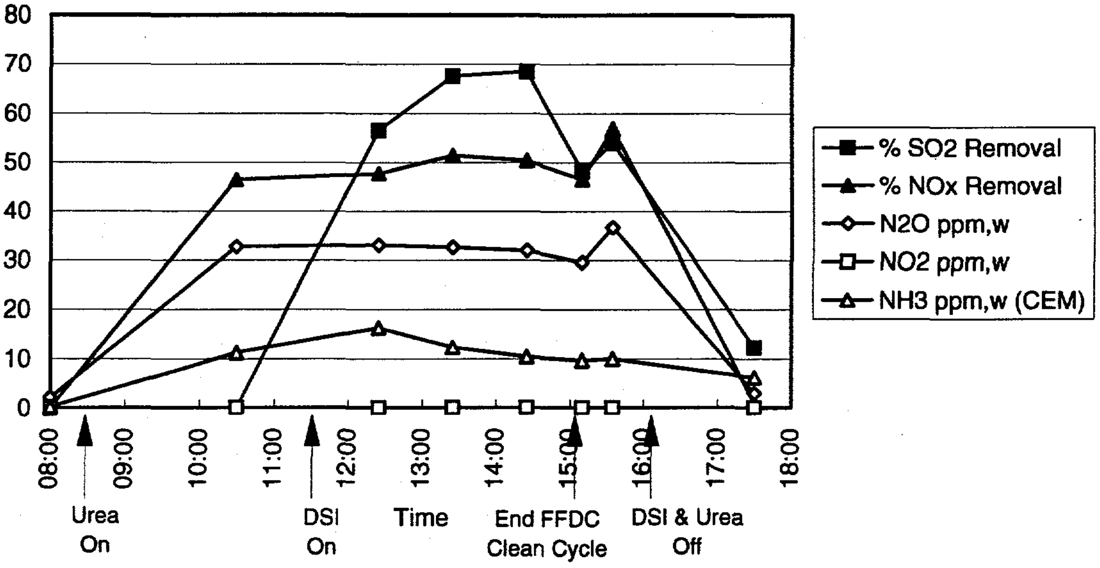


Figure 17. Time History of Integrated Test with Sodium Bicarbonate Injection (2Na/S = 1.1, N/NO$_x$ = 1.1, 100 MWe, 4 Mills in Service)

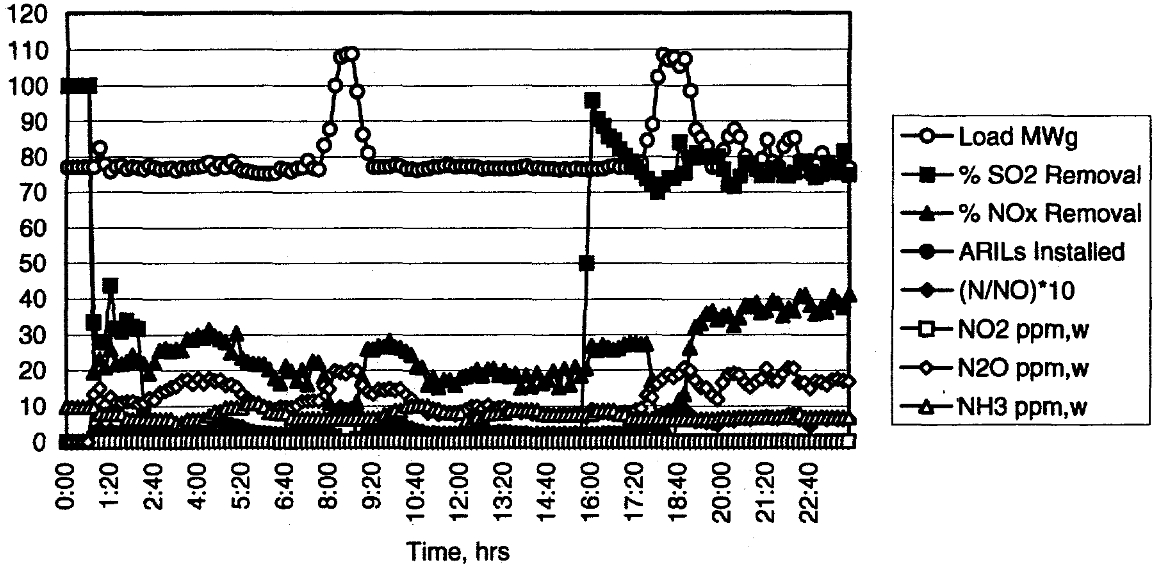


Figure 18. Long-Term Integrated Load-Following Test Results (February 29, 1996)

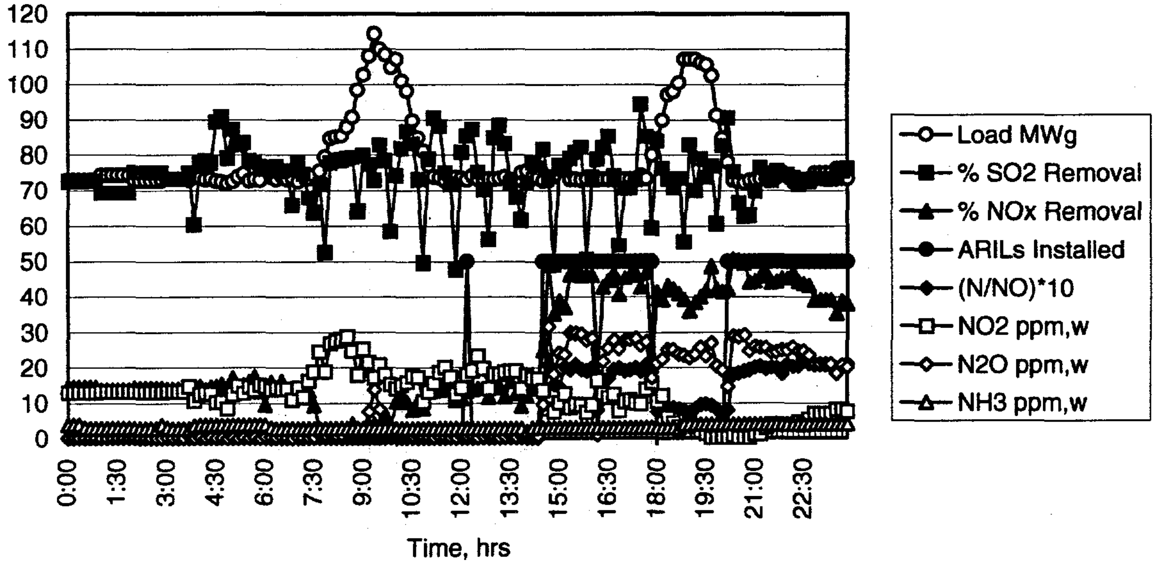


Figure 19. Long-Term Integrated Load Following Test Results (March 4, 1996)

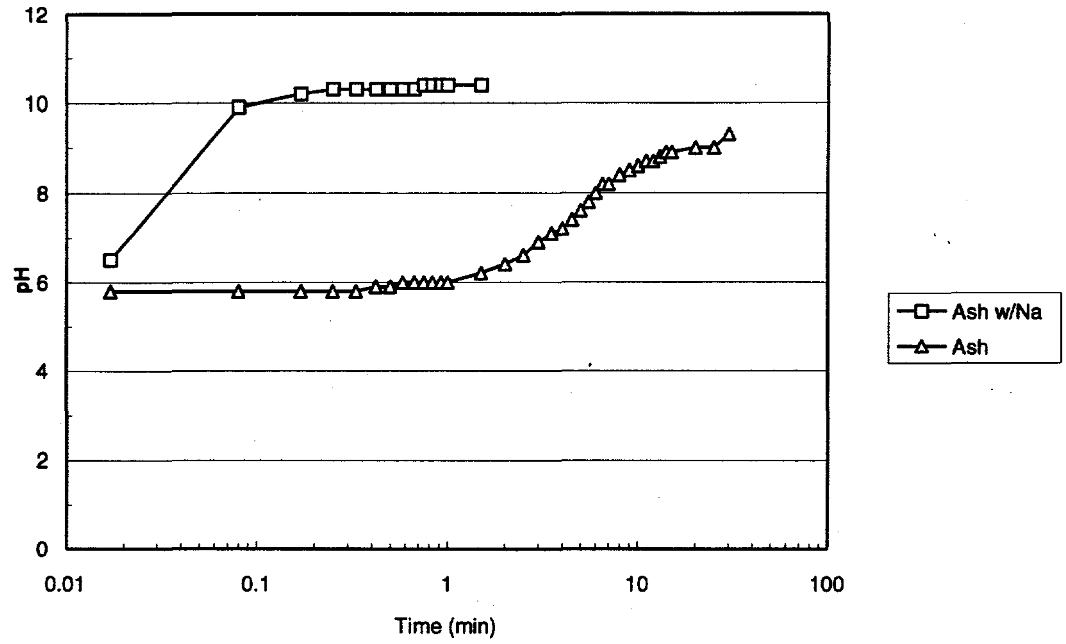


Figure 20. pH versus Time for Coal Ash and Coal Ash/Sodium Mixture
(0.5 grams of ash in 200 ml of H_2O)

PURE AIR'S BAILLY SCRUBBER
A FOUR-YEAR RETROSPECTIVE

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January 7 - 10, 1997
Tampa, Florida

ABSTRACT

Pure Air's Advanced Flue Gas Desulfurization (AFGD) Clean Coal Project has completed four highly successful years of operation at NIPSCO's Bailly Station. As part of their program, Pure Air has concluded a six-part study of system performance. This paper will summarize the results of the demonstration program, including AFGD performance on coals ranging from 2.0 - 4.5% sulfur. The paper will highlight novel aspects of the Bailly facility, including pulverized limestone injection, air rotary sparger for oxidation, wastewater evaporation system and the production of PowerChip® gypsum. Operations and maintenance which have lead to the facility's notable 99.47% availability record will also be discussed. A project company, Pure Air on the Lake Limited Partnership, owns the AFGD facility. Pure Air was the turn key contractor and Air Products and Chemicals, Inc. is the operator of the AFGD system.

Introduction

Pure Air's innovative AFGD project was selected in Round II of the Department of Energy's Clean Coal Technology Program. The project is located at Northern Indiana Public Service Company's (NIPSCO) Bailly Station in Chesterton, Indiana, about 40 miles southeast of Chicago, Illinois. The $151.3 million project involves the retrofit construction and the first three years of operation of a single module AFGD system serving two coal-fired boilers.

On 2 June 1992, the AFGD system began to process flue gas, thus becoming the first commercial scrubber in the country meeting the requirements of the new U.S. Clean Air Act Amendments of 1990 and the first "own and operate" AFGD system in the world. In its first two years of operation, the facility has achieved an availability rate of 99.47% while having the ability to scrub more than 95% of the sulfur dioxide (SO_2) emissions from two coal-fired boilers.

The specific technical objectives of the Pure Air's AFGD demonstration project are the following:

- Effectively demonstrate high reliability and SO_2 removal capability of a large single module absorber serving multiple boilers
- High velocity co-current absorber
- In-situ forced oxidation producing commercial grade gypsum
- Direct dry limestone injection system
- High efficiency air rotary sparger to enhance oxidation
- Wastewater evaporation system (WES)
- "Own and Operate" business arrangement

In addition to demonstrating the above objectives, a series of six testing periods will be performed by Pure Air and NIPSCO over a period of three years to demonstrate the operation of the facility using coal with a wide range of sulfur contents. Each demonstration test will last approximately 5 to 6 weeks and will test coals of specific sulfur content outlined below:

- 2.0 - 2.5% sulfur coal
- 2.5 - 3.0% sulfur coal
- 3.0 - 3.5% sulfur coal
- 3.5 - 4.0% sulfur coal
- 4.0 - 4.5% sulfur coal
- Optimum sulfur coal

The objective of the DOE demonstration test plans are to evaluate the effects of liquid-to-gas (L/G) ratio, calcium-to sulfur ratio and oxidation capability of the air rotary sparger on the overall system performance. This includes SO_2 removal efficiency, slurry composition and by-product gypsum quality.

Plant Description

NIPSCO's Bailly Generating Station consists of two coal-fired boilers; Unit #7 rated at 183 MWG capacity began commercial operation in 1962 and Unit #8 rated at 345 MWG capacity began commercial operation in 1968. Both Units #7 and #8 burn Midwestern high sulfur coal. The AFGD facility includes one resin-lined absorber module to process the flue gas generated from Units #7 and #8 and the required ancillary systems. The absorber is a co-current grid-packed tower with two levels of slurry distribution and an integral reaction tank.

The flue gas generated from Units #7 and #8 is combined and sent to a single co-current absorber. Upon entering the absorber module, the flue gas is saturated by contacting a CO_3 enriched gypsum slurry as it passes through a 3 to 4 meter open-faced grid. The absorber grid provides the required surface area for the flue gas and slurry to react so that greater than 95% of incoming SO_2 is removed in the absorber section of the AFGD system. The cleaned flue gas then passes through a two-stage mist eliminator where liquid and solids droplets are removed prior to exiting the scrubber. *(See Figure 1.)*

The co-current design allows the flue gas and liquid slurry to flow in the same direction, and features a large gas-liquid disengagement zone above the absorber tank. This permits gas velocity through the absorber of up to 20 fps. The use of the high gas velocity co-current absorber allows Pure Air to design one single module for the 528 MWG Bailly Generating Station. There is no spare or back-up module. Instead, the scrubber is designed for a very high level of availability while removing 95% or more of the SO_2, without the use of performance enhancing chemical additives. Therefore, a high degree of system reliability will be demonstrated. The Pure Air design for the Bailly AFGD system uses a non-pressurized slurry distribution system. This system requires approximately 30% less recirculation pump power than conventional counter-current spray towers. Also, since the fountain-like flow does not generate a fine mist, the mist eliminator loading is reduced by as much as 95% compared to counter-current systems.

An important aspect of Bailly's AFGD system is its in-situ forced oxidation using the air rotary sparger (ARS). The ARS has several advantages over the conventional stationary spargers such as higher oxygen utilization, overall reduction of oxidation air and agitation power, and reduced maintenance cost of the oxidation system.

The Bailly Station has very limited space available for the AFGD facility. In addition to the single module, another space-saving feature used in the Bailly AFGD design is a pulverized dry limestone injection system. The direct injection of pulverized limestone eliminates the need for on-site wet grinding systems, thereby reducing space requirements and capital costs. Unlike conventional wet ball mill systems, the dry limestone injection system does not required ball mills, tanks, pumps, and other associated equipment.

Another feature of the Bailly AFGD demonstration involves wastewater treatment. In addition to a conventional wastewater treatment system, Pure Air has provided a wastewater evaporation system (WES) for the Bailly FGD system. In the WES, a portion of the chloride purge stream is sprayed in the ductwork ahead of the Unit #8 precipitator. The purpose of this unique system is to evaporate the high chloride purge stream in the path of the hot flue gas prior to ESP. Calcium chloride solids remain after the water has evaporated and these solids, along with the fly ash, are collected and removed in the precipitator.

In addition to reducing emissions, this AFGD system converts the SO_2 into a high purity (on average between 96% and 97%) commercial grade of synthetic gypsum. This synthetic gypsum is used by United States Gypsum Company to produce wallboard at its East Chicago, Indiana plant. This makes East Chicago the only facility in North America to produce wallboard from 100% FGD gypsum and the sixth U.S. Gypsum plant to use synthetic gypsum on a sustained basis. The sale of this synthetic gypsum is covered under a long term agreement entered in June 1990 between NIPSCO and United States Gypsum Company. Since production began, U.S. Gypsum has taken NIPSCO's entire output of synthetic gypsum and is providing NIPSCO with a reliable alternative to landfill disposal.

At the Bailly Station, Pure Air is also demonstrating a unique gypsum agglomeration process known as PowerChip gypsum. There are many different FGD technologies that produce gypsum. Most make a product similar in consistency to wet sand. This type of gypsum is difficult and time consuming to unload from rail cars or trucks. The material handling problems with conventional FGD gypsum are numerous, as it has poor processing flexibility and contains agglomerates which must continuously be broken and dispersed. It requires an extensive drying process, using expensive, energy consuming drying equipment, before it can be used. For some users, the solution may be to purchase FGD gypsum that has already been dried. A technology from Pure Air is helping utilities produce PowerChip gypsum, a product with all the performance and purity benefits of FGD gypsum and all the handling benefits of natural gypsum. The particle size distribution of milled PowerChip gypsum much more closely resembles that of natural gypsum than does conventional unprocessed FGD gypsum. This makes PowerChip gypsum just as easy to use as natural rock gypsum. PowerChip gypsum is produced in relatively dry, consistently sized chips which will not freeze together in cold weather. It requires no special storage systems, nor expensive drying or conveying equipment.

The PowerChip agglomeration process utilizes a compression mill at an optimum compacting force with an exclusive curing time and temperature relationship that reformulates and modifies the physical structure of FGD by-product gypsum. This technology produces stable, semi-dry, agglomerated flakes of calcium sulfate dihydrate (gypsum) which range between 1/8" to 1/16" in thickness and 3/8" to 1-1/4" in length and width. The production rate of the PowerChip demonstration facility at the Bailly Station is 7 tons per hours. This material can be handled, transported and stored with existing infrastructure designed for natural rock gypsum without costly modifications associated with handling other forms of FGD gypsum.

Bailly AFGD Performance

Maintenance

The key components to achieving the 99.47% availability record over the three years of operation are the operating/maintenance philosophy of the Pure Air Bailly facility coupled with the incorporation of technological modifications. Pure Air embraces a modified reliability centered maintenance philosophy which began in the project design phase. Through critical mode analysis, equipment is identified that, if down, would have the largest impact on availability. In-line spares have been incorporated for these critical pieces of equipment. The operating technicians, who have both operating and maintenance responsibilities, are trained to focus on equipment important to availability, which has no installed spare. Operating technicians use predictive maintenance techniques to determine which equipment needs attention and to what extent. A team then determines the cost of repair versus replacement. Key equipment and spare parts identified and purchased in the design phase using operating history, are on hand locally at the facility. Pure Air also maintains a centrally located spare parts inventory which is available through a mainframe application within the parent company, Air Products and Chemicals. As such, virtually all critical parts and equipment are either on line or readily accessible. As part of the reliability centered maintenance system, a computerized maintenance system is utilized. This tool provides tracking on every piece of equipment and process system in terms of costs and maintenance history. Available to all personnel within the facility, this information is used to facilitate total impact operating and maintenance decisions.

Material Construction

Another major modification to the AFGD system was the implementation of new alloy technology. C-276 alloy over carbon steel clad material replaced existing alloy wallpaper construction within the absorber tower wet/dry interface. The installation of this technology advancement resulted in significant maintenance cost reductions.

WES Modifications

The original WES system consisted of a series of high pressure nozzles spraying wastewater in the path of flue gas ahead of the Unit #8 electrostatic precipitator. After a series of testing with the high pressure nozzles, it was determined that the high pressure nozzle did not perform satisfactorily and excessive amounts of solids accumulated in the duct work. It was decided to replace the high pressure nozzles with two-fluid nozzles, which provide better droplet size distribution and more control of liquid being evaporated. With the installation of a two-fluid nozzle, the WES has been running for the past eight months resulting in zero liquid discharged for Unit #8. Inspection of the WES duct work reveals no solids accumulation or corrosion in the duct work.

DOE Demonstration Test Results

As stated earlier, the performance of the AFGD system will be evaluated over a three year period. A total of five demonstration tests will be conducted (Test I - Test V) during this period. The first three years of this demonstration have been completed and all the DOE tests have been completed. A summary of the completed tests in chronological order is provided below.

DOE Demonstration Test III. The first series of DOE demonstration tests started in August 1992 burning Indiana coal containing approximately 3.2% sulfur coal (*see Table 1*). This is the normal coal for NIPSCO's Bailly Generating Station. A total of 31 days of testing were conducted in a six week period. The AFGD system was tested at 100%, 67% and 33% of it design capacity. The primary variables tested were liquid-to-gas (L/G) ratio, calcium-to-sulfur (stoichiometric) ratio and ARS oxidation capability. The results of the 3.2% sulfur coal test have been previously issued *(2)*.

DOE Demonstration Test IV. This DOE demonstration test started in May 1993 and lasted for approximately six weeks. The coal used for this test had a sulfur content of approximately 3.8% and heating value of 11,000 BTU/lb *(Table 1)*.

A total of 48 tests were conducted during this demonstration. The variables studied during this demonstration were flue gas velocity (boiler load varied from 100% to 33% of its maximum capacity), liquid-to-gas (L/G) ratio, and calcium-to-sulfur (stoichiometric) ratio. The parametric study included the effect of the above variables on SO_2 removal and on gypsum quality.

Stoichiometric Ratio (SR): For wet limestone scrubbers, SR is defined as moles of total calcium (or calcium carbonate) fed to the FGD system per moles of SO_2 removed. To evaluate the effect of SR in the system performance, the absorber calcium carbonate level was varied from 50 to 130 mmole/l while maintaining 75% and 94% of design recirculation flow at 100% boiler load *(Figure 2)*. As illustrated in this figure, the SO_2 removal efficiency increases from 94% to 97% by increasing SR from 1.03 to 1.08 moles/mole at 94% of design recirculation rate. The effect of SR on SO_2 removal efficiency was also tested at 67% and 33% boiler load. The results of the testing at 67% boiler load are presented in *Figure 3*. As illustrated in this figure, 98% SO_2 removal was achieved at SR of 1.065 while operating with only 92% of design L/G.

Liquid-to-Gas (L/G) Ratio: The L/G was fluctuated by varying the number of the operating recirculation pumps. At 100% load, the recirculation flow was varied from 75% to 94% of total liquid flow while maintaining relatively constant stoichiometric ratio of 1.03 and 1.08 moles/mole (moles of calcium per mole of SO_2 removed).

As expected, SO_2 removal increased with increasing recirculation flow rate (L/G). For example, at 100% boiler load and stoichiometric ratio of 1.03, SO_2 removal increased from about 90.5% to 94% by increasing absorber recirculation flow from 75% to 94% of its design value *(Figure 4)*. The performance of the system is further improved from about 94% to 97% at higher SR (1.08) over the same range of L/G. Much better system performance was achieved under similar conditions at reduced (67%) boiler load *(Figure 5)*.

Gypsum Quality: The gypsum purity during the 3.8% sulfur coal demonstration test period averaged 96.7%. All other gypsum parameters are presented in *Table 2*.

DOE Demonstration Test V. This test consisted of NIPSCO burning coal with greater than 4.5% sulfur for a period of at least 100 hours. During this period, the boiler was kept under full load (100%) conditions. The L/G and the SR were set so that the system would maintain at least 95% pure gypsum. This DOE demonstration test started in June 1994 and lasted for 16 days. The coal used for this test had a sulfur content of approximately 4.7% and a heating value of 12,700 BTU/lb *(Table 1)*. The variables studied during this demonstration were calcium-to-sulfur (stoichiometric) ratio and liquid-to-gas (L/G) ratio.

Stoichiometric Ratio (SR): To evaluate the effect of SR on the system performance, the absorber calcium carbonate level was varied from 50 to 130 mmole/l while maintaining 75% and 93% of design recirculation flow *(Figure 6)*. As indicated in this figure, 95% of SO_2 removal efficiency was achieved at SR of 1.05 (moles/mole) and 93% of design L/G.

Liquid to Gas (L/G) Ratio: During the 4.7% sulfur coal and 100% boiler load test, data was collected to evaluate the system performance at constant stoichiometric ratio (1.035 and 1.05) while varying the recirculation flow rate from 75% to 100% of its design rate. Again, as expected, SO_2 removal increased with increasing L/G. With 4.7% sulfur coal and a stoichiometric ratio of 1.05, SO_2 removal efficiency increased form 92.5% to 96% by increasing absorber recirculation flow from 75% to 100% of its design rate *(Figure 7)*.

Gypsum Quality: Similar to 3.8% sulfur coal, very high gypsum purity was produced during the 4.7% sulfur coal testing *(Table 2)*.

DOE Demonstration Test I. This demonstration test started in July 1994 and lasted for approximately 6 weeks. The coal used for this test had a sulfur content of approximately 2.25% and a heating value of 11,932 BTU/lb *(Table 1)*.

A total of 42 tests were conducted during the demonstration, which consisted of 20 tests at full boiler load, 15 tests at 67% boiler load and 8 tests at 33% boiler load. The variables studied during this demonstration test were flue gas velocity, liquid-to-gas (L/G) ratio, calcium-to-sulfur ratio (stoichiometric) ratio and their effect on gypsum quality.

Stoichiometric Ratio (SR): To determine the effect of SR on SO_2 removal efficiency, the absorber calcium carbonate level was varied from 24 to 90 mmole/l (corresponding to SR of 1.015 to 1.056 moles of calcium carbonate fed per mole of SO_2 removed) while maintaining 70% and 85% of design recirculation flow rate at 100% boiler load *(Figure 8)*. As illustrated in this figure, SO_2 removal efficiency increased with increasing absorber SR; in fact, the AFGD system could easily remove 97% of the inlet SO_2 at stoichiometric ratio of 1.055 with only 85% of design recirculation flow rate. 92% SO_2 removal efficiency is achievable with SR of only about 1.015 moles/mole at the same recirculation flow rate.

As expected, reducing the gas velocity from 100% to 67% of boiler load improves the system performance *(Figure 9)*. For example, increasing the system SR from 1.014 to 1.03 at 84% of design L/G, resulted in increasing SO_2 removal from 94% to 97% at 67% boiler load.

Liquid-to-Gas (L/G) Ratio: The effect of the L/G on SO_2 removal was tested by varying the number of the operating recirculation pumps while maintaining constant stoichiometric ratio. With 2.25% sulfur coal, the recirculation flow rate was varied from 70% to 85% of the design flow rate at SR of 1.03 and 1.053 moles/mole at 100% boiler load *(Figure 10)* and at SR of 1.014 and 1.03 moles/mole at 67% boiler load *(Figure 11)*. As presented in *Figures 10 and 11*, SO_2 removal efficiency increased with increasing L/G at constant SR. For example, increasing recirculation rate from 70% to 85% at the SR of 1.03 increased the SO_2 removal from 91% to 94% at 100% boiler load, and from 95% to 97% at 67% boiler load.

Gypsum Quality: The AFGD system produced very high quality gypsum *(Table 2)* during the 2.25% sulfur coal testing.

DOE Demonstration Test II. This demonstration test started in September 1994 and lasted for approximately four weeks. The coal used for this test was a blend of Captain coal containing 3.2% sulfur and Credero mined coal containing 2.25%. This blended coal had a sulfur content of approximately 2.75% *(Table 1)*.

A total of 29 tests were conducted during this demonstration, which consisted of 14 tests at 100% boiler load, 8 tests at 67% boiler load, and 7 tests at 33% boiler load. The variables studied during this demonstration test were flue gas velocity, liquid-to-gas (L/G) ratio, calcium-to-sulfur (stoichiometric) ratio and their effect on gypsum quality.

Stoichiometric Ratio (SR): To evaluate the effect of SR on SO_2 removal efficiency, the absorber calcium carbonate level was varied from 25 to 100 mmole/l (corresponding to SR of 1.015 to 1.065 moles of calcium carbonate fed per mole of SO_2 removed) while maintaining 80%, 88% and 100% of design recirculation flow rate at 100% boiler load *(Figure 12)*. As illustrated in this figure, SO_2 removal efficiency increases with increasing stoichiometric ratio. For example, at 100% of design recirculation flow rate, increasing the absorber SR from 1.015 to 1.065 resulted in increased SO_2 removal from 94% to 96%.

As expected, reducing the gas velocity from 100% to 67% improves the system performance *(Figure 13)*. For example, at 67% boiler load, increasing the SR from 1.02 to 1.04 at 88% of design L/G, resulted in increasing SO_2 removal from 96.5% to 97.5%.

Liquid-to-Gas (L/G) Ratio: The effect of the L/G on SO_2 removal was tested by varying the number of the operating recirculation pumps while maintaining constant stoichiometric ratio with 2.75% sulfur coal. The recirculation flow was varied from 80% to 100% of the design flow rate at SR of 1.015, 1.03, 1.04 and 1.055 moles/mole at 100% boiler load *(Figure 14)* and at SR of 1.025 and 1.04 moles/mole at 67% boiler load *(Figure 15)*. Again as illustrated in *Figures 14 and 15*, SO_2 removal efficiency increased with increasing L/G at constant SR.

Gypsum Quality: The AFGD system produced very high quality gypsum *(Table 2)* during the 2.75% sulfur coal testing.

DOE Demonstration Test VI. This demonstration test started in March 1995 and continued until May 1995. The overall objective of this demonstration test was to evaluate the system performance while the boilers burning coal with an optimum level of sulfur for a period of sixty days. The optimum level of sulfur was identified as a result of the previous testing.

The coal used for this test had a sulfur content of approximately 3% sulfur and heating value of 10,970 Btu/lb (Table 1). A total of 52 tests were conducted during this demonstration and all of the tests were operated at full boiler load. The variables studied during this demonstration test were liquid-to-gas (L/G) ratio, calcuim-to-sulfur (stoichiometric) ratio and their effect on gypsum quality.

Stoichiometric Ratio (SR): To evaluate the effect of SR on the system performance, the absorber calcium carbonate level was varied from 40 to 116 mmol/l (corresponding to SR of 1.025 to 1.078) while maintaining 84%, 93% and 96% of design recirculation flow rate at 100% boiler load (Figure 16). As illustrated in this figure, the SO2 removal of efficiency increases from 94% to 96% by increasing SR from 1.03 to 1.06.

Liquid-to-Gas (L/G) Ratio: At 100% load, the recirculation flow was varied from 65% to 93% while maintaining relatively constant stoichiometric ratio of 1.03, 1.045 and 1.065 (moles of calcium per moles of SO2 removed). As expected, SO2 removal increased with increasing recirculation flow rate. For example, at 100% load and SR of 1.065, SO2 removal increased from 94.5% to 96% by increasing absorber recirculation from 60% to 100% of its design value (Figure 17).

Gypsum Quality: The AFGD system produced very high quality gypsum (Table 2) during the optimum sulfur level coal testing.

Summary of Results

The demonstration of the AFGD system at the Bailly Generating facility has established this technology as an efficient and reliable means of removing SO_2. The overall system SO_2 removal efficiency results from the first four DOE demonstration tests have been analyzed and several relationships have been determined. First, at constant stoichiometric ratio, SO_2 removal efficiency increases with the absorber recirculation rate *(Figure 18)*.

The rate increase in SO_2 removal efficiency is highest at low sulfur coal (2.25% sulfur) and decreases as coal sulfur content increases to 4.5%. Next, at moderately low calcium stoichiometric ratio, very high levels of SO_2 removal are achieved at two thirds of design L/G ratio *(Figure 19)*. As illustrated in this figure, the AFGD system is capable of achieving higher than 95% reduction in SO_2 at 76% of design L/G for all sulfur coal tested.

To date, operations have gone well. The scrubber has already exceeded its target of demonstrating 95+% SO_2 removal efficiency, while producing a commercial grade gypsum by-product and meeting all system operating and contractual requirements *(Tables 3 through 5)*.

FIGURE 1

Advanced Flue Gas Desulfurization Process Flow Diagram

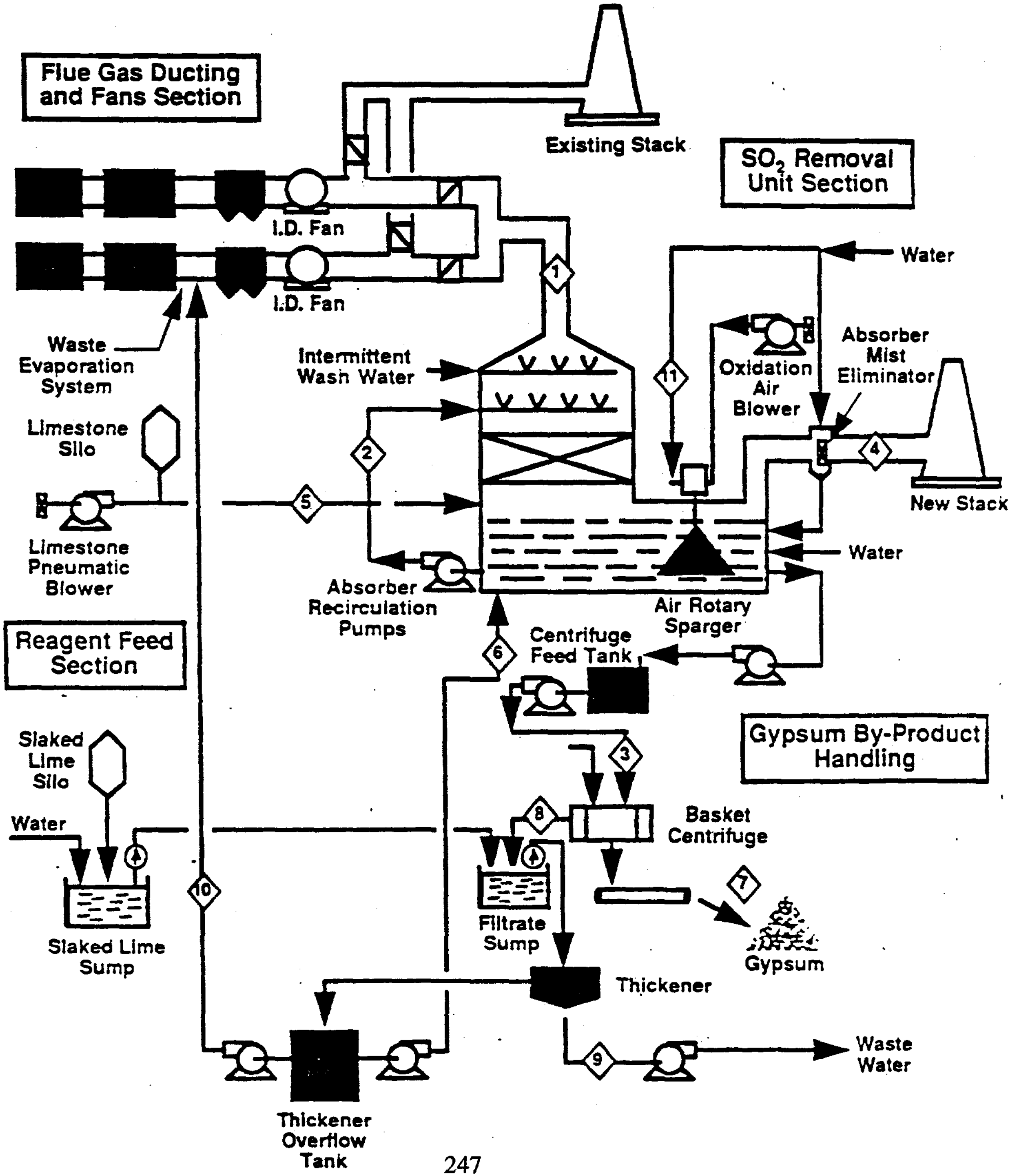

FIGURE 2

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

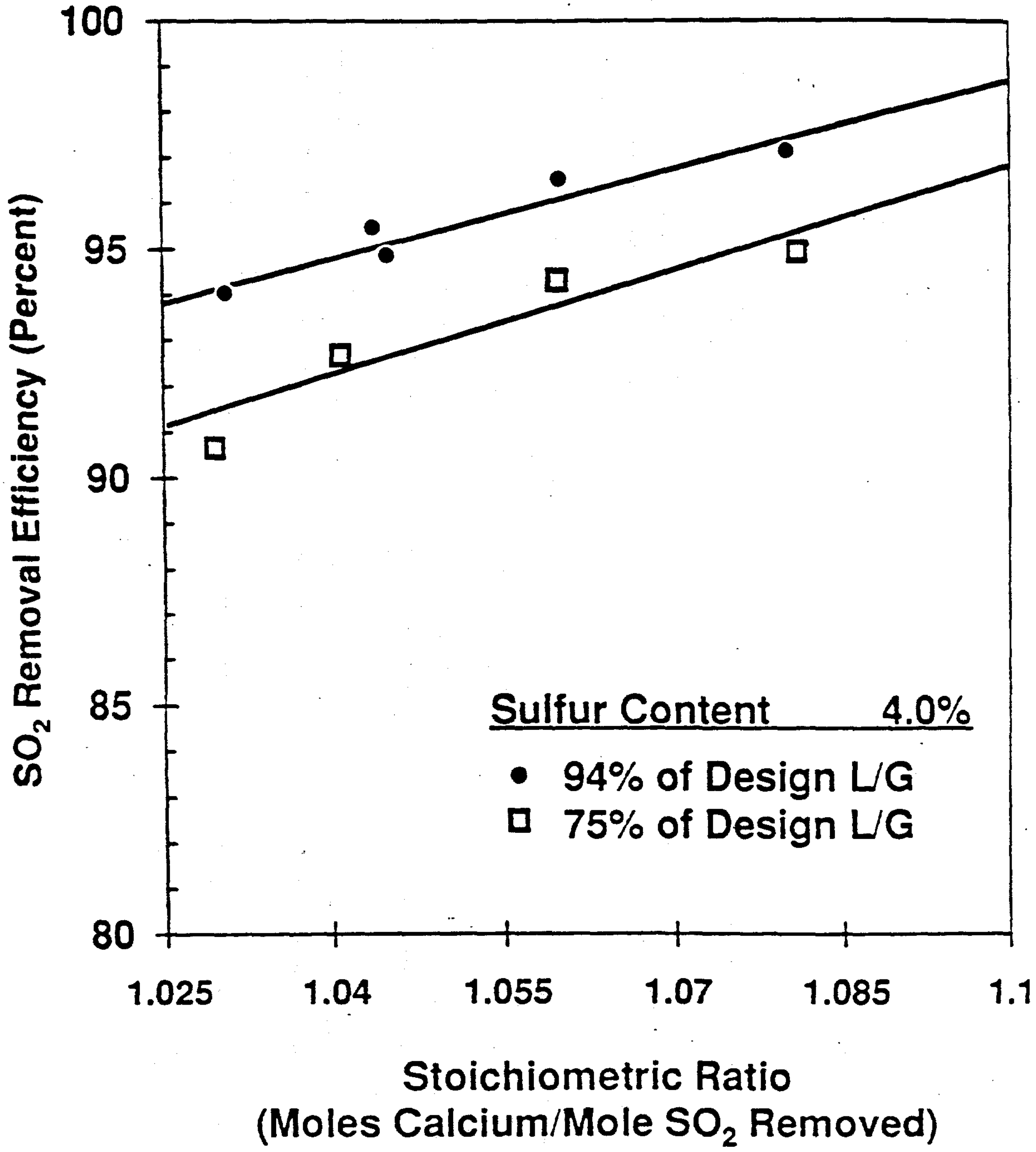

FIGURE 3

SO_2 Removal Performance at Bailly AFGD

(67 Percent Boiler Load)

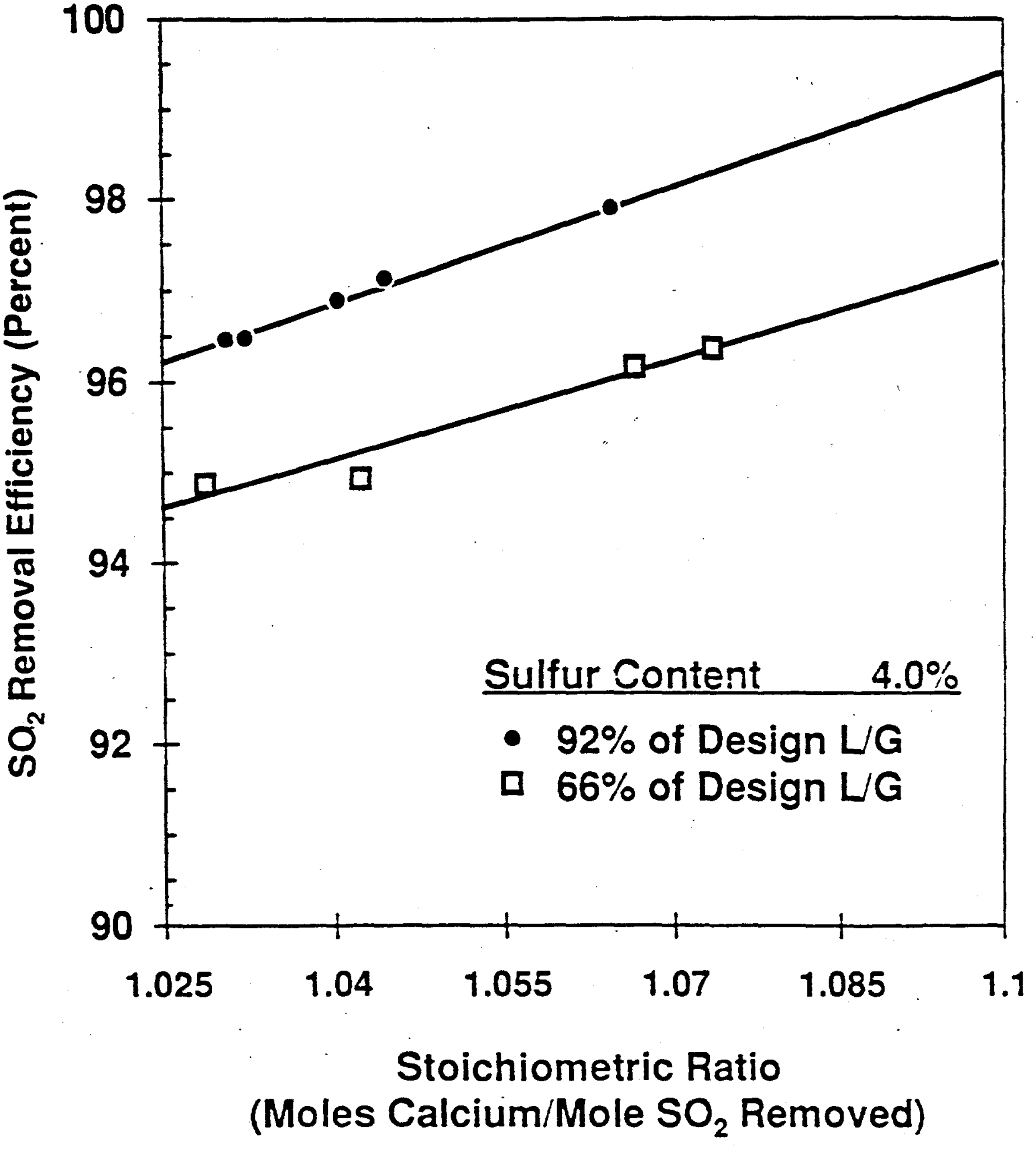

FIGURE 4

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

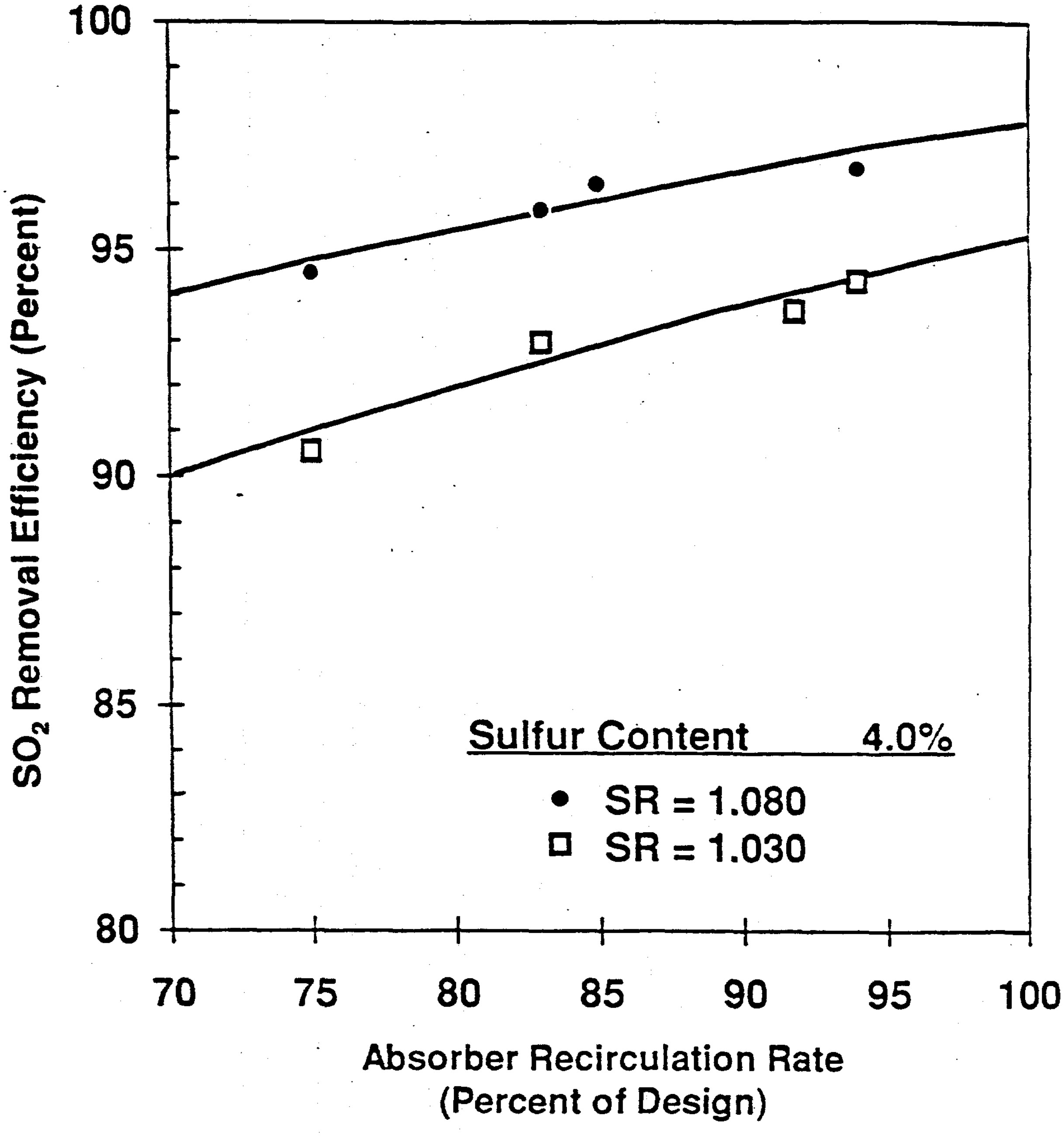

FIGURE 5

SO_2 Removal Performance at Bailly AFGD

(67 Percent Boiler Load)

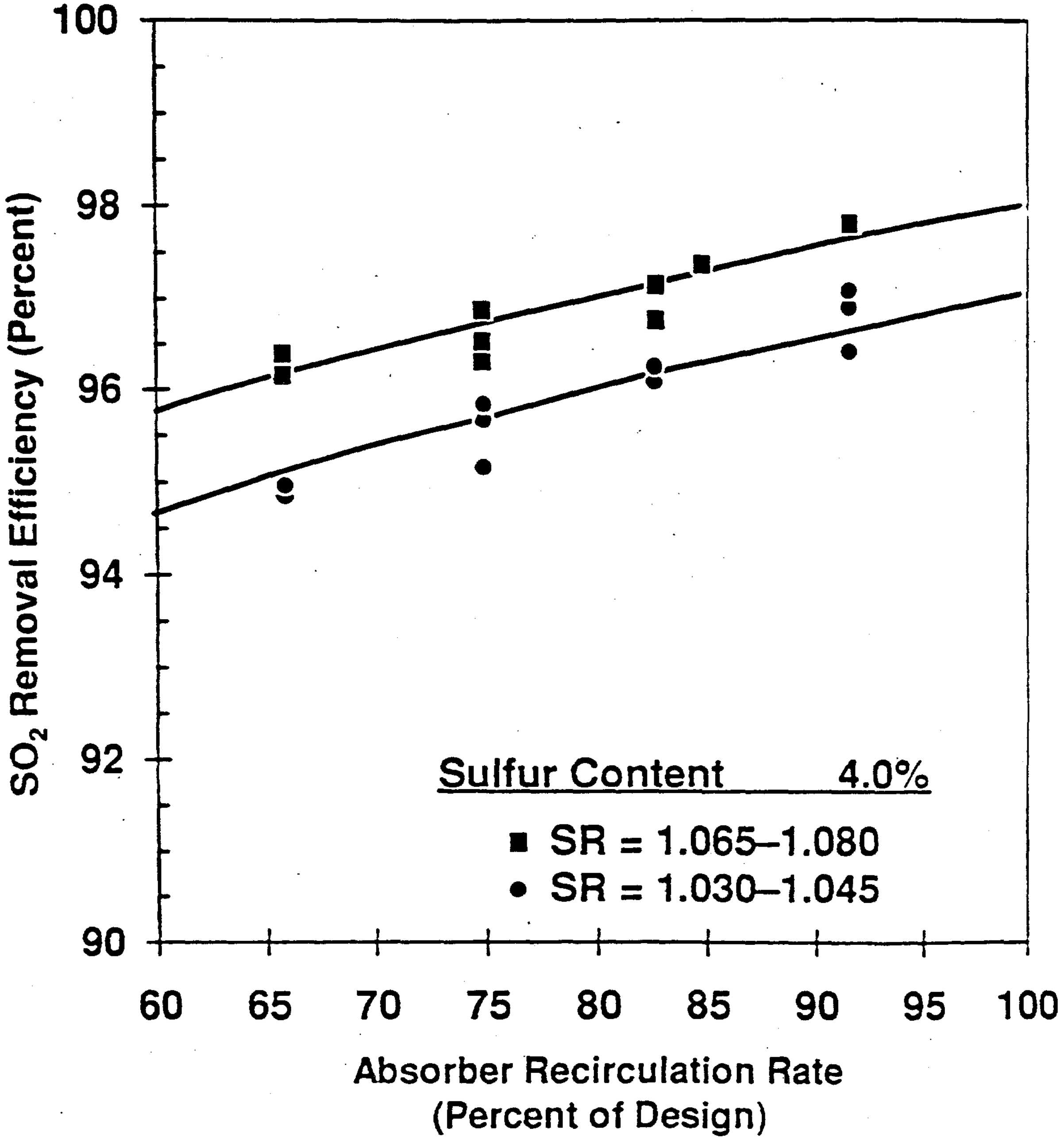

FIGURE 6

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

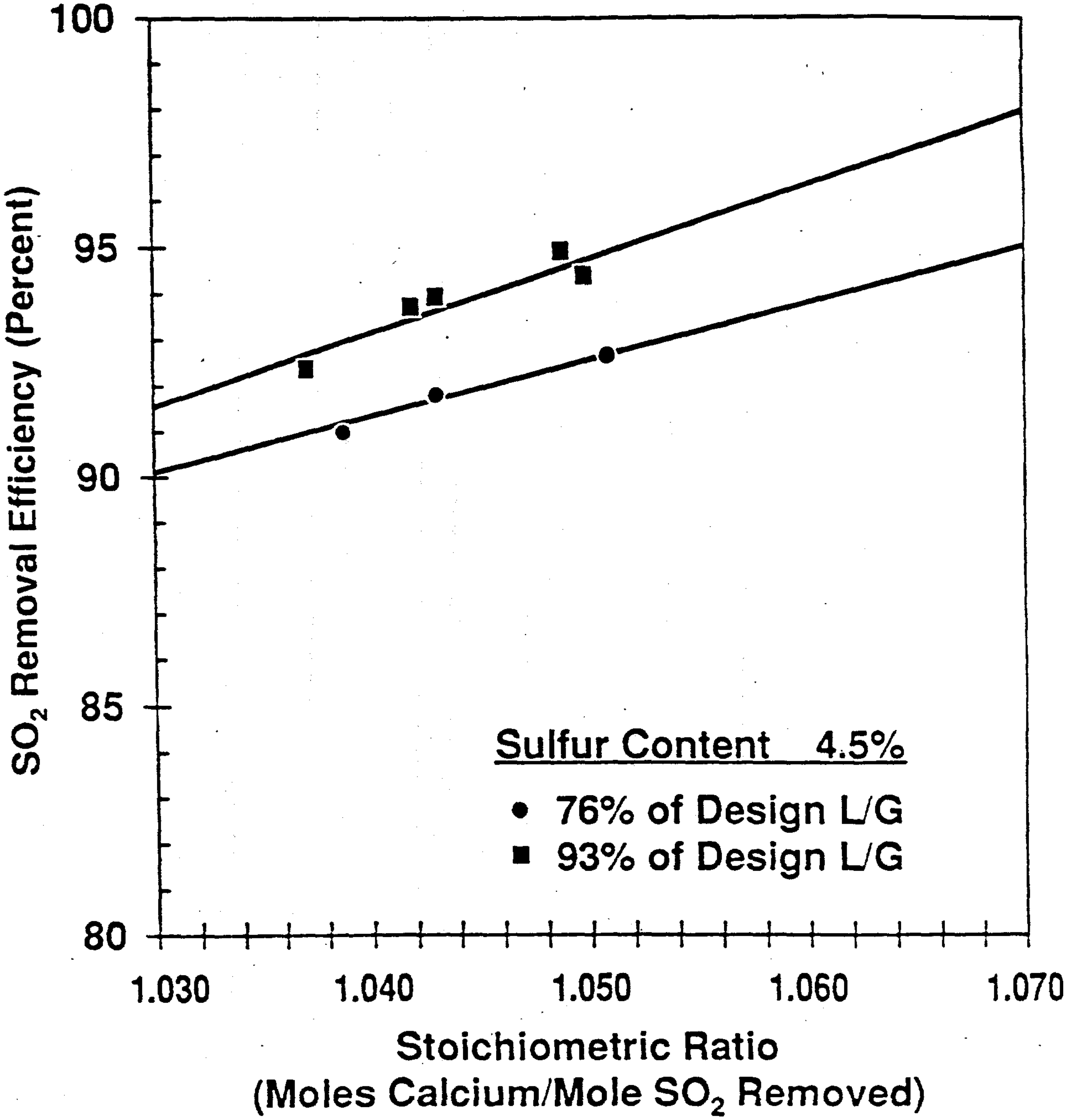

FIGURE 7

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

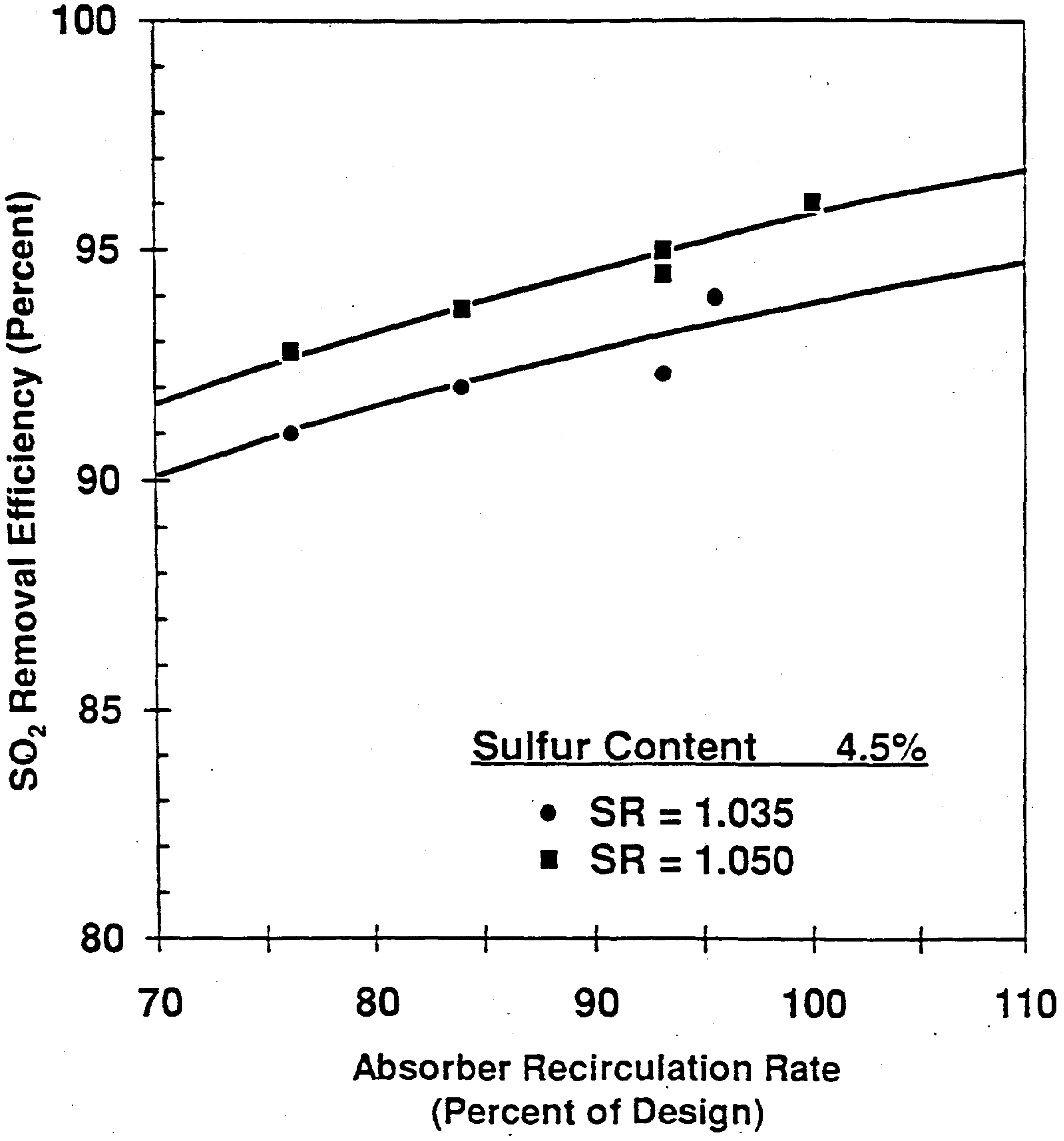

FIGURE 8

SO_2 Removal Performance at Bailly AFGD
(100 Percent Boiler Load)

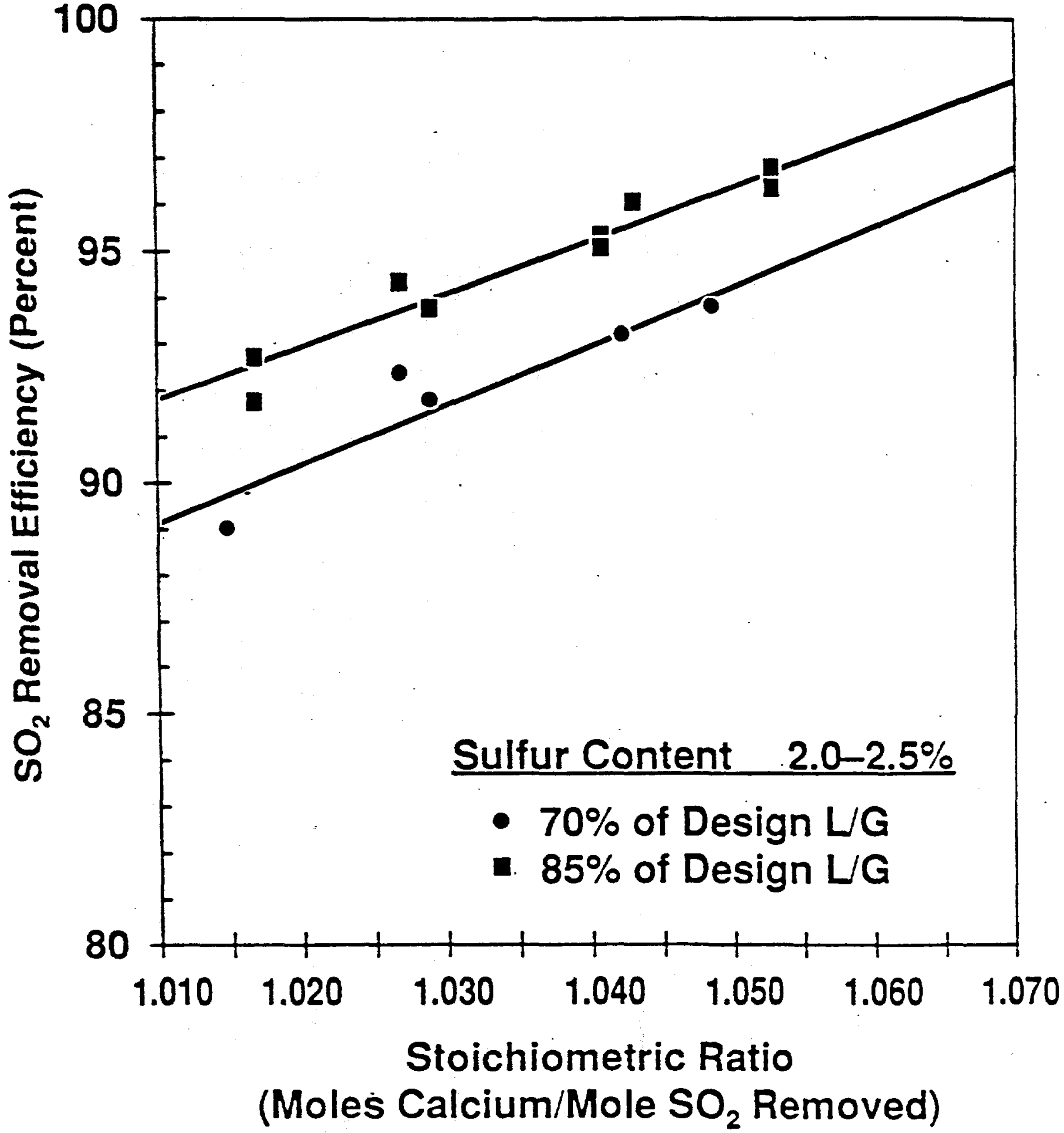

FIGURE 9

SO_2 Removal Performance at Bailly AFGD

(67 Percent Boiler Load)

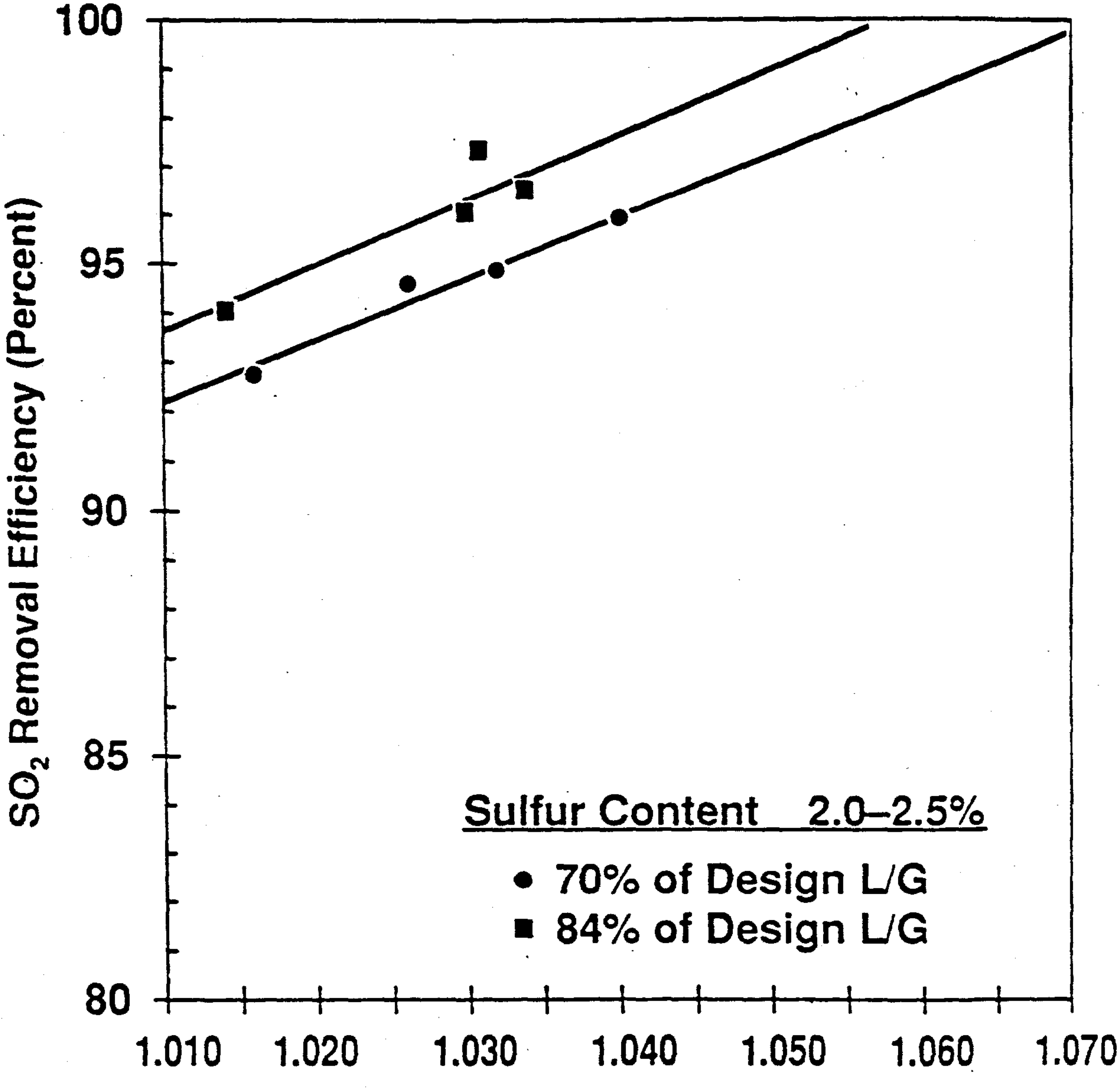

FIGURE 10

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

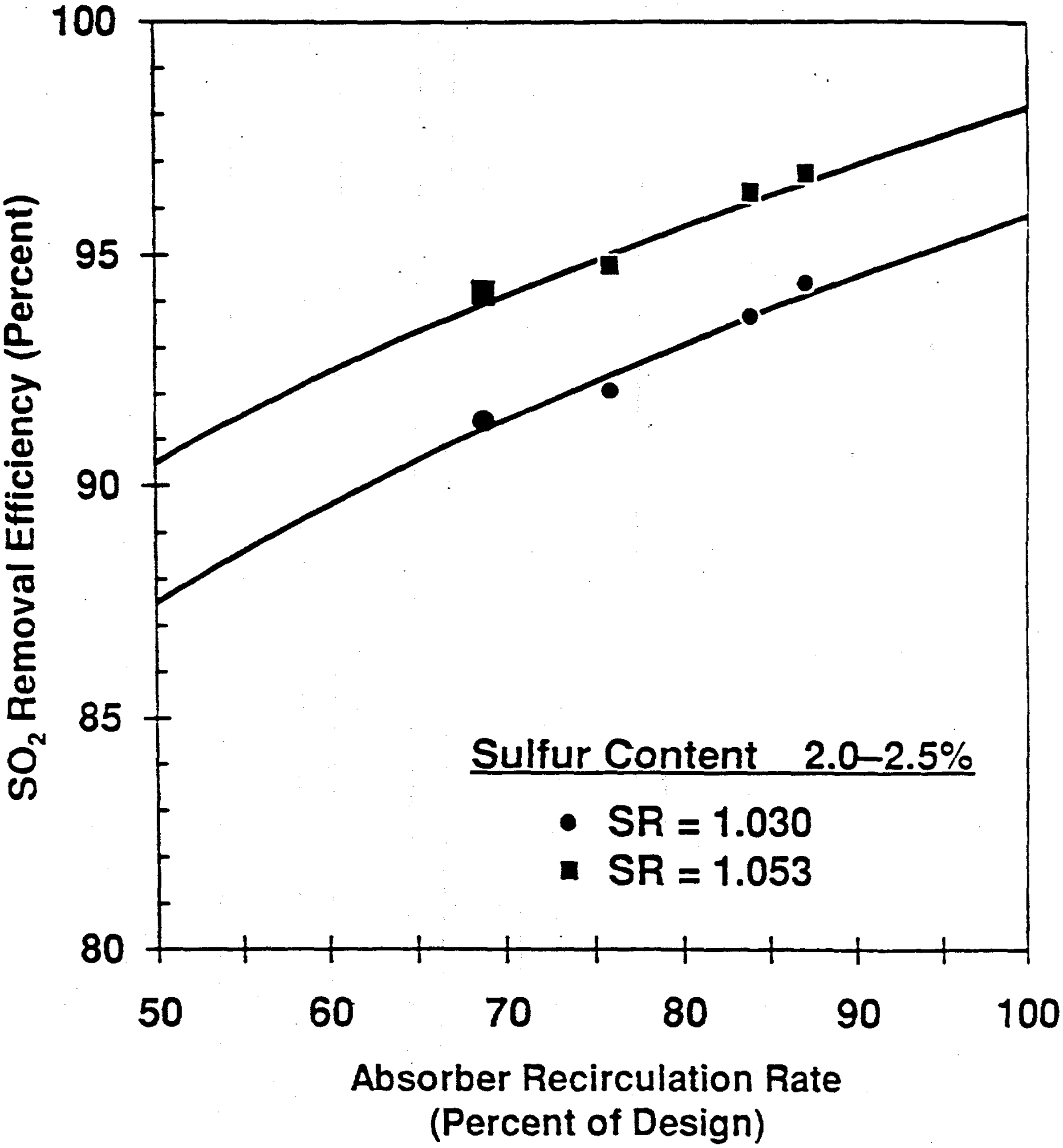

FIGURE 11

SO_2 Removal Performance at Bailly AFGD

(67 Percent Boiler Load)

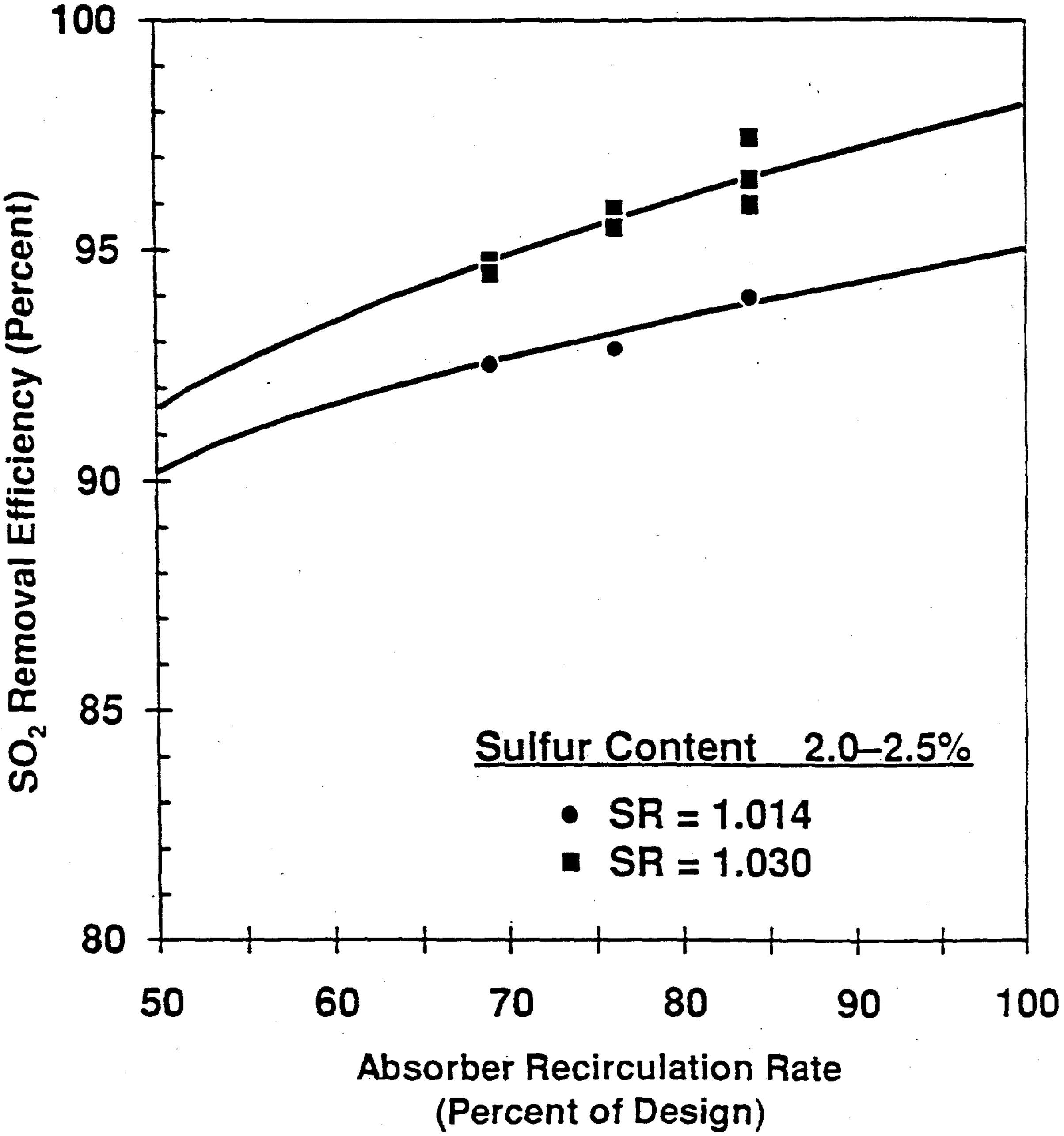

FIGURE 12

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

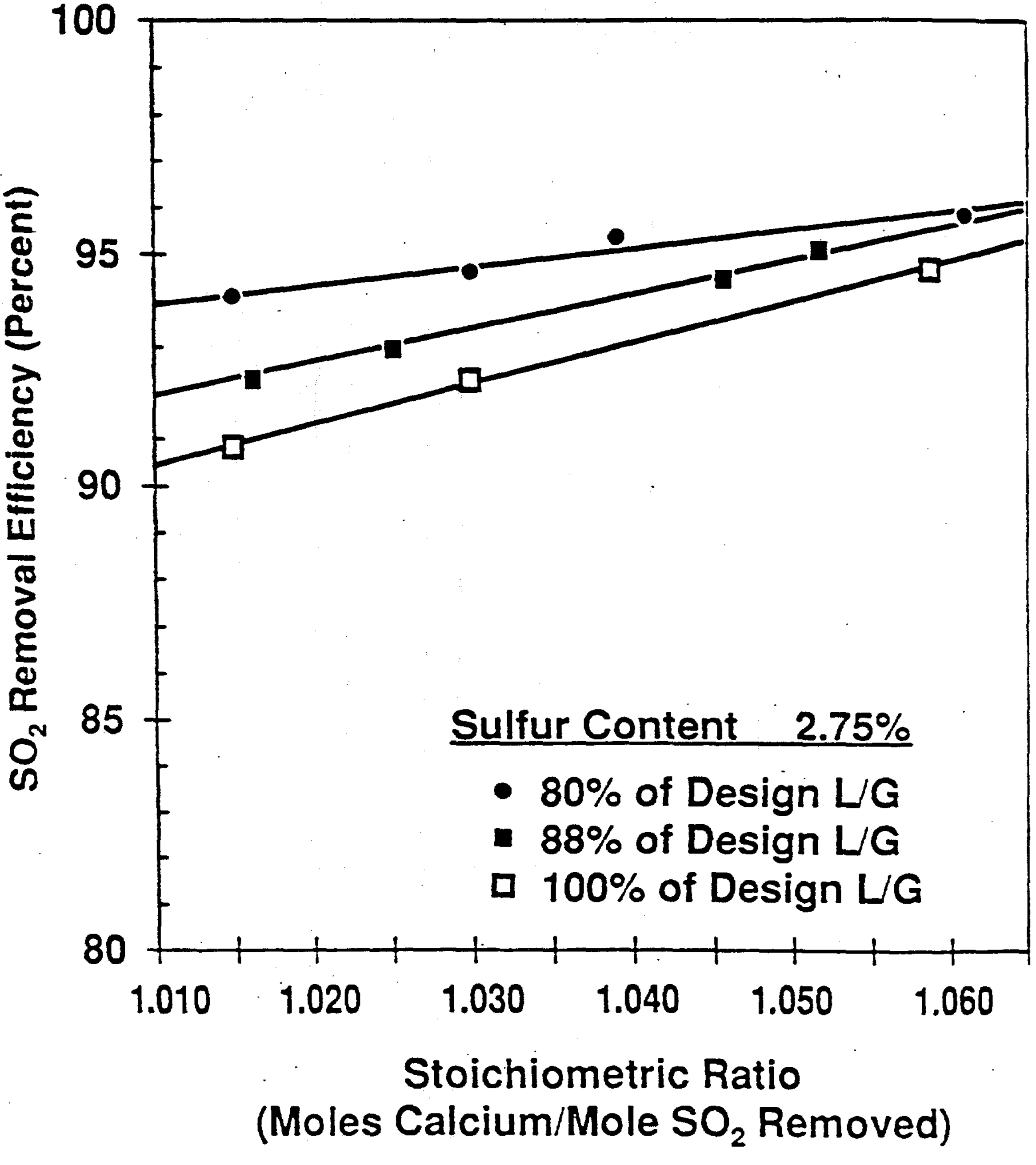

FIGURE 13

SO_2 Removal Performance at Bailly AFGD

(67 Percent Boiler Load)

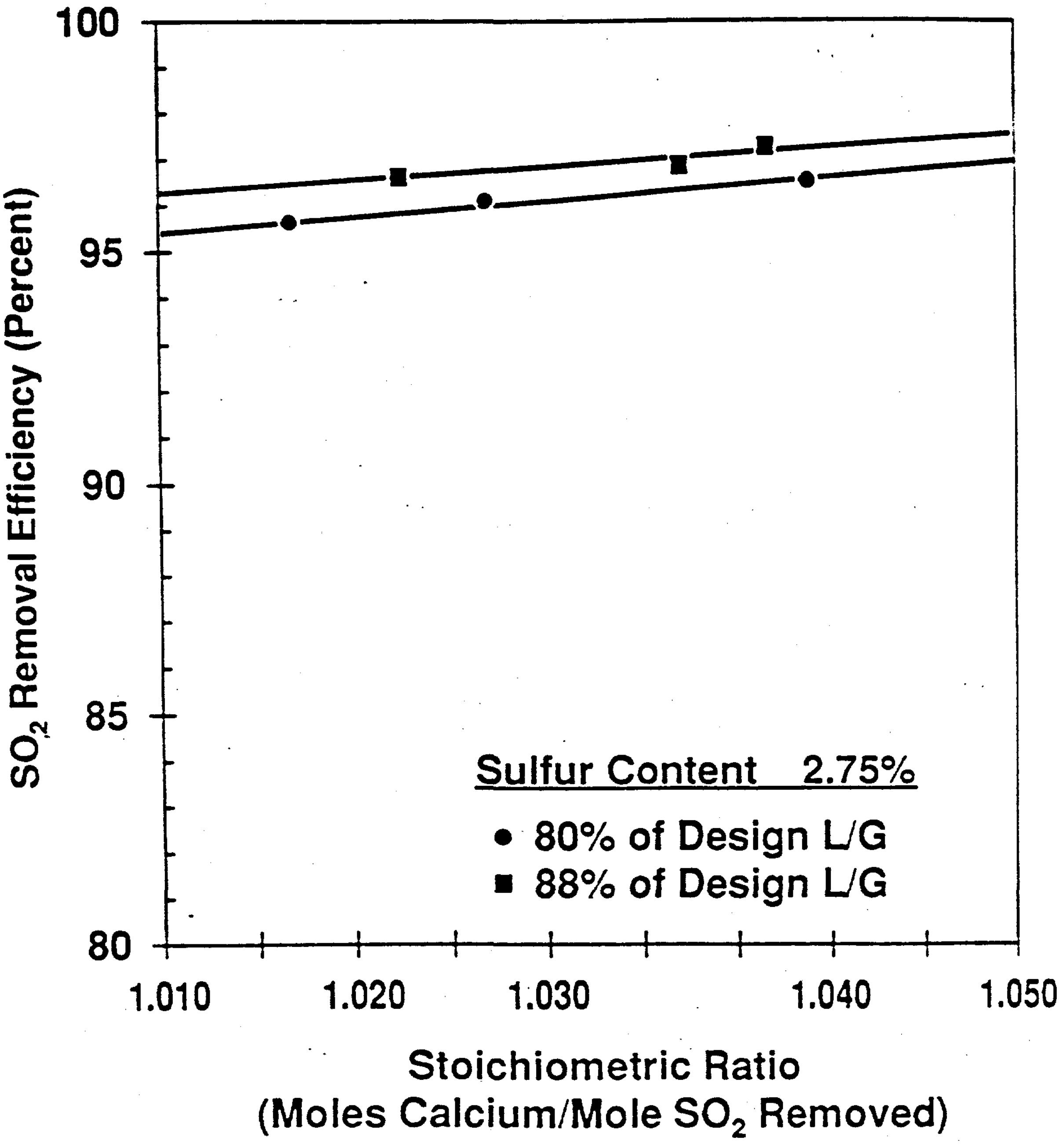

FIGURE 14

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

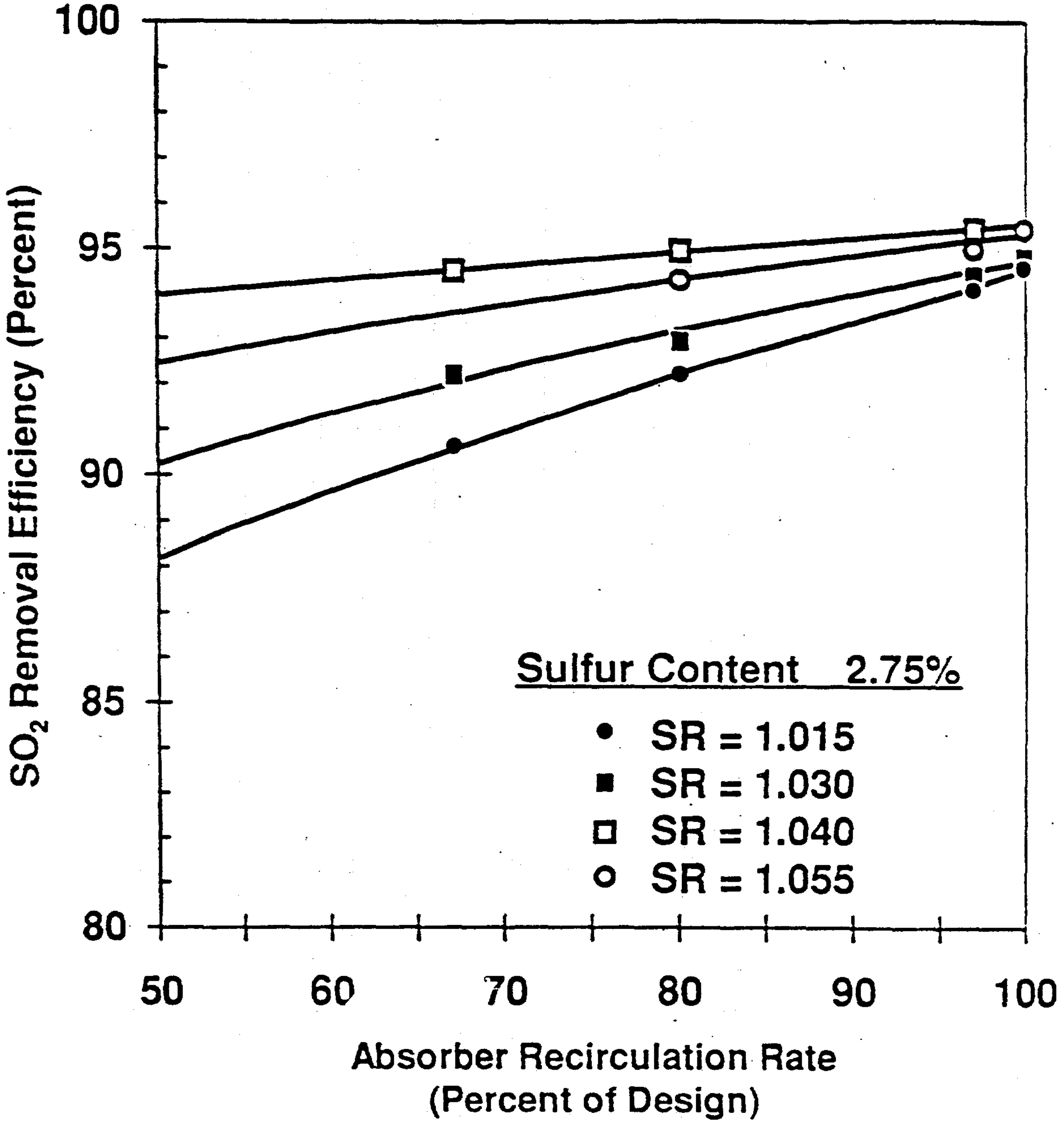

FIGURE 15

SO_2 Removal Performance at Bailly AFGD

(67 Percent Boiler Load)

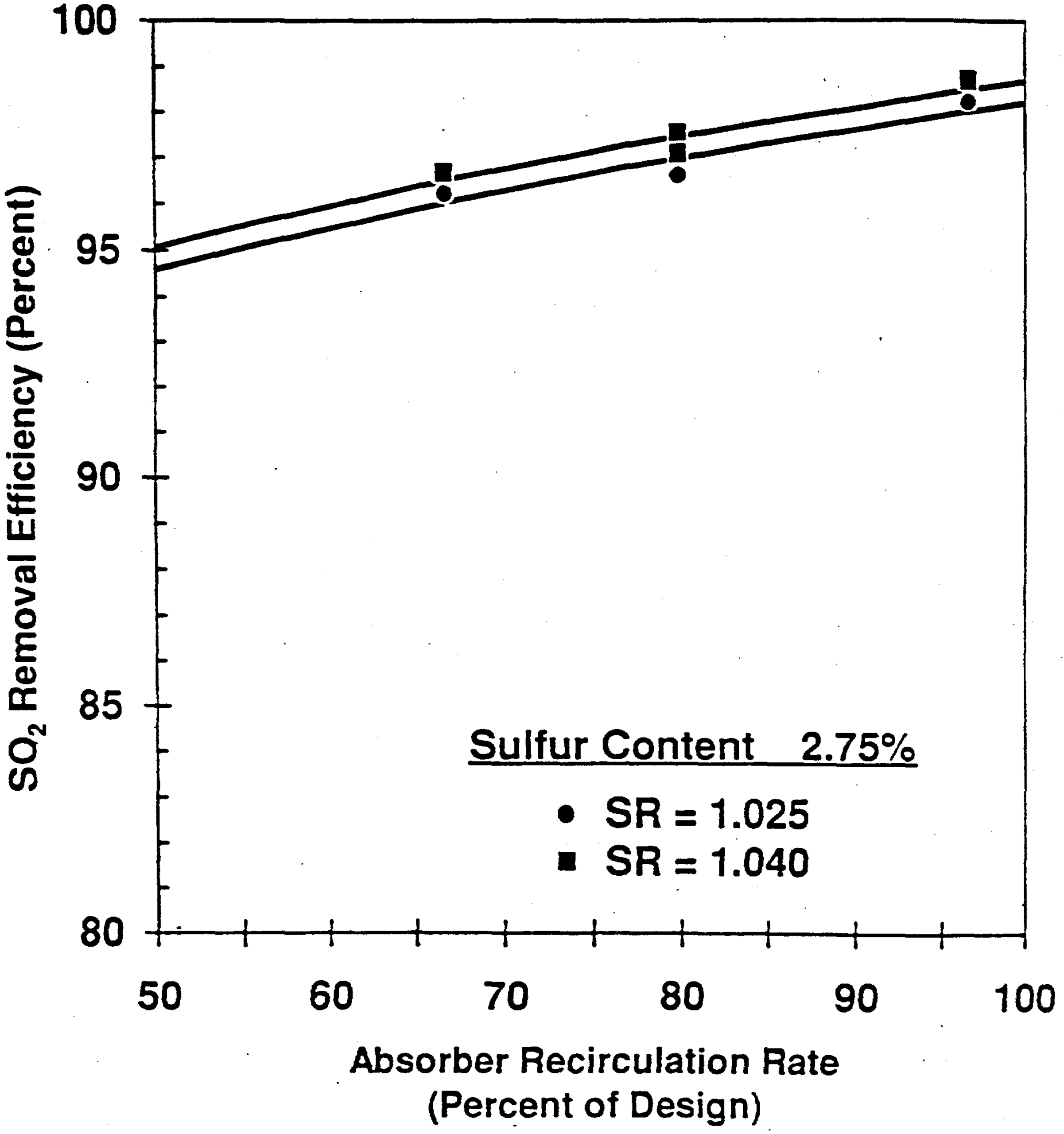

FIGURE 16

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

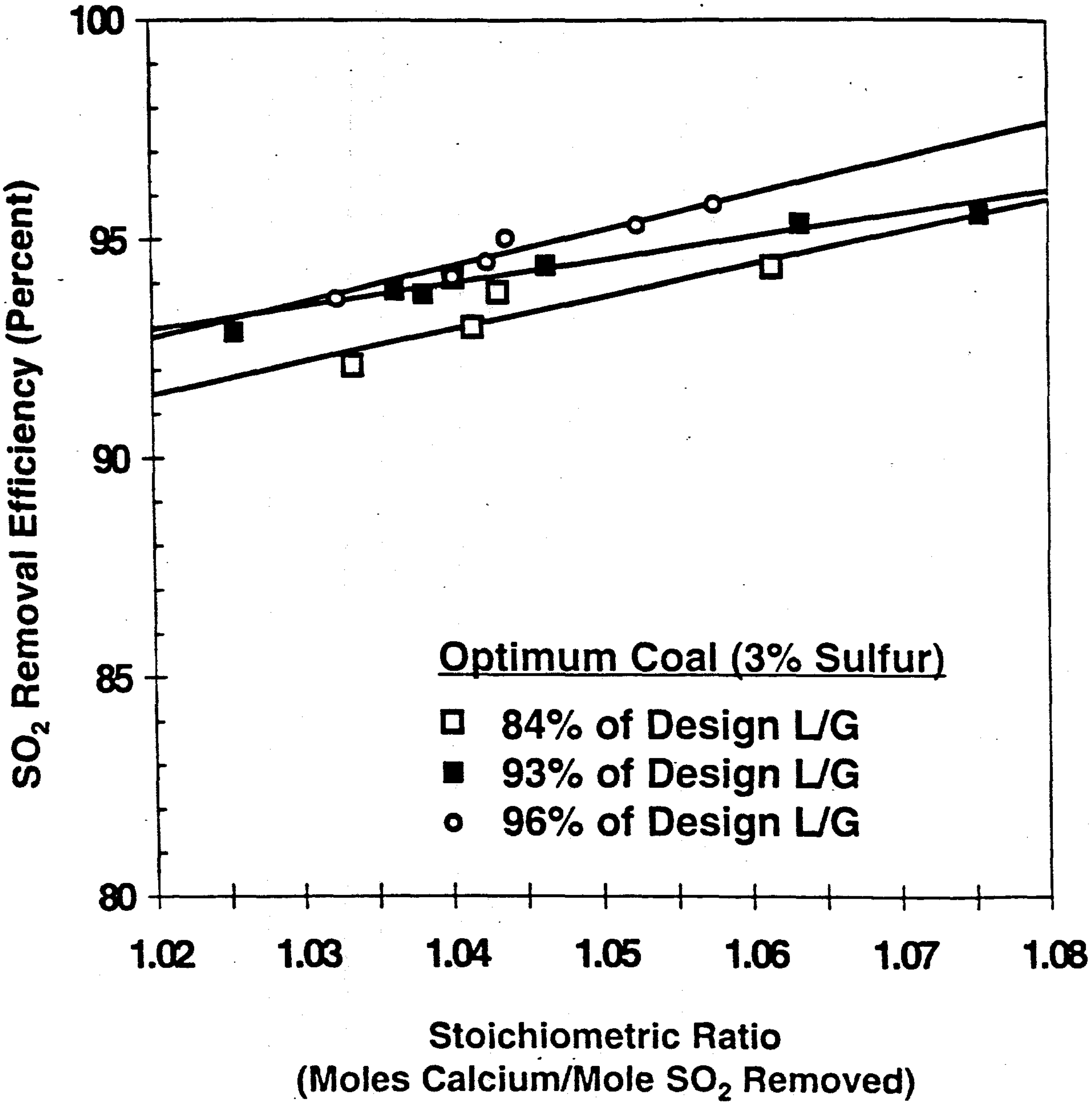

FIGURE 17

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

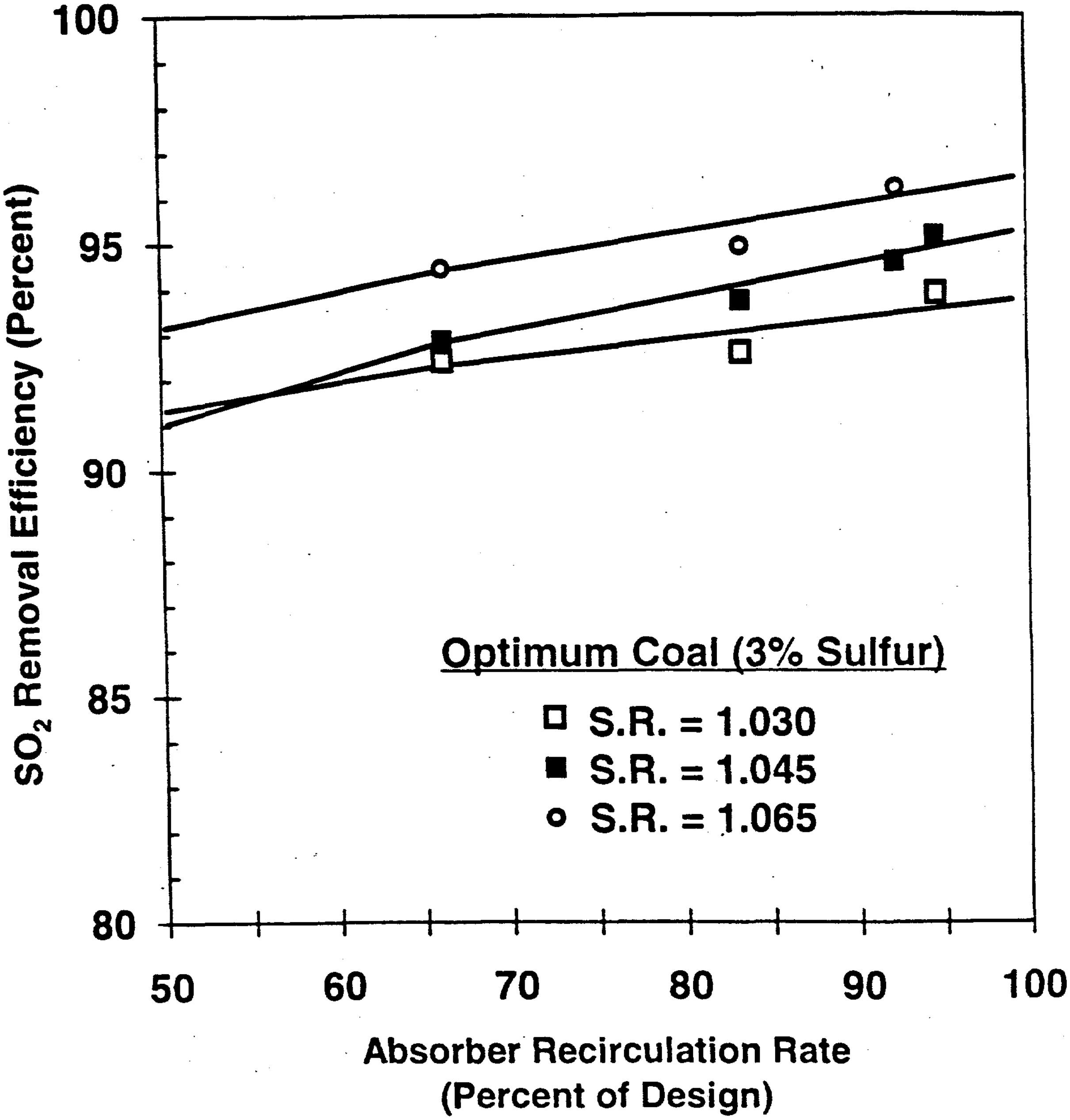

FIGURE 18

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

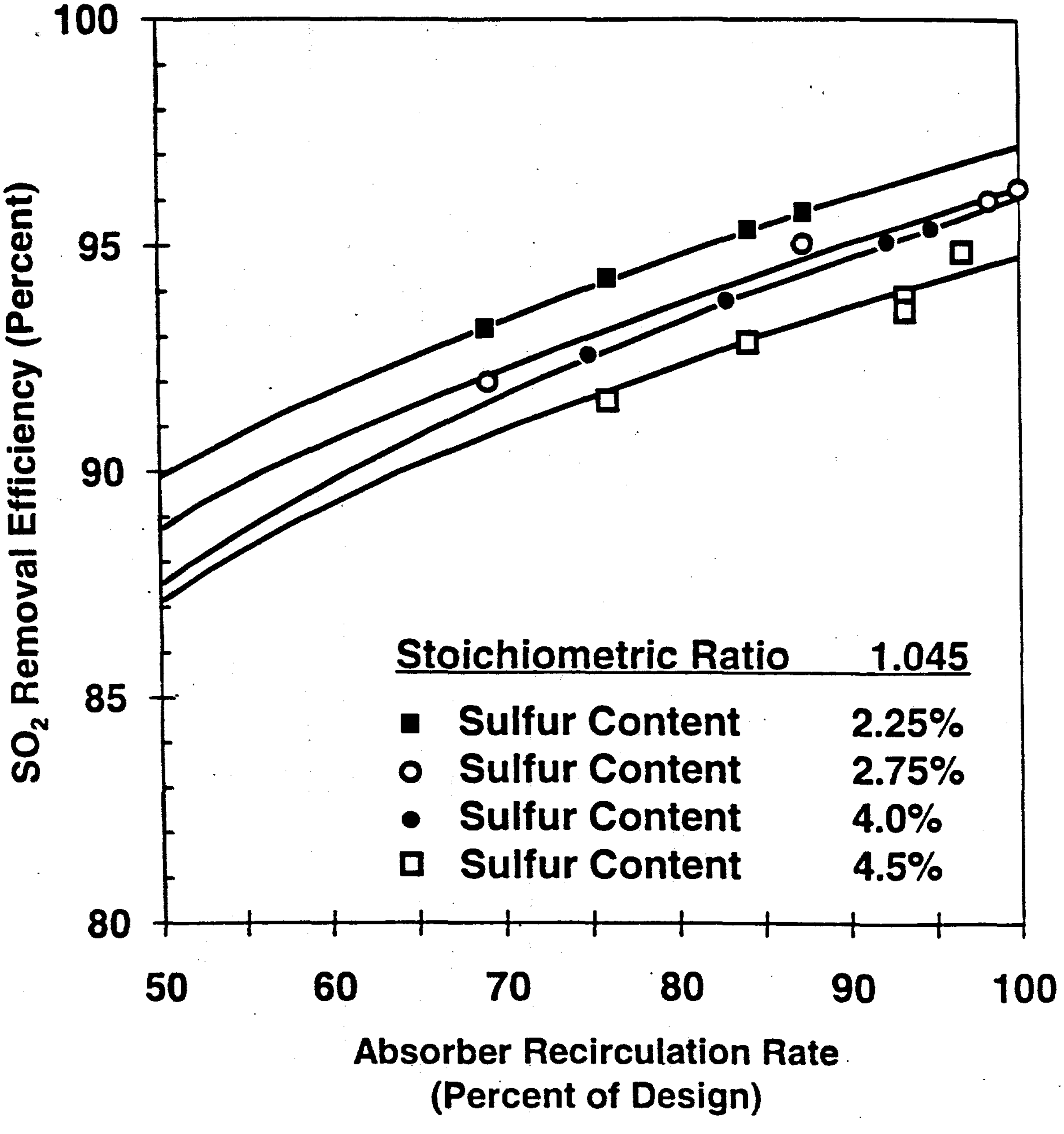

FIGURE 19

SO_2 Removal Performance at Bailly AFGD

(100 Percent Boiler Load)

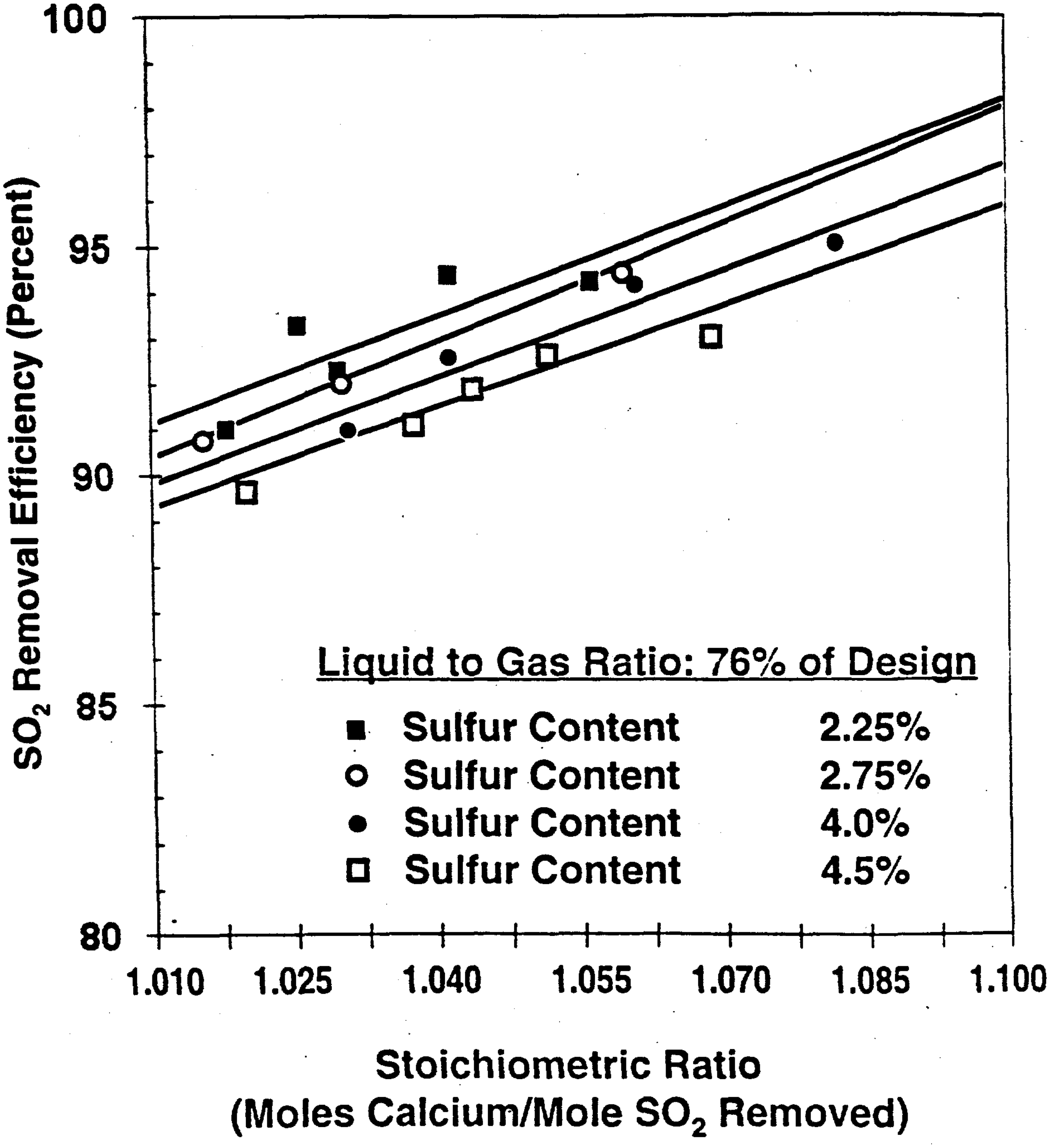

TABLE 1

COAL ANALYSIS

	DOE Demo I (S=2.25%)	DOE Demo II (S=2.75%)	DOE Demo III (S=3%)	DOE Demo IV (S=4%)	DOE Demo V (S=4.5%)	DOE Demo VI S=3%)
Carbon	66.56	61.61	62.1	59.14	69.32	61.1
Hydrogen	4.5	4.39	4.09	4.37	4.94	4.32
Nitrogen	1.44	1.23	1.22	1.26	1.17	1.16
Sulfur	2.21	2.91	3.21	3.79	4.73	3.00
Oxygen	6.71	7.45	8.19	7.19	5.63	7.40
Chlorine	0.14	0.1	0.06	0.03	0.07	0.07
Moisture	8.6	12.92	11.14	13.72	4.74	13.05
Ash	9.53	9.63	10.1	10.7	9.3	9.93
Btu/lb	11,932	11,022	10,874	11,000	12,700	10,972

TABLE 2

GYPSUM ANALYSIS

Ultimate Analysis (As Rec'd Weight %)

	DOE Demo I (S=2.25%)	DOE Demo II (S=2.75%)	DOE Demo III (S=3%)	DOE Demo IV (S=4%)	DOE Demo V (S=4.5%)	DOE Demo VI (S=3%)
Gypsum ($CaSO_4 \cdot 2H_2O$, %)	96.7 - 99.7	96.3 - 99.4	94.6 - 98.8	93.5 - 97.3	95.6 - 99.7	97.4 - 99.1
Calcium Carbonate ($CaCO_3$, %)	0.7 - 2.8	0.4 - 2.8	1.5 - 3.7	0.4 - 4.5	1.6 - 2.9	1.4 - 3.5
Chloride (Cl, ppm)	<20 - 37	<20 - 38	<20 - 38	9 - 148	<20 - 37	20 - 107
Magnesium Oxide (MgO, %)	0.04 - 0.17	0.04 - 0.2	0.08 - 0.23	0.08 - 0.51	0.08 - .21	0.03 - 0.16
Free Moisture (%)	4.6 - 7.8	4.3 - 8.4	3.7 - 8.4	4.2 - 8.8	5.8 - 9.6	5.4 - 7.9

TABLE 3

OPERATIONS SUMMARY

	Expected	Achieved
SO_2 Emissions	90% removal or 0.6 lb/MMBtu, whichever is less stringent	Averaged 94% (during DOE test up to 98+%, or 0.382 lb/MMBtu)
Power Consumption 24 hour average (kw)	<8,650	5,275
Facility Pressure Drop 24 hour average (IWC)	<13.5	3.23
Particulate Emissions (g/SCFD)	no net increse	0.04 inlet 0.0071 outlet
Availability (%)	95	99.47
Gypsum Moisture (%)	<10	6.64
Gypsum Chloride (ppm)	<120	33
Gypsum Purity (%)	93	97.2
Average Water Consumption (GPM)	3,000	1,560
Average Wastewater Flow (GPM)	275	81

TABLE 4

WALLBOARD-GRADE GYPSUM SPECIFICATIONS

	Expected	Four Year Average
Gypsum Purity (wt. % dry)		
$CaSO_4 \cdot 2H_2O$	>93.0	97.4
$CaSO_3 \cdot 1/2H_2O$	<2.0	0.08
SiO_2	<2.5	0.5
Fe_2O_3	<3.5	0.22
R_2O_3	—	0.3
Chlorides (ppm)	<120	30
Free H_2O (wt %)	<10	6.5
Mean Particle Size (microns)	>20	50

TABLE 5

WATER REQUIREMENTS

	Expected	Four Year Average
Supply Water Flow (gpm)	<3,000	1,500
Wastewater pH	6.0 to 9.0	7.5 to 9.0
Wastewater Total Suspended Solids (ppm)	<30	<18
Wastewater Dissolved Solids (ppm)		
Chlorides (Cl)	<30,000	8,960
Sulfates (SO_4^{+2})	<2,500	<2,500
Fluorides (F)	<1,100	23
Total Dissolved Solids	<100,000	15,000

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DEMONSTRATION OF SCR TECHNOLOGY FOR THE CONTROL OF NO_x EMISSIONS FROM HIGH-SULFUR COAL-FIRED UTILITY BOILERS

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ABSTRACT

This paper describes the completed Innovative Clean Coal Technology project which demonstrated SCR technology for reduction of flue gas NO_x emissions from a utility boiler burning U.S. high-sulfur coal. The project was sponsored by the U.S. Department of Energy, managed and co-funded by Southern Company Services, Inc. on behalf of the Southern Company, and also co-funded by the Electric Power Research Institute and Ontario Hydro. The project was located at Gulf Power Company's Plant Crist Unit 5 (a 75 MW tangentially-fired boiler burning U.S. coals that had a sulfur content ranging from 2.5 - 2.9%), near Pensacola, Florida. The test program was conducted for approximately two years to evaluate catalyst deactivation and other SCR operational effects. The SCR test facility had nine reactors: three 2.5 MW (5000 scfm), and six 0.2 MW (400 scfm). Eight reactors operated on high-dust flue gas, while the ninth reactor operated on low-dust flue gas. The reactors operated in parallel with commercially available SCR catalysts obtained from suppliers throughout the world. Long-term performance testing began in July 1993 and was completed in July 1995. A brief test facility description and the results of the project are presented in this paper.

PROJECT GOALS

Although SCR is widely applied in Japan and western Europe, numerous technical uncertainties are associated with applying SCR to U.S. coals. These uncertainties include:

(1) potential catalyst deactivation due to poisoning by trace metal species present in domestic coals but not present, or present at much lower concentrations, in fuels from other countries;

(2) performance of the technology and effects on the balance-of-plant equipment in the presence of high amounts of SO_2 and SO_3 (e.g., plugging of downstream equipment with ammonia-sulfur compounds); and

(3) performance of a wide variety of SCR catalyst compositions, geometries and manufacturing methods at typical U.S. high-sulfur coal-fired utility operating conditions.

These uncertainties were explored by constructing and operating a series of small-scale SCR reactors and simultaneously exposing different commercially available SCR catalysts to a common flue gas derived from the combustion of high-sulfur U.S. coal. First, long-term SCR catalyst performance was evaluated for two years under realistic operating conditions similar to those found in U.S. pulverized-coal-fired utility boilers. Deactivation rates for exposed catalysts were documented to determine life and associated process economics. Second, short-term parametric tests were performed during which SCR operating conditions were adjusted above and below design values to observe catalyst performance. The performance of air preheaters was also observed to evaluate the effects of SCR operating conditions on heat transfer, pressure drop and boiler efficiency. Third, honeycomb- and plate-type SCR catalysts of various commercial compositions from the U.S., Japan, and Europe were evaluated. Tests with these catalysts were used to expand knowledge of the performance of various catalyst types under U.S. utility operating conditions with high-sulfur coal.

SCR TEST FACILITY DESCRIPTION

The SCR demonstration facility was located at Gulf Power Company's Plant Crist in Pensacola, Florida. The facility treated a flue gas slip-stream from Unit 5, a commercially operating 75-MW (nameplate) unit, firing U.S. coals with a sulfur content ranging from 2.5 - 2.9%. Unit 5 is a tangentially-fired, dry bottom boiler with hot- and cold-side electrostatic precipitators (ESPs) for particulate control. The SCR test facility consisted of nine reactors operating in parallel for comparisons of commercially available SCR catalysts obtained from suppliers throughout the world. With all reactors in operation, the amount of combustion flue gas that could be treated was 17,400 scfm or 12% of Unit 5's capacity (about 8.7 MWe). Table 1 shows the average test facility flue gas constituent concentrations measured over the life of the project. Also shown are representative particulate loadings.

Table 1: Test Facility Inlet Flue Gas Composition

Constituent	Average	Data Source
Unit Load (MW)	63	continuous data average over project life
NO_x (ppmv at actual O_2)	314	continuous data average over project life
O_2 (%)	4.8	continuous data average over project life
CO_2 (%)	14.7	continuous data average over project life
SO_2 (ppm)	1570	continuous data average over project life
SO_3 (ppm)	4	average of parametric test data
HCl (ppm)	104	start-up/commissioning test data
NH_3 (ppm)	< 0.4	start-up/commissioning test data
Particulate (gr/dscf)	3.1	estimated by interpolation using average unit load and high/low load start-up/commissioning data

The process flow diagram for the SCR test facility is shown in Figure 1. The facility consisted of three large SCR reactors (2.5 MW, 5000 scfm) and six smaller SCR reactors (0.2 MW, 400 scfm). Eight of the nine reactors were designed to treat flue gas containing full particulate loading (high dust) extracted from the inlet duct of the hot-side ESP, while one small reactor was designed to treat flue gas extracted from the hot-side ESP outlet (low dust). Only seven of the eight high-dust reactors were operated for catalyst long-term and parametric testing.

Each reactor train was equipped with an electric duct heater to independently control flue gas temperature and a venturi flow meter to measure the flue gas flow rate. An economizer bypass line maintained a minimum flue gas temperature of 620 °F to the high-dust reactors. Anhydrous ammonia was independently metered to a stream of heated dilution air and was injected via nozzles into the flue gas upstream of each SCR reactor. The flue gas and ammonia passed through the SCR reactors, which had the capacity to contain up to four catalyst layers.

For the large reactor trains, the flue gas exited the reactor and entered a specially modified pilot scale air preheater (APH). This modified APH was used to better simulate full scale APH's for improved extrapolation of results to commercial scale. The APHs were incorporated in the project to evaluate the effects of the SCR process on APH deposit formation and the effects of the deposits on APH performance and operation.

All reactor trains, except the low-dust train, had a cyclone downstream of the SCR reactor to protect the induced draft (ID) fans from particulates. The exhaust from all of the SCR reactors was combined into a single manifold and reinjected into the host boiler's flue gas stream ahead of the cold-side ESP. The preheated air from the APH on the large reactors was also combined into a single manifold and returned to the host boiler draft system at the existing host APH outlet. All particulate removed from the test facility was combined with ash from the host unit's ESP and sent to an ash disposal area.

The test facility examined the performance of eight SCR catalysts (one reactor was idled due to the withdrawal of a project participant), each with its own particular geometry and physical and chemical design. Each catalyst supplier was given great latitude in designing their particular catalyst offering for the project. Suppliers were required to meet the following general criteria.

- design catalyst baskets to match predetermined reactor dimensions
- provide a maximum of four catalyst layers
- insure a maximum base-line SO_2 oxidation rate of 0.75%
- insure a maximum base-line NH_3 slip at the reactor exit of 5 ppm
- maintain 80% NO_x reduction while meeting NH_3 slip requirements
- design for 2-year life meeting performance criteria

The particular design characteristics of each catalyst follow in Table 2. The number of catalyst beds, general type of catalyst (honeycomb or plate) and reactor in which the catalyst was tested are provided. In addition, the project life flue gas exposure time for each of the catalysts is shown.

Table 2. Catalyst Design Specifications

Parameter	Grace Noxeram	NSKK	Siemens	Grace Synox	Cormetech High-Dust	Haldor Topsoe	Hitachi Zosen	Cormetech Low-dust
Reactor	A	B	C	D	E	F	G	J
Dust Level	High	High	High	High	High	High	High	Low
Composition	V-W/Ti	V-W/ Ti/Si	V/Ti	V/Ti/Si	V-W/Ti	NA[6]	V/Ti	V-W/Ti
Type[5]	HC	HC	Plate	HC	HC	Plate	Plate	HC
Pitch, mm (opening/wall thick.)	7.5 (6.1/1.4)	7.0	5.0	7.5 (6.1/1.4)	7.1 (6.0/1.1)	DNX-16 6.4 hyd.D.	5.5	3.7 (3.2/0.5)
Void Fraction, %	65	70	81	65	71	73	90	72
Density, lb/ft^3	39±3.5	25	37[1]	18±1.5	37	16	23[1]	32
Geom. Surf. Area, m^2/m^3	430	470	383	430	470	455	420	910
Gas Flow, Nm^3/hr	8500	8500	8500	680	680	680	680	680
Gas Flow, scfm	5000	5000	5000	400	400	400	400	400
Catalyst Volume, m^3	3.1	3.026	2.30	0.19	0.245	0.189	0.27	0.097
GHSV @ 0°C, hr^{-1}	2742	2809	3692	3579	2776	3600	2500	7033
Gas Flow Parametric Range (as % of design) Min. Max.	63% 126%	91% 127%	60% 150%	66% 131%	60% 150%	50% 100%	NA NA	60% 150%
Cross Sec. Area, m^2 [2]	1.080	1.08	1.106 (1.164)	0.09	0.081	0.094	0.09	0.081
Super.Lin.Veloc.,Nm/s[3] Min. Max	2.186 2.76	2.186 2.5	2.135 3.203	2.10 2.76	2.34 3.0	2.00 2.5	2.1 2.5	2.34 3.5
No. of Beds	3	3	2	3	3	3	3	2
Temperature, °F	700	700	700	700	700	700	700	700
Temp. Parametric Range (°F) Min. Max.	660 750	644 842	617 806	660 750	644 790	617 752	626 752	644 790
% SO_2 Oxidation	0.75	0.5	< 0.6	< 0.75	<0.75	0.68	0.7	< 0.75
Inlet NO_x, ppmv (wet)	400	400	417	400	400	400	400	400
NH_3/NO_x	0.813	0.811	0.817	0.813	0.81	0.81	0.85	0.81
NH_3/NO_x Parametric Range Min. Max.	0.6 1.0	0.6 1.0	0.6 1.2	0.6 1.0	0.6 1.2	0.6 1.2	0.7 1.1	0.6 1.2
% deNO_x Activity	80	80	80	80	80	80	80	80
NH_3 Slip, ppmv (wet)	< 5	5	5	< 5	<5	< 5	< 5	< 5
Particulates, g/Nm^3	6 - 8	6 - 8	6-8	6 - 8	6-8	6-8	6 - 8	NA
Pressure Drop, in. H_2O	3.85[4]	2.62	1.28	2.60[4]	2.6	1.73	< 4	3.5
Actual Flue Gas Exposure Time (hrs.)	11012	11859	11632	10151	10151	10175	7293	5363

1. Includes basket, otherwise catalyst density only.
2. Catalyst only. Value in parentheses includes basket.
3. Velocity based on cross-sectional area of catalyst, not basket.
4. For Grace, ΔP includes baskets
5. HC = Honeycomb
6. NA = Not Available

Fuel

One of the primary purposes of this project was to evaluate the performance of SCR technology on U.S. coals. To insure that an adequate database of coal composition was generated, daily as-burned coal samples were acquired from the host boiler. Monthly composites were then used to carefully track both primary and trace coal constituents. Where possible, neutron activation analysis (INAA) was used to augment other measurement techniques.

The coal supply during the project consisted of eastern bituminous coals with sulfur contents greater than or equal to 2.5%. When coal monitoring began (March 1993), Peabody coal having a 2.9% sulfur content was being burned. In September 1993, however, a coal miner's strike and buy-out of the Peabody contract caused a change in coal supply. During September and October, Kerr-McGee and Taft coals were purchased. The Kerr-McGee coal was a blend of Illinois #5 and #6 mines while the Taft coal was from Walker and Jefferson counties in Alabama. For November and December, coal from the Illinois basin in Western Kentucky was purchased from Jader to be burned. All of these coals had a sulfur content of approximately 2.5%. In January 1994, a new contract was set up to purchase high sulfur coal (2.7% sulfur) for the host unit from Old Ben Franklin and Kerr-McGee coal companies. The Old Ben coal from the Illinois #6 mine and Kerr-McGee from the Herring #6 mine continued to be the coal source throughout the remainder of the project.

A summary of the coal analyses performed over the project life follows in Tables 3 and 4. The analyses presented contain data measured by Alabama Power Company Laboratories (APCo) using atomic absorption and ion-coupled plasma techniques (primary constituents and some trace metals) and data measured by the University of Missouri-Columbia, Research Reactor Center using neutron activation analysis (INAA) to supplement trace metals testing In cases where a constituent was measured by both laboratories, the data give a basis of comparison for the two analytical methods.

Table 3. Proximate Coal Analysis Data

Test	APCo Method	Units	APCo data
Moisture, Total	ASTM D 3302	% by Wt.	10.87
Ash	ASTM D 3180	% by Wt.	9.30
Gross Cal. Value	ASTM D 3180	Btu/lb	13268
Sulfur, Total	ASTM D 3180	% by Wt.	2.58
Sulfur lb/MMBtu	ASTM D 3180	lb/MMBtu	1.95
Carbon	ASTM D 3180	% by Wt.	74.82
Hydrogen	ASTM D 3180	% by Wt.	5.00
Nitrogen	ASTM D 3180	% by Wt.	1.58
Oxygen	ASTM D 3180	% by Wt.	6.73
Carbon, Fixed	ASTM D 3180	% by Wt.	52.83
Volatile Matter	ASTM D 3180	% by Wt.	37.88

Table 4. Elemental Coal Analysis Data

Element	Units	APCo data	INAA data
Aluminum	wt %	1.09	0.97
Antimony	mg/kg	<1.0	0.74
Arsenic	mg/kg	3.2	7.6
Barium	mg/kg	40	58
Beryllium	mg/kg	3	
Cadmium	mg/kg	< 1.0	
Calcium	wt %	0.24	0.16
Cerium	mg/kg		12.1
Cesium	mg/kg		0.97
Chlorine	mg/kg	1767	
Chromium	mg/kg	19	16.7
Cobalt	mg/kg	7	3.61
Copper	mg/kg	9	
Dysprosium	mg/kg		0.92
Europium	mg/kg		0.23
Fluorine	mg/kg	56	
Hafnium	mg/kg		0.55
Iron	wt %	1.08	1.04
Lanthanum	mg/kg		5.6
Lead	mg/kg	12	
Lithium	mg/kg	9	
Lutetium	mg/kg		0.12
Magnesium	wt %	0.06	
Manganese	mg/kg	24	24.6
Mercury	mg/kg	0.07	0.11
Molybdenum	mg/kg	7.8	9.3
Neodymium	mg/kg		6.2
Nickel	mg/kg	15	<28
Phosphorus	wt %.	0.02	
Potassium	wt %	0.20	0.18
Rubidium	mg/kg		12.5
Samarium	mg/kg		1.21
Scandium	mg/kg		2.61
Selenium	mg/kg	<2.0	1.91
Silica	wt %	2.27	
Sodium	wt %	0.06	0.05
Strontium	mg/kg		<38
Tantalum	mg/kg		0.13
Terbium	mg/kg		0.14
Thorium	mg/kg		1.72
Titanium	wt %	0.06	0.06
Uranium	mg/kg		2.61
Vanadium	mg/kg	41	39.5
Ytterbium	mg/kg		0.50
Zinc	mg/kg	39	47
Zirconium	mg/kg		34

TEST PLAN

The project test plan was divided into five main categories as follows. Start-up and commissioning tests were performed during the initial start-up of the test facility beginning in early 1993 and continuing until the beginning of long term operations (July 1993). The remaining categories of tests were performed during the two year period of long term operations.

1) **Start-up and Commissioning Tests**
2) **Long-Term Performance Tests**
3) **Short-Term Parametric Tests**
4) **Air Preheater Tests**
5) **Miscellaneous Tests**

Start-Up and Commissioning Tests

The start-up and commissioning of the test facility was of extreme importance to the overall success of the project. Tests were performed to demonstrate that parameters such as ash loading, particle size distribution, flue gas flow, velocity distribution, and flue gas constituents (including trace metals), etc. were comparable between the reactors and were representative of the host unit flue gas parameters. Also, the start-up and commissioning tests showed that the specifications for the test facility design had been met. These basic design criteria follow in Table 5.

Table 5. Test Facility Design Criteria

Parameter	Design Specification
Flue Gas Flow Rate	Control to within ± 2% for large reactors, ± 5% for small reactors
NH_3/NO_x Ratio	Control to within 0.005 of desired NH_3/NO_x
Flue Gas Velocity Distribution	Not more than ± 10% deviation in flow velocity across the cross-section of each individual reactor
NH_3 Distribution	Not more than ± 5% deviation in ammonia distribution
Particulate Concentration	Not more than ± 5% deviation between total mass loading between individual reactors, not more than ±10 % deviation in particle size distribution

Long-Term Catalyst Performance Testing

Long term catalyst performance testing focused on the performance of the catalysts/reactors at baseline conditions evaluating long-term performance parameters such as catalyst deactivation, ammonia slip, and reactor pressure drop. This evaluation depended on continuous or periodical measurement of various operational parameters to determine performance. The reactor baseline conditions under which these long-term tests were performed are identical to the baseline conditions shown for the short term parametric tests in Table 6.

Catalyst Pressure Drop, Fouling, and Erosion

Reactor pressure drops were measured continuously using the test facility's distributed control/data acquisition system. The data shown in Figure 2 represent pressure drop in inches of water column across the catalyst beds in each of the reactors in the test facility. The pressure drop created by the dummy beds has been excluded to more closely represent actual catalyst pressure drop. In practice, full scale installations experience very little pressure drop across the flow straightening grid, unlike the test facility which required significantly higher dummy bed pressure drops to meet design specifications. To produce a meaningful plot, the data has been limited to periods of operation at or near baseline conditions. Periods where no data is shown represent periods when either the particular reactor or entire test facility was not in operation.

The general trend of the plot shows a fairly steady pressure drop over time for most of the test facility catalysts. This is encouraging because it indicates that sootblower operations were providing the cleaning necessary to prevent long term fouling of the catalysts. Sootblowing at the test facility was normally carried out on each of the catalyst beds and dummy beds every eight hours. The large reactors utilized a traversing rake steam sootblowing system. Steam was injected at approximately 250 psig at a distance of roughly 2 ft. above each of the catalyst layers and dummy layer. The small reactors were manually sootblown using a compressed air lance. Interestingly, due to the delicate nature of the Hitachi Zosen catalyst, this catalyst was not sootblown. This did not, however, adversely affect the fouling of this catalyst, as the extremely low facial area of the catalyst prevented fouling to a great degree.

Catalyst fouling and erosion was tracked through the analysis of reactor pressure drops and periodic visual inspections. In commercial SCRs, which have relatively infrequent visual inspections of the catalyst layers, the primary indication of catalyst fouling is by the monitoring of reactor pressure drop. In addition, monitoring of NO_x and/or slip distributions at the reactor outlet can give an indication of catalyst fouling. Erosion in commercial SCRs is tracked through visual inspection and by periodic sampling of the catalyst at which time erosion can be determined in the laboratory. Severe erosion may also be detected by an analysis of catalyst bed pressure drops.

Since catalyst sampling at the test facility required quarterly shut-down of the reactors, visual inspection of all catalyst beds was made at least on a quarterly basis. In practice, other outages were also used to make visual inspections, resulting in a quite frequent inspection schedule. These inspections usually indicated some fouling near the reactor walls. Problems due to this fouling were greatly mitigated on the large reactors by the design of the catalyst baskets. Catalyst suppliers had insured a smooth transition of flow from the reactor dimensions to the actual catalyst open face dimensions, thus preventing significant fouling due to wall effects. The small reactors exhibited more fouling near the reactor walls, primarily due to the small size of these reactors thus exacerbating the wall effects. In any case, all high-dust catalysts demonstrated an ability to handle the dust loadings without significant impacts to performance. In the case of the low-dust reactor, more fouling was noted. This was not a result of catalyst design, however, but was the result of test facility design. Long duct runs with increased heating requirements and a less than optimum take-off scoop all contributed to large particulate material being passed through the low-dust reactor system. The catalyst was not designed for these conditions and the fluctuation in this catalyst's pressure drop should not be construed as demonstrating the inapplicability of a low-dust configuration.

An important addition to each of the catalyst beds was a screen used to assist in the prevention of catalyst fouling. Screens placed several inches above the catalyst face with a mesh opening roughly equal to or slightly smaller than the catalyst channel openings proved to be extremely helpful in preventing channel blockage. The screens helped by breaking up large ash particles into small pieces which could pass through the catalyst rather than lying on the catalyst face, thereby blocking channels. Large ash particles that did not break up merely rested on the screen until they could be removed through periodic cleaning. Since the screens were placed several inches above the catalyst face, flow could redistribute around the large ash pieces, thus minimizing adverse effects.

The harsh environment in which catalysts in coal fired applications must operate make high physical strength and durability a necessity for catalysts. Erosion of the catalyst can be worsened by areas of high velocity or high particulate as well as by erosion from sootblowing operations. Severe erosion in certain areas of a particular catalysts is accelerated by the fact that eroded areas become the preferred flow path due to localized reduced pressure drop, thereby increasing the erosion potential in that particular area. This phenomenon seems to be more prevalent with honeycomb catalysts, however, leading edge hardening of the catalyst can help to mitigate facial erosion. In the case of some plate catalysts such as the Siemens, the catalysts support is steel, preventing significant erosion past the face of the catalyst. In the case of the Haldor Topsoe catalyst, a face hardening procedure appeared to enhance this catalyst's ability to withstand facial erosion. Little facial erosion was noted on the Hitachi Zosen catalyst likely due to the extremely small cross-sectional area of exposed catalyst face (due to extremely thin plates). The majority of erosion on the honeycomb catalysts appeared to be due to aggressive sootblowing procedures (based on the erosion pattern). One exception to this finding however, was the Grace Synox catalyst. This special-design catalyst represents a significant deviation from other honeycomb catalysts currently marketed. The bimodal design resulted in a relatively soft material which showed significant erosion during the test program mainly indicated by the thinning of the catalyst walls. Also, some dummy bed material initially used in the test facility was considerably less durable than the catalysts. This material exhibited the phenomenon previously discussed where erosion problems were exacerbated by channeling.

Catalyst deNO$_x$ Activity and Deactivation

As expected, the deNO$_x$ activity of the test catalysts varied greatly with respect to bulk volume. However, by varying bulk volume, all catalysts in the test program met the long-term performance criteria. DeNO$_x$ activity is an important parameter because the bulk volume (and weight) control to a large degree the number of catalysts beds and physical size and structure of an SCR reactor and the resulting capital cost. Generally, as NO$_x$ reduction increases to high levels (>90%) ammonia slip levels increase dramatically. The specific design of the reactor is important in this observation. A reactor could, of course, be designed for 90% NO$_x$ reduction while maintaining very low values of ammonia slip. This however, results in a high capital cost due to the disproportionately increased amount of catalyst required. High slip values at high NO$_x$ reduction are not only due to catalyst volume and design, but also to the physical design of the reactor. As NO$_x$ reduction nears 100%, non-idealities in the reactor design, such as NO$_x$ distribution, ammonia distribution, and velocity distribution become extremely important and eventually control the maximum NO$_x$ reduction that can be obtained for a specific reactor.

The catalyst suppliers were given a great deal of latitude in specifying the volume of catalyst for their respective reactors as previously mentioned. In some cases suppliers chose to minimize catalyst volume by designing a system which met, but did not exceed (improve upon) performance requirements. In other cases, catalyst suppliers chose to increase catalyst volume to insure performance margin and thus improve performance. Honeycomb catalysts are generally considered to have a high surface area when compared to plate-type catalysts (on a bulk volume basis) and would thus generally require less volume (although weight may be greater). However, offerings in the test program demonstrated the ability of plate-type catalysts to meet specifications with similar or less volume than honeycomb configurations. Thus, it appears that installed catalyst design margin can be as significant to overall catalyst volume as is the basic geometrical (honeycomb/plate) design.

One of the fundamental purposes of the test program was to address the potential catalyst deactivation associated with U.S. high-sulfur coal applications. Although a considerable experience base was present for foreign coals, little was known about potential catalyst poisons and the resulting catalyst deactivation of U.S. coals prior to this study. The primary method of tracking this deactivation was through laboratory tests performed by the catalyst suppliers. Catalyst samples were removed from the reactors on roughly a quarterly basis and sent to the respective laboratories for testing. Results were then reported directly to SCS. Data are presented in Figure 3 at the base condition for which the most data is available. This base condition corresponds to a temperature of 700 °F, an area velocity of 50 Nm^3/m^2h, an inlet NO_x concentration of 400 ppm, an NH_3/NO_x ratio of 1.0 and an SO_2 concentration of 2000 ppm. These deactivation trends are in keeping with expectations of the catalyst suppliers based on other experience, primarily foreign installations. Trace analysis of potential catalyst poisons such as arsenic and sodium were performed at the individual supplier's laboratories to correlate these concentrations to the measured deactivation rates. Interestingly, poison levels (especially arsenic) on the catalysts were higher than would be expected for the measured amount of deactivation.

Long-Term Ammonia Slip

The general result of catalyst deactivation is the increase in ammonia slip from the reactor over time assuming deNO_x efficiency is held constant. Catalyst deactivation, however, is not always the source of increases in ammonia slip. Other factors such as maldistributions in NO_x and ammonia, fouling, and erosion can all contribute to increases in ammonia slip over time. As in the test facility, most commercial SCRs operate at constant deNO_x efficiency. This is necessary to meet emission regulations, but results in ammonia slip increases as reactor operation becomes less than optimum and the catalyst deactivates. Unfortunately, this ammonia slip is often poorly tracked in commercial SCRs, and thus the overall "health" of an SCR is not known. Consequently, the test facility monitoring program did closely track ammonia slip from the reactors at base-line conditions as well as other parametric conditions to adequately characterize the reactor/catalyst performance. As expected, ammonia slip increased over the life of the project. Figure 4 shows this general trend using the combination of all test facility catalysts. Data has been limited to operation at or very near base-line conditions. It should be noted that the depicted increases in ammonia slip are due not only to catalyst deactivation, but also due to degradation in reactor-specific parameters such as NH_3/NO_x distribution, velocity distribution, etc. Careful monitoring of these parameters, however, helped to minimize their contribution to the overall performance degradation of the reactor. The plot shows very low ammonia slip at the start of the project (<2 ppm for all catalysts with an average of less than 1ppm). As exposure time increased, however, ammonia slip increased, up to an average of roughly 3 ppm, with some catalysts at or near the design limit of 5 ppm. This demonstrates maturity in catalyst design and also shows that deactivation beyond that normally expected did not occur (based on prior world-wide experience). Figure 5 shows average ammonia slip for each of the catalysts on an individual basis using data taken over the life of the project at or near the base-line operating condition. It should be cautioned that ammonia slip is a direct function of catalyst volume and economics must be addressed in determining the most appropriate catalyst for a particular installation. Is should also be noted that catalyst exposure times differ significantly is some cases (see Table 2). The data presented represent average ammonia slip over the life of the project at conditions very near baseline. High and low measurement variability is shown. This is a measure of repeatability over the project life. The high values shown should not be construed as end-of-project ammonia slip values and therefore do not indicate a failure of the particular catalyst to meet design specifications.

Intermediate NO_x Reduction

Intermediate NO_x reduction was tracked throughout the life of the project using NO_x and ammonia measurements taken upstream and downstream of the first catalyst bed at various parametric conditions during the five parametric test sequences. In general, deactivation based on NO_x measurements is difficult to see in a small scale facility due to the relatively slight decline in activity, especially considering measurement accuracy. However, first bed NO_x reduction measurements are most likely to exhibit an effect from catalyst deactivation than are other NO_x measurements such as over-all NO_x reduction. Since intermediate NO_x reduction is a strong function of the NH_3/NO_x ratio, it can be difficult to discern trends since in practice the actual NH_3/NO_x ratio is difficult to hold constant. However, a large number of points at or near the same NH_3/NO_x ratio can be used to track decreases in catalyst activity. Also shown in Figure 4 is the intermediate NO_x reduction versus catalyst exposure time at base-line conditions. An average value of all reactors has been used to increase the total amount of data available. The plot exhibits a very slight decrease in NO_x reduction over the nearly 12,000 hours of catalyst exposure time. This indicates that on average, the catalysts were not deactivating rapidly as is supported by the laboratory activity tests previously discussed.

Short-Term Parametric Testing

Five sequences of parametric tests were performed on the test facility catalysts during the project life. These parametric tests were designed to examine ammonia slip, $deNO_x$ efficiency (intermediate ammonia), SO_2 oxidation, N_2O formation, NO_x and ammonia distributions, fly ash composition and loading, velocity distributions, and particulate distributions at varying conditions. Table 6 shows the general variations in temperature, flow rate, and NH_3/NO_x ratio that were tested. These parametric tests are most important in helping to define the applicable operating boundaries of the SCR reactors and their responses to changes in the various parameters.

Table 6. Parametric Conditions

Parameter	Minimum	Base-Line	Maximum
Temperature, (°F)	620	700	750
NH_3/NO_x molar ratio	0.6	0.8	1.0
Space velocity, % of design flow	60	100	150
Flow rate (scfm) -large reactor	3000	5000	7500
-small reactor	240	400	600

Ammonia Slip

In general, ammonia slip, not NO_x reduction is the controlling factor in the long term operation of a commercial SCR. Unless extremely severe deactivation has occurred, required NO_x reduction can almost always be achieved if no restriction is placed on ammonia slip. In practice, this is not acceptable due to balance-of-plant impacts and, consequently, limits of ammonia slip are placed on commercial applications, usually ≤ 5 ppm. The specifications for the test facility catalysts required the base-line ammonia slip not exceed 5 ppm through the end of the two year test period.

In addition to the ammonia slip depending on catalyst exposure time, it is a strong function of other parameters such as flow rate, temperature, NH_3/NO_x distribution, and especially NH_3/NO_x ratio (NO_x reduction). The catalyst design margin partly controls the ability of a particular installation to handle changes in these parameters without exceeding desired ammonia slip levels.

In addition, the catalyst design plays a role in its ability to handle parametric changes and consequently some catalysts appear to be more susceptible to increases in ammonia slip due to changes in parametric conditions than do others.

Flow Rate Effects

Theoretically, from a kinetic standpoint, NO_x reduction should be inversely proportional to changes in flow rate. In practice, improvements in mass transfer tend to mitigate some of the effects of increased flow rate and in general flows could be increased to 150% of design without the catalyst exceeding the 5 ppm limit at 80% NO_x reduction and design temperature. This ability depends on installed margin and design, although sensitivity to this parameter is not excessive.

Temperature Effects

Temperature increases would be expected to significantly improve reactivity based on a kinetic model of the catalysts. However, in practice, mass transfer limitations become controlling in commercial SCRs and improvements in test facility catalyst activity were not very significant above 700 °F. Most catalysts exhibited fairly significant improvements in overall performance as temperatures were increased from 620 °F to 700 °F, but relatively little improvement was noted as the temperature was increased from 700 °F to 750 °F. This shows that the benefits of high temperature operation probably do not outweigh the heat rate penalties involved in operating the SCR at the higher temperature. The reader should be cautioned that these performance conclusions are based on catalyst designed for a 700 °F base-line operating temperature. Catalyst suppliers are quite mature in their design capabilities and would likely be able to take advantage of a higher temperature operating environment in a specialized design and thus realize more significant improvements in performance.

NH_3/NO_x Ratio Effects

Changes in the NH_3/NO_x ratio and consequently NO_x reduction generally produce the most significant changes in ammonia slip. Decreases in NO_x reduction below 80%, down to values of roughly 60% were examined in the test facility. In most cases the ammonia slip at the low NO_x reduction was at or near the ammonia slip detection limit of 1 ppm. As NO_x reduction was increased above 80%, ammonia slip also increased. Slip values were fairly reasonable up to NO_x reduction levels of approximately 90%. As NO_x reduction increased past 90%, however, ammonia slip levels increased dramatically.

SO_2 Oxidation

Unfortunately, the catalytically active species that result in $deNO_x$ activity often contribute to SO_2 oxidation activity. Since increased SO_3 is detrimental to equipment downstream of the SCR, these two reactions tend to bound the catalyst design. In general, as requirements to minimize SO_2 oxidation relax, $deNO_x$ activity per volume of catalyst can be increased. The upper bound for SO_2 oxidation for the test facility catalysts was set at 0.75% at base-line conditions. The oxidation of SO_2 was measured in the test facility at various operating conditions in addition to the base-line measurements. The average SO_2 oxidation rates for each of the test facility catalysts are shown in Figure 6. These data reflect base-line conditions over the life of the project. All of the catalysts were within the design limits, with most exhibiting oxidation rates well below the specified limit. One notable exception is the Siemens catalyst which had a high activity with minimum catalyst volume, resulting in an SO_2 oxidation rate very close to the design value. In practice, some benefits may be realized by decreasing SO_2 oxidation past some arbitrarily chosen point, but these decreases are generally made at the expense of increasing catalyst volume. As with $deNO_x$ activity, the SO_2 oxidation margin chosen by the individual catalyst suppliers likely contributed significantly to the overall catalyst volume. It is difficult to determine the point of diminishing return for decreasing SO_2 oxidation, as little full scale long-term operational data is available for U.S. boilers. Although differences in SO_2 oxidation characteristics were very apparent between the catalyst suppliers, these differences do not necessarily denote a difference in catalyst design expertise between the suppliers, but more likely reflect a philosophical decision as to the catalyst offering made. In practice, all suppliers would likely be able to meet an applications specific SO_2 oxidation requirements (assuming flue gas constituents do not contribute to the oxidation characteristics). Unlike ammonia slip (or catalyst $deNO_x$ activity), the SO_2 oxidation rate of a catalyst is not expected to change significantly with time. This was the general finding based on measurements at the test facility as well as with the catalyst supplier laboratory tests. Since SO_2 oxidation remains relatively constant over time for a given catalyst volume, balance-of-plant effects must be evaluated in terms of the maximum SO_2 oxidation that may occur in later years with the addition of catalyst layers.

Flow Rate Effects

Due to the reaction order and the relatively constant SO_2 concentration, SO_2 oxidation should be inversely proportional to reactor flow rate (i.e. space velocity, linear velocity). However, most of the test facility catalysts exhibited a fairly constant SO_2 oxidation rate with respect to flow. The Siemens catalyst, with a somewhat higher SO_2 oxidation rate did exhibit a more linear response to flow rate. It is believed that physical phenomenon in the test facility such as SO_3 deposition within the reactor may have masked some of the effects on SO_2 oxidation from flow rate changes.

Temperature Effects

The oxidation of SO_2 is normally a much stronger function of temperature than of flow rate. Theoretically the catalysts should exhibit an exponential relationship of SO_2 oxidation to temperature. However, the measurements in the test facility showed that this relationship was more linear than exponential. Little difference was noted in SO_2 oxidation between 620 and 700 °F. However, SO_2 oxidation did increase more significantly between 700 and 750 °F. Figure 7 shows average SO_2 oxidation for the test facility catalysts at high temperature, with fairly significant increases in oxidation rate over base-line values previously shown.

NH_3/NO_x Ratio Effects

The rate of SO_2 oxidation is not expected to be a direct function of the NH_3/NO_x ratio, since ammonia does not play a direct role in the SO_2 oxidation reaction. However, other extraneous factors can create apparent changes in SO_2 oxidation rate with changes in NH_3/NO_x ratio. This is mainly due to physical effects such as precipitation of ammonium bisulfate. Increases in the NH_3/NO_x ratio change the ammonium bisulfate formation patterns between the measurement points and can result in an apparent change in SO_2 oxidation rate. For this reason, most SO_2 oxidation tests were made at 80% NO_x reduction.

Other Parametric Tests

Other tests were performed as part of the parametric sequences. These included evaluation of N_2O formation, NO_x and ammonia distribution analysis, fly ash composition and loading, velocity distributions and particulate distributions. The majority of these tests were performed at base-line conditions and were performed as part of an ongoing assessment of reactor operating conditions and performance. The results of these tests showed no significant changes in N_2O levels across the SCR reactors and that parameters such as particulate and velocity distributions remained within the original design specifications.

Air Preheater Tests

The three test facility air preheaters consisted of a two layer (APH A) and a three layer (APH B) Ljungstrom® air preheater and a heat-pipe or Q-pipe® (APH C) all provided by ABB Air Preheater, Inc. (ABB). The performance of the air preheaters was evaluated using manual tests conducted during the parametric test sequences, long-term continuous tests, and evaluations made by ABB utilizing laboratory analyses of deposit samples, basket materials, wash water, and by visual inspections. The following results were reported by ABB.

The thermal performance of the three air preheaters was measured by the drop in number of transfer units (Ntu) from the initial values to the final values. The thermal performance dropped by about 11%, 7%, and 25% for APHs A, B, and C respectively. The fact that APH B seemed to deteriorate less than the others is attributed to the fact that this air preheater received significantly less ammonia slip as an average over the project life (partially due to differences in reactor operating conditions) than did the other two air preheaters. It is believed that APH C deteriorated most because it was a recuperator, and was therefore more sensitive to a given amount of fouling than a regenerator.

As one would expect, the gas side pressure drops were more sensitive to the degree of fouling and plugging than were the Ntu values. In general, all three air preheaters showed steady increases in gas-side pressure drop (ΔP_g) during the test period, punctuated by occasional spikes which may have been caused by system upsets such as sootblower failure. In general, the high ΔP's could be reduced by aggressive cleaning methods, including sootblowing at 4 hour intervals, thorough water washing, and occasional increases in the gas outlet temperature. It was not possible, however, to maintain the original, clean ΔP of any of the air preheaters. The air and gas Euler (Eu) numbers (defined as $g_c(-dp)/\rho V^2$) for air preheater A increased by 145% and 115%, respectively, from the beginning of the test to the end. For air preheater B, the increases in both Eu numbers were in the 50-55% range.

Although the 3-layer air preheater appeared to perform better than the 2-layer air preheater, it cannot be concluded that the 3-layer design is superior to the 2-layer design. This is because the 2-layer air preheater received much more ammonia slip than the 3-layer air preheater -- possibly as much as four to five times more. Given this significant difference in operating condition, the 2-layer preheater performed remarkably well, and might logically have done better than the 3-layer design if the concentration of ammonia into the two regenerators had been equal.

The Q-pipe® air preheater seemed to steadily lose thermal performance with time during most of the test period. It is possible that some part of this performance loss may have been due to the loss of some of the air heater pipes in the unit as a result of the sootblower eroding the tube wall and the consequent loss of heat transfer fluid from those pipes. However, the majority of the drop in Ntu is believed to be attributable to fouling of the finned tubes on the gas side.

Corrosion tests were performed on various heat transfer surface materials used in APHs A and B. It was determined that enameled heat transfer surfaces should be used for Ljungstrom® air heaters when ammonia and sulfur compounds are both present in the gas stream. Other conclusions from the study were as follows.

1) Ammonium bisulfate or its corrosion products were shown to be a major constituent in the Ljungstrom® air preheater deposits at temperatures less than the ammonium bisulfate formation point.

2) The magnitude of the ammonia slip had a significant effect on the corrosion losses and deposit accumulations in the units.

3) The enameled heat transfer surface was an order of magnitude superior to the low-alloy corrosion resistant and carbon steel materials with regard to corrosion losses and deposit accumulations.

4) The enameled heat transfer surface exhibited superior cleanability compared to the low-alloy corrosion resistant and carbon steel materials.

Table 7 summarizes the changes in Ntu, air-side Euler number (Eu_a), and gas-side Euler number (Eu_g) for the three air preheaters. Worst case increases in Eu_g varied from 185% up to 320%. Worst case increases in air-side pressure drop (ΔP_a) varied from 245% up to 345%. Final air and gas Eu numbers were about 115-145% higher than initial values on air preheater A, and about 50-55% higher on air preheater B. This is roughly consistent with the declines in Ntu (final vs. initial) of 11% for air preheater A and only 7% for air preheater B, since the air preheater with the high increase in Eu (air preheater A) also had the greater loss of Ntu.

Table 7. Air Preheater Performance Data

Air Heater	Initial Value (I)	Worst Case Value (WC)	Final Case Value (F)	% Change WC vs. I	% Change F vs. I
Ntu values					
A	3.75	3.1	3.3	-17%	-12%
B	3.50	2.8	3.3*	-20%	-7%
C	2.40	1.4	1.8	-42%	-25%
Gas-side Eu Number					
A	35	140	75	+300%	+115%
B	28	118	43	+320%	+55%
C	63	180	135	+185%	+115%
Air-side Eu Number					
A	40	137	98	+245%	+145%
B	35	155	53	+345%	+50%

* Excludes questionable data from months 25-26

Miscellaneous tests

Waste Stream Impacts

A concern associated with the implementation of SCR technology is the effect that it may have on plant waste streams. This is thought to be primarily due to the adsorption of slip ammonia on fly ash. A special study was conducted to examine ammonia volatilization, ammonia extraction, and the effect of ammonia on metals extraction from fly ash.

Ammonia Extractability and Volatilization

Almost no ammonia volatilizes from SCR fly ash until a significant amount of water has been absorbed by the ash. A plausible mechanism for the apparent volatilization is that enough water must be gained by the ash to form a moist layer with a pH high enough to evolve gas-phase ammonia from the ammonium compounds on the ash. In closed-container experiments, nearly all of the ammonia on the ash evolved to the gas phase when wet. Ammonia concentrations in enclosed spaces depend on the ammonia concentration of the ash, the volume of air surrounding the ash and the presence of a humid atmosphere

Tests were performed to examine the extractability of ammonia from fly ash using aqueous solutions of various pH. It was determined that the extraction does depend upon solution pH. All or nearly all of the ammonia present was extracted in buffered solutions at pH 4.7, and pH 6.2, but not all was recovered in alkaline unbuffered extracts. In the pH 6.2 buffer, however, the completeness of extraction seemed to fall off somewhat as the ratio of ash to buffer increased. At 3g of ash per 50 ml of pH 6.2 buffer, the amount of ammonia extracted was about 200 μg/g whereas at 1 g per 50 ml, the amount was near 250 μg/g.

Tests were also performed to examine the particle size dependency of fly ash on ammonia adsorption using a series cyclone in-situ ash separation method. Ammonia concentration in the ash was much higher for the smaller particle sizes, but most of the total ammonia was found to reside with larger particles simply because these comprise the vast majority of the ash mass. The implication is that little slip ammonia will exit the process in the gas phase when high efficiency particulate emission controls are in place. This is due to much of the ammonia being in the solid phase at the air heater exit (≥ 50%) and to most of the ammonia being associated with the larger particle sizes which are most readily collected. In addition, the cooler temperatures downstream of the air preheater allow for easy deposition of ammonia by-products throughout the plant duct-work and in other pollution control processes (if any) prior to the stack.

Metal Extractability from SCR Fly Ash

Test facility ash samples were subjected to extraction with water, and extracts were analyzed for each of 28 metals. This was done to ascertain whether exposure of the fly ash to ammonia vapor caused an enhancement of the metal's extractabilities under conditions resembling those that might exist in an ash pond. The tested metals were Ag, Al, As, B, Ba, Be, Ca, Cd, Ce, Co, Cr, Cu, Fe, Hg, Mg, Mn, Mo, Ni, P, Pb, Rb, Sb, Se, Sn, Sr, Tl, V, and Zn. Of these 28 metals included in the study, only 17 could be detected in the fly-ash extracts (Ag, Be, Ce, Cu, Hg, Mn, Ni, Pb, Sn, Tl, and Fe were not detected). Of the 17 detectable metals, only barium underwent an increase in extractability following exposure to ammonia. The magnitude of the increases was found to depend directly on the magnitude of the NH_3/NO_x ratio in the SCR unit (ammonia slip), however, the increase was slight for all NH_3/NO_x ratios tested. Of the 16 additional metals that could be detected in the fly ash extracts, none displayed what were considered to be genuine enhancements in extractability, and several exhibited decreases in extractability as a result of exposure of the fly ash to ammonia.

TCLP Analysis

Toxicity Characteristics Leaching Procedure (TCLP) analyses were performed on fly ash samples from the test facility at various times throughout the project life. These samples were taken as grab samples from the cyclone ash hoppers from each of the high-dust SCR reactors. The reactors were operated at base-line conditions at least 12 hours prior to sampling. All hoppers were cleaned of ash prior to the period of base-line operation to insure that all ash acquired in the sampling was produced during reactor base-line operation. In general the TCLP analyses showed non-detectable amounts of metal species. In those cases where detectable amounts were present, little or no change was noted between the SCR hopper samples and the host unit reference sample. It was concluded from this study that the SCR process does not significantly affect the TCLP results of fly ash.

Ammonia Partitioning

As part of the three series of air preheater tests, the gas/solid phase partitioning of ammonia was determined. These tests showed that a large portion of the gas-phase ammonia present at the reactor exit adsorbed onto the fly ash as the flue gas was cooled through the air preheaters. In general, roughly one half of the total ammonia present was adsorbed onto the fly ash on a mass basis. Ash ammonia concentrations are normally reported on a ppm by weight (μg/g) basis as compared to gas-phase concentration reported on a ppm by volume basis. Using these conflicting units, the ash phase concentration was roughly 100 times that of the gas phase. In other words, an ammonia slip value of 1 ppm by volume gave roughly 100 μg/g (or ppm by weight) of ammonia on the fly ash downstream of the air preheater (this finding compares well with similar foreign experience). It should be cautioned that these results are highly variable and are strongly dependant on temperature regimes, fly ash mass loading, and fly ash composition as well as the ammonia slip level. However, solid-phase ammonia concentrations can be correlated to ammonia slip levels if large amounts of data are used to help prevent skewing of the results due process variations other than ammonia slip. This data is useful only in the determination of long term trends and would not be adequate for the evaluation of short-term slip such as during a parametric or compliance test.

ECONOMIC EVALUATION

The estimated capital and O&M costs from an economic evaluation of applying SCR technology at full scale to a new facility (coal-fired boiler utilizing high-sulfur, U.S. indigenous coal) are reported in this section. This information is not applicable to retrofit situations due to site-specific costs associated with space limitations, fan or air preheater upgrades, or major ductwork redesign. In addition, this economic analysis is not meant to supplant the need to perform site-specific financial analyses when evaluating SCR technology to a specific new facility. There will be project-specific constraints, sensitivity analyses, and market forces which no generalized economic analysis will be able to capture. Rather, the information reflects a macro-economic analysis of SCR technology based on historical data measured at the test facility, empirical laboratory data generated as a result of the test program, and consolidation of operational lessons learned, tempered with the knowledge of the current commercial market trends. This analysis is in a draft form and dollar values presented are likely to change during review and subsequent modification.

Base Case

The economic estimates for the base case evaluation are founded upon the application of a high-dust, hot-side SCR (i.e., located between the boiler economizer outlet and the air preheater inlet) to a new coal fired installation utilizing high-sulfur domestic coal. The technical design premises used to prepare the economic analysis were selected to be representative of actual or anticipated plant configurations and NO_x control requirements currently being permitted or likely to be permitted on coal-fired boilers in the United States. Thus, defining assumptions were selected in an effort to have broad utility applicability.

The base case represents a new, base-load 250 MW pulverized-coal power plant typical of the majority of new coal-fired projects currently under development, construction, or recently declared in commercial operation. The unit size of 250 MW is consistent with future trends of new domestic power plants. The plant is located in a rural area with minimal space limitations. The fuel is a high-sulfur bituminous Illinois No. 6 coal.

The plant will utilize a single, balanced-draft, pulverized-coal fired boiler complete with all required auxiliary equipment. The boiler will be designed to produce approximately 1,610,000 lb/hr of main steam at turbine inlet conditions of 2400 psig and 1000 °F. Utilizing current generation low NO_x combustion systems, the boiler is assumed to produce a NO_x emission rate of

0.35 lb/MBTU. For purposes of this study, it is assumed that tangentially-fired boilers and wall-fired boilers are interchangeable with respect to all thermal performance and flue gas constituents.

The flue gas exits the boiler and enters a single, hot-side SCR. Flue gas flow is vertically downward through the reactor. The SCR is located directly above the air preheater and is designed as a "universal" reactor able to accept either (or both) plate type catalyst and honeycomb type catalyst. Ammonia injection will utilize stand-alone dilution air fans rather than combustion air.

A single, trisector, Ljungstrom® regenerative air preheater will be utilized to reclaim heat from the flue gas stream and transfer that heat to the primary and secondary air. Physical features of the air preheater are typical of what is commercially offered as a "deNO$_x$" air preheater. The heater transfer surface arrangement will include hot, intermediate, and cold sections.

Sulfur dioxide will be removed using a lime spray dryer FGD system. A reverse gas, fabric filter baghouse will be used which will collect the dried reaction products from the spray dryer as well as the fly ash produced in the boiler by the combustion of coal.

The SCR reactor for this analysis includes three catalyst support layers plus a flow straightener (dummy bed). At time zero, two of the three catalyst support layers are loaded with catalyst. To optimize catalyst life, a spare (empty) catalyst support elevation inside the reactor is provided. The spare layer allows catalyst suppliers to develop optimized catalyst management plans which increase catalyst utilization. Thus, a fresh catalyst layer can be added to the reactor after the guarantee period when the ammonia slip begins to exceed the guaranteed limit. The activity of the new catalyst combined with the residual activity of the existing catalyst restores the performance of the SCR and extends the next addition/replacement outage beyond the initial guarantee interval.

The catalyst management plan developed for the economic analysis is based on composite empirical laboratory data developed on the test facility catalysts and uses a 16,000 hour (2 year) catalyst life guarantee period. After the initial guarantee period of 2 years, a new layer of catalyst is added to the reactor, thus, taking advantage of the residual activity in the initial layers to "boost" the performance of the SCR. The next addition of catalyst is required in project year 6 when one of the initial layers is replaced. Thus, catalyst is added and/or replaced after 2, 6, 9, 12, 15, 18, 21, 24, and 27 years for a total of 9 times during the 30 year evaluation period. It should be noted that catalyst life projections of 30 years are speculative at best due to uncertainty of catalyst structural properties with long-term exposure to flue gas and thermal cycling.

The base case design criteria are shown below in Table 8. Also included in the table are charge and levelization factors as well as fixed and variable O & M assumptions and unit costs.

Table 8. Base Case Design Criteria

Parameter	Specification
Type of SCR	Hot-Side
Number of SCR Reactors	One
Reactor Configuration	3 catalyst support layers + 1 dummy layer
Initial Catalyst Load	2 of 3 layers loaded, 1 spare layer
Required Range of Operation	35% to 100% boiler load
NOx concentration at Inlet	0.35 lb/MBtu
Design NOx reduction	60%
Flue Gas Temperature at SCR Inlet	700 °F
Flue Gas Pressure at SCR Inlet	-5 in. W.G.
Design Ammonia Slip	5 ppm
Guaranteed Catalyst Life	2 years (16,000 hours)
SO_2 Oxidation	0.75% (initial catalyst load)
Maximum Pressure Drop	6 in. W.G. (fully loaded reactor)
Velocity Distribution	$\Delta V/V_{mean}$ <10% over 90% of reactor area $\Delta V/V_{mean}$ <20% over remaining 10% of area
Ammonia Distribution	$\Delta C/C_{mean}$ <10%
Temperature Distribution	ΔT < 10 °C max. deviation from mean
Anhydrous Ammonia Cost	$ 250/ton
SCR Catalyst Cost	$ 400/ft^3
SCR Catalyst Guarantee Period	2 years
SCR Catalyst Escalation	3.0%
Power Cost	30 mills/kWh
ID Fan Efficiency	75%
SCR Draft Loss (fully loaded reactor)	3.0 in. W.G.
Ductwork Draft Loss	0.75 in. W.G.
Ammonia Injection Grid Draft Loss	0.75 in. W.G.
Unrecoverable Air Preheater Draft Loss	1.0 in. W.G.
Operating Labor Man-Hour Rate	$ 23/hr
Maintenance Factor (% of total process capital)	2.0%
Current Dollar Analysis (1996 dollars):	
Capital Charge Factor	0.150
O & M Cost Levelization Factor	1.362
Constant Dollar Analysis	
Capital Charge Factor	0.116
O & M Cost Levelization Factor	1.000

Capital, O&M, and Levelized Cost vs. Unit Size

In order to examine the change in SCR costs vs. unit size, additional capital and O&M estimates were repeated for a 125 MW unit and 700 MW unit. To maintain consistency with the 250 MW base case unit, an SCR removal efficiency of 60% NO_x reduction was assumed. Where possible, consistent (or identical) assumptions were made with regard to the 125 MW and 700 MW units.

Results are shown in Table 9 for the base case (250 MW), 125 MW, and 700 MW unit sizes for an SCR with a NO_x removal efficiency of 60%. On a $/kw vs. unit size basis, the total capital requirement of the SCR system shows a trend of decreasing unit cost ($/kw) with increasing unit size indicating significant economies of scale. Total capital requirement ranges from $61/kw for the 125 MW unit to $45/kw for the 700 MW unit.

Table 9. Capital, O&M, and Levelized Cost vs. Unit Size
(60 % NOx Removal)

Unit Size	125 MW	250 MW	700 MW
Total capital requirement ($)	7,602,000	13,415,000	31,327,000
Total capital requirement ($/kw)	61	54	45
First year fixed operating cost ($)	213,000	312,000	614,000
First year variable operating cost ($)	367,000	733,000	2,053,000
Current Dollar Analysis			
Levelized Cost (mills/kWh)	2.89	2.57	2.22
Levelized Cost ($/ton)	2,811	2,500	2,165
Constant Dollar Analysis			
Levelized Cost (mills/kWh)	2.09	1.85	1.59
Levelized Cost ($/ton)	2,037	1,802	1,547

Capital, O&M, and Levelized Cost vs. NO_x Removal Efficiency

In addition to the 250 MW base case NO_x removal efficiency of 60%, two additional NO_x reduction cases at 40% and 80% were calculated to examine the impact on cost. Tabular results of this analysis are shown below in Table 10.

Table 10. Capital, O&M, and Levelized Cost vs. NO_x Removal Efficiency
(250 MW Unit Size)

NO_x Reduction	40%	60%	80%
Total capital requirement ($)	12,974,000	13,415,000	14,142,000
Total capital requirement ($/kw)	52	54	57
First year fixed operating cost ($)	305,000	312,000	324,000
First year variable operating cost ($)	621,000	733,000	857,000
Current Dollar Analysis			
Levelized Cost (mills/kWh)	2.39	2.57	2.79
Levelized Cost ($/ton)	3,502	2,500	2,036
Constant Dollar Analysis			
Levelized Cost (mills/kWh)	1.74	1.85	2.00
Levelized Cost ($/ton)	2,536	1,802	1,460

O&M Cost vs. Inlet NO_x Concentration

Many new boiler installations face difficult decisions on how to best optimize overall NO_x reduction requirements using a combination of low NO_x burners and SCR. While maximizing combustion NO_x reductions can allow lower SCR variable O&M costs, it typically comes at the expense of increased LOI in the fly ash, and hence, lower plant cycle efficiency. Optimizing the burners to minimize LOI often leads to higher NO_x concentrations entering the SCR and, hence, higher variable O&M costs to achieve a permitted outlet NO_x emission limit. Results showing levelized cost vs. SCR inlet NO_x concentration for a 250 MW unit operating at 60% NO_x removal are summarized in Table 11.

Table 11. Levelized Cost vs. SCR Inlet NO_x Concentration

(250 MW @ 60% NO_x Removal)

Inlet NO_x (lb/MMBtu)	**.45**	**.40**	**.35**	**.30**	**.25**
Current Dollar Analysis					
Levelized Cost (mills/kWh)	2.61	2.59	2.57	2.55	2.53
Levelized Cost ($/ton)	1,977	2,205	2,500	2,894	3,446
Constant Dollar Analysis					
Levelized Cost (mills/kWh)	1.88	1.87	1.85	1.84	1.82
Levelized Cost ($/ton)	1,425	1,590	1,802	2,086	2,483

CONCLUSIONS

SCR Catalysts

All of the catalysts in the test program proved to be acceptable for the application and met design specifications. Significant differences, however, were noted in catalyst parameters such as volume, weight, activity (both $deNO_x$ and SO_2 oxidation), fouling, and pressure drop. No one catalyst can be considered superior since each may possess advantages particular to the characteristics of a proposed installation. For instance, pressure drop may not be a critical factor in a new-plant installation, but may be critical in the catalyst selection for a retrofit situation where a fan retrofit lies in the balance. Similarly, weight may be the controlling factor for an in-duct installation, while other installations may be particularly concerned with SO_2 oxidation. In short, careful examination of catalyst characteristics are necessary to choose the optimum catalyst for a particular installation.

Deactivation

The deactivation trends of the catalysts were within expected ranges based on operating experience in European and Japanese installations. Catalyst poison concentrations were somewhat higher than prior experience. However, no unusual acceleration in deactivation was noted and it appears that at least for the coals tested during this project, the resulting deactivation is similar in significance to other world-wide installations.

Volume/DeNO$_x$ Activity

Total catalyst volume and weight are strong functions not only of catalyst design (surface area, activity, etc.), but of installed margin. In some cases catalyst suppliers chose to minimize catalyst volume by designing a system which met, but did not exceed performance requirements. In other cases, catalyst suppliers chose to increase catalyst volume to insure performance margin and thus improve performance. Honeycomb catalysts are generally considered to have higher surface area when compare to plate-type catalysts (on a bulk volume basis) and would thus generally require less volume (although weight may be greater). However, offerings in the program have demonstrated the ability of plate-type catalyst to meet specifications with similar or less bulk volume than honeycomb configurations. Thus, it appears that installed catalyst design margin can be as significant to bulk catalyst volume as is the basic geometrical (honeycomb/plate) design.

SO_2 Oxidation

While some catalysts essentially met the SO_2 oxidation requirement, others improved upon it greatly. In practice some benefits may be realized by decreasing SO_2 oxidation past some arbitrarily chosen point, but those decreases are generally made at the expense of increasing catalyst volume. As with deNO$_x$ activity, the SO_2 oxidation design margin chosen by the individual catalyst suppliers likely contributed to the overall catalyst volume. It is difficult to determine the point of diminishing return for decreasing SO_2 oxidation as little full scale long-term operational data is available for U.S. boilers. Differences in SO_2 oxidation characteristics were very apparent between the catalysts. However, these differences do not necessarily denote a difference in catalyst design expertise between suppliers, but more likely reflect a philosophical decision as to the catalyst offering made. In practice, all suppliers would likely be able to meet a particular application's SO_2 oxidation requirements.

Pressure Drop

Catalyst pressure drops are generally a function of catalyst geometry and volume. Honeycomb catalysts commonly have less open area than plate catalysts, resulting in increased pressure drop per unit length of catalyst. In some cases, this is mitigated partially by less volume of honeycomb catalyst required compared to plate catalysts. In addition, honeycomb pressure drops can be modified by adjusting the wall thickness of the catalyst. In the case of the Siemens plate, the pressure drop was considerably lower than the other catalysts especially considering the fact that Siemens utilized only two catalyst beds to meet the program requirements without substantial margin. On a per volume basis, the Haldor and Hitachi catalysts also had low pressure drop, but the increased volume, compared to the Siemens catalyst, resulted in an overall pressure drop more similar to the honeycomb catalysts in the program. Thus, the overall reactor pressure drop must be considered when evaluating catalysts and neither the basic geometry nor the catalyst volume are exclusively controlling.

Fouling and Erosion

The fouling characteristics of SCR catalysts are important to the successful long-term operation of the reactor. In practice, the pressure drop across a catalyst increases slightly from new conditions once the catalyst is placed in service. The pressure loss, however, should remain relatively consistent after the initial start-up of the reactor. Steadily increasing pressure drops over time indicate that a catalyst is either improperly designed for the application or that appropriate sootblowing steps are not being taken. Small reactors similar to those present in the test facility represent worst case scenarios for catalyst fouling due to the wall effects and space constraints preventing optimum sootblowing. However, the test facility pressure drop data show relatively level pressure drop over time for the catalysts. This indicates that the sootblowing procedures used in the test facility were effective. The plate-type configurations exhibited somewhat less fouling potential than did the honeycomb configurations (based on visual inspection) although quantification is difficult, but both configurations were acceptable for the application.

Erosion of the test facility catalysts was not considered to be a significant problem. Most of the erosion in the test facility is thought to be due at least in part to the aggressive sootblowing operations. The primary conclusion is that erosion is likely not the controlling factor in catalyst life. However, catalyst design does play a major role in erosion susceptibility. In both honeycomb and plate catalysts, the substrate support material as well as the primary catalytically active material contribute to the overall erosion resistance of the particular offering.

Air Preheater Performance

As expected, the study concluded that the SCR process exacerbates performance degradation of air preheaters mainly due to ammonia slip and subsequent by-product formation. It is recommended that enameled heat transfer surfaces be used to aid in corrosion resistance and cleaning. Air preheater performance is likely a direct function of the magnitude of ammonia slip, however, it is undetermined if deleterious effects have an ammonia slip threshold value. A comparison of regenerator versus recuperator type air preheaters showed that regenerators tend to outperform recuperators in SCR applications in terms of both thermal performance and fouling.

Low/High Dust Configuration

The installation of the reactor J low-dust configuration allowed a comparison of low-dust to high-dust applications. The test results were somewhat inconclusive due to fouling problems associated with the design of the test facility low-dust reactor and extraction scoop, biasing the physical testing. The results did show that catalyst parameters such as deactivation, NO_x reduction capability, and SO_2 oxidation were similar to the high-dust configuration. The results also showed that comparable NO_x reduction could be accomplished with significantly less volume of catalyst (primarily due to the higher specific surface area of the low-dust catalyst). Physical characteristics such as fouling and pressure drop, however, were much more difficult to determine. The results showed that a low-dust configuration could be very sensitive to upsets in boiler/ESP operation. The small channel openings in the low-dust configuration present a particular problem during upset conditions where large amounts of particulate are forced through the catalyst. Since it appears that relatively short periods of operation at such upset conditions could be catastrophic to the catalyst, process design must unfortunately focus on mitigating the effects of upset conditions. It is unclear as to the degree that upsets would be experienced on full scale installations as compared to the test facility experience. It is also unclear how these full-scale upsets would compare in severity to the test facility upsets. The study basically concludes that the low-dust configuration worked well for the conditions for which it was designed, but to the degree that excursions from these design conditions occur, the catalyst could be adversely impacted.

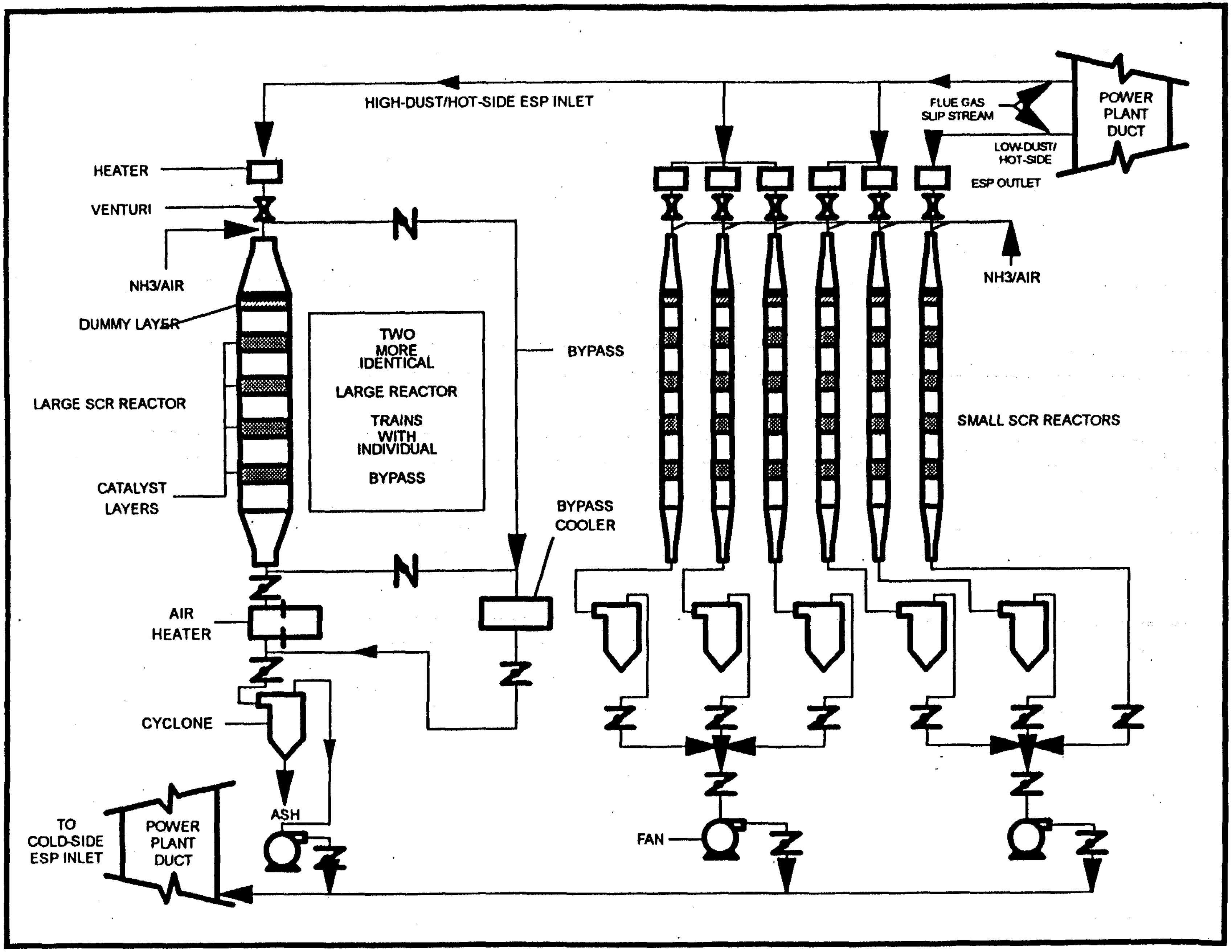

Figure 1. SCR Demonstration Facility-Process Flow Diagram

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Figure 2. Catalyst Bed Pressure Drop vs. Project Date

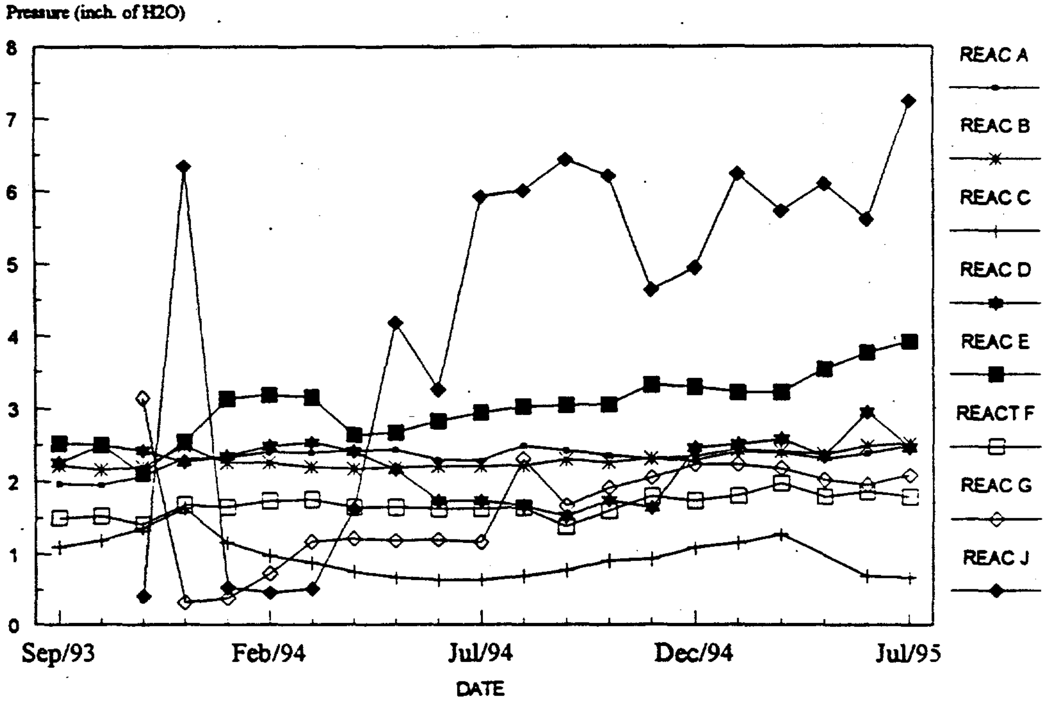


Figure 3. K/Ko vs. Exposure Time

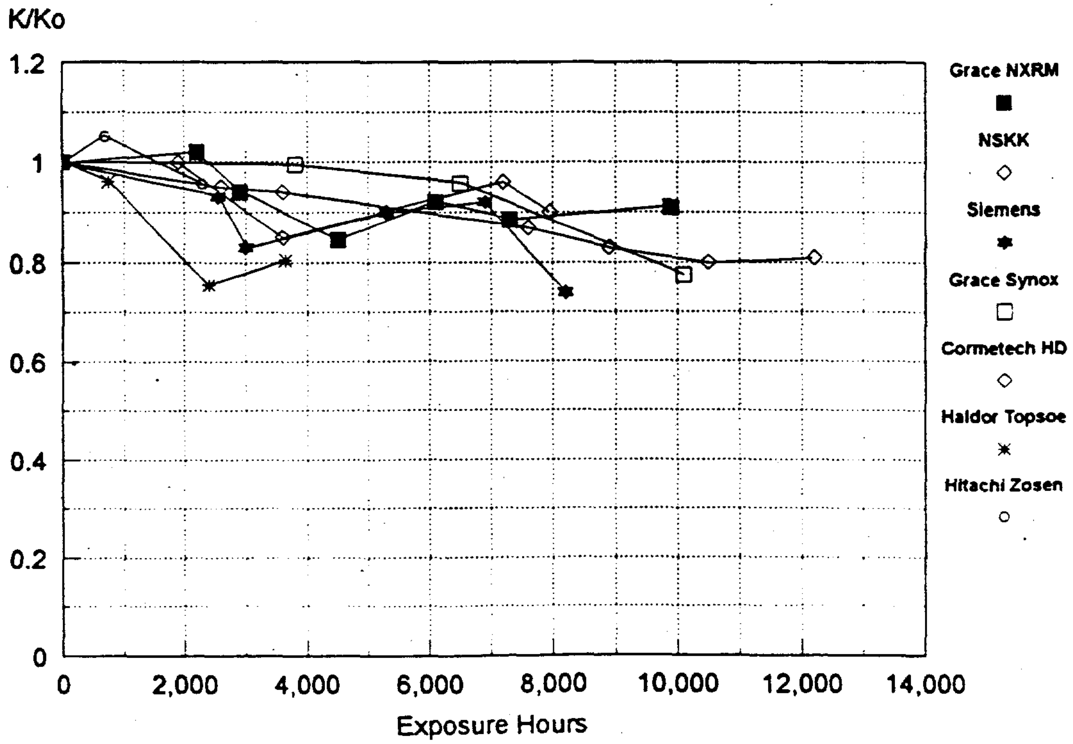


Figure 4. Intermediate NOx Reduction and Ammonia Slip vs. Time

(base-line conditions)

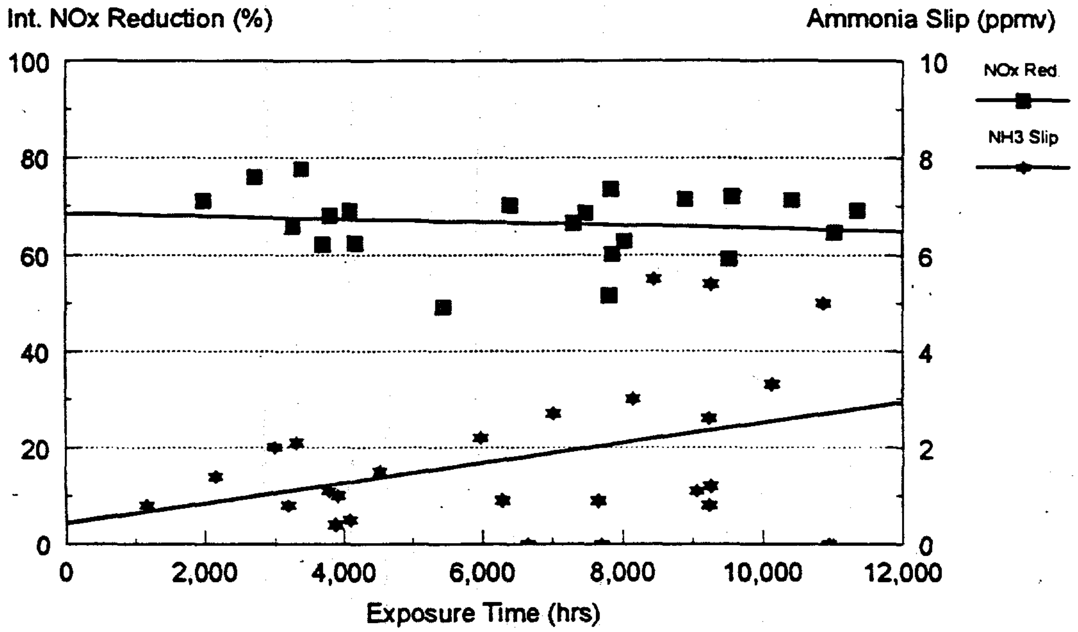
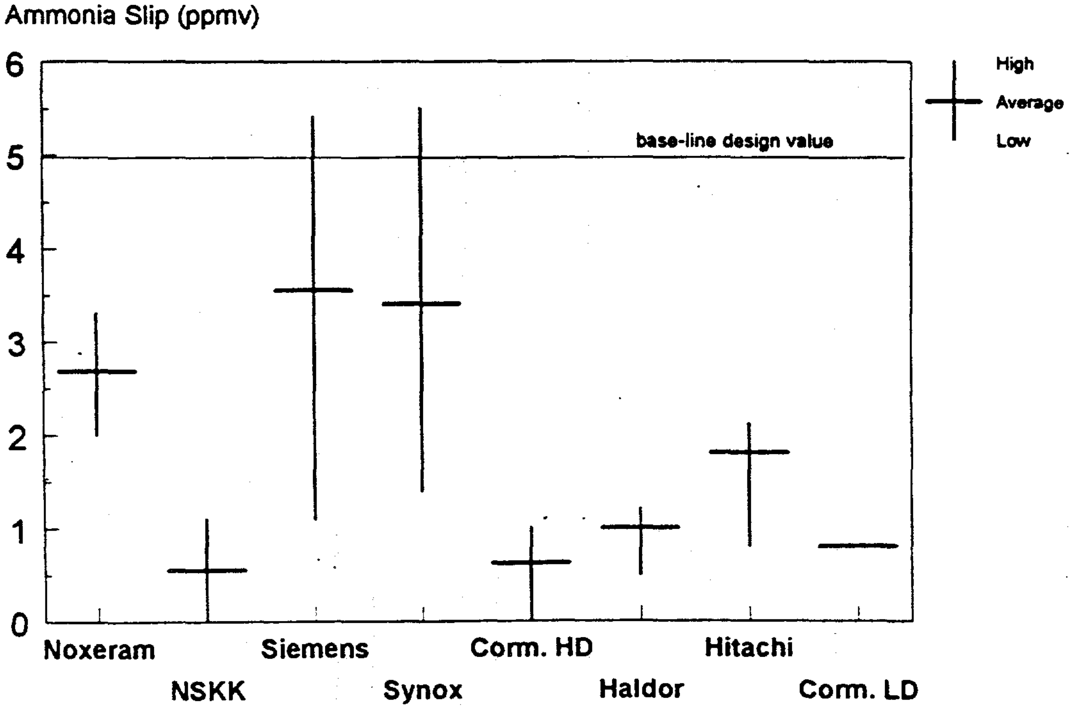


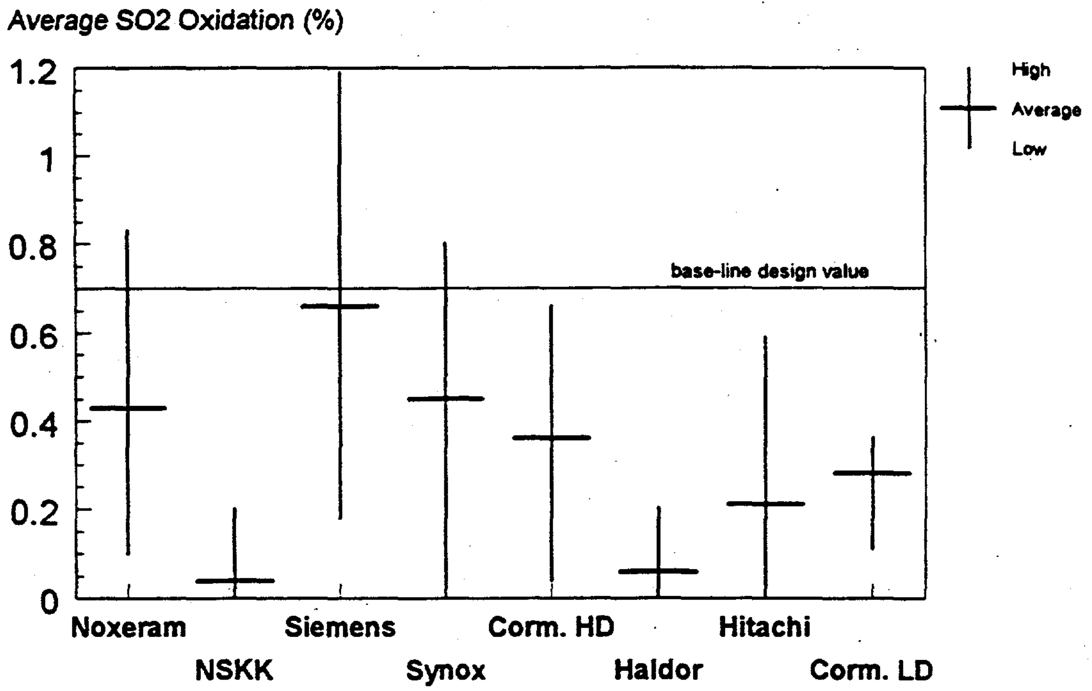
Figure 5. Average Base-line Ammonia Slip



NH3/NOx=0.8, 700 F, Base-line Flow

Figure 6. Average SO2 Oxidation Rate

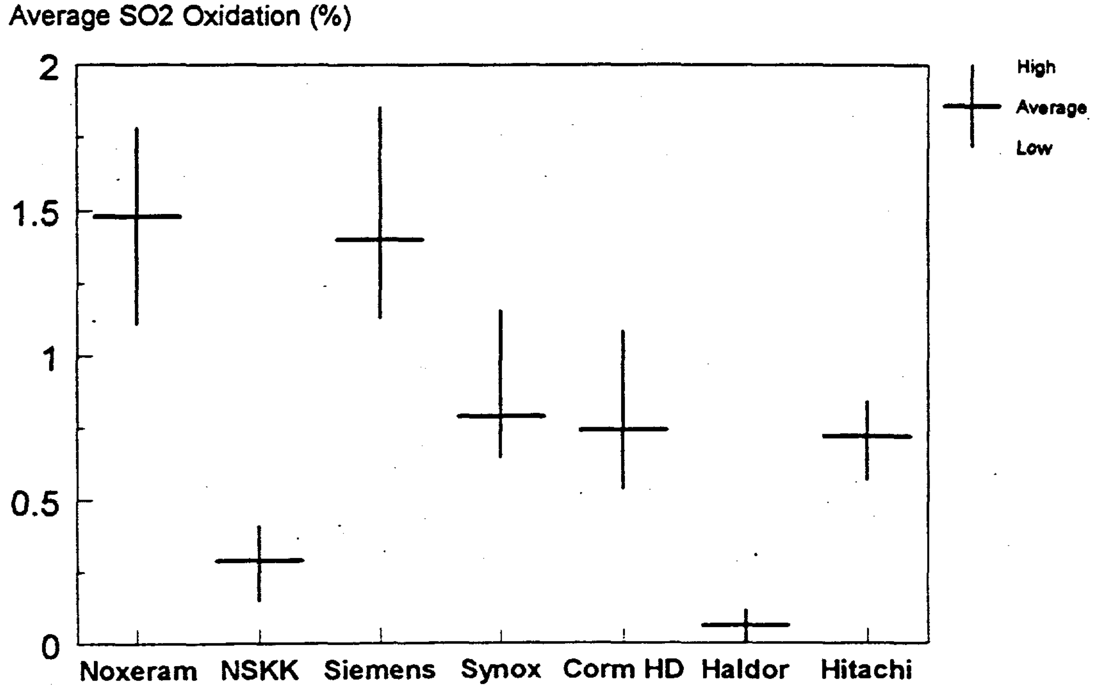
(Base-Line)



NH3/NOx=0.8, 700 F, Design Flow

Figure 7. Average SO2 Oxidation Rate

(High Temperature)



NH3/NOx=0.8, High Temp, Design Flow

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This project has been a collective effort on the part of many individuals and organizations too numerous to name individually. In lieu of a detailed listing, we would like to recognize the following organizations and their employees for their contribution to the success of this project; U.S. Department of Energy, Southern Company Services, Inc., Electric Power Research Institute, Ontario Hydro, Gulf Power Company, Southern Research Institute, ABB Air Preheater, Inc., Spectrum Systems, Inc., ICS, Inc., Cormetech, Inc., Haldor Topsoe A/S, Hitachi Zosen, Nippon Shokobai, Siemens, and W.R. Grace & Co.

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500 MW DEMONSTRATION OF ADVANCED WALL-FIRED COMBUSTION TECHNIQUES FOR THE REDUCTION OF NITROGEN OXIDE EMISSIONS FROM COAL-FIRED BOILERS

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ABSTRACT

This paper discusses the technical progress of a U. S. Department of Energy Innovative Clean Coal Technology project demonstrating advanced wall-fired combustion techniques for the reduction of nitrogen oxide (NOx) emissions from coal-fired boilers. The primary objective of the demonstration is to determine the long-term NOx reduction performance of advanced overfire air (AOFA), low NOx burners (LNB), and advanced digital control/optimization methodologies applied in a stepwise fashion to a 500 MW boiler. The focus of this paper is to report (1) on the installation of three on-line carbon-in-ash monitors and (2) the design and results to date from the advanced digital control/optimization phase of the project.

INTRODUCTION

This paper discusses the technical progress of one of the U. S. Department of Energy's Innovative Clean Coal Technology (ICCT) projects demonstrating advanced combustion techniques for the reduction of nitrogen oxide (NOx) emissions from wall-fired boilers. The demonstration is being conducted at Georgia Power Company's Plant Hammond Unit 4, a 500 MW, pre-NSPS (New Source Performance Standards), wall-fired boiler. Plant Hammond is located near Rome, Georgia, northwest of Atlanta.

The Hammond project is being managed by Southern Company Services, Inc. (SCS) on behalf of the project co-funders: the Southern Company, the U. S. Department of Energy (DOE), and the Electric Power Research Institute (EPRI). In addition to SCS, Southern includes the five electric operating companies: Alabama Power, Georgia Power, Gulf Power, Mississippi Power, and Savannah Electric and Power. SCS provides engineering and research services to the Southern electric system. The ICCT program is a jointly funded effort between DOE and industry to move the most promising advanced coal-based technologies to the commercial marketplace. The goal of ICCT projects is the demonstration of commercially feasible, advanced coal-based technologies that have already reached the "proof-of-concept" stage. The ICCT projects are jointly funded endeavors between the government and the private sector in which the industrial participant contributes at least 50 percent of the total project cost. The DOE is participating through the Office of Clean Coal Technology at the Pittsburgh Energy Technology Center (PETC).

The primary objective of the demonstration is to determine the long-term NOx reduction performance of advanced overfire air (AOFA), low NOx burners (LNB), and advanced digital control/optimization methodologies applied in a stepwise fashion to a 500 MW boiler. Short-term tests of each technology are also being performed to provide engineering information about emissions and performance trends [1,2,3,4].

Following a brief unit and technology review, this paper focuses on the design and results to date from the advanced digital control/optimization phase of the project.

UNIT AND TECHNOLOGY REVIEW

Georgia Power Company's Plant Hammond Unit 4 is a Foster Wheeler Energy Corporation (FWEC) opposed wall-fired boiler, rated at 500 MW gross, with design steam conditions of 2500 psig and 1000/1000°F superheat/reheat temperatures, respectively. The unit was placed into commercial operation on December 14, 1970. Prior to the LNB retrofit in 1991, six FWEC Planetary Roller and Table type mills provided pulverized eastern bituminous coal (12,900 Btu/lb, 33% VM, 53% FC, 72% C, 1.7% S, 1.4% N, 10% ash) to 24 pre-NSPS, Intervane burners. The burners are arranged in a matrix of 12 burners (4W x 3H) on opposing walls with each mill supplying coal to four burners per elevation (Figure 1).

During a spring 1991 unit outage, the Intervane burners were replaced with FWEC Controlled Flow/Split Flame (CF/SF) burners. In the CF/SF burner, secondary combustion air is divided between inner and outer flow cylinders. A sliding sleeve damper regulates the total secondary air flow entering the burner and is used to balance the burner air flow distribution. An adjustable outer register assembly divides the burner's secondary air into two concentric paths and also imparts some swirl to the air streams. The secondary air that traverses the inner path, flows across an adjustable inner register assembly that, by providing a variable pressure drop, apportions the flow between the inner and outer flow paths. The inner register also controls the degree of additional swirl imparted to the coal/air mixture in the near throat region. The outer air flow enters the furnace axially, providing the remaining air necessary to complete combustion. An axially movable inner sleeve tip provides a means for varying the primary air velocity while maintaining a constant primary flow. The split flame nozzle segregates the coal/air mixture into four concentrated streams, each of which forms an individual flame when entering the furnace. This segregation minimizes mixing between the coal and the primary air, assisting in the staged combustion process.

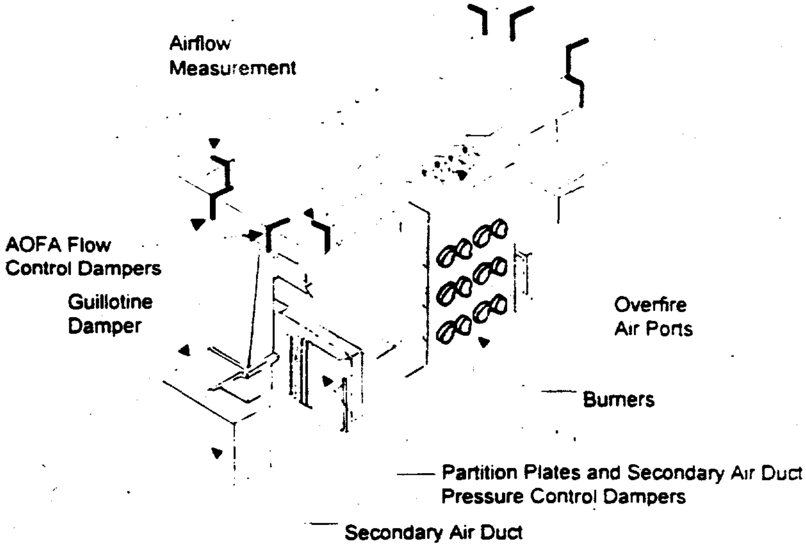

Figure 1. Hammond Unit 4 Furnace Layout

As part of this demonstration project, the unit was also retrofit with an Advanced Overfire Air (AOFA) system. The FWEC design diverts air from the secondary air ductwork and incorporates four flow control dampers at the corners of the overfire air windbox and four overfire air ports on both the front and rear furnace walls. Due to budgetary and physical constraints, FWEC designed an eight port AOFA system more suitable to the project and unit than the twelve port system originally proposed.

The Unit 4 boiler was designed for pressurized furnace operation but was converted to balanced draft operation in 1977. The unit is equipped with a coldside ESP and utilizes two regenerative secondary air preheaters and two regenerative primary air heaters. During the course of the ICCT demonstration, the unit was retrofitted with six Babcock & Wilcox MPS 75 mills (two each during the spring 1991, spring 1992, and fall 1993 outages).

REVIEW OF PRIOR TESTING

Baseline, AOFA, LNB, and LNB+AOFA test phases have been completed (Table 1). Short-term and long-term baseline testing was conducted in an "as-found" condition from November 1989 through March 1990. Following retrofit of the AOFA system during a four-week outage in spring 1990, the AOFA configuration was tested from August 1990 through March 1991. The

FWEC CF/SF low NOx burners were then installed during a seven week outage starting on March 8, 1991 and continuing to May 5, 1991. Following optimization of the LNBs and ancillary combustion equipment by FWEC personnel, LNB testing was commenced during July 1991 and continued until January 1992. Testing in the LNB+AOFA configuration was completed during August 1993. During both the LNB and LNB+AOFA, there were significant increases (when compared to baseline) in precipitator fly ash loading and gas flow rate and also, increases in fly ash LOI which adversely impacted stack particulate emissions and forced the unit to be load limited [5].

Table 1. Project Schedule

Phase	Description	Date	Status
0	Pre-Award Negotiations		
1	Baseline Characterization	8/89 - 4/90	Completed
2	Advanced Overfire Air Retrofit (AOFA) & Characterization	4/90 - 3/91	Completed
3A	Low NOx Burner Retrofit (LNB) & Characterization	3/91 - 1/92	Completed
3B	LNB+AOFA Characterization	1/92 - 8/93	Completed
4	Digital Controls/Optimization Retrofit & Characterization	8/93 - 12/96*	In Progress
5	Final Reporting and Disposition	9/95 - 12/96*	In Progress

* Indicates change from original work breakdown structure. Final schedule dependent upon availability of unit.

A summary of the baseline, AOFA, LNB, and LNB+AOFA long-term NOx emissions data for Hammond Unit 4 are shown in Figure 2. Baseline testing was performed in an "as-found" condition. For the AOFA, LNB, and LNB+AOFA test phases, following optimization of the unit by FWEC personnel, the unit was operated according to FWEC instructions provided in the design manuals. As shown, the AOFA, LNBs, and LNB+AOFA provide a long-term *full load* NOx reduction of 24, 48, and 68 percent, respectively. The *load-weighted* average of NOx emissions reductions was 14, 48, and 63 percent, respectively, for AOFA, LNBs, and LNB+AOFA test phases. Although the LNB plus AOFA NOx level represents a 67 percent reduction from baseline levels, a substantial portion of the incremental change in NOx emissions between the LNB and LNB+AOFA configurations is the result of operational changes and is not the result of the AOFA system [6].

The *time-weighted* average of NOx emissions for the baseline, AOFA, LNB, LNB+AOFA test phases are shown in Table 2. Since NOx emissions are generally dependent on unit load, the NOx values shown in this table are influenced by the load dispatch of the unit during the corresponding test frame. Also shown in this table are the 30 day and annual achievable emission limits as determined during these test periods. The 30-day rolling average achievable emission limit is defined as the value that will be exceeded, on average, no more than one time per ten years. For the annual average, a compliance level of 95 percent was used in the calculation.

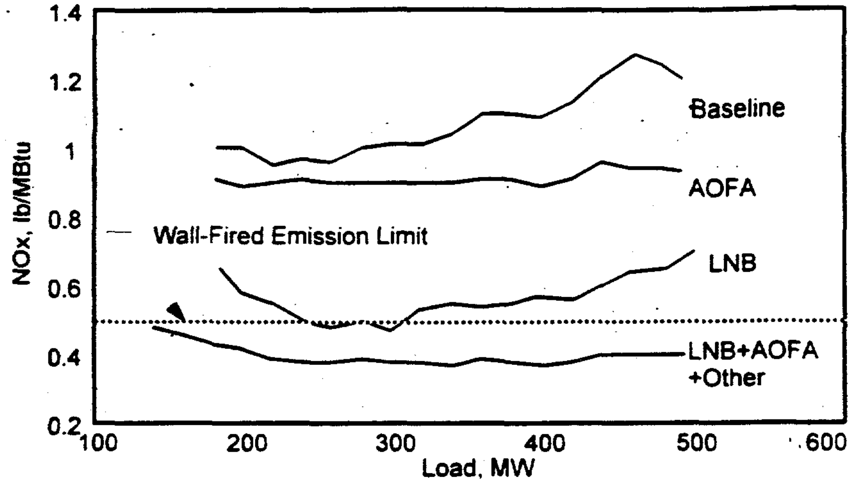

Figure 2. Long-Term NOx Emissions vs. Load Characteristic

Table 2. Long-Term NOx Emissions

Unit Configuration →	Baseline		AOFA		LNB		LNB-AOFA	
Parameter ↓	Mean	RSD,%	Mean	RSD,%	Mean	RSD,%	Mean	RSD,%
Number of Daily Avg. Values	52	-	86	-	94	-	63	-
Load (MW)	407	9.4	386	17.9	305	17.7	293	23.9
NOx Emissions (lb/MBtu)	1.12	9.5	0.92	8.6	0.53	13.7	0.41	12.9
O2 Level (percent at stack)	5.8	11.7	7.3	12.6	8.4	7.7	8.73	16.3
NOx 30 Day AEL (lb/MBtu)	1.24	-	1.03	-	0.64	-	0.51	-
NOx Annual AEL (lb MBtu)	1.13	-	0.93	-	0.55	-	0.42	-

AEL = Achievable Emission Limit; RSD = Relative Standard Deviation

EVALUATION OF ON-LINE CARBON-IN-ASH ANALYZERS

A subsidiary goal of the Wall-Fired project is the evaluation of advanced instrumentation as applied to combustion control. Based on this goal, several on-line carbon-in-ash monitors are being evaluated as to their:

- Reliability and maintenance,
- Accuracy and repeatability, and
- Suitability for use in the control strategies being demonstrated at Hammond Unit 4.

This instrumentation has the potential to allow faster response to changes in boiler conditions, resulting in benefits to boiler efficiency and ash monitoring.

Three units are currently installed at this site: (1) Applied Synergistics FOCUS, (2) CAMRAC Corporation CAM, and (3) Clyde-Sturtevant SEKAM. The SEKAM unit samples from two locations at the economizer outlet while the CAM unit samples from a single location at the precipitator inlet. The FOCUS unit is a non-extractive system that utilizes two cameras located above the nose of the furnace. The following paragraphs briefly describe these devices.

CAMRAC

The CAM (Carbon-in-Ash-Monitor) was developed by GAI Consultants during the 1980's for the CAMRAC company. Financial support was provided by several utilities throughout the United States as well as the Electric Power Research Institute. This instrument offers automatic monitoring of unburned carbon in combustion products. As of June 1996, a total of 5 units have been installed at various locations. The system has been tested on Georgia Power Company's Plant Hammond Unit 4 as well as Alabama Power Company's Plant Gaston Unit 4. Other locations outside the Southern Company include Pennsylvania Electric Company's Conemaugh Station. Allegheny Power Systems' Harrison Station, Philadelphia Electric's Eddystone Station and Duquesne Light's Cheswick Station.

The CAM's operating principle is based on the microwave absorptive properties of carbon particles in ash. Approximately 5 grams of fly ash are automatically extracted isokinetically from the duct and placed in a small collection cell. Microwaves at a frequency of 2450 MHz are passed through the collection cell. A power level of less than 150 milliwatts is used. Relative microwave absorbance between carbon and carbon-free ash is used to determine the carbon content of the sample. In other words. the power into the collection cell minus the transmitted and reflected power is equivalent to the power absorbed by carbon in the ash sample. An internal calibration curve is used to convert this absorbance to percent carbon. According to the manufacturer. the measured carbon level is independent of the coal being burned. The system sends results of the sample analysis to the plant control room for combustion performance optimization. Following analysis of the sample. the sample is returned to the combustion duct.

Although the unit can accommodate extraction locations on the cold gas side of the air heaters. CAMRAC's preferred sampling location is between the economizer and the air heaters. An adaptive sampler is operated by CAM so that isokinetic sampling conditions are maintained at all load levels. For single point sampling, the collection cell is purged and a new ash sample is collected for analysis every five to ten minutes. Multipoint designs with up to ten samplers are available for additional accuracy in duct LOI characterization. although sample cycle time is greater for multipoint sampling.

Figure 3 shows the CAM unit as installed at Hammond 4. A schematic of a typical CAM arrangement is shown as Figure 4. Table 3 provides other aspects of the system including size. accuracy and cost. These specifications were taken from CAM material and system dimensions.

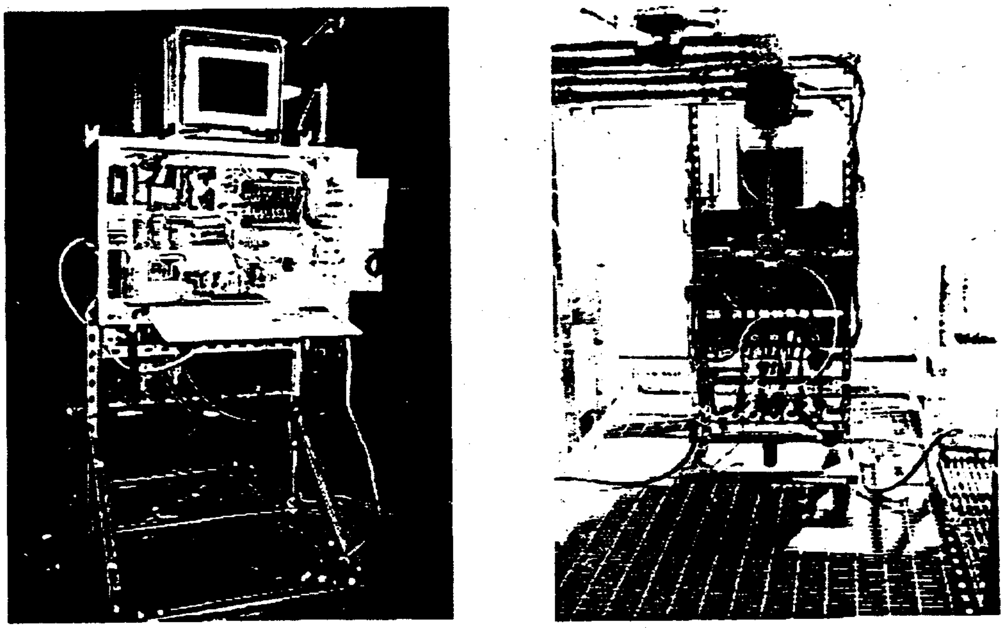

Figure 3. CAM On-Line Analyzer

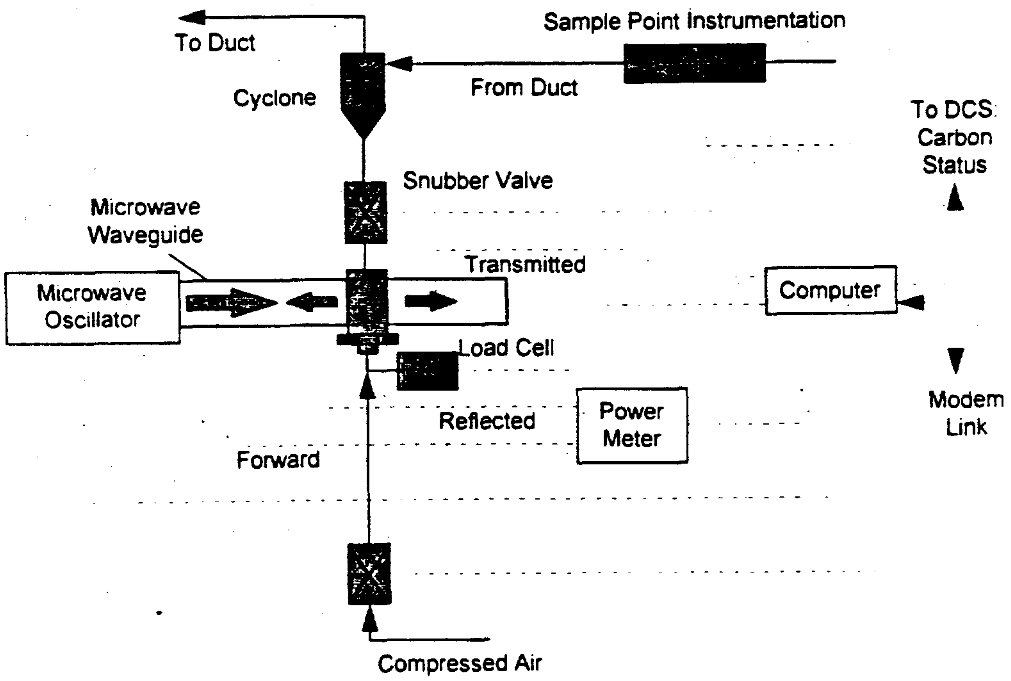

Figure 4. CAM General Arrangement

Table 3. CAM Performance Parameters

Carbon-in-Ash-Monitor (CAM)	
Operating Principle	Absorbance of microwave energy
Instrument Size (WxDxH)	8 ft x 1 ft x 6 ft
Mobility	instrumentation: medium sampling device: medium
Sample Size	~5 grams
Quoted Accuracy	± 0.5% (absolute) below 5% carbon ± 10% (relative) above 5% carbon
Analysis Display	actual % carbon of collected sample
Response Time	~ 5 minutes
Normal Maintenance	calibrate pressure cells - 6 months calibrate load cell - 6 months replace air filters @ intervals based on site air quality
Cost	$50-100 K depending on options
Contact	Mr. Anthony DiGioia or Mr. Phil Glogowski CAMRAC Company, Inc. 570 Beatty Road Monroeville, Pennsylvania 15146 (412) 856-3200 phone (412) 856-4970 fax

CLYDE-STURTEVANT

The SEKAM unit was developed by the United Kingdom's Central Electric Generating Board (CEGB) in conjunction with Sturtevant Engineering Systems. Ownership was later transferred to Clyde-Sturtevant Engineering. Commercial production of the current monitor began in late 1990. As of June 1996, total installations worldwide were 40 instruments. In the United States, the system has been tested at four locations including Georgia Power Company's Plant Hammond Unit 4, Carolina Power and Light Roxboro Station, PEPCO in Alexandria, Virginia and on a fluidized bed combustion unit at AES Thames in Connecticut.

The operating principle associated with SEKAM involves trapping fly ash in a glass cell and measuring its capacitance which varies inversely with carbon content. Ash is collected super-isokinetically from the flue gas stream using multiple probes positioned in the flue gas stream. The ash is separated from the gas by a cyclone and is then deposited into a rectangular glass chamber, known as a Kajaani cell, that is located between two capacitance sensors. Ash passes through the vertical glass chamber on a plug flow basis rather than a batch basis. Here the sample capacitance is measured and converted to percent unburned carbon based upon a correlation curve of carbon vs. capacitance. Upon completion of analysis, a portion of the cell is purged to allow a small amount of newly-collected ash to enter the system. In this way, the system displays percent unburned carbon as a rolling average. The total cycle time varies with ash loading. Cycle times of 15 minutes (full load) to greater than two hours (low loads) have been observed.

Photos of the unit as installed at Hammond are shown in Figure 5. A schematic of a general SEKAM arrangement is shown in Figure 6. Information provided in system specifications is shown in Table 4. Some aspects mentioned include accuracy, mobility and cost as provided in marketing literature.

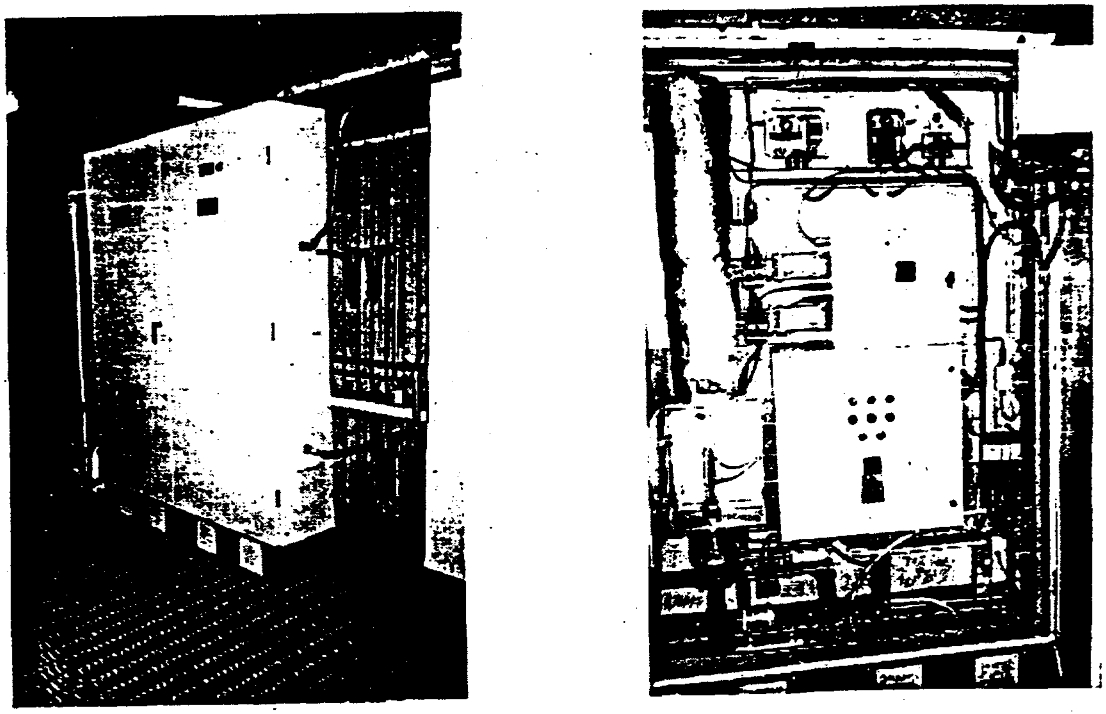

Figure 5. SEKAM On-Line Analyzer

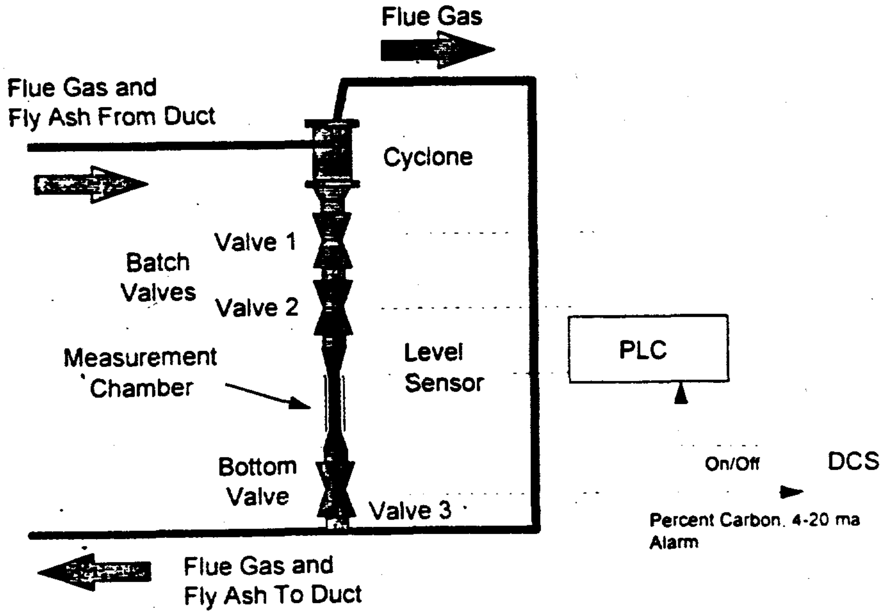

Figure 6. SEKAM General Arrangement

Table 4. SEKAM Performance Parameters

	SEKAM
Operating Principle	Capacitance
Instrument Size (WxDxH)	6 ft x 6 ft x 8.5 ft
Mobility	instrumentation: low sampling device: low
Sample Size	~ 375 grams
Quoted Accuracy	± 1.2%
Analysis Display	% carbon or LOI
Response Time	~ 15 minutes (full load)
Normal Maintenance	replace seals - 6 months
Cost	$45-50 K
Contact	Mr. Peter Wilson 5732 Rebel Drive Charlotte, North Carolina 28210 (704) 556-1555 phone (704) 556-0136 fax

APPLIED SYNERGISTICS

Applied Synergistics' FOCUS (Furnace On-line CombUstion System) Unburned Carbon Module is a non-intrusive device that provides a continuous real-time indication of on-line unburned carbon levels in fly ash. Presently, there are cameras installed at six locations worldwide. Testing on the system has been conducted at Georgia Power Company's Plant Hammond Unit 4. Dairyland Power Cooperative's Genoa Unit 3, Baltimore Gas & Electric Company's Brandon Shores Station, and Potomac Electric Power Company's Morgantown Station.

The FOCUS operating principle is based on the premise that unburned carbonaceous material exiting the furnace will be hotter than the surrounding gases and carbon-free ash. Therefore, the carbon-laden particles will emit higher levels of radiant energy in the infrared range. Infrared video cameras installed along the wall of the furnace will record these hotter particles as white spots. These images are then processed to determine the number of traverses in counts per minute. Site-specific equations enable the processor to predict LOI as a function of counts per minute, load, and excess O_2.

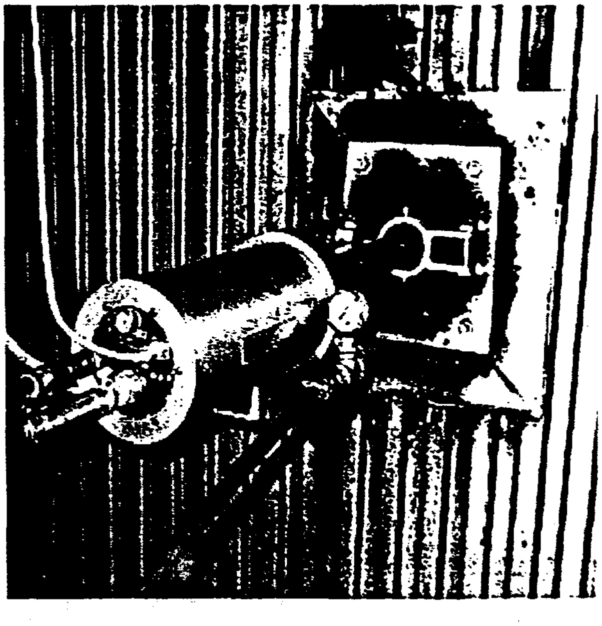

Figure 7. FOCUS On-Line Analyzer

Figure 7 is a photograph of a camera from the FOCUS set-up at Hammond. A schematic of a typical FOCUS arrangement showing major elements of the system is shown in Figure 8. Table 5 presents various aspects of the system including accuracy, instrument size, and cost as provided in company material and system specifications.

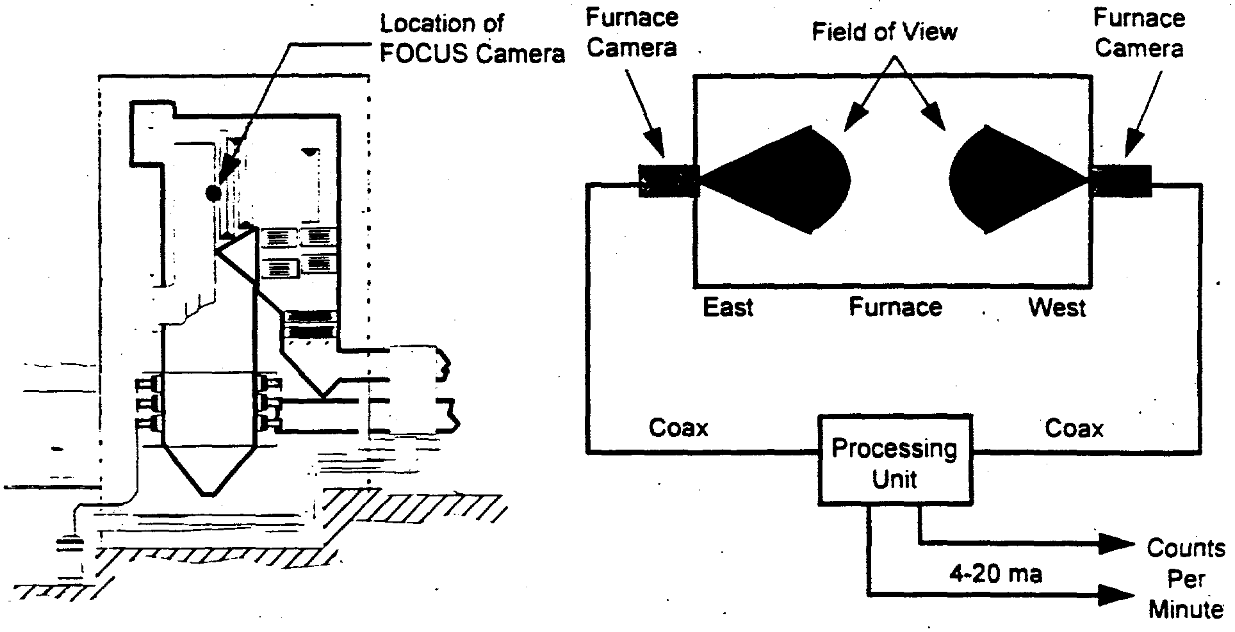

Figure 8. FOCUS General Arrangement

Table 5. FOCUS Performance Parameters

Furnace On-line CombUstion System	
Operating Principle	infrared detection
Instrument Size (WxDxH)	1.0' x 8.9' x 1.0' *(one camera)*
Mobility	instrumentation: medium monitoring device: medium
Sample Size	none collected
Quoted Accuracy	1.1% standard error of LOI
Analysis Display	bar graph of counts for 24 hour period
Response Time	less than one minute
Normal Maintenance	replace compressed air filters @ intervals dependent on site air quality
Cost	$40-55 K depending on # of cameras installed
Contact	Mr. Randy Carter Applied Synergistics, Inc. 3831 Old Forest Road, Suite 6 Lynchburg, Virginia 24501 (804) 385-6102 phone (804) 385-0714 fax

Goals of the equipment evaluation at Plant Hammond were as follows: to provide data for additional calibration of instruments demonstrated at this site (CAM, FOCUS, and SEKAM); to compare accuracy of instrument readings versus laboratory determinations of ash samples (except FOCUS); to determine the response time of analyzers to changes in boiler conditions; to compare isokinetic duct conditions to instrument readings and ESP hopper samples; and to estimate availability and durability of instruments using current information on equipment problems (type and duration).

Equipment Set-Up

CAM, SEKAM and FOCUS on-line LOI analyzers were installed at Georgia Power Company's Plant Hammond Unit 4. Each was placed in a different location as described below. CAM was set up for single point isokinetic sampling with the ash samples being extracted from the "B" side duct between the air heater and the cold side ESP. SEKAM collects samples from the ductwork between the economizer and the air heater. Two sample probes are used to simultaneously extract ash from both the "A" and "B" sides. FOCUS has two cameras positioned in the superheat region of the boiler. One camera provides counts from the east wall ("A" side) and the other from the west wall ("B" side) of the furnace. During the test program, data from the instruments was archived on the unit's digital control system and later retrieved for analysis.

Test Conditions

A series of tests were conducted in July 1995 (Test 150 & 151) and in February 1996 (Test 152 & 153) in an attempt to evaluate the performance of the CAM, SEKAM, and FOCUS units. Composite duct samples were collected on the "A" and "B" side at the precipitator inlet during each test with each composite sample consisting of ash extracted from 15 locations per side. A duct traverse was conducted at low, normal, and high oxygen levels while the unit was running at nominal 300, 400 and 500 MW loads. In addition to the composite duct samples collected during the duct traverse, ESP hopper samples were collected from the front row of hoppers on A and B sides during each test. SEKAM, CAM, and FOCUS unit readings were also recorded via the digital control system.

Accuracy

Instrument accuracies were compared in two ways. First, composite duct samples for each test were compared to average unit readings taken during the same testing period. The second method involved placing ash samples from other plants into the units for evaluation, then comparing the instrument and lab values. Since the FOCUS unit does not collect a sample, the latter method was used for the CAM and SEKAM analyzers only.

Test data in Figures 9 through 11 gives an indication of the accuracy of each instrument when compared to the appropriate isokinetic duct composite samples. Since the ash samples collected by the instruments were not analyzed, measurement errors include that resulting from non-representative ash sampling and inherent instrument inaccuracies. It should also be noted that the isokinetic results are themselves just estimates of the actual carbon levels in the duct.

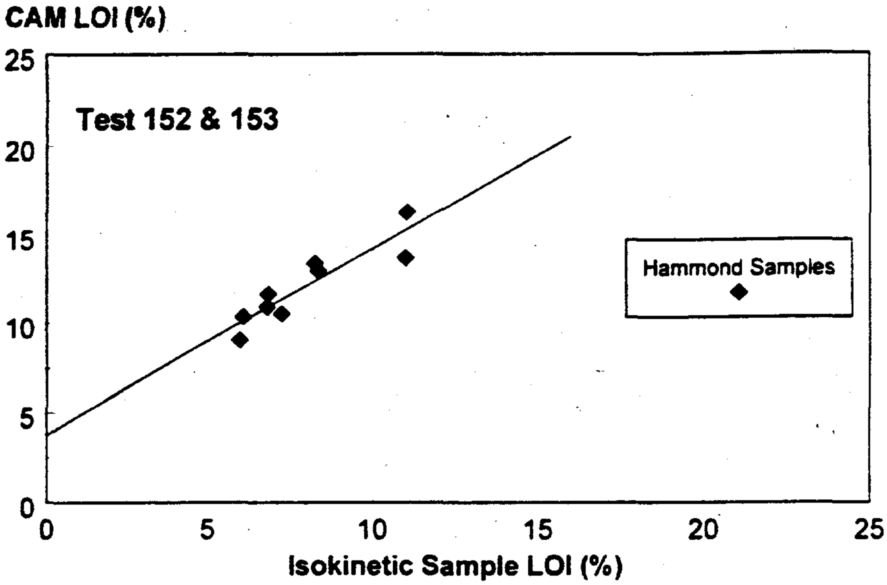

Figure 9. CAM versus Isokinetic LOI

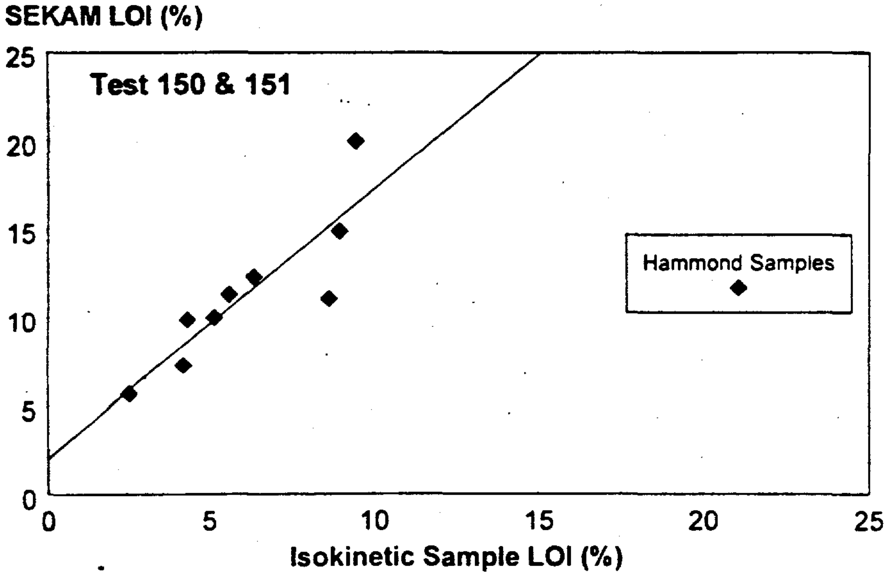

Figure 10. SEKAM versus Isokinetic LOI