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INNOVATIVE CLEAN COAL TECHNOLOGY (ICCT)

**DEMONSTRATION OF SELECTIVE CATALYTIC REDUCTION (SCR)
TECHNOLOGY FOR THE CONTROL OF NITROGEN OXIDE (NO_x)
EMISSIONS FROM HIGH-SULFUR COAL-FIRED BOILERS**

**Technical Progress Report
First & Second Quarters 1994**

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Section 1

SUMMARY

The objective of this project is to demonstrate and evaluate commercially available Selective Catalytic Reduction (SCR) catalysts from U. S., Japanese and European catalyst suppliers on a high-sulfur U. S. coal-fired boiler. SCR is a post-combustion nitrogen oxide (NO_x) control technology that involves injecting ammonia into the flue gas generated from coal combustion in a boiler. The flue gas containing ammonia is then passed through a reactor that contains a specialized catalyst. In the presence of the catalyst, the ammonia reacts with NO_x to form nitrogen and water vapor.

Although SCR is widely practiced in Japan and Europe on gas-, oil-, and low-sulfur coal-fired boilers, there are several technical uncertainties associated with applying SCR to U. S. coals. These uncertainties include:

- (1) potential catalyst deactivation due to poisoning by trace metal species present in U. S. coals that are not present in other fuels.
- (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 .
- (3) performance of a wide variety of SCR catalyst compositions, geometries and methods of manufacture under typical high-sulfur coal-fired utility operating conditions.

These uncertainties are being explored by operating a series of small-scale SCR reactors and simultaneously exposing different SCR catalysts to flue gas derived from the combustion of high sulfur U. S. coal.

The demonstration is being performed at Gulf Power Company's Plant Crist Unit No. 5 (75 MW capacity) near Pensacola, Florida. The project is funded by the U. S. Department of Energy (DOE), Southern Company Services, Inc. (SCS on behalf of the entire Southern electric system), the Electric Power Research Institute (EPRI), and Ontario Hydro. SCS is the participant responsible for managing all aspects of this project.

The project is being conducted in the following three phases:

- Phase I - Permitting, Environmental Monitoring Plan and Preliminary Engineering
- Phase II - Detailed Design Engineering and Construction
- Phase III - Operation, Testing, Disposition and Final Report

The project was in the operation and testing phase during this reporting period.

During the first and second quarters of 1994, all the reactors were operated in long-term base operating conditions, i.e. at an effective ammonia/NO_x ratio near 0.81 for 80% NO_x removal. In addition to the long-term base operating conditions, the second sequence of parametric tests was completed for Reactors A - G by May 1994. No parametric tests were performed on Reactor J, since there was no catalyst in the reactor until late in this reporting period. The catalyst vendors, assigned reactors, and the total number of operating hours with exposure to flue gas for each reactor at the end of the first and second quarters in 1994 are as follows:

<u>Catalyst supplier</u>	<u>Reactor designation</u>	<u>Operating hours</u>	
		<u>1st qtr</u>	<u>2nd qtr</u>
W. R. Grace	A (large)	4562	6031
Nippon Shokubai	B (large)	5464	6946
Siemens	C (large)	5412	6912
W. R. Grace	D (small)	3665	5145
Cormetech	E (small)	3725	5205
Haldor Topsoe	F (small)	3725	5205
Hitachi Zosen	G (small)	851*	2330
Empty	H (small)	-- **	--
Cormetech	J (small/low-dust)	-- ***	695

* Replacement catalyst loaded in February.

** Reactor left empty after Engelhard withdrew from program.

*** Cormetech catalyst loaded in April to replace Engelhard.

Operational highlights during this reporting period are provided in the following text. The test facility resumed operation on January 3, 1994, after a catalyst sampling and maintenance outage over the holidays at the end of December 1993. Except for a brief three hour boiler outage on Crist Unit 5, the SCR test facility remained on-line until shut down over February 14 to reload catalyst in Reactor G. (During the last quarter of 1993, catalyst in Reactor G was severely damaged, apparently from air sootblowing procedures destroying the sample coupons located

above the catalyst beds. The coupons subsequently fell and damaged the main catalyst layers. The catalyst supplier, Hitachi Zosen, provided replacement catalyst elements and based upon vendor recommendation, sootblowing has ceased for Reactor G.) The catalyst coupons which had been destroyed in Reactors F and G were replaced during this outage.

The test facility was shut down at the end of March for a catalyst sampling outage. Catalyst sampling was performed on all the previously operating reactors (Reactors A - G) and no problems were noted during the sampling. All samples were shipped to the respective catalyst supplier for laboratory analysis.

Upon resuming operation, the SCR test facility remained on-line the entire month of April and through May 11, when the facility was shut down for a scheduled boiler outage. Prior to the outage, the replacement catalyst for Reactor J, the low-dust train, was loaded and operation of Reactor J began on April 21 and continued until the outage in May. The second parametric testing sequence and the air heater testing were completed prior to the outage. Testing across the inlet and outlet of each air heater included the following: particulate and velocity traverses, ammonia partitioning between solid/gas phase, SO_3 and HCl measurements.

The May outage was used to meet facility maintenance needs. The electric heater from Reactor H train was relocated to the original heater location at the Reactor J inlet to compensate for the extra heat loss in the over 100 foot length of ductwork between the ESP outlet and reactor inlet. (The original Reactor J electric heater was relocated to the ESP outlet.) Extreme wear was found on the first row of elements of the electric heater on A train. The eroded heater elements were left in the bundle as shields, and spare elements were wired in their place. Inspection, maintenance, and/or repair were also performed for the reactor fans, air booster fan, and ductwork on the suction side of the fans; service water pump and air compressor; the ammonia storage, handling, and injection system; and the NO_x analyzers, analyzer pumps, operator work stations on the gas analyzer system, and the oxygen analysis system.

Also during this outage, one replacement section of baskets was supplied by ABB with different materials of construction. These were loaded into the two rotary air heaters, and the removed baskets were weighed. There were significant deposits in the cold end basket of air heater A. During the inspection, the A cold end sootblower was found to have two plugged nozzles that contributed to the fouling. Both the rotary air heaters were water washed with a high pressure system.

The SCR facility resumed operations on Friday, June 2, 1994. The facility was taken off line on Friday, June 17, because of a host boiler outage to resolve air heater problems. The SCR facility was started up again on Monday, June 20. After returning to service and after receiving the high-pressure water wash during the May outage, the two rotary air heaters experienced reduced pressure drops at design flows.

During June 1994 the flue gas venturis and ammonia flow controllers were recalibrated and velocity distributions were performed across all reactors. Particulate distribution and resistivity sampling, and SO₂/SO₃ balances were completed across the air heaters. Ammonia balances were underway at the end of this reporting period in June 1994.

As noted in the previous progress reports, problems were experienced with NO_x measurements in the presence of ammonia, apparently with catalytic reactions proceeding in the sampling system, thus reducing NO_x before the sample reached the analyzers. A series of traps and filters were previously installed in the sample line to capture the ash, water vapor and acid condensate in order to improve the accuracy of the analyzer system. The original dilution probe, made of Inconel, for measuring NO_x at intermediate levels in the reactors appeared to produce up to 30% NO_x reduction in the presence of ammonia, thus giving low NO_x readings at the analyzer. The previous technical progress report noted that in December 1993 the probe was re-tested, after being modified with 316 SS, with and without ceramic probe tips. The results were still the same, with or without the ceramic probe tips, i.e., about 30% NO_x reduction across the probe. Work continued to resolve this issue with plans formulated to evaluate a glass-lined tip. In January 1994 the development of a glass-lined tip was completed. On February 1 and 2, 1994, the revised dilution probe was tested, but the results were still not acceptable. The best results seen in all the testing were still at least a 10 percent reduction in NO_x readings, but as much as 60 to 80% reduction had been noted. At the end of this reporting period, negotiations were underway regarding final equipment disposition/payment for resolution of the failure of this equipment to meet the specifications under which this system was procured.

Precision mass flow control valves supplied by Sierra Instrumentation were originally installed to control the ammonia vapor flow rates for injection into the reactors. These controllers were affected by liquid in the flow stream, pressure variations, trash in the line, and also the orientation of the controller itself. Actions taken to correct this situation included installing coalescent filters on the ammonia supply lines to each control valve, reorienting the controllers, replacing the

ammonia header pressure regulator, cleaning each controller, and recalibrating and verifying with other instruments, but the controllers on the large reactors were still unacceptable.

One of the original ammonia flow controllers was replaced with another vendor's controller (Brooks) for a trial period of two months. The evaluation was concluded in December 1993 and based on the better performance of the Brooks' equipment, the decision was made to replace the original ammonia flow controllers for the large reactors. By the end of February 1994, all the large reactor ammonia flow controllers were replaced with the Brooks' model. These new controllers improved repeatability and have experienced much fewer malfunctions than those originally purchased. The original controllers for the small reactors have not experienced the level of problems as those on the large reactors, and there are no plans to replace these controllers.

Near the end of 1993, Engelhard, citing "commercial and strategic considerations," withdrew from the project. Thus, operation of Reactors H and J ceased in the fourth quarter of 1993. In considering the alternatives for replacement of the Engelhard catalysts, it was decided to approach the catalyst suppliers already participating in this project for offerings to immediately replace the low-dust catalyst in Reactor J. It was also decided not to seek a replacement for the high-dust catalyst in Reactor H. The decision to replace only the low-dust catalyst was based on the higher priority for the low-dust catalyst versus the high-dust catalyst, and the greater value of performing additional tests with the seven existing high-dust catalysts than a reduced testing scope with eight high-dust catalysts.

Solicitation, evaluation and selection of catalyst replacements were initiated in January 1994. A low-dust catalyst offering from Cormetech was selected as the replacement in March. The catalyst was loaded into Reactor J and operation began on April 21, 1994. There were only a few weeks of operation prior to the outage in May.

During the brief operational period prior to the mid-May 1994 outage, reaching and maintaining desired upper ranges of flue gas temperature were still difficult. Because of the decision to not replace the catalyst in Reactor H, the electric heater from Reactor H was relocated during the May outage to the original heater location at the Reactor J inlet to compensate for the extra heat loss in the over 100 foot length of ductwork between the ESP outlet and reactor inlet. During the June operations, improved temperature control was achieved as a result of the two heaters located on this train. No parametric tests were performed on this catalyst during this reporting period.

A project review meeting was held for all involved parties on May 25-26 at Plant Crist in which general progress since December 1993 was discussed, and specific vendor data were discussed in separate meetings with each vendor.

Data have been collected for ammonia/NO_x ratios, flue gas temperatures, air heater performance data, and catalyst specific performance parameters (reactor pressure drop, NO_x reduction, ammonia slip, and SO₂ oxidation). The voluminous data collected during this reporting period is not reproduced here in the Summary of this report. Instead, please see Task 1.3.2 - Process Evaluation under PROJECT STATUS, later in this report, for the actual data. The general findings from the generated data are summarized below. Please note that full interpretation and comparison of data among the various catalysts (previous results, results from this reporting period, and future test results) are being left for inclusion in the project's final report. One major reason for this action is the difference in number of operating hours among the catalysts during this and all reporting periods. Any comparison would only be meaningful if compared on a similar operating hour basis. The operating hours differ among the catalysts as a result of extended start-up and initial start-up testing requirements, plus required replacement of catalysts or catalyst coupons due to failures or substitution to replace a catalyst that was withdrawn from the project.

Overall reactor pressure drops (inlet to outlet, including dummy layer and all catalysts) generally ranged from 2 to 5 inches of water. The Siemens catalyst consistently displayed the lowest pressure drop during the quarter at about 2 inches of water. The Cormetech low-dust catalyst in Reactor J had the highest pressure drop of between 5 and 7 inches of water, with one brief excursion to 9 inches of water resulting from the initial start-up problems after the May outage.

The long term NO_x reduction data indicates the average performance of the catalyst at or near the design operating conditions of 0.8 ammonia-to-NO_x ratio, 5000 or 400 SCFM flow rate, and 700°F reactor temperature. Although presented in results by quarter in the body of this report, the average values for NO_x reduction efficiencies over the entire first half of 1994 range from 75 to 80 percent with ranges of 0.76 to 0.80 in average ammonia-to-NO_x ratios.

The parametric tests conducted during this reporting period are composed of thirteen reactor operating conditions defined by variations in the flue gas flow rate, temperature, and ammonia-to-NO_x ratio. The test conditions are listed in Table S-1. The particular measurements (intermediate ammonia, slip ammonia, SO₂, SO₃, HCl, N₂O, and velocity and mass concentration profiles) are also shown in this table at the various test conditions. Although similar test conditions are

TABLE S-1. PARAMETRIC TEST CONDITIONS

Flue gas flow rate Large / Small reactor	Flue gas temperature	NH ₃ /NO _x Ratio	Measurements
KSCFM	°F		
3.0 / 0.24	620	0.6	intermediate NH ₃
3.0 / 0.24	620	0.8	SO ₂ /SO ₃
5.0 / 0.40	620	0.8	slip NH ₃
7.5 / 0.60	620	0.6	intermediate NH ₃
7.5 / 0.60	620	0.8	intermediate & slip NH ₃
7.5 / 0.60	620	1.0	intermediate & slip NH ₃
3.0 / 0.24	700	0.8	SO ₂ /SO ₃
5.0 / 0.40	700	0.8	* intermediate & slip NH ₃ , SO ₂ /SO ₃ , N ₂ O, HCl, velocity & mass concentration profile
7.5 / 0.60	700	0.8	slip NH ₃ , SO ₂ /SO ₃
3.0 / 0.24	750	0.8	SO ₂ /SO ₃
5.0 / 0.40	750	0.8	SO ₂ /SO ₃
7.5 / 0.60	750	0.6	intermediate NH ₃
7.5 / 0.60	750	0.8	slip NH ₃ & SO ₂ /SO ₃ **

*Design conditions. **An extra intermediate NH₃ was run on a few reactors.

indicated, the measurements were not taken simultaneously (e.g., ammonia and SO₂ data for the same conditions were not collected during the exact same test run, but rather under similar process conditions at different times).

The trend for ammonia slip data versus flow rate, as expected, shows increasing ammonia slip with increasing reactor flow rate. The ammonia slip is usually relatively minor, indicating the ability of the catalyst to withstand significant increases in flow while maintaining ammonia slip limits. Since the overall reaction rate is partly controlled by mass transfer limitations, improvements in bulk mass transfer coefficients are likely mitigating the effect of increased flow on slip ammonia increases. This result demonstrates the ability of an SCR system to follow load variations while maintaining design specifications.

Generally, at low temperature and high flow rate (worst case scenario), ammonia slip sharply increases as the ammonia-to-NO_x ratio approaches 1.0. This finding is in keeping with published data of this type. At ammonia-to-NO_x ratios near 1.0, non-idealities in the reactor system force the catalyst to slip ammonia since areas are present in the reactor where NO_x is the limiting reagent.

In going from 620 to 700°F, some improvement (decrease) in ammonia slip is noted, likely due to improvements in the reaction rate with increasing temperature. Only slight improvements are noted with increasing temperature above 700°F. This may possibly be due to mass transfer limitations that become controlling at these higher temperatures. In general, significant improvements in ammonia slip are not realized at temperatures above 700°F. Losses in boiler efficiency would probably outweigh any improvements in ammonia slip from an SCR reactor operating at temperatures near 750°F.

In general, decreasing SO₂ oxidation is noted with increasing flow rate. This finding is expected since the reaction rate is predicted to be inversely proportional to flow rate according to published rate data and information provided by catalyst suppliers. Some catalysts produced a relatively flat profile, indicating that either chemical or physical characteristics of the catalyst and reactor are masking the linear effect.

SO₂ oxidation increased with rises in temperature. Published information describing the effects of temperature on SO₂ oxidation and information obtained from the catalyst suppliers indicates that increases in SO₂ oxidation are exponential with respect to temperature. The exact shape of the trends with the project data is somewhat unclear since fairly large measurement variability exists in the test facility and other phenomena such as SO₃ deposition occur in the test facility reactors. The general trend of increasing SO₂ oxidation is observed, however, and the results provide a good basis for determining the potential for SO₃ formation across the SCR reactor at various operating temperatures.

Measurements of N₂O at the reactor inlets and outlets indicated that N₂O levels did not increase across the SCR catalytic reactors.

Section 2

INTRODUCTION

The Innovative Clean Coal Technology (ICCT) Program is designed to demonstrate clean coal technologies that are capable of retrofitting or repowering existing facilities to achieve significant reduction in sulfur dioxide (SO_2) and/or nitrogen oxides (NO_x) emissions. The technologies selected for demonstration are capable of being commercialized in the 1990s and are expected to be more cost effective than current technologies.

This ICCT project is jointly funded by the U. S. Department of Energy, the Electric Power Research Institute (EPRI), Ontario Hydro, and by Southern Company Services (SCS) on behalf of the entire Southern electric system. The project's objective is to demonstrate the selective catalytic reduction (SCR) process that removes nitrogen oxides (NO_x) from the flue gas of boilers that burn U.S. high-sulfur coal. The SCR technology involves the injection of NH_3 into the flue gas and the subsequent catalytic reduction of NO_x by NH_3 to produce molecular nitrogen (N_2) and water vapor.

A simplified SCR process flow diagram with major equipment is shown in Figure 1. Specifically, hot flue gas leaving the economizer section of the boiler is ducted to the SCR reactor. Prior to entering the reactor, NH_3 is injected into the flue gas at a sufficient distance upstream of the reactor to provide for complete mixing of the NH_3 and flue gas. The quantity of NH_3 can be adjusted and it reacts with the NO_x from the flue gas. The flue gas leaving the catalytic reactor enters the air preheater where it transfers heat to the incoming combustion air. Provisions are made for ash removal from the bottom of the reactor since some fallout of fly ash is expected. Duct work is also provided to bypass some flue gas around the economizer during periods when the boiler is operating at reduced load. This is done to maintain the temperature of the flue gas entering the catalytic reactor at the proper reaction temperature of about 700°F . The flue gas leaving the air preheater goes to the electrostatic precipitator (ESP) where fly ash is removed. The ESP is part of the existing plant and is generally unaffected by the SCR system except as higher SO_3 content affects the electrical resistivity of the fly ash or if NH_4HSO_4 co-precipitates with the fly ash.

Figure 1. Flow Diagram of Typical SCR Installation.

The SCR technology is in commercial use in Japan and western Europe on gas-, oil-, and on low-sulfur, coal-fired power plants. The first utility applications of SCR catalyst technology started in Japan in 1977 for oil- and gas-fired boilers and subsequently in 1979 for coal-fired boilers. As of 1986, ninety utility boilers in Japan had been equipped with SCR catalyst technology including twenty-two coal-fired boilers. These coal-fired boilers represent a combined capacity in excess of 6500 MWe and are typically fired with a low-ash, low-sulfur coal.

In addition to Japanese experience, several countries in western Europe (most notably Germany and Austria) have passed stringent NO_x emission regulations that have all but mandated the installation of SCR. Prior to commercial SCR installations in Germany, utility companies demonstrated several types of SCR facilities in prototype demonstration programs similar to this ICCT project. Over 50 SCR pilot plants were built and operated in western Europe. These pilot plants ranged from 19 to 6200 SCFM and provided the data base that led to commercialization of the SCR technology in western Europe.

Previously completed U. S. work with the SCR process on utility boilers consists of three projects which were carried out in the late 1970s and early 1980s. One of these was carried out on a natural gas fired boiler by Southern California Edison. Another project consisted of a pilot test conducted for the EPA at Georgia Power's Plant Mitchell. This pilot plant treated a 1000 ACFM (0.5 MWe) slip stream of flue gas resulting from the combustion of low- to medium-sulfur coal. A third pilot-scale project, carried out at the Public Service Company of Colorado's Arapaho Station, treated a 5000 ACFM (2.5 MWe) slip stream of flue gas resulting from the combustion of U. S. low-sulfur coal.

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

- (1) potential catalyst deactivation due to poisoning by trace metal species present in U. S. coals that are not present in other fuels.
- (2) performance of the technology and effects on the balance-of-plant equipment in the presence of high amounts of SO₂ and SO₃.
- (3) performance of a wide variety of SCR catalyst compositions, geometries and methods of manufacture under typical high-sulfur coal-fired utility operating conditions.

These uncertainties are being explored by operating a series of small-scale SCR reactors and simultaneously exposing different SCR catalysts to flue gas derived from the combustion of high sulfur U. S. coal.

The first uncertainty above is being handled by evaluating SCR catalyst performance for two years under realistic operating conditions found in U. S. pulverized coal utility boilers. The deactivation rates for the catalysts exposed to flue gas from high sulfur U. S. coal will be documented to determine accurate catalyst life, and thus, accurate process economics.

The second uncertainty above is being explored by performing parametric testing and through the installation/operation of air preheaters downstream of the larger reactors. During parametric testing, operating conditions are adjusted above and below design values to observe deNO_x performance and ammonia slip as functions of the change in operating conditions. Air preheater performance is observed to evaluate effects from SCR operation upon heat transfer, and therefore, upon boiler efficiency.

The third uncertainty is being handled by using honeycomb- and plate-type SCR catalysts from U. S., Japan and Europe of various commercial compositions. Results from the tests with these catalysts expands our knowledge of performance on a variety of SCR catalysts under U. S. utility operating conditions with high-sulfur coal.

The intent of this project is to demonstrate commercial catalyst performance, proper operating conditions, and catalyst life for the SCR process. This project is also demonstrating the technical and economic viability of SCR, while reducing NO_x emissions by at least 80%.

The project is being conducted at Gulf Power Company's Plant Crist Unit 5, a commercially operating 75 MW unit, located in Pensacola, Florida, on U. S. coals with a sulfur content near 3.0%. Unit 5 is a tangential-fired, dry bottom boiler, with hot and cold side ESP for particulate control. The SCR process used in this demonstration is designed to treat a slip-stream of flue gas and features multiple reactors installed in parallel. With all reactors in operation, the maximum amount of combustion flue gas that can be treated is 17,400 standard cubic feet per minute (SCFM) which is roughly equivalent to 8.7 MWe.

The SCS facility is a slip-stream SCR test facility consisting of three 2.5 MWe (5000SCFM) SCR reactors and six 0.20 MWe (400SCFM) reactors that operate in parallel for side-by-side

comparisons of commercially available SCR catalyst technologies obtained from vendors throughout the world. Figure 2 presents a simplified process flow diagram for the SCR test facility. The large (2.5 MWe) SCR reactors contain commercially available SCR catalysts as offered by SCR catalyst suppliers. These reactors are coupled with small-scale air preheaters to evaluate the long-term effects of SCR reaction chemistry on air preheater deposit formation and the deposits' effects on an air preheater. The small reactors are used to test additional commercially available catalysts. This demonstration facility size is adequate to develop performance data to evaluate SCR capabilities and costs that are applicable to boilers using high-sulfur U. S. coals.

The demonstration project is organized into three phases: (1) Phase I - Permitting, Environmental Monitoring Plan and Preliminary Engineering; (2) Phase II - Detail Design Engineering and Construction; and (3) Phase III - Operation, Testing, Disposition, and Final Report. The cooperative agreement was signed June 14, 1990, and the project completion date is now projected to be at the end of 1995. The original total estimated project costs were \$15.6 million, but project growth increased the expected total cost to \$23 million. The co-funders are SCS (\$10 million), DOE (\$9.4 million), EPRI (\$2.9 million) and Ontario Hydro (\$0.75 million).

Figure 2. Prototype SCR Demonstration Facility-Process Flow Diagram

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Section 3

PROJECT DESCRIPTION

Within the three phases of the project, the following tasks are to be conducted to effectively demonstrate the SCR process:

Phase I - Permitting, Environmental Monitoring Plan and Preliminary Engineering

- Task 1.1.1 - Prototype Plant Permitting Activities
- Task 1.1.2 - Develop Environmental Monitoring Program
- Task 1.1.3 - Preliminary Engineering
- Task 1.1.4 - Engineering and Construction Contracts Scope Development
- Task 1.1.5 - Project Management and Reporting

Phase II - Detail Design Engineering and Construction

- Task 1.2.1 - Detailed Design Engineering
- Task 1.2.2 - Construction
- Task 1.2.3 - Operation Staff Training
- Task 1.2.4 - Planning for Detailed Testing
- Task 1.2.5 - Start-Up/Shakedown
- Task 1.2.6 - Project Management and Reporting

Phase III - Operations, Testing, Disposition and Final Report

- Task 1.3.1 - SCR Demonstration Facility Operations and Maintenance
- Task 1.3.2 - Process Evaluation
- Task 1.3.3 - Environmental Data Management and Reporting
- Task 1.3.4 - Economic Evaluation
- Task 1.3.5 - Dismantling/Disposition
- Task 1.3.6 - Project Management and Reporting

Section 4 PROJECT STATUS

Progress during January - June, 1994, is summarized below for each of the on-going tasks in the Scope of Work.

PHASE III - OPERATIONS, TESTING, DISPOSITION, AND FINAL REPORT

Task 1.3.1 - SCR Demonstration Facility Operations and Maintenance

The catalyst vendors, assigned reactors, and the total number of operating hours with exposure to flue gas for each reactor at the end of the first and second quarters in 1994 are as follows:

<u>Catalyst supplier</u>	<u>Reactor designation</u>	<u>Operating hours</u>	
		<u>1st qtr</u>	<u>2nd qtr</u>
W. R. Grace	A (large)	4562	6031
Nippon Shokubai	B (large)	5464	6946
Siemens	C (large)	5412	6912
W. R. Grace	D (small)	3665	5145
Cormetech	E (small)	3725	5205
Haldor Topsoe	F (small)	3725	5205
Hitachi Zosen	G (small)	851*	2330
Empty	H (small)	-- **	--
Cormetech	J (small/low-dust)	-- ***	695

* Replacement catalyst loaded in February (see discussion later in report).

** Reactor left empty after Engelhard withdrew from program (see discussion later in report).

*** Cormetech catalyst loaded in April to replace Engelhard (see discussion later in report).

During the first and second quarters of 1994, all the reactors were operated in long-term base operating conditions, i.e. at an effective ammonia/NO_x ratio near 0.81 for 80% NO_x removal. In addition to the long-term base operating conditions, the second sequence of parametric tests was completed for Reactors A - G by May 1994. No parametric tests were performed on Reactor J since there was no catalyst in the reactor until late in this reporting period. (This is discussed in

more detail later in this section under "Low Dust Reactor".) Operational highlights during the reporting period are given below.

The test facility resumed operation on January 3, 1994, after a catalyst sampling and maintenance outage over the holidays at the end of December 1993. After resuming operation of the test facility, complete calibrations of the venturis and ammonia flow controllers were performed. By the end of January, most of the ammonia work on the second parametric sequence had been completed on Reactors A - C. Except for a brief three hour boiler outage on Crist Unit 5, the SCR test facility remained on-line until shut down over February 14 to reload catalyst in Reactor G. (During the last quarter of 1993, catalyst in Reactor G was severely damaged, apparently from air sootblowing procedures destroying the sample coupons located above the catalyst beds. The coupons subsequently fell and damaged the main catalyst layers. The catalyst supplier, Hitachi Zosen, provided replacement catalyst elements and based upon vendor recommendation, sootblowing has ceased for Reactor G.) The catalyst coupons which had been destroyed in Reactors F and G were also replaced during this outage.

There were several unscheduled boiler outages on Crist Unit 5 due to reheater leaks which caused delays in the testing schedule during March. Some of this lost time was made up by the end of the month due to test crew efforts. The test facility was shut down at the end of March for a catalyst sampling outage. Catalyst sampling was performed on all the previously operating reactors (Reactors A - G) and no problems were noted during the sampling. Although scheduled to take two days, the actual sampling was accomplished in one working day and the test facility was returned to service a day early. All samples were shipped to the respective catalyst supplier for laboratory analysis.

Upon resuming operation after the catalyst sampling at the end of March, the SCR test facility remained on-line the entire month of April and through May 11, when the facility was shut down for a scheduled boiler outage. Prior to the outage, the replacement catalyst was loaded for Reactor J, the low-dust train. Operation of Reactor J began on April 21 and continued until the outage (see catalyst replacement evaluation results later in this section). The second parametric testing sequence and the air heater testing were completed prior to the outage. Testing across the inlet and outlet of each air heater included the following: particulate and velocity traverses, ammonia partitioning between solid/gas phase, SO_3 and HCl measurements.

The May outage was used to meet facility maintenance needs. The electric heater from Reactor H train was relocated to the original heater location at the Reactor J inlet to compensate for the extra heat loss in the over 100 foot length of ductwork between the ESP outlet and reactor inlet. (The original Reactor J electric heater was relocated to the ESP outlet.) Extreme wear was found on the first row of elements of the electric heater on A train. The eroded heater elements were left in the bundle as shields, and spare elements were wired in their place. The reactor fans, air booster fan, and ductwork on the suction side of the fans were washed out. Maintenance was performed on the service water pump and air compressor. The ammonia storage, handling, and injection system was inspected and minor repairs were made. Two NO_x analyzers, several analyzer pumps, and two operator work stations on the gas analyzer system were repaired. The oxygen analysis system had also been giving some erroneous readings, primarily attributed to routine aging of the zirconium cells. The most important oxygen measurements were corrected and work was planned to replace other aged cells.

Also during this outage, one replacement section of baskets was supplied by ABB with different materials of construction and these were loaded into the two rotary air heaters, and the removed baskets were weighed. There was a notable amount of deposits in the cold end basket of air heater A. During the inspection, the A cold end sootblower was found to have two plugged nozzles that contributed to the fouling. Both the rotary air heaters were water washed with a high pressure system.

The SCR facility resumed operations on Friday, June 2, 1994. The facility was taken off line on Friday, June 17, because of a host boiler outage to resolve air heater problems. The SCR facility was started up again on Monday, June 20. After returning to service and after receiving the high-pressure water wash during the May outage, the two rotary air heaters experienced reduced pressure drops at design flows: gas-side pressure drops were reduced by 3 inches of H₂O for A air heater (from about 9 inches) and by 2 inches of H₂O for B air heater (from 5.5 inches).

During June 1994 the flue gas venturis and ammonia flow controllers were recalibrated and velocity distributions were performed across all reactors. Particulate distribution and resistivity sampling, and SO₂/SO₃ balances were completed across the air heaters. Ammonia balances were underway at the end of this reporting period in June 1994.

Southern Research Institute (SRI) had previously maintained a staff of six people, supplementing their regular staff of three people with two additional testing crew and an additional analytical

chemist. This doubling of staff was required to complete the preliminary parametric testing on all reactors before the end of December 1993. Beginning in January 1994, SRI's staff was reduced to four (one supervisor, two test crew, and one analytical chemist) for handling the remaining parametric test sequences. Supplemental staff will be used as circumstances dictate.

Several other major experiences encountered during operation are highlighted below.

- Dilution/Extraction Gas Sampling/Monitoring System

The SCR test facility uses a dilution/extraction sampling system from Lear Seigler Measurement Controls Corporation (LSMCC, now Monitor Labs). The system consists of thirteen dilution/extraction probes for the measurement of NO_x , SO_2 , CO_2 , and CO . There are twenty-six Yokagawa in-situ probes for the measurement of oxygen. Using dry air as a dilution medium, normal dilution ratios range from 30:1 to 250:1. NO_x analysis is performed using an LSMCC ML8840 chemiluminescence NO_x analyzer with a detection limit of 2 ppb resulting in a flue gas detection limit of approximately 0.25 ppm. CO is measured using a LSMCC model ML8830 infrared CO analyzer with a detection limit of 0.1 ppm resulting in a flue gas detection limit of approximately 3 ppm. CO_2 is measured using a Siemens Ultimat 5E non-dispersive infrared CO_2 analyzer. SO_2 is measured using a LSMCC model ML8850 fluorescence SO_2 analyzer with a detection limit of 1 ppb resulting in a flue gas detection limit of 0.1 ppm. Oxygen is measured using in-situ zirconium oxide cell technology.

As noted in the previous progress reports, problems were experienced with NO_x measurements in the presence of ammonia, apparently with catalytic reactions proceeding in the sampling system, thus reducing NO_x before the sample reaches the analyzers. A series of traps and filters were previously installed in the sample line to capture the ash, water vapor and acid condensate in order to improve the accuracy of the analyzer system. The original dilution probe, made of Inconel for measuring NO_x at intermediate levels in the reactors, appeared to produce up to 30% NO_x reduction in the presence of ammonia, thus giving low NO_x readings at the analyzer. The previous report noted that in December 1993 the probe was re-tested, after being modified with 316 SS, with and without ceramic probe tips. The results were still the same, with or without the ceramic probe tips, i.e., about 30% NO_x reduction across the probe. Work continued to resolve this issue with plans formulated to evaluate a glass-lined tip. In January 1994 the development of a glass-lined tip was completed. On February 1 and 2, 1994, the revised dilution probe was tested. The results were still not acceptable. The best results seen in all the testing were still at

least a 10 percent reduction in NO_x readings, but as much as 60 to 80% reduction had been noted. The results of the February tests are included in Appendix A. At the end of this reporting period, negotiations were underway regarding final equipment disposition/payment for resolution of the failure of this equipment to meet the specifications under which this system was procured.

- Ammonia Injection Flow Control

Precision mass flow control valves supplied by Sierra Instrumentation were installed to control the ammonia vapor flow rates for injection into the reactors. These controllers were affected by liquid in the flow stream, pressure variations, trash in the line, and also the orientation of the controller itself. These controllers were calibrated on nitrogen and scaled to read ammonia flow. Although initial results indicated accurate flow control, subsequent measurements indicated that actual ammonia flow was 10 to 25 percent higher than the controllers were indicating. Action taken to correct this situation included installing coalescing filters on the ammonia supply lines to each control valve, reorienting the controllers, replacing the ammonia header pressure regulator, cleaning each controller, and recalibrating and verifying with other instruments.

After the above changes, the results for the controllers on the large reactors were still unacceptable. One of the original ammonia flow controllers was replaced with another vendor's controller (Brooks) for a trial period of two months. The evaluation was concluded in December 1993, and based on the better performance of the Brooks' equipment, the decision was made to replace the original ammonia flow controllers for the large reactors. By the end of February 1994, all the large reactor ammonia flow controllers were replaced with the Brooks' model. These new controllers improved repeatability and have experienced much fewer malfunctions than those originally purchased. The original controllers for the small reactors have not experienced the same level of problems as those on the large reactors, and no plans were made to replace the ammonia flow controllers for the small reactors.

- Catalyst Replacement on Low Dust Reactor

Near the end of 1993, Engelhard cited "commercial and strategic considerations" for withdrawing from the project. Thus, operation of two small reactors, Reactors H and J, were ceased in the fourth quarter of 1993 because of this decision. In considering the alternatives for replacement of the Engelhard catalysts, it was decided to approach the catalyst suppliers already participating in this project for offerings to immediately replace the low-dust catalyst in Reactor J. It was decided

not to seek a replacement for the high-dust catalyst in Reactor H. The reasoning for these decisions was as follows:

Although there may be considerably less market for low-dust SCR catalyst, the catalyst by Engelhard was the only low-dust catalyst in the project. Replacement would maintain the broader applications of the project results. There continues to be interest by all co-funders on the low-dust alternative SCR configuration.

The test plan has been reduced from the originally conceived scope because the actual testing manpower and testing time requirements were greater than anticipated. Originally, it was thought that eight parametric test sequences could be performed on each catalyst during the two-year operating duration of this demonstration, with well over 30 different test conditions in each test sequence. In actual practice, the first parametric test sequence, which had a lower number of total tests (at 9 different test conditions), required six months to be completed for all operating reactors. With this number of reactors and tests, a total of only three sets of parametric tests could be performed for each catalyst during the demonstration. Replacing only one of the two Engelhard catalysts allowed extra flexibility in the testing for the remaining catalysts. Modification of the test plan to reduce the number of tests conducted in each test sequence from the original plans and replacement of only one of the two catalysts would allow the completion of a total of five parametric test sequences to be performed during the project on the remaining eight catalysts.

The decision to only contact participating project catalyst suppliers was a timing decision. Selection of a supplier already on the project would significantly reduce the time required to obtain a replacement catalyst and actually begin testing. If a catalyst vendor outside the project were selected, the time required to secure a new contract (with flowdown provisions and confidentiality agreements) would probably have reduced the length of testing opportunity by 3 to 6 months.

In summary, the decision to replace only the low-dust catalyst was based on the higher priority for the low-dust catalyst versus the high-dust catalyst, and the greater value of performing additional tests with the seven existing high-dust catalysts than a reduced testing scope with eight high-dust catalysts.

Solicitation, evaluation and selection of catalyst replacements were initiated in January 1994. A low-dust catalyst offering from Cormetech was selected as the replacement in March. The catalyst was loaded into Reactor J and operation began on April 21, 1994. There were only a few weeks of operation prior to the outage in May. No parametric tests were performed on this catalyst during this reporting period. The first parametric tests on this catalyst were to be completed during the next reporting period.

- Low Dust Reactor Operation

As reported in the previous progress report, several changes were made during the fourth quarter of 1993 on the low-dust reactor, Reactor J, to prevent recurrence of the severe plugging of the first catalyst layer experienced after only a few hours of operation during its first start-up after catalyst loading. One of the resulting design changes previously described and implemented was increasing the reactor heater capacity and moving the heater to just downstream of the isolation damper to raise the minimum gas temperature throughout the entire length of ductwork between the scoop and reactor inlet.

Prior to the May 1994 outage, the replacement low-dust catalyst from Cormetech was received and loaded into the reactor in April 1994. During the brief operational period prior to the mid-May 1994 outage, reaching and maintaining desired upper ranges of flue gas temperature were still difficult. Based on the decision mentioned above to not replace the catalyst in Reactor H, the electric heater from Reactor H train was relocated during the May outage to the original heater location at the Reactor J inlet to compensate for the extra heat loss in the over 100 foot length of ductwork between the ESP outlet and reactor inlet. During the June operations, improved temperature control was achieved as a result of the two heaters located on this train.

Upon resuming operation of the low-dust reactor, Reactor J, after the May outage, pressure drop rapidly increased, caused by deposits breaking loose from the ductwork and depositing on the wire-mesh screens covering the catalysts. The catalyst baskets were removed and vacuumed, and the reactor was operated without catalyst until deposits were cleared from the system. The catalyst was reloaded and new screens were installed which matched the pitch of the new catalyst which had been loaded just prior to the May outage. The pressure drop returned to normal after these measures were taken.

- Reactor and Air Preheater Sootblowing

During normal operation, reactor and air heater pressure drops do slowly increase, but these were usually controlled by sootblowing. The large reactors are equipped with retracting lances with a perpendicular array of nozzles mounted that blow steam across the catalyst as the lance traverses across the reactor. The steam sootblowers on the air preheaters are stationary and use steam. Much work went into eliminating the condensate from the sootblowing steam supply piping before the sootblowers extend into the reactors, including an extra steam isolation valve on each sootblower, using a process steam condensate trap on each reactors steam supply header, and adding warm-up vents to assure the piping is hot enough to prevent condensation. The small reactors are blown by air lance inserted through a ball valve installed above each basket layer. The typical schedules and procedures for sootblowing of the reactors and air preheaters are shown below.

Sootblowing

Schedules:

- Large Reactors - every 8 hours
- Small Reactors - every 8 or 12 hours
- Air Preheaters - every 4 hours

Procedures:

- Warm-up lines by venting the top of verticals
- Condensate drains under each vertical and air preheater
- Monitoring pressure drops-before/after

As noted in the previous report, based upon damage to the catalyst in Reactor G from sootblowing practices and subsequent catalyst vendor recommendation, sootblowing was discontinued for Reactor G. This practice produced no significant effects through this reporting period on reactor pressure drop. Also at a vendor's request, sootblowing frequency was reduced from three to two times per day for Reactor D beginning in February 1994 to decrease the rate of erosion noted. After one month, an increase above former values in the reactor pressure drop during operation at the higher flue gas flow rates was observed. The sootblowing schedule for Reactor D was re-evaluated at the end of March. Initial study indicated that the previously noted erosion may result from a manufacturing problem, and while further review was begun, normal sootblowing schedule was resumed after March 1994.

- Measurement Accuracy/Repeatability

The majority of manual measurements are made using triplicate samples. This ensures the quality of the data and helps to remove operating variability from the results. Since triplicate measurements usually require several hours to complete, they also help to evaluate and ensure that steady-state operation has been achieved for the particular test conditions. Each of the triplicate samples are analyzed in duplicate in the laboratory. Thus, six values are acquired for each average reported measured value. Detection limits for ammonia slip measurements are normally approximately 1 ppm with analytical repeatability to within 0.1 ppm. Intermediate ammonia detection limits are somewhat higher, namely 1 to 2 ppm with analytical repeatability to within 1 ppm. SO₂ measurements normally have detection limits of 5 ppm with analytical repeatability to within 1% of the measured value. SO₃ detection limits are normally 0.4 ppm with an analytical repeatability to within 0.4% of the measured value.

Task 1.3.2 - Process Evaluation

- Coal Chemistry

Coal samples were analyzed monthly for major components and certain trace elements. The monthly results, yearly averages, and project-to-date averages are shown in Table 1.

- Flue Gas Composition

Several flue gas constituents are measured continuously by the gas analyzer system (described previously) at the SCR test facility inlet. Table 2 shows the constituent concentrations in terms of average values, normal high values, and normal low values over the reporting period. These quarterly values were determined using daily averages, daily highs, and daily lows during periods that the host boiler was on-line (operating at greater than 40 MWe).

Table 1: Plant Crist Unit 5 Coal Analyses - 1994 Monthly, As-Burned Composites
Alabama Power Company Results, Dry Basis

SCS ID No.			94-042	94-072	94-114	94-159	94-177	94-241	1994	*Total	*Total	1993
Test	Method	Units	January	February	March	April	May	June	Average	Average	Std. Dev.	Average
Moisture, Total	ASTM D 3302	% by Wt.	9.30	9.72	9.73	11.76	11.46	12.00	10.66	10.53	0.91	10.45
Ash	ASTM D 3180	% by Wt.	9.84	9.39	9.69	8.86	8.92	8.65	9.23	9.47	0.55	9.62
Gross Calorific Value	ASTM D 3180	Btu/lb	13389	13448	13336	13334	13277	13242	13338	13271	102	13231
Sulfur, Total	ASTM D 3180	% by Wt.	2.80	2.88	2.78	2.60	2.78	2.68	2.75	2.78	0.20	2.79
Sulfur, lb/MMBtu	ASTM D 3180	lb/MMBtu	2.09	2.14	2.08	1.95	2.09	2.02	2.06	2.09	0.15	2.11
Carbon	ASTM D 3180	% by Wt.	75.27	75.08	75.31	75.06	75.10	74.69	75.09	74.82	0.49	74.67
Hydrogen	ASTM D 3180	% by Wt.	4.97	4.94	4.99	4.91	4.96	5.04	4.97	5.01	0.06	5.03
Nitrogen	ASTM D 3180	% by Wt.	1.59	1.58	1.59	1.59	1.57	1.56	1.58	1.58	0.04	1.58
Oxygen	ASTM D 3180	% by Wt.	5.53	6.13	5.64	6.98	6.67	7.38	6.39	6.34	0.51	6.31
Carbon, Fixed	ASTM D 3180	% by Wt.	54.00	54.15	53.10	53.41	53.02	53.21	53.48	52.31	1.34	51.61
Volatile Matter	ASTM D 3180	% by Wt.	36.16	36.46	37.21	37.73	38.06	38.14	37.29	38.22	1.31	38.77
Aluminum	ASTM D 3682	% by Wt.	1.09	1.13	1.09	1.06	1.11	1.09	1.10	1.09	0.09	1.09
Antimony	ASTM D 3683	mg/kg	2.0	<1.0	0.9	1.3	0.4	0.6	<1.0	<1.0	---	1
Arsenic	Slurry/GFAAS	mg/kg	4.8	2.0	2.6	1.7	1.1	1.6	2.3	3.0	2.1	3.4
Barium	ASTM D 3683	mg/kg	33	15	38	41	17	29	29	44	21	53
Beryllium	ASTM D 3683	mg/kg	3	3	3	2	2	3	3	3	0	3
Cadmium	ASTM D 3683	mg/kg	<1.00	<1.00	0.30	0.30	0.19	0.21	<1.00	<1.00	---	<1
Calcium	ASTM D 3682	% by Wt.	0.23	0.18	0.22	0.23	0.22	0.21	0.22	0.25	0.04	0.26
Chlorine	ASTM D 4208	mg/kg	858	1116	1759	3403	2236	2092	1911	1525	703	1294
Chromium	ASTM D 3683	mg/kg	21	26	20	27	21	28	24	21	4	19
Cobalt	ASTM D 3683	mg/kg	8	4	6	7	9	9	7	7	2	8
Copper	ASTM D 3683	mg/kg	12	4	10	8	9	10	9	9	3	9
Fluorine	ASTM D 3761	mg/kg	<1	41	87	98	63	70	60	54	30	50
Iron	ASTM D 3682	% by Wt.	1.33	1.17	1.34	1.11	1.17	1.16	1.21	1.13	0.14	1.08
Lead	ASTM D 3683	mg/kg	13	14	14	6	7	19	12	12	5	12
Lithium	ASTM D 3683	mg/kg	7	8	6	15	4	9	8	7	3	7
Magnesium	ASTM D 3682	% by Wt.	0.08	0.07	0.07	0.05	0.05	0.05	0.06	0.06	0.01	0.05
Manganese	ASTM D 3682	mg/kg	29	35	28	22	22	26	27	25	4	23
Mercury	ASTM D 3684	mg/kg	0.09	0.10	0.08	0.07	0.08	0.08	0.08	0.08	0.05	0.09
Molybdenum	ASTM D 3683	mg/kg	5.0	13.0	5.2	3.5	12.3	1.7	6.8	7.7	4.5	8
Nickel	ASTM D 3683	mg/kg	17	9	17	19	17	17	16	15	3	15
Phosphorus	ASTM D 3682	% by Wt.	0.01	0.01	0.01	0.01	<0.01	<0.01	0.01	0.03	0.03	0.04
Potassium	ASTM D 3682	% by Wt.	0.29	0.19	0.18	0.15	0.16	0.15	0.19	0.21	0.06	0.23
Selenium	Slurry/GFAAS	mg/kg	2.0	<2.0	1.2	1.6	1.5	1.3	<2.0	<2.0	---	<2
Silica	ASTM D 3682	% by Wt.	2.58	2.50	2.21	2.09	2.16	2.10	2.27	2.31	0.20	2.33
Sodium	ASTM D 3682	% by Wt.	0.06	0.05	0.06	0.06	0.07	0.05	0.06	0.06	0.02	0.06
Titanium	ASTM D 3682	% by Wt.	0.06	0.06	0.06	0.06	0.06	0.06	0.06	0.06	0.01	0.06
Vanadium	ASTM D 3683	mg/kg	50	26	49	34	40	47	41	46	9	49
Zinc	ASTM D 3683	mg/kg	36	31	36	28	30	37	33	33	6	33

*Total Includes Average and Standard Deviation from March '93 through Latest Month Shown
< Less Than Detection Limit.

TABLE 2. TEST FACILITY GAS CONCENTRATIONS

<u>January - March 1994</u>			
<u>Constituent</u>	<u>Average</u>	<u>High</u>	<u>Low</u>
Unit #5 Load (MW)	64	84	40
Inlet NO _x (ppm)	322	376	274
Inlet O ₂ (%)	4.4	7.2	2.4
Inlet CO ₂ (%)	16.2	19.0	12.7
Inlet CO (ppm)	63	386	17
Inlet SO ₂ (ppm)	1690	1935	1355

<u>April - June 1994</u>			
<u>Constituent</u>	<u>Average</u>	<u>High</u>	<u>Low</u>
Unit #5 Load (MW)	68	85	45
Inlet NO _x (ppm)	334	396	281
Inlet O ₂ (%)	4.3	7.7	2.1
Inlet CO ₂ (%)	16.4	19.0	12.9
Inlet CO (ppm)	21	182	0.7
Inlet SO ₂ (ppm)	1583	1821	1004

- Mass Concentration and Particle Size Distribution

The mass concentration and particle size distribution measurements made to confirm that all reactors were receiving fly ash similar in concentration and particle size to that of the fly ash in the main plant ductwork, were reported in the previous quarterly progress report for April - September 1993. Mass concentration was measured for each reactor during parametric testing. These were made for profile characterizations rather than reactor to reactor comparisons. Simultaneous measurements were not made during this reporting period. These mass concentration results are shown later under the individual catalyst/reactor performance sections.

- Fly Ash Chemistry and Resistivity

Ash mineral analyses and fly ash resistivity data were reported previously in the April - September, 1993, progress report.

- Ammonia/NO_x Ratio

The ammonia-to-NO_x ratio is normally maintained to give approximately 80% reduction in NO_x across the reactors. This gives a theoretically required ammonia-to-NO_x ratio of roughly 0.83 assuming that the flue gas NO_x concentration consists of 5% NO₂. However, oxygen inleakage can cause an increase in ammonia-to-NO_x ratio, if the NO_x values are measured upstream of the flue gas metering device. This is the case in the SCR reactors. As a result, an ammonia-to-NO_x ratio of 0.83 gives an effective ammonia-to-NO_x ratio of approximately 0.88 in the SCR reactors. Consequently, the ammonia-to-NO_x ratio is normally set at 0.74 to 0.76, resulting in an effective ammonia-to-NO_x ratio of about 0.81, giving a NO_x reduction of approximately 80% (based on 1% of the NO_x as NO₂, which is the approximate value for the SCR facility). Due to calibration problems in the ammonia injection system and calibration drift in the venturi flow measurement, the long term effective ammonia-to-NO_x ratio of the SCR reactors ranged from 0.75 to 0.80 during these two quarters of testing. Parametric testing required variations in the ammonia-to-NO_x ratio. During these tests, the effective ammonia-to-NO_x ratio was varied from a high of approximately 1.19 to a low of approximately 0.51 on each reactor. These parametric tests were conducted over relatively short periods of time and represent only a small fraction of the total reactor operating time.

- Ammonia Distribution

Ammonia distribution within the individual SCR reactors is extremely important in controlling ammonia slip and maintaining reactor efficiency. Assuming a perfectly smooth NO_x distribution, it is desirable to maintain an ammonia distribution that is as even as possible. The general criteria set for the pilot plant is to maintain ammonia distribution to within 10% of the average. Ammonia distributions were measured during the previous quarter downstream of the dummy bed on each reactor at nine equally spaced points on the large reactors and four on the small reactors. Ammonia distribution measurements indicated that all distributions were within the 10% limit, with averages comparing favorably with the predicted value.

- Flue Gas Temperature

The SCR facility is equipped with an economizer by-pass duct. This allows for high temperature flue gas extracted upstream of the host unit economizer to be mixed with flue gas extracted

downstream of the host unit economizer. By adjusting the relative flows of these two components of the pilot plant feed gas, the temperature to the pilot plant can be adjusted. Under normal operation, the flue gas temperature to the pilot plant is maintained at 650° F. This is possible when the host unit is running at relatively high loads; however, at low unit load, the pilot plant inlet gas temperature often drops below 650°F, even with full use of the economizer by-pass duct. Also, under some circumstances, while the host unit is operating at very high load, the feed gas to the pilot plant may exceed 650°F, even with no economizer by-pass gas being used. The average flue gas temperature to the test facility, as well as the daily high and low temperatures are shown below for this reporting period. As with the previously shown gas concentration data, these data are constructed using daily averages, daily highs, and daily lows, during periods of on-line host boiler operation.

Test Facility Inlet Gas Temperature (° F)

1994

<u>Date</u>	<u>Average</u>	<u>High</u>	<u>Low</u>
Jan.-Mar.	633	655	602
Apr.-Jun.	654	666	634

Each reactor is equipped with a flue gas heater which maintains strict control over the temperature of the flue gas entering the reactors. Under normal operating conditions the flue gas is maintained at 700°F at the entrance to the reactors. This requires that the gas be heated from approximately 650°F to 700°F under normal operating conditions. Of course, heat loss in the system requires some additional flue gas heating over the 50°F temperature difference noted. Under parametric conditions, the flue gas temperature to the reactors is varied from a low of approximately 620°F to a high of 750°F. Under these conditions, the economizer bypass duct flow rate may be adjusted to assist the heaters in obtaining the appropriate temperature. Under high temperature parametric conditions, heat loss through the system is more apparent, and flue gas temperatures of 780°F just downstream of the heater may be required to give 750°F at the reactor inlet. Lower temperature parametric conditions do not show as extreme a temperature loss between the heaters and the inlet to the reactors. As a result, the heater exit temperature is

normally much closer to the reactor inlet gas temperature during low temperature parametric tests and normal operating conditions.

- Air Preheater Performance Data

The three large reactors of the SCR facility are each equipped with an air preheater. Reactor A is equipped with a trisector Lungstrum type rotary air preheater supplied by ABB Air Preheater, Inc. of Wellsville, New York. Reactor B is equipped with a bisector rotary air preheater nearly identical to the A preheater, also supplied by ABB. Reactor C is equipped with an ABB Q-pipe which is a heat pipe design utilizing toluene and naphthalene as the working fluids. The original design of the SCR facility included air preheater bypass ducting which allowed the air preheaters to be bypassed during any condition other than normal operating conditions. This was done to ensure that the air preheater's long-term fouling characteristics were not skewed by extreme conditions during some of the short term parametric tests. The large reactor fan design requires relatively cool gas (less than 350°F). To accommodate this restriction, the air preheater bypass ducting was equipped with heat exchangers which were designed to cool the flue gas in place of the air preheaters. Unfortunately, the design of the by-pass heat exchangers caused immediate fouling upon use, making it unsatisfactory for the application. Consequently, the SCR facility is forced to use the three large reactor air preheaters at all times when on-line to maintain proper flue gas conditions for the large reactor fans. As a result, the air preheaters are exposed to the harsh conditions created by some of the parametric tests. However, these test periods are very short compared to the overall operating time at standard conditions, and it is assumed that overall fouling characteristics of the air preheaters are not greatly affected by the current operational requirements. Tables 3 and 4 show the average operating parameters for the three air preheaters over the reporting period, and these values include any parametric test conditions during the specific time period.

An initial air heater test series was completed prior to the May 1994 outage. This testing included measurement of the concentrations of particulates NH_3 , SO_2/SO_3 , and HCl . Ammonia samples were segregated into gas- and solid-phase fractions to characterize the partitioning of NH_3 across the air heaters. The fly ash was further analyzed to determine particle size distributions, ash mineralogy, and ash resistivity. The results of this testing are given in Appendix B. However, these results are very preliminary and, further testing will be performed. This and future air heater testing results will be summarized in the project final report, and full interpretation of results will be delayed until all the air heater testing and operations are completed.

TABLE 3.

AVERAGE AIR PREHEATER OPERATIONAL PARAMETERS, JAN. - MAR. 1994

<u>PARAMETER</u>	<u>REACTOR A</u>	<u>REACTOR B</u>	<u>REACTOR C</u>
GAS FLOW RATE (SCFM)	4998	5004	5103
AIR FLOW RATE (SCFM)	4491	3943	6770
INLET GAS TEMP. (°F)	677	672	667
EXIT GAS TEMP. (°F)	319	598	297
INLET AIR TEMP. (°F)	72	72	72
EXIT AIR TEMP. (°F)	605	598	474
GAS SIDE PRESS. DROP ("H ₂ O)	7.26	5.90	1.63
AIR SIDE PRESS. DROP ("H ₂ O)	2.51	2.34	NA
AIR/GAS DIFF. PRESS. ("H ₂ O)	0.50	0.50	NA
INLET GAS O ₂ (% WET)	5.02	4.94	5.43
EXIT GAS O ₂ (% WET)	8.37	7.40	NA

TABLE 4.

AVERAGE AIR PREHEATER OPERATIONAL PARAMETERS, APRIL - JUNE 1994

<u>PARAMETER</u>	<u>REACTOR A</u>	<u>REACTOR B</u>	<u>REACTOR C</u>
GAS FLOW RATE (SCFM)	4966	4947	4953
AIR FLOW RATE (SCFM)	5080	4463	7721
INLET GAS TEMP. (°F)	673	673	675
EXIT GAS TEMP. (°F)	305	338	299
INLET AIR TEMP. (°F)	88	88	88
EXIT AIR TEMP. (°F)	594	591	459
GAS SIDE PRESS. DROP ("H ₂ O)	7.67	4.74	1.85
AIR SIDE PRESS. DROP ("H ₂ O)	2.97	2.14	NA
AIR/GAS DIFF. PRESS. ("H ₂ O)	0.50	0.50	NA
INLET GAS O ₂ (% WET)	4.64	4.15	4.65
EXIT GAS O ₂ (% WET)	7.02	7.32	NA

- Catalyst-Specific Performance Parameters

The long term NO_x reduction and parametric test results characterizing the performance for Reactors A - G for this reporting period are discussed in the following sections. The parametric tests conducted during this reporting period are composed of thirteen reactor operating conditions defined by variations in the flue gas flow rate, temperature, and ammonia-to-NO_x ratio. The test conditions are listed in Table 5. The particular measurements (intermediate ammonia, slip ammonia, SO₂, SO₃, HCl, N₂O, and velocity and mass concentration profiles) are also shown in this table at the various test conditions. Although similar test conditions are indicated, the measurements were not taken simultaneously (e.g., ammonia and SO₂ data for the same conditions were not collected during the exact same test run, but rather under similar process conditions at different times). Parametric test data were not taken on Reactor J because replacing the catalyst withdrawn by Engelhard occurred so late in this reporting period.

Overall reactor pressure drops (inlet to outlet, including dummy layer and all catalysts) are shown in Figure 3 for Reactors A - G during January - March, 1994 and in Figure 4 for Reactors A - J for April - June, 1994. In general, the pressure drops ranged from 2 to 5 inches of water. The Siemens catalyst consistently displayed the lowest pressure drop at about 2 inches of water. The Cormetech low-dust catalyst in Reactor J had the highest pressure drop of between 5 and 7 inches of water, with one brief excursion to 9 inches of water resulting from the initial start-up problems after the May outage, as described earlier in this report.

- Reactor A

Table 6 shows the parametric test data on intermediate ammonia, slip ammonia, and sulfur dioxide oxidation collected during this reporting period for the Grace Noxeram catalyst. All the ammonia data are corrected to the reactor inlet oxygen concentration. The long term NO_x reduction is also given in this table as an average over the operating periods shown, i.e., for January - March and April - June of 1994. The long term NO_x reduction data indicate the average performance of the catalyst at or near the design operating conditions of 0.8 ammonia-to-NO_x ratio, 5000 SCFM flow rate, and 700°F reactor temperature.

TABLE 5. PARAMETRIC TEST CONDITIONS

Flue gas flow rate Large / Small reactor	Flue gas temperature	NH ₃ /NO _x Ratio	Measurements
<u>KSCFM</u>	<u>°F</u>		
3.0 / 0.24	620	0.6	intermediate NH ₃
3.0 / 0.24	620	0.8	SO ₂ /SO ₃
5.0 / 0.40	620	0.8	slip NH ₃
7.5 / 0.60	620	0.6	intermediate NH ₃
7.5 / 0.60	620	0.8	intermediate & slip NH ₃
7.5 / 0.60	620	1.0	intermediate & slip NH ₃
3.0 / 0.24	700	0.8	SO ₂ /SO ₃
5.0 / 0.40	700	0.8	* intermediate & slip NH ₃ , SO ₂ /SO ₃ , N ₂ O, HCl, velocity & mass concentration profile
7.5 / 0.60	700	0.8	slip NH ₃ , SO ₂ /SO ₃
3.0 / 0.24	750	0.8	SO ₂ /SO ₃
5.0 / 0.40	750	0.8	SO ₂ /SO ₃
7.5 / 0.60	750	0.6	intermediate NH ₃
7.5 / 0.60	750	0.8	slip NH ₃ & SO ₂ /SO ₃ **

*Design conditions. **An extra intermediate NH₃ was run on a few reactors.

The intermediate ammonia measurements were made after the first catalyst bed and at conditions thought to give the best kinetic information. The NO_x removals reported with the intermediate ammonia measurements are computed from the measured ammonia concentration using standard material balance techniques.

The ammonia slip data given in Table 6 are presented below in three plots: ammonia slip versus flow rate, ammonia-to-NO_x ratio, and temperature. Figure 5 shows ammonia slip versus flow rate for roughly 80% NO_x reduction. As expected, the trend shows increasing ammonia slip with increasing reactor flow rate. The ammonia slip is, however, relatively minor indicating the ability of the catalyst design to withstand significant increases in flow while maintaining ammonia slip limits. Since a portion of the overall reaction rate is due to mass transfer limitations,

Figure 3

Pressure Drop vs. Time

1st Quarter 1994

Pressure (inch. of H₂O)

Figure 4

Pressure Drop vs. Time

2nd Quarter 1994

Pressure (inch. of H2O)

TABLE 6. REACTOR A DATA

INTERMEDIATE AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	INT. NH ₃ (ppmv)	INT. NO _x REDUCTION (%)
2999	622	2.451	280	0.561	17.4	49.8
7161	625	2.240	252	0.561	30.0	44.2
7179	625	2.808	269	0.773	41.2	62.0
6998	626	3.633	311	0.978	45.7	83.1
4999	705	3.234	295	0.765	31.7	65.8
6234	751	4.826	351	0.568	28.1	48.8
6351	747	5.313	253	0.847	28.7	73.4

SLIP AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	SLIP NH ₃ (ppmv)
7000	625	6.328	312	0.800	3.8
6986	625	6.328	333	0.895	6.6
7092	625	4.638	287	1.059	21.7
4997	704	5.531	294	0.803	2.0
6297	704	6.316	316	0.799	3.0
5991	756	6.154	321	0.852	2.1

SULFUR DIOXIDE OXIDATION PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	OUTLET O ₂ (%)	INLET SO ₂ (ppmv)	NH ₃ /NO _x RATIO	PRED. SO ₃ IN (ppmv)	SO ₃ OUT	SO ₃ formed (ppmv)	OXID. RATE (%)
3005	624	3.261	1817	0.739	9.3	8.4	-0.9	-0.05
3002	705	3.611	1727	0.739	9.7	22.5	12.8	0.77
5001	710	2.666	1890	0.771	8.0	9.9	-17.9	0.10
6654	704	3.135	1787	0.781	8.0	10.0	2	0.11
2999	754	3.984	1691	0.740	9.3	23.8	14.5	0.89
4997	756	3.770	1685	0.777	9.3	27.6	18.3	1.11
6166	757	3.221	1725	0.765	9.3	15.4	6.1	0.36

LONG TERM NO_x REDUCTION

TIME PERIOD	FLOW RATE (SCFM)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	OUTLET NO _x (ppmv)	NO _x RED. (%)
JAN. - MAR.	5000	348	0.76	71	76
APRIL - JUNE	4992	347	0.78	66	78

Figure 5

Ammonia Slip vs. Flow Rate

1st & 2nd Quarter 1994

AMMONIA SLIP (ppm)

80% NO_x Reduction, 700° F, GRACE NOXERAM

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improvements in bulk mass transfer coefficients are likely mitigating the effect of increased flow on slip ammonia increases. This plot demonstrates the ability of an SCR system to follow load variations dictated by the host boiler while maintaining design specifications.

Figure 6 shows ammonia slip versus ammonia-to- NO_x ratio at low temperature and high flow rate (worst case scenario). The plot shows sharp increases in ammonia slip as the ammonia-to- NO_x ratio approaches 1.0. This finding is in keeping with published data of this type. At ammonia-to- NO_x ratios near 1.0, non-idealities in the reactor system force the catalyst to slip ammonia, since areas are present in the reactor where NO_x is the limiting reagent. Again, this plot is for the worst case scenario and ammonia slip values are likely to be lower for higher temperature or lower flow rate.

Ammonia slip versus temperature for high flow and roughly 80% NO_x reduction is plotted in Figure 7. Some improvement (decrease) in ammonia slip is noted between 620 and 700°F, likely due to improvements in the kinetic reaction rate with increasing temperature. Only slight improvements are noted with increasing temperature above 700°F. This may possibly be due to mass transfer limitations that have become controlling at these higher temperatures. In general, the plot demonstrates that in terms of ammonia slip, significant improvements are not realized with temperatures above 700°F. Losses in boiler efficiency would probably outweigh any improvements that may be obtained in ammonia slip by designing an SCR reactor to operate at temperatures near 750°F.

The plot for SO_2 oxidation versus flow rate at high temperature for the combination of all catalyst beds in the reactor is shown in Figure 8. In general, decreasing SO_2 oxidation is observed with increasing flow rate. This is expected since reaction rate is inversely proportional to flow rate according to published rate data and information provided by catalyst suppliers.

Figure 9 shows SO_2 oxidation versus temperature at low flow rate for the combination of all catalyst beds in the reactor. The trend of the plot shows a fairly linear increase in SO_2 oxidation with increasing temperature. Published information describing the effects of temperature on SO_2 oxidation and information obtained from the catalyst suppliers indicates that increases in SO_2 oxidation should be exponential with temperature. The exact shape of the plot is somewhat

Figure 6

Ammonia Slip vs. NH3/NOx Ratio

1st & 2nd Quarter 1994

AMMONIA SLIP (ppm)

620° F, High Flow, GRACE NOXERAM

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Figure 7

Ammonia Slip vs. Temperature

1st & 2nd Quarter 1994

80% NOx Reduction, High Flow, GRACE NOXERAM

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Figure 8

SO₂ Oxidation vs. Flow Rate

1st & 2nd Quarter 1994

80% NO_x Reduction, 750 °F, GRACE

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Figure 9

SO2 Oxidation vs. Temperature

1st & 2nd Quarter 1994

SO2 OXIDATION (%)

80% NOx Reduction, Low Flow, GRACE NOXERAM

s:/projects/doe/crist/doc/jdm/para2/so3vstea.pre

unclear since fairly large measurement variability exists in the test facility, and other phenomena, such as SO₃ deposition, occur in the test facility reactors. The general trend of increasing SO₂ oxidation is expected, however, and the plot provides a good basis for determining the potential for SO₃ formation across the SCR reactor at various operating temperatures.

The SO₂ oxidation data are corrected to reactor outlet oxygen concentrations. The value for sulfur trioxide produced in the reactor is based on the measured outlet sulfur trioxide concentration and estimated sulfur trioxide reactor inlet values. The reactor inlet sulfur trioxide values are estimated using past measurements of inlet sulfur trioxide based on host boiler load. In many cases, especially those involving low temperature conditions, there was no net increase in sulfur trioxide concentration across the reactor. In fact, some operating conditions showed a loss in sulfur trioxide across the reactor in addition to the loss that normally occurs without catalyst present. (Tables showing SO₂ oxidation rates present reactor flow rates as calculated for the reactor exit, since SO₃ is measured at this point. For consistency with other figures, the plots of SO₂ oxidation versus flowrate are based on reactor inlet flowrates. In practice, differences between inlet and outlet flowrates are slight and do not greatly affect the overall analysis.)

The pressure drop across the reactor is another important criteria for SCR usage because of its effect on fan horsepower requirements, the single largest energy need for the SCR process. Typical reactor pressure drop is shown in Figures 3 and 4.

Mass concentrations (five-point) and flue gas velocity (nine-point) profiles were conducted near design operating condition (700°F, 100% flow rate [5000 scfm]) at the reactor outlet and reactor inlet, respectively. The mass concentration profile data are given in Figure 10 in four different units of concentration and as mass emission rate. The average mass concentration was 3.49 ± 0.25 gr/dscf (with a relative standard deviation of 7.2%); and the average emission rate was 5.7 lb/MBtu. The flue gas velocity profile is presented in Figure 11. The average velocity was 13.3 ± 1.4 ft/sec. The velocity on the front side of the reactor was slightly higher than that on the back side.¹

HCl concentration was measured at the design operating condition at the reactor outlet. The average HCl concentration (at 3% O₂, dry) was 241 ± 11 ppm(v). The N₂O concentrations were measured at the reactor inlet [2.4 ppm(v)] and at the reactor outlet [1.2 ppm(v)] (both measurements are dry at 3% O₂).¹

Reactor A

Catalyst Layer 4 (Reactor outlet)

Mass Concentration Profile
gr/acf

Average = 1.32 ± 0.10 gr/acf

Mass Concentration Profile
gr/dscf

Average = 3.49 ± 0.25 gr/dscf

Mass Concentration Profile
g/acm

Average = 3.02 ± 0.23 g/acm

Mass Concentration Profile
g/dscm

Average = 7.98 ± 0.56 g/dscm

Mass Emission Rate Profile
lb/MBtu

Average = 5.7 ± 0.4 lb/MBtu

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Figure 10. Five presentations of the Reactor A mass concentration profile.

Reactor A

Catalyst Layer 1 Inlet

Velocity Profile ft/s

Average = 13.3 ± 1.4 ft/s

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Figure 11. Reactor A flue gas velocity profile.

- Reactor B

Table 7 shows the parametric test data on intermediate ammonia, slip ammonia, and sulfur dioxide oxidation collected during this reporting period for the NSKK catalyst. All the ammonia data are corrected to reactor inlet oxygen concentration. The long term NO_x reduction is also given in this table as an average over the operating periods shown, i.e., for January - March and April - June of 1994. The long term NO_x reduction data indicate the average performance of the catalyst at or near the design operating conditions of 0.8 ammonia-to-NO_x ratio, 5000 SCFM flow rate, and 700°F reactor temperature. As can be noted in the second quarter data averages, analytical errors and drift in calibrations led to values falsely indicating NO_x reduction being higher than the ammonia-to-NO_x ratio.

The intermediate ammonia measurements were made after the first catalyst bed and at conditions thought to give the best kinetic information. The NO_x removals reported with the intermediate ammonia measurements are computed from the measured ammonia concentration using standard material balance techniques.

The ammonia slip data given in Table 7 are presented below in three plots: ammonia slip versus flow rate, ammonia-to-NO_x ratio, and temperature. Figure 12 shows ammonia slip versus flow rate for roughly 80% NO_x reduction. As expected, the trend shows increasing ammonia slip with increasing reactor flow rate. The ammonia slip is, however, relatively minor, indicating the ability of the catalyst to withstand significant increases in flow while maintaining ammonia slip limits. Since the overall reaction rate is controlled partly by mass transfer limitations, improvements in bulk mass transfer coefficients are likely mitigating the effect of increased flow on slip ammonia increases. This plot demonstrates the ability of an SCR system to follow host boiler load variations while maintaining design specifications.

Figure 13 shows ammonia slip versus ammonia-to-NO_x ratio at low temperature and high flow rate (worst case scenario). The plot shows sharp increases in ammonia slip as the ammonia-to-NO_x ratio approaches 1.0. This finding is in keeping with published data of this type. At ammonia-to-NO_x ratios near 1.0, non-idealities in the reactor system force the catalyst to slip ammonia since areas are present in the reactor where NO_x is the limiting reagent. Again, this plot is for the worst case scenario and ammonia slip values are likely to be lower for higher temperature or lower flow rate.

TABLE 7. REACTOR B DATA

INTERMEDIATE AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ / NO _x RATIO	INT. NH ₃ (ppmv)	INT. NO _x REDUCTION (%)
2996	626	4.965	354	0.564	5.5	54.8
6690	632	3.323	280	0.560	54.4	36.5
7232	623	5.912	353	0.786	54.2	63.2
7271	623	1.914	258	0.828	86.5	49.2
4999	706	2.531	272	0.763	19.4	69.2
6808	755	4.736	289	0.573	22.8	49.4
6630	757	4.699	269	0.792	26.1	69.5

SLIP AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ / NO _x RATIO	SLIP NH ₃ (ppmv)
7000	625	6.154	319	0.880	0.2
7003	628	6.154	335	0.889	1.6
6948	623	5.809	335	1.015	5.0
4992	706	5.066	300	0.781	0.4
6599	707	6.341	317	0.798	0.6
6503	755	5.895	328	0.839	0.6

SULFUR DIOXIDE OXIDATION PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	OUTLET O ₂ (%)	INLET SO ₂ (ppmv)	NH ₃ /NO _x RATIO	PRED. SO ₃ IN	SO ₃ OUT	SO ₃ formed (ppmv)	OXID. RATE (%)
3006	624	2.397	1926	0.743	9.3	3.9	-5.4	-0.28
2999	710	2.013	2116	0.761	8.9	21.8	12.9	0.62
4999	706	2.435	1902	0.752	9.3	7.2	-2.1	-0.11
6427	704	2.949	1696	0.769	10.1	4.7	-5.4	-0.32
5000	750	2.365	1914	0.770	9.3	18.8	9.5	0.50
4955	753	2.365	1815	0.760	9.7	12.4	2.7	0.15
6604	752	3.128	1679	0.752	9.7	14.2	4.5	0.27

LONG TERM NO_x REDUCTION

TIME PERIOD	FLOW RATE (SCFM)	INLET NO _x (ppmv)	NH ₃ NO _x RATIO	OUTLET NO _x (ppmv)	NO _x RED. (%)
JAN. - MAR.	5070	344	0.77	71	76
APRIL - JUNE	4963	359	0.77	64	81

Figure 12

Ammonia Slip vs. Flow Rate

1st & 2nd Quarter 1994

AMMONIA SLIP (ppm)

80% NOx Reduction, 700 °F, NSKK

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Figure 13

Ammonia Slip vs. NH₃/NO_x Ratio

1st & 2nd Quarter 1994

620 °F, High Flow, NSKK

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Ammonia slip versus temperature for high flow and roughly 80% NO_x reduction is plotted in Figure 14. Some improvement (decrease) in ammonia slip is noted between 620 and 700°F, likely due to improvements in the reaction rate with increasing temperature. Only slight improvements are noted with increasing temperature above 700°F. This may be due to mass transfer limitations that become controlling at these higher temperatures. In general, the plot demonstrates that in terms of ammonia slip, significant improvements are not realized at temperatures above 700°F. Losses in boiler efficiency would probably outweigh any improvements in ammonia slip for an SCR reactor operating at temperatures near 750°F.

The plot for SO₂ oxidation versus flow rate at high temperature is shown in Figure 15. In general, decreasing SO₂ oxidation is observed with increasing flow rate. This finding is expected since the reaction rate is inversely proportional to flow rate, according to published rate data and information provided by catalyst suppliers.

Figure 16 shows SO₂ oxidation versus temperature at low flow rate. The plot shows an increase in SO₂ oxidation with increasing temperature. Published information, describing the effects of temperature on SO₂ oxidation and information obtained from the catalyst suppliers, indicates that SO₂ oxidation follows an exponential relationship with temperature. The exact shape of the plot is somewhat unclear, since fairly large measurement variability exists in the test facility, and other phenomena, such as SO₃ deposition, occur in the test facility reactors. The extremely low oxidation rate of the catalyst further exacerbates this problem. A general trend of increasing SO₂ oxidation with temperature is expected, however, and the plot provides a good basis for determining the potential for SO₃ formation across the SCR reactor at various operating temperatures.

The SO₂ oxidation data are corrected to reactor outlet oxygen concentrations. The value for sulfur trioxide produced in the reactor is based on the measured outlet sulfur trioxide concentration and estimated sulfur trioxide reactor inlet values. The reactor inlet sulfur trioxide values are estimated using past measurements of inlet sulfur trioxide based on host boiler load. In many cases, especially those involving low temperature conditions, there was no net increase in sulfur trioxide concentrations across the reactor. In fact, some operating conditions showed a loss in sulfur trioxide across the reactor in addition to the loss that normally occurs without catalyst present. (Tables showing SO₂ oxidation rates quote reactor flow rates as calculated for

Figure 14

Ammonia Slip vs. Temperature

1st & 2nd Quarter 1994

AMMONIA SLIP (ppm)

80% NOx Reduction, High Flow, NSKK

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Figure 15

SO2 Oxidation vs. Flow Rate

1st & 2nd Quarter 1994

SO2 OXIDATION (%)

80% NOx Reduction, 750 °F, NSKK

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Figure 16

SO₂ Oxidation vs. Temperature

1st & 2nd Quarter 1994

SO₂ OXIDATION (%)

80% NO_x Reduction, Low Flow, NSKK

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the reactor exit, since SO_3 is measured at this point. For consistency with other figures, the plots of SO_2 oxidation versus flowrate are based on reactor inlet flowrates. In practice, differences between inlet and outlet flowrates are slight and do not greatly affect the overall analysis.)

The pressure drop across the reactor is another important criteria for SCR usage because of its effect on fan horsepower requirements, the single largest energy need for the SCR process. Typical reactor pressure drop is shown in Figures 3 and 4.

Mass concentrations (five-point) and flue gas velocity (nine-point) profiles were conducted near design operating condition (700°F , 100% flow rate [5000 scfm]) at the reactor outlet and reactor inlet, respectively. The mass concentration profile data are given in Figure 17 in four different units of concentration and as mass emission rate. The average mass concentration was 3.22 ± 0.52 gr/dscf (with a relative standard deviation of 16.1%); and the average emission rate was 5.2 lb/MBtu. The flue gas velocity profile is presented in Figure 18. The average velocity was 12.2 ± 0.5 ft/sec.¹

HCl concentration was measured at the design operating condition at the reactor outlet. The average HCl concentration (at 3% O_2 , dry) was 158 ± 36 ppm(v). The N_2O concentrations were measured at the reactor inlet [2.4 ppm(v)] and at the reactor outlet [1.0 ppm(v)] (both measurements are dry at 3% O_2).¹

Reactor C

Table 8 shows the parametric test data on intermediate ammonia, slip ammonia, and sulfur dioxide oxidation collected during this reporting period for the Siemens catalyst. All the ammonia data are corrected to reactor inlet oxygen concentration. The long term NO_x reduction is given in this table as an average over the operating periods shown, i.e., for January - March and April - June of 1994. The long term NO_x reduction data indicate the average performance of the catalyst at or near the design operating conditions of 0.8 ammonia-to- NO_x ratio, 5000 SCFM flow rate, and 700°F reactor temperature. As can be noted in the second quarter data averages, analytical errors and drift in calibrations led to values falsely indicating NO_x reduction being higher than the ammonia-to- NO_x ratio.

Reactor B Catalyst Layer 4 (Reactor outlet)

Mass Concentration Profile
gr/acf

Average = 1.21 ± 0.22 gr/acf

Mass Concentration Profile
gr/dscf

Average = 3.22 ± 0.52 gr/dscf

Mass Concentration Profile
g/acm

Average = 2.78 ± 0.50 g/acm

Mass Concentration Profile
g/dscm

Average = 7.36 ± 1.20 g/dscm

Mass Emission Rate Profile
lb/MBtu

Average = 5.2 ± 0.9 lb/MBtu

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Figure 17. Five presentations of the Reactor B mass concentration profile.

Reactor B

Catalyst Layer 1 Inlet

Velocity Profile ft/s

Average = 12.2 ± 0.5 ft/s

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Figure 18. Reactor B flue gas velocity profile.

TABLE 8. REACTOR C DATA

INTERMEDIATE AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	INT. NH ₃ (ppmv)	INT. NO _x REDUCTION (%)
2990	625	3.468	291	0.564	1.5	55.9
7447	629	3.584	300	0.768	19.5	70.3
7487	625	5.167	364	0.811	23.7	74.6
7428	623	4.975	272	1.018	47.2	84.5
5003	703	5.059	350	0.982	12.4	94.6
7001	745	3.227	277	0.562	12.8	51.6

SLIP AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	SLIP NH ₃ (ppmv)
5007	623	4.344	293	0.832	2.2
7441	622	5.911	331	0.801	4.8
7477	624	4.609	266	1.017	10.6
4995	704	4.170	342	0.783	1.1
7408	707	4.814	365	0.782	2.3
7202	743	6.191	301	0.832	3.0

SULFUR DIOXIDE OXIDATION PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	OUTLET O ₂ (%)	INLET SO ₂ (ppmv)	NH ₃ /NO _x RATIO	PRED. SO ₃ IN (ppmv)	SO ₃ OUT	SO ₃ formed (ppmv)	OXID. RATE (%)
2999	621	3.060	1891	0.770	8.0	7.5	-0.5	-0.03
3020	709	3.408	1804	0.775	8.0	17.4	9.4	0.54
5002	706	3.671	1693	0.754	9.7	19.8	10.1	0.62
7497	704	3.423	1790	0.752	9.3	10.1	0.8	0.04
3002	756	4.210	1670	0.781	9.3	50.5	41.2	2.55
5000	755	3.877	1689	0.775	9.3	39.7	30.4	1.85
7299	749	3.579	1891	0.743	9.7	26.9	17.2	0.95

LONG TERM NO_x REDUCTION

TIME PERIOD	FLOW RATE (SCFM)	INLET NO _x (ppmv)	NH ₃ / NO _x RATIO	OUTLET NO _x (ppmv)	NO _x RED. (%)
JAN. - MAR.	5101	348	0.77	75	75
APRIL - JUNE	5000	354	0.78	59	81

The intermediate ammonia measurements were made after the first catalyst bed and at conditions thought to give the best kinetic information. The NO_x removals reported with the intermediate ammonia measurements are computed from the measured ammonia concentration using standard material balance techniques.

The ammonia slip data given in Table 8 are presented below in three plots: ammonia slip versus flow rate, ammonia-to- NO_x ratio, and temperature. Figure 19 shows ammonia slip versus flow rate for roughly 80% NO_x reduction. As expected, the trend shows increasing ammonia slip with increasing reactor flow rate. The ammonia slip is, however, relatively minor indicating the ability of the catalyst to withstand significant increases in flow while maintaining ammonia slip limits. Since the overall reaction rate is partly controlled by mass transfer limitations, the improvements in bulk mass transfer coefficients are likely mitigating the effect of increased flow on slip ammonia increases. This plot demonstrates the ability of an SCR system to follow host boiler load variations while maintaining design specifications.

Figure 20 shows ammonia slip versus ammonia-to- NO_x ratio at low temperature and high flow rate (worst case scenario). The plot shows sharp increases in ammonia slip as the ammonia-to- NO_x ratio approaches 1.0. This finding is in keeping with published data of this type. At ammonia-to- NO_x ratios near 1.0, non-idealities in the reactor system force the catalyst to slip ammonia since areas are present in the reactor where NO_x is the limiting reagent. Again, this plot is for the worst case scenario and ammonia slip values are likely to be lower for higher temperature or lower flow rate.

Ammonia slip versus temperature for high flow and roughly 80% NO_x reduction is plotted in Figure 21. Some improvement (decrease) in ammonia slip is noted between 620 and 700°F, likely due to improvements in the reaction rate with increasing temperature. The slight increase shown between the 700 and 750°F points is likely due to measurement variability and is not considered significant. In this case, the plot should not be construed as demonstrating increases in ammonia slip with increasing temperature above 700°F. It is expected that the 700 and 750°F values are roughly equivalent which may be due in part to mass transfer limitations that become controlling at these higher temperatures. In general, the plot demonstrates that in terms of ammonia slip, significant improvements are not realized at temperatures above 700°F. Losses in boiler efficiency would probably outweigh any improvements in ammonia slip from an SCR reactor operating at temperatures near 750°F.

Figure 19

Ammonia Slip vs. Flow Rate

1st & 2nd Quarter 1994

AMMONIA SLIP (ppm)

80% NOx Reduction, 700 °F, SIEMENS

s:/projects/doe/crist/doc/jdm/para2/slpvsflc.pre

Figure 20
Ammonia Slip vs. NH₃/NO_x Ratio
1st & 2nd Quarter 1994

AMMONIA SLIP (ppm)

620 °F, High Flow, SIEMENS

s:/projects/doe/crist/doc/jdm/para2/slpvsrc.pre

Figure 21

Ammonia Slip vs. Temperature

1st & 2nd Quarter 1994

80% NO_x Reduction, High Flow, SIEMENS

s:/projects/doe/crist/doe/jdm/para2/slpvstc

The plot for SO₂ oxidation versus flow rate at high temperature for the combination of all catalyst beds in the reactor is shown in Figure 22. In general, decreasing SO₂ oxidation is noted with increasing flow rate. This finding is expected since the reaction rate is predicted to be inversely proportional to flow rate according to published rate data and information provided by catalyst suppliers.

Figure 23 shows SO₂ oxidation versus temperature at low flow rate for the combination of all catalyst beds in the reactor. The trend of the plot shows a fairly linear increase in SO₂ oxidation with increasing temperature. Published information describing the effects of temperature on SO₂ oxidation and information obtained from the catalyst suppliers indicates that increases in SO₂ oxidation should be exponential with temperature. The exact shape of the plot is somewhat unclear, since fairly large measurement variability exists in the test facility and other phenomena, such as SO₃ deposition, occur in the test facility reactors. The general trend of increasing SO₂ oxidation is expected, however, and the plot provides a good basis for determining the potential for SO₃ formation across the SCR reactor at various operating temperatures.

The SO₂ oxidation data are corrected to reactor outlet oxygen concentrations. The value for sulfur trioxide produced in the reactor is based on the measured outlet sulfur trioxide concentration and estimated sulfur trioxide reactor inlet values. The reactor inlet sulfur trioxide values are estimated using past measurements of inlet sulfur trioxide based on host boiler load. In many cases, especially those involving low temperature conditions, there was no net increase in sulfur trioxide concentration across the reactor. In fact, some operating conditions showed a loss in sulfur trioxide across the reactor in addition to the loss that normally occurs without catalyst present. (Tables showing SO₂ oxidation rates present reactor flow rates as calculated for the reactor exit, since SO₃ is measured at this point. For consistency with other figures, the plots of SO₂ oxidation versus flowrate are based on reactor inlet flowrates. In practice, differences between inlet and outlet flowrates are slight and do not greatly affect the overall analysis.)

The pressure drop across the reactor is another important criteria for SCR usage because of its effect on fan horsepower requirements, the single largest energy need for the SCR process. Typical reactor pressure drop is shown in Figures 3 and 4.

Figure 22

SO2 Oxidation vs. Flow Rate

1st & 2nd Quarter 1994

80% NOx Reduction, 750 °F, SIEMENS

s:/projects/doe/crist/doc/jdm/para2/so3vsfc.pre

Figure 23

SO₂ Oxidation vs. Temperature

1st & 2nd Quarter 1994

SO₂ OXIDATION (%)

80% NO_x Reduction, Low Flow, SIEMENS

s:/projects/doe/crist/doc/jdm/para2/so3vstec.pre

Mass concentrations (five-point) and flue gas velocity (nine-point) profiles were conducted near design operating condition (700°F, 100% flow rate [5000 scfm]) at the reactor outlet and reactor inlet, respectively. The mass concentration profile data are given in Figure 24 in four different units of concentration and as mass emission rate. The average mass concentration was 3.27 ± 0.29 gr/dscf (with a relative standard deviation of 8.9%); and the average emission rate was 5.5 lb/MBtu. The flue gas velocity profile is presented in Figure 25. The average velocity was 12.4 ± 1.0 ft/sec.¹

HCl concentration was measured at the design operating condition at the reactor outlet. The average HCl concentration (at 3% O₂, dry) was 211 ± 14 ppm(v). The N₂O concentrations were measured at the reactor inlet [2.4 ppm(v)] and at the reactor outlet [1.7 ppm(v)] (both measurements are dry at 3% O₂).¹

- Reactor D

Table 9 shows the parametric test data on intermediate ammonia, slip ammonia, and sulfur dioxide oxidation collected during this reporting period for the Grace Synox catalyst. All the ammonia data are corrected to reactor inlet oxygen concentration. The long term NO_x reduction is given in this table as an average over the operating periods shown, i.e., for January - March and April - June of 1994. The long term NO_x reduction data indicate the average performance of the catalyst at or near the design operating conditions of 0.8 ammonia-to-NO_x ratio, 400 SCFM flow rate, and 700°F reactor temperature.

The intermediate ammonia measurements were made after the first catalyst bed and at conditions thought to give the best kinetic information. The NO_x removals reported with the intermediate ammonia measurements are computed from the measured ammonia concentration using standard material balance techniques.

The ammonia slip data given in Table 9 are presented below in three plots: ammonia slip versus flow rate, ammonia-to-NO_x ratio, and temperature. Figure 26 shows ammonia slip versus flow rate for roughly 95% NO_x reduction. As expected, the trend shows increasing slip with increasing reactor flow rate. A fairly significant increase in ammonia slip is noted with flow rate. However, this ammonia slip is not likely to be nearly as significant at lower NO_x removal rates. Since the overall reaction rate is partly controlled by mass transfer limitations, improvements in bulk mass transfer coefficients are likely mitigating the effect of increased flow on slip ammonia increases.

Reactor C

Catalyst Layer 4 (Reactor outlet)

Mass Concentration Profile
gr/acf

Average = 1.21 ± 0.10 gr/acf

Mass Concentration Profile
gr/dscf

Average = 3.27 ± 0.29 gr/dscf

Mass Concentration Profile
g/acm

Average = 2.76 ± 0.24 g/acm

Mass Concentration Profile
g/dscm

Average = 7.49 ± 0.66 g/dscm

Mass Emission Rate Profile
lb/MBtu

Average = 5.5 ± 0.5 lb/MBtu

s:\ptd\maj\maj\reactor.pre

Figure 24. Five presentations of the Reactor C mass concentration profile.

Reactor C

Catalyst Layer 1 Inlet

Velocity Profile ft/s

Average = 12.4 ± 1.0 ft/s

s:\pdt\maj\maj\reactor.pre

Figure 25. Reactor C flue gas velocity profile.

TABLE 9. REACTOR D DATA

INTERMEDIATE AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	INT. NH ₃ (ppmv)	INT. NO _x REDUCTION(%)
239	625	5.425	267	0.771	20.0	69.6
601	624	4.319	294	0.706	59.1	50.5
600	624	4.475	277	0.915	88.7	59.5
602	624	4.440	282	1.149	147.2	62.7
403	705	4.473	289	0.989	54.8	79.9
598	753	4.094	294	0.711	44.2	56.1

SLIP AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	SLIP NH ₃ (ppmv)
382	623	7.505	324	1.050	0.9
597	627	5.416	304	0.930	25.6
602	627	4.589	263	1.192	50.0
399	705	4.284	231	1.028	3.0
588	705	3.851	232	0.941	8.5
599	705	5.165	321	0.903	11.2

SULFUR DIOXIDE OXIDATION PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	OUTLET O ₂ (%)	INLET SO ₂ (ppmv)	NH ₃ /NO _x RATIO	PRED.SO ₃ IN (ppmv)	SO ₃ OUT	SO ₃ formed (ppmv)	OXID. RATE (%)
240	625	6.414	1896	1.045	9.3	2.1	-7.2	-0.44
240	709	6.176	1742	1.010	10.1	17.3	7.2	0.46
400	705	5.901	1629	0.996	9.7	18.5	8.8	0.58
600	704	5.090	1746	0.912	9.7	8.1	-1.6	-0.09
240	756	6.688	1604	1.060	10.1	21.0	10.9	0.74
400	753	5.079	1831	1.033	9.3	21.0	11.7	0.65
600	746	4.706	1825	0.926	8.9	21.6	12.7	0.73

LONG TERM NO_x REDUCTION

TIME PERIOD	FLOW RATE (SCFM)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	OUTLET NO _x (ppmv)	NO _x RED. (%)
JAN. - MAR.	399	345	0.77	68	73
APRIL - JUNE	404	345	0.78	78	77

Figure 26

Ammonia Slip vs. Flow Rate

1st & 2nd Quarter 1994

95% NOx Reduction, 700 °F, GRACE SYNOX

s:/projects/doe/crist/doc/jdm/para2/slpvsfld.pre

Figure 27 shows ammonia slip versus ammonia-to-NO_x ratio at low temperature and high flow rate (worst case scenario). The plot shows sharp increases in ammonia slip as the ammonia-to-NO_x ratio approaches 1.0. This finding is in keeping with published data of this type. At ammonia-to-NO_x ratios near 1.0, non-idealities in the reactor system force the catalyst to slip ammonia since areas are present in the reactor where NO_x is the limiting reagent. Again, this plot is for the worst case scenario and ammonia slip values are likely to be lower for higher temperature or lower flow rate.

Ammonia slip versus temperature for high flow and roughly 90% NO_x reduction is plotted in Figure 28. Some improvement (decrease) in ammonia slip is noted between 620 and 700°F, likely due to improvements in the reaction rate with increasing temperature. The slight increase shown between the 700 and 750°F points is likely due to measurement variability and is not considered significant. In this case, the plot should not be construed as demonstrating increases in ammonia slip with increasing temperature above 700°F. It is expected that the 700 and 750°F values are roughly equivalent which may be due in part to mass transfer limitations that become controlling at these higher temperatures. In general, the plot demonstrates that in terms of ammonia slip, significant improvements are not realized at temperatures above 700°F. Losses in boiler efficiency would probably outweigh any improvements in ammonia slip from an SCR reactor operating at temperatures near 750°F.

The plot for SO₂ oxidation versus flow rate at high temperature is shown in Figure 29. The plot is relatively flat showing little change in SO₂ oxidation rate. Published rate equations and catalyst supplier information indicates that the general form of the plot should be linear with SO₂ oxidation decreasing with flow rate. In this case, the relatively flat profile indicates that either chemical or physical characteristics of the catalyst and reactor are masking this linear effect.

Figure 30 shows SO₂ oxidation versus temperature at low flow rate for the combination of all catalyst beds in the reactor. The trend of the plot shows a fairly linear increase in SO₂ oxidation with increasing temperature. Published information describing the effects of temperature on SO₂ oxidation and information obtained from the catalyst suppliers indicates that increases in SO₂ oxidation should be exponential with temperature. The exact shape of the plot is somewhat unclear, since fairly large measurement variability exists in the test facility, and other phenomena, such as SO₃ deposition, occur in the test facility reactors. The general trend of increasing SO₂ oxidation is expected, however, and the plot provides a good basis for determining the potential for SO₃ formation across the SCR reactor at various operating temperatures.

Figure 27
Ammonia Slip vs. NH₃/NO_x Ratio
1st & 2nd Quarter 1994

Figure 28

Ammonia Slip vs. Temperature

1st & 2nd Quarter 1994

90% NOx Reduction, High Flow, GRACE SYNOX

s:/projects/doe/crist/doc/jdm/para2/slpvstd.pre

Figure 29

SO₂ Oxidation vs. Flow Rate

1st & 2nd Quarter 1994

98% NO_x Reduction, 750 °F, GRACE SYNOX

s:/projects/doe/crist/doc/jdm/para2/so3vsfd.pre

Figure 30

SO₂ Oxidation vs. Temperature

1st & 2nd Quarter 1994

98% NO_x Reduction, Low Flow, GRACE SYNOX

s:/projects/doe/crist/doc/jdm/para2/so3vsted.pre

The SO₂ oxidation data are corrected to reactor outlet oxygen concentrations. The value for sulfur trioxide produced in the reactor is based on the measured outlet sulfur trioxide concentration and estimated sulfur trioxide reactor inlet values. The reactor inlet sulfur trioxide values are estimated using past measurements of inlet sulfur trioxide based on host boiler load. In many cases, especially those involving low temperature conditions, there was no net increase in sulfur trioxide concentration across the reactor. In fact, some operating conditions showed a loss in sulfur trioxide across the reactor in addition to the loss that normally occurs without catalyst present. (Tables showing SO₂ oxidation rates present reactor flow rates as calculated for the reactor exit, since SO₃ is measured at this point. For consistency with other figures, the plots of SO₂ oxidation versus flowrate are based on reactor inlet flowrates. In practice, differences between inlet and outlet flowrates are slight and do not greatly affect the overall analysis.)

The pressure drop across the reactor is another important criteria for SCR usage because of its effect on fan horsepower requirements, the single largest energy need for the SCR process. Typical reactor pressure drop is shown in Figures 3 and 4.

Mass concentrations (two-point) and flue gas velocity (three-point) profiles were conducted near design operating condition (700°F, 100% flow rate [400 scfm]) at the reactor outlet and reactor inlet, respectively. The mass concentration profile data are given in Figure 31 in four different units of concentration and as mass emission rate. The average mass concentration was 3.72 ± 0.58 gr/dscf (with a relative standard deviation of 15.6%); and the average emission rate was 7.1 lb/MBtu. The flue gas velocity profile is presented in Figure 32. The average velocity was 13.2 ± 0.1 ft/sec.¹

HCl concentration was measured at the design operating condition at the reactor outlet. The average HCl concentration (at 3% O₂, dry) was 160 ± 8 ppm(v). The N₂O concentrations were measured at the reactor inlet [1.8 ppm(v)] and at the reactor outlet [1.8 ppm(v)] (both measurements are dry at 3% O₂).¹

- Reactor E

Table 10 shows the parametric test data on intermediate ammonia, slip ammonia, and sulfur dioxide oxidation collected during this reporting period for the Cormetech high dust catalyst. All the ammonia data are corrected to reactor inlet oxygen concentration. The long term NO_x

Reactor D

Catalyst Layer 4 (Reactor outlet)

Mass Concentration Profile
gr/acf

Average = 1.47 ± 0.25 gr/acf

Mass Concentration Profile
gr/dscf

Average = 3.72 ± 0.58 gr/dscf

Mass Concentration Profile
g/acm

Average = 3.36 ± 0.57 g/acm

Mass Concentration Profile
g/dscm

Average = 8.51 ± 1.33 g/dscm

Mass Emission Rate Profile
lb/MBtu

Average = 7.1 ± 1.1 lb/MBtu

s:\ptd\vmaj\vmaj\reactor.pre

Figure 31. Five presentations of the Reactor D mass concentration profile.

Reactor D

Catalyst Layer 1 Inlet

Velocity Profile ft/s

Average = 13.2 ± 0.1 ft/s

s:\p\td\maj\maj\reactor.pre

Figure 32. Reactor D flue gas velocity profile.

TABLE 10. REACTOR E DATA

INTERMEDIATE AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	INT. NH ₃ (ppmv)	INT. NO _x REDUCTION (%)
238	627	3.025	280	0.569	7.1	54.3
600	623	2.332	321	0.528	40.1	40.3
599	627	5.278	372	0.729	47.3	60.1
600	625	4.723	372	0.926	111.0	62.8
400	703	6.832	362	0.733	18.8	68.1
600	741	4.333	349	0.545	23.6	47.8

SLIP AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	SLIP NH ₃ (ppmv)
400	625	3.446	328	0.751	0.8
599	626	3.384	305	0.776	1.2
597	627	2.750	313	0.943	5.9
401	706	3.698	327	0.750	1.0
600	704	2.598	288	0.727	0.5
595	756	3.078	325	0.738	1.0

SULFUR DIOXIDE OXIDATION PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	OUTLET SO ₂ (%)	INLET SO ₂ (ppmv)	NH ₃ /NO _x RATIO	PRED. SO ₃ IN (ppmv)	SO ₃ OUT	SO ₃ formed (ppmv)	OXID. RATE (%)
240	625	3.493	1876	0.751	9.3	11.0	1.7	0.10
244	700	4.171	1695	0.740	10.5	15.5	5	0.31
400	708	3.631	1717	0.731	10.1	21.1	11	0.66
583	708	2.785	1935	0.729	9.7	19.4	9.7	0.53
240	755	3.943	1810	0.740	10.1	28.6	18.5	1.08
389	755	3.435	1887	0.724	9.7	29.3	19.6	1.08
601	756	3.374	1836	0.718	9.7	30.5	20.8	1.17

LONG TERM NO_x REDUCTION

TIME PERIOD	FLOW RATE (SCFM)	INLET NO _x (ppmv)	NH ₃ / NO _x RATIO	OUTLET NO _x (ppmv)	NO _x RED. (%)
JAN. - MAR.	391	343	0.75	76	73
APRIL - JUNE	402	331	0.78	52	82

reduction is given in this table as an average over the operating periods, i.e., for January - March and April - June of 1994. The long term NO_x reduction data indicate the average performance of the catalyst at or near the design operating conditions of 0.8 ammonia-to-NO_x ratio, 400 SCFM flow rate, and 700°F reactor temperature. As can be noted in the second quarter data averages, analytical errors and drift in calibrations led to values falsely indicating NO_x reduction being higher than the ammonia-to-NO_x ratio.

The intermediate ammonia measurements were made after the first catalyst bed and at conditions thought to give the best kinetic information. The NO_x removals reported with the intermediate ammonia measurements are computed from the measured ammonia concentration using standard material balance techniques.

The ammonia slip data given in Table 10 are presented below in three plots: ammonia slip versus flow rate, ammonia-to-NO_x ratio, and temperature. Figure 33 shows ammonia slip versus flow rate for roughly 80% NO_x reduction. As expected, the trend shows increasing ammonia slip with increasing reactor flow rate. The ammonia slip is, however, relatively minor indicating the ability of the catalyst to withstand significant increases in flow while maintaining ammonia slip limits. Since the overall reaction rate is partly controlled by mass transfer limitations, improvements in bulk mass transfer coefficients are likely mitigating the effect of increased flow on slip ammonia increases. This plot demonstrates the ability of an SCR system to follow host boiler load variations while maintaining design specifications.

Figure 34 shows ammonia slip versus ammonia-to-NO_x ratio at low temperature and high flow rate (worst case scenario). The plot shows sharp increases in ammonia slip as the ammonia-to-NO_x ratio approaches 1.0. This finding is in keeping with published data of this type. At ammonia-to-NO_x ratios near 1.0, non-idealities in the reactor system force the catalyst to slip ammonia since areas are present in the reactor where NO_x is the limiting reagent. Again, this plot is for the worst case scenario and ammonia slip values are likely to be lower for higher temperature or lower flow rate.

Ammonia slip versus temperature for high flow and roughly 80% NO_x reduction is plotted in Figure 35. Some improvement (decrease) in ammonia slip is noted between 620 and 700°F, likely due to improvements in the reaction rate with increasing temperature. The slight increase shown between the 700 and 750°F points is likely due to measurement variability and is not considered significant. In this case, the plot should not be construed as demonstrating increases in ammonia

Figure 33
Ammonia Slip vs. Flow Rate
1st & 2nd Quarter 1994

Figure 34

Ammonia Slip vs. NH₃/NO_x Ratio

1st & 2nd Quarter 1994

AMMONIA SLIP (ppm)

620 °F, High Flow, CORMETECH

s:/projects/doe/crist/doc/jdm/para2/slpvrte.pre

Figure 35

Ammonia Slip vs. Temperature

1st & 2nd Quarter 1994

80% NO_x Reduction, High Flow, CORMETECH

s:/projects/doe/crist/doc/jdm/para2/slpvste.pre

slip with increasing temperature above 700°F. It is expected that the 700 and 750°F values are roughly equivalent which may be due in part to mass transfer limitations that become controlling at these higher temperatures. In general, the plot demonstrates that in terms of ammonia slip, significant improvements are not realized at temperatures above 700°F. Losses in boiler efficiency would probably outweigh any improvements in ammonia slip from an SCR reactor operating at temperatures near 750°F.

The plot for SO₂ oxidation versus flow rate at high temperature is shown in Figure 36. The plot is relatively flat showing little change in SO₂ oxidation rate. Published rate equations and catalyst supplier information indicates that the general form of the plot should be linear with SO₂ oxidation decreasing with flow rate. In this case, the relatively flat profile indicates that either chemical or physical characteristics of the catalyst and reactor are masking this linear effect.

Figure 37 shows SO₂ oxidation versus temperature at low flow rate. The plot shows a fairly linear increase in SO₂ oxidation with increasing temperature. Published information, describing the effects of temperature on SO₂ oxidation and information obtained from the catalyst suppliers, indicates that SO₂ oxidation follows an exponential relationship with temperature. The exact shape of the plot is somewhat unclear, since fairly large measurement variability exists in the test facility and other phenomena, such as SO₃ deposition, occur in the test facility reactors. The general trend of increasing SO₂ oxidation with temperature is expected, however, and the plot provides a good basis for determining the potential for SO₃ formation across the SCR reactor at various operating temperatures.

The SO₂ oxidation data are corrected to reactor outlet oxygen concentrations. The value for sulfur trioxide produced in the reactor is based on the measured outlet sulfur trioxide concentration and estimated sulfur trioxide reactor inlet values. The reactor inlet sulfur trioxide values are estimated using past measurements of inlet sulfur trioxide based on host boiler load. (Tables showing SO₂ oxidation rates quote reactor flow rates as calculated for the reactor exit, since SO₃ is measured at this point. For consistency with other figures, the plots of SO₂ oxidation versus flowrate are based on reactor inlet flowrates. In practice, differences between inlet and outlet flowrates are slight and do not greatly affect the overall analysis.)

The pressure drop across the reactor is another important criteria for SCR usage because of its effect on fan horsepower requirements, the single largest energy need for the SCR process. Typical reactor pressure drop is shown in Figures 3 and 4.

Figure 36

SO2 Oxidation vs. Flow Rate

1st & 2nd Quarter 1994

SO2 OXIDATION (%)

80% NOx Reduction, 750 °F, CORMETECH

s:/projects/doe/crist/doc/jdm/para2/so3vsfe.pre

Figure 37

SO₂ Oxidation vs. Temperature

1st & 2nd Quarter 1994

SO₂ OXIDATION (%)

80% NO_x Reduction, Low Flow, CORMETECH

s:/projects/doe/crist/doc/jdm/para2/so3vstee.pre

Mass concentrations (two-point) and flue gas velocity (three-point) profiles were conducted near design operating condition (700°F, 100% flow rate [400 scfm]) at the reactor outlet and reactor inlet, respectively. The mass concentration profile data are given in Figure 38 in four different units of concentration and as mass emission rate. The average mass concentration was 5.25 ± 0.84 gr/dscf (with a relative standard deviation of 16%); and the average emission rate was 9.68 lb/MBtu. The flue gas velocity profile is presented in Figure 39. The average velocity was 12.5 ± 1.3 ft/sec.¹

HCl concentration was measured at the design operating condition at the reactor outlet. The average HCl concentration (at 3% O₂, dry) was 196 ± 9 ppm(v). The N₂O concentrations were measured at the reactor inlet [1.8 ppm(v)] and at the reactor outlet [1.8 ppm(v)] (both measurements are dry at 3% O₂).¹

- Reactor F

Table 11 shows the parametric test data on intermediate ammonia, slip ammonia, and sulfur dioxide oxidation collected during this reporting period for the Haldor Topsoe catalyst. All the ammonia data are corrected to reactor inlet oxygen concentration. The long term NO_x reduction is given in this table as an average over the operating periods shown, i.e., for January - March and April - June of 1994. The long term NO_x reduction data indicate the average performance of the catalyst at or near the design operating conditions of 0.8 ammonia-to-NO_x ratio, 400 SCFM flow rate, and 700°F reactor temperature. As can be noted in the second quarter data averages, analytical errors and drift in calibrations led to values falsely indicating NO_x reduction being higher than the ammonia-to-NO_x ratio.

The intermediate ammonia measurements were made after the first catalyst bed and at conditions thought to give the best kinetic information. The NO_x removals reported with the intermediate ammonia measurements are computed from the measured ammonia concentration using standard material balance techniques.

The ammonia slip data given in Table 11 are presented below in three plots: ammonia slip versus flow rate, ammonia-to-NO_x ratio, and temperature. Figure 40 shows ammonia slip versus flow rate for roughly 80% NO_x reduction. As expected, the trend shows increasing ammonia slip with increasing reactor flow rate. The ammonia slip is, however, relatively minor indicating the ability of the catalyst to withstand significant increases in flow while maintaining ammonia slip limits.

Reactor E

Catalyst Layer 4 (Reactor outlet)

Mass Concentration Profile
gr/acf

Average = 2.03 ± 0.35 gr/acf

Mass Concentration Profile
gr/dscf

Average = 5.25 ± 0.84 gr/dscf

Mass Concentration Profile
g/acm

Average = 4.64 ± 0.80 g/acm

Mass Concentration Profile
g/dscm

Average = 12.02 ± 1.93 g/dscm

Mass Emission Rate Profile
lb/MBtu

Average = 9.68 ± 1.55 lb/MBtu

s:\ptd\maj\maj\reactor.pre

Figure 38. Five presentations of the Reactor E mass concentration profile.

Reactor E

Catalyst Layer 1 Inlet

Velocity Profile ft/s

Average = 12.5 ± 1.3 ft/s

s:\ptd\maj\maj\reactor.pre

Figure 39. Reactor E flue gas velocity profile.

TABLE 11. REACTOR F DATA

INTERMEDIATE AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	INT. NH ₃ (ppmv)	INT. NO _x REDUCTION (%)
245	622	2.880	310	0.544	13.0	50.2
594	626	2.266	293	0.539	38.1	40.9
599	625	5.348	373	0.729	71.9	53.6
599	625	4.514	372	0.930	96.9	66.9
400	706	2.682	312	0.747	39.0	62.2
594	755	6.724	368	0.525	37.1	42.4
600	754	2.704	318	0.741	51.8	57.7

SLIP AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	SLIP NH ₃ (ppmv)
397	624	5.835	380	0.708	0.8
600	626	3.991	339	0.722	3.6
597	625	2.212	281	0.929	19.5
400	708	3.455	359	0.699	0.5
601	705	1.837	286	0.714	2.6
594	706	2.730	337	0.730	3.1

SULFUR DIOXIDE OXIDATION PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	OUTLET O ₂ (%)	INLET SO ₂ (ppmv)	NH ₃ /NO _x RATIO	PRED. SO ₃ IN (ppmv)	SO ₃ OUT	SO ₃ formed (ppmv)	OXID. RATE (%)
240	623	3.828	1840	0.726	10.1	7.5	-2.6	-0.15
240	708	3.640	1910	0.717	9.7	17.9	8.2	0.46
400	706	3.531	1723	0.729	10.1	13.4	3.3	0.20
605	704	3.408	1722	0.726	10.1	12.6	2.5	0.15
246	755	2.928	1848	0.721	8.9	7.9	-1	-0.05
400	757	2.619	1848	0.606	8.9	10.0	1.1	0.06
597	756	2.941	1804	0.725	8.9	10.9	2	0.12

LONG TERM NO_x REDUCTION

TIME PERIOD	FLOW RATE (SCFM)	INLET NO _x (ppmv)	NH ₃ / NO _x RATIO	OUTLET NO _x (ppmv)	NO _x RED. (%)
JAN. - MAR.	398	346	0.76	81	71
APRIL - JUNE	402	354	0.76	56	78

Figure 40
Ammonia Slip vs. Flow Rate
1st & 2nd Quarter 1994

Since the overall reaction rate is partly controlled by mass transfer limitations, improvements in bulk mass transfer coefficients are likely mitigating the effect of increased flow on slip ammonia increases. This plot demonstrates the ability of an SCR system to follow host boiler load variations while maintaining design specifications.

Figure 41 shows ammonia slip versus ammonia-to- NO_x ratio at low temperature and high flow rate (worst case scenario). The plot shows sharp increases in ammonia slip as the ammonia-to- NO_x ratio approaches 1.0. This finding is in keeping with published data of this type. At ammonia-to- NO_x ratios near 1.0, non-idealities in the reactor system force the catalyst to slip ammonia since areas are present in the reactor where NO_x is the limiting reagent. Again, this plot is for the worst case scenario and ammonia slip values are likely to be lower for higher temperature or lower flow rate.

Ammonia slip versus temperature for high flow and roughly 80% NO_x reduction is plotted in Figure 42. Some improvement (decrease) in ammonia slip is noted between 620 and 700°F, likely due to improvements in the reaction rate with increasing temperature. The slight increase shown between the 700 and 750°F points is likely due to measurement variability and is not considered significant. In this case, the plot should not be construed as demonstrating increases in ammonia slip with increasing temperature above 700°F. It is expected that the 700 and 750°F values are roughly equivalent which may be due in part to mass transfer limitations that become controlling at these higher temperatures. In general, the plot demonstrates that in terms of ammonia slip, significant improvements are not realized at temperatures above 700°F. Losses in boiler efficiency would probably outweigh any improvements in ammonia slip from an SCR reactor operating at temperatures near 750°F.

The plot for SO_2 oxidation versus flow rate at high temperature is shown in Figure 43. The plot is relatively flat showing little change in SO_2 oxidation rate. Published rate equations and catalyst supplier information indicates that the general form of the plot should be linear with SO_2 oxidation decreasing with flow rate. In this case, the relatively flat profile indicates that either chemical or physical characteristics of the catalyst and reactor are masking this linear effect.

Figure 44 shows SO_2 oxidation versus temperature at low flow rate. Published information, describing the effects of temperature on SO_2 oxidation and information obtained from the catalyst suppliers, indicates that SO_2 oxidation follows an exponential relationship with temperature. The exact shape of the plot is somewhat unclear, since fairly large measurement variability exists in the

Figure 41
Ammonia Slip vs. NH₃/NO_x Ratio
1st & 2nd Quarter 1994

Figure 42
Ammonia Slip vs. Temperature
1st & 2nd Quarter 1994

80% NO_x Reduction, High Flow, HALDOR TOPSOE

s:/projects/doe/crist/doc/jdm/para2/slpvstf.pre

Figure 43

SO₂ Oxidation vs. Flow Rate

1st & 2nd Quarter 1994

SO₂ OXIDATION (%)

s:/projects/doe/crist/doc/jdm/para2/so3vsff.pre

Figure 44

SO₂ Oxidation vs. Temperature

1st & 2nd Quarter 1994

SO₂ OXIDATION (%)

80% NO_x Reduction, Low Flow, HALDOR TOPSOE

s:/projects/doe/crist/doc/jdm/para2/so3vstef.pre

test facility, and other phenomena, such as SO_3 deposition, occur in the test facility reactors. The extremely low oxidation rate of the catalyst further exacerbates this problem, in fact, only one data point shows any detectable amount of SO_2 oxidation with this value being very close to the detection limits. The plot is valuable in the context of showing that this particular catalyst has very low SO_2 oxidation potential.

The SO_2 oxidation data are corrected to reactor outlet oxygen concentrations. The value for sulfur trioxide produced in the reactor is based on the measured outlet sulfur trioxide concentration and estimated sulfur trioxide reactor inlet values. The reactor inlet sulfur trioxide values are estimated using past measurements of inlet sulfur trioxide based on host boiler load. In many cases, especially those involving low temperature conditions, there was no net increase in sulfur trioxide concentrations across the reactor. In fact, some operating conditions showed a loss in sulfur trioxide across the reactor in addition to the loss that normally occurs without catalyst present. (Tables showing SO_2 oxidation rates quote reactor flow rates as calculated for the reactor exit, since SO_3 is measured at this point. For consistency with other figures, the plots of SO_2 oxidation versus flowrate are based on reactor inlet flowrates. In practice, differences between inlet and outlet flowrates are slight and do not greatly affect the overall analysis.)

The pressure drop across the reactor is another important criteria for SCR usage because of its effect on fan horsepower requirements, the single largest energy need for the SCR process. Typical reactor pressure drop is shown in Figures 3 and 4.

Mass concentrations (two-point) and flue gas velocity (three-point) profiles were conducted near design operating condition (700°F, 100% flow rate [400 scfm]) at the reactor outlet and reactor inlet, respectively. The mass concentration profile data are given in Figure 45 in four different units of concentration and as mass emission rate. The average mass concentration was 7.87 ± 1.05 gr/dscf (with a relative standard deviation of 13.3%); and the average emission rate was 14.2 lb/MBtu. The flue gas velocity profile is presented in Figure 46. The average velocity was 9.5 ± 3.6 ft/sec.¹

HCl concentration was measured at the design operating condition at the reactor outlet. The average HCl concentration (at 3% O_2 , dry) was 140 ± 5 ppm(v). The N_2O concentrations were measured at the reactor inlet [2.4 ppm(v)] and at the reactor outlet [1.6 ppm(v)] (both measurements are dry at 3% O_2).¹

Reactor F

Catalyst Layer 4 (Reactor outlet)

Mass Concentration Profile
gr/acf

Average = 3.08 ± 0.41 gr/acf

Mass Concentration Profile
gr/dscf

Average = 7.87 ± 1.05 gr/dscf

Mass Concentration Profile
g/acm

Average = 7.1 ± 0.9 g/acm

Mass Concentration Profile
g/dscm

Average = 18.0 ± 2.4 g/dscm

Mass Emission Rate Profile
lb/MBtu

Average = 14.2 ± 1.9 lb/MBtu

s:\ptd\maj\maj\reactor.pre

Figure 45. Five presentations of the Reactor F mass concentration profile.

Reactor F

Catalyst Layer 1 Inlet

Velocity Profile ft/s

Average = 9.5 ± 3.6 ft/s

s:\pfd\maj\maj\reactor.pre

Figure 46. Reactor F flue gas velocity profile.

- Reactor G

Table 12 shows the parametric test data on intermediate ammonia, slip ammonia, and sulfur dioxide oxidation collected during this reporting period for the Hitachi Zosen catalyst. All the ammonia data are corrected to reactor inlet oxygen concentration. The long term NO_x reduction is given in this table as an average over the operating periods shown, i.e., for January - March and April - June of 1994. The long term NO_x reduction data indicate the average performance of the catalyst at or near the design operating conditions of 0.8 ammonia-to-NO_x ratio, 400 SCFM flow rate, and 700°F reactor temperature.

The intermediate ammonia measurements were made after the first catalyst bed and at conditions thought to give the best kinetic information. The NO_x removals reported with the intermediate ammonia measurements are computed from the measured ammonia concentration using standard material balance techniques.

The ammonia slip data given in Table 12 are presented below in three plots: ammonia slip versus flow rate, ammonia-to-NO_x ratio, and temperature. Figure 47 shows ammonia slip versus flow rate for roughly 80% NO_x reduction. As expected, the trend shows increasing ammonia slip with increasing reactor flow rate. The ammonia slip is, however, relatively minor indicating the ability of the catalyst to withstand significant increases in flow while maintaining ammonia slip limits. Since the overall reaction rate is partly controlled by mass transfer limitations, improvements in bulk mass transfer coefficients are likely mitigating the effect of increased flow on slip ammonia increases. This plot demonstrates the ability of an SCR system to follow host boiler load variations while maintaining design specifications.

Figure 48 shows ammonia slip versus ammonia-to-NO_x ratio at low temperature and high flow rate (worst case scenario). The plot shows sharp increases in ammonia slip as the ammonia-to-NO_x ratio approaches 1.0. This finding is in keeping with published data of this type. At ammonia-to-NO_x ratios near 1.0, non-idealities in the reactor system force the catalyst to slip ammonia since areas are present in the reactor where NO_x is the limiting reagent. Again, this plot is for the worst case scenario and ammonia slip values are likely to be lower for higher temperature or lower flow rate.

TABLE 12. REACTOR G DATA

INTERMEDIATE AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	INT. NH ₃ (ppmv)	INT. NO _x REDUCTION (%)
242	626	3.638	361	0.514	5.5	49.9
600	625	2.728	374	0.511	19.6	45.9
600	626	2.724	362	0.708	38.1	60.2
402	676	1.928	324	1.083	72.2	86.1
401	705	2.023	344	0.695	15.2	65.1
601	757	2.322	355	0.506	19.9	45.0

SLIP AMMONIA PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	INLET O ₂ (%)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	SLIP NH ₃ (ppmv)
400	625	1.404	272	0.711	0.8
591	626	1.385	265	0.700	1.1
600	264	2.079	369	0.895	16.2
400	704	2.438	331	0.684	0.8
589	707	1.423	304	0.719	2.2
593	755	2.488	354	0.704	2.2

SULFUR DIOXIDE OXIDATION PARAMETRIC TEST DATA

FLOW RATE (SCFM)	TEMP. (°F)	OUTLET O ₂ (%)	INLET SO ₂ (ppmv)	NH ₃ /NO _x RATIO	PRED. SO ₃ IN (ppmv)	SO ₃ OUT	SO ₃ formed (ppmv)	OXID. RATE (%)
241	624	3.439	1899	0.721	9.3	6.9	-2.4	-0.13
240	707	3.314	1890	0.721	9.3	21.0	11.7	0.66
400	704	3.196	1851	0.711	10.1	20.8	10.7	0.59
590	707	2.706	1885	0.715	9.3	19.1	9.8	0.53
241	753	3.452	1693	0.725	9.7	31.3	21.6	1.33
400	758	2.954	1804	0.714	9.7	22.6	12.9	0.74
600	750	2.889	1773	0.720	9.7	21.2	11.5	0.67

LONG TERM NO_x REDUCTION

TIME PERIOD	FLOW RATE (SCFM)	INLET NO _x (ppmv)	NH ₃ /NO _x RATIO	OUTLET NO _x (ppmv)	NO _x RED. (%)
APRIL - JUNE	399	345	0.80	51	80

Figure 47

Ammonia Slip vs. Flow Rate

1st & 2nd Quarter 1994

AMMONIA SLIP (ppm)

80% NOx Reduction, 700 °F, HITACHI ZOSEN

s:/projects/doe/crist/doc/jdm/para2/slpvsflg.pre

Figure 48

Ammonia Slip vs. NH3/NOx Ratio

1st & 2nd Quarter 1994

Ammonia slip versus temperature for high flow and roughly 80% NO_x reduction is plotted in Figure 49. Some improvement (decrease) in ammonia slip is noted between 620 and 700°F, likely due to improvements in the reaction rate with increasing temperature. Only slight improvements are noted with increasing temperature above 700°F. This may be due to mass transfer limitations that become controlling at these higher temperatures. In general, the plot demonstrates that in terms of ammonia slip, significant improvements are not realized at temperatures above 700°F. Losses in boiler efficiency would probably outweigh any improvements in ammonia slip from an SCR reactor operating at temperatures near 750°F.

The plot for SO₂ oxidation versus flow rate at high temperature is shown in Figure 50. In general, decreasing SO₂ oxidation is noted with increasing flow rate. This finding is expected, since the reaction rate is inversely proportional to flow rate, according to published rate data and information provided by catalyst suppliers.

Figure 51 shows SO₂ oxidation versus temperature at low flow rate. The plot shows a fairly linear increase in SO₂ oxidation with increasing temperature. Published information describing the effects of temperature on SO₂ oxidation and information obtained from the catalyst suppliers, indicates that SO₂ oxidation follows an exponential relationship with temperature. The exact shape of the plot is somewhat unclear, since fairly large measurement variability exists in the test facility, and other phenomena, such as SO₃ deposition occur in the test facility reactors. The general trend of increasing SO₂ oxidation with temperature is expected, however, and the plot provides a good basis for determining the potential for SO₃ formation across the SCR reactor at various operating temperatures.

The SO₂ oxidation data are corrected to reactor outlet oxygen concentrations. The value for sulfur trioxide produced in the reactor is based on the measured outlet sulfur trioxide concentration and estimated sulfur trioxide reactor inlet values. The reactor inlet sulfur trioxide values are estimated using past measurements of inlet sulfur trioxide based on host boiler load. In many cases, especially those involving low temperature conditions, there was no net increase in sulfur trioxide concentrations across the reactor. In fact, some operating conditions showed a loss in sulfur trioxide across the reactor in addition to the loss that normally occurs without catalyst present. (Tables showing SO₂ oxidation rates quote reactor flow rates as calculated for the reactor exit, since SO₃ is measured at this point. For consistency with other figures, the plots of SO₂ oxidation versus flowrate are based on reactor inlet flowrates. In practice, differences between inlet and outlet flowrates are slight and do not greatly affect the overall analysis.)

Figure 49

Ammonia Slip vs. Temperature

1st & 2nd Quarter 1994

Figure 50

SO₂ Oxidation vs. Flow Rate

1st & 2nd Quarter 1994

SO₂ OXIDATION (%)

80% NO_x Reduction, 750 °F, HITACHI ZOSEN

s:/projects/doe/crist/doc/jdm/para2/so3vsfg.pre

Figure 51

SO₂ Oxidation vs. Temperature

1st & 2nd Quarter 1994

SO₂ OXIDATION (%)

80% NO_x Reduction, Low Flow, HITACHI ZOSEN

s:/projects/doe/crist/doc/jdm/para2/so3vsteg.pre

The pressure drop across the reactor is another important criteria for SCR usage because of its effect on fan horsepower requirements, the single largest energy need for the SCR process. Typical reactor pressure drop is shown in Figures 3 and 4.

A flue gas velocity (three-point) profile was conducted at the reactor inlet. The flue gas velocity profile is presented in Figure 52. The average velocity was 6.9 ± 1.3 ft/sec.¹ The velocity was lower near the back of the reactor and is reflected in the rather large standard deviation. This velocity was lower than other reactors because the flow rate was lower (200 scfm) than the normal flow rate for the other small reactors (usually 400 scfm) when velocity measurements are taken.

HCl concentration was measured at the design operating condition at the reactor outlet. The average HCl concentration (at 3% O₂, dry) was 157 ± 2 ppm(v). The N₂O concentrations were measured at the reactor inlet [2.4 ppm(v)] and at the reactor outlet [1.6 ppm(v)] (both measurements are dry at 3% O₂).¹

Task 1.3.6 - Project Management and Reporting

The Management Information System, developed for tracking overall budget and schedule information, was used to monitor budget and schedule and to help fulfill DOE reporting requirements. Monthly progress reports were submitted to DOE.

A project review meeting was held for all involved parties on May 25-26 at Plant Crist in which general progress since December 1993 was discussed, and specific vendor data were discussed in separate meetings with each vendor.

Project information was presented at DOE's Second Annual Historically Black Colleges and Universities/Private Sector/Energy Research and Development Technology Transfer Symposium in April 1994 and at the EPRI NO_x Controls for Utility Boilers Workshop in May 1994.

The Cooperative Agreement was amended to reflect approval of the Continuation Application which requested DOE's matching funds for Budget Period 2 to continue the project to completion. Budget Period 2 provides DOE funding for Phase III, which includes operations, testing, disposition, and final report.

Reactor G

Catalyst Layer 1 Inlet

s:\ptd\maj\maj\reactor.pre

Figure 52. Reactor G flue gas velocity profile.

Section 5

PLANNED ACTIVITIES

During the last two quarters of 1994, the following activities are planned:

- Complete the next set of parametric tests (third series) on all reactors/catalysts and begin the fourth sequence of parametric tests. Meanwhile, continue operations at long term conditions for the reactors/catalysts when not undergoing parametric testing.
- Take catalyst samples and submit for laboratory analysis.
- Resolve final equipment disposition and payment resulting from the failure of the sampling system for NO_x at intermediate levels in the reactors where ammonia concentration is still high, relative to outlet reactor slip ammonia levels.
- Monitor reactor D catalyst to evaluate the erosion on this catalyst being greater in comparison to the other catalysts.
- Maintain and repair equipment as needed to continue satisfactory operations. Monitor and clean air heaters as needed.
- Reconfigure control system for more accurate ammonia injection control.
- Present paper on project status at DOE's Third Annual Clean Coal Technology Conference.
- Hold status meeting in December 1994.

REFERENCES

1. Testing and Analytical Services for the Innovative Clean Coal Technology Demonstration of Selective Catalytic Reduction (SCR) Technology for the Control of Nitrogen Oxide (NO_x) Emissions from High Sulfur Coal, Task 4: Second Long-Term Parametric Test, Final Report Parts 1 - 7 (Reactors A - G), Final SRI Report to SCS, November 3, 1995.

APPENDIX A

Memorandum on SCR/DOE Gas Analysis System Status

InterOffice Memo

To: Scott Hinton, Doug Maxwell
From: Jimmy L. Horton
Date: February 10, 1994
Subject: SCR/DOE Gas Analysis System Status

On February 1 & 2, 1994, we completed testing of the revised EPM dilution probe for the AB-803 system. The results are attached. The probe will not operate acceptably within the specifications that the AB-803 system was purchased for. The best results we have seen with .73 NH₃/NO_x is a 10% reduction of the NO_x reading. We have seen as much as 60-80% reduction of the readings.

We began testing this system with the original inconel probe when it was discovered that the readings would drop significantly upon the initiation of the NH₃ injection into the process. The first suspicion was the nickel content in the inconel probes. EPM Environmental supplied us with a 316L SS probe for testing. It first appeared that with this probe we could remove the SS enclosure around the glass orifice and eliminate any measurement errors. This prompted LSMCC and EPM to simultaneously begin development of alternate probe tips. LSMCC developed a ceramic tip which was tested in early December, 1993. The tests revealed a 30% reduction with or without the tip installed. EPM completed their development of a glasslined tip in January, 1994. Tests in early February revealed a reduction of 10-20%.

The specifications that were written by SCS and agreed upon by Lear Siegler Measurement Controls Corp. for the AB-803 system require that it operate with NH₃ concentrations as high as 300ppm. It will not do this.

I have indicated this "failure to perform" to Jim Pankiewicz of EPM Environmental, Chris Lavery and Steve Devita of Monitor Labs (formerly Lear Siegler Measurement Controls Corp.) by phone. I also expressed to them that we have expended more manpower than desired to accomodate the extensive testing required to this point.

[REDACTED]

At present, the other gas analysis systems are operating satisfactorily.

CC: James Gibson, King Knight, Jim Witt

SCR/DOE Project

SCR SS Probe Test Runs With Newly
Developed Glasslined Tip by EPM

2/1/94

NH3 Inject. @ .73 NH3/NOx

Time	SP-801	SP-811	SP-829	% Red.	
15:10	355	267	354	25	Start NH3
15:13	353	267	354	24	
15:16	352	260	153	26	
15:19	354	253	131	29	
15:22	354	243	111	31	
15:25	354	242	102	32	
15:28	356	240	102	33	
15:31	358	240	101	33	
15:34	355	240	99	32	
15:37	351	240	95	32	
15:40	354	240	93	32	Stop NH3
15:43	355	240	94	32	
15:45	354	241	168	32	
15:48	355	243	356	32	
15:51	358	253	378	29	
15:54	363	261	343	28	
16:04	347	257	332	26	
16:07	360	265		26	
16:10	359	262		27

SCR/DOE Project

SCR SS Probe Test Runs With Newly
Developed Glasslined Tip by EPM

2/2/94

NH3 Inject. @ .73 NH3/NOx

Time	SP-801	SP-811	SP-829	% Red.	
08:30	308	259		16	Start NH3
08:33	299	253		15	
08:36	311	261		16	
08:39	266	237		11	
08:42	289	223		23	
08:45	307	231		25	
08:58	306	289		6	
08:51	333	234		30	
08:54	325	233		28	
08:57	315	226		28	
09:00	315	224		29	
09:03	314	223		29	
09:06	316	223		29	
09:09	312	222		29	
09:12	311	218		30	
09:15	312	217		30	Stop NH3
09:34	312	216		31	
09:37	315	214		32	
09:40	320	214		33	
09:43	321	214		33	
09:46	321	218		32	
09:49	322	233		28	
09:52	319	237		26	
09:55	316	241		24	
09:57	314	245		22	
10:01	318	252		21	
10:05	318	256		19	
10:08	317	258		19	
10:11	317	260		18	
10:14	315	260		17	
10:17	315	261		17	
10:20	317	263		17	

SCR/DOE Project @ Crist Steam Plant

APPENDIX B

Testing and Analytical Services for the Innovative Clean Coal Technology Demonstration of Selective Catalytic Reduction (SCR) Technology for the Control of Nitrogen Oxide (NO_x) Emissions from High Sulfur Coal, Final Report, Task 4: Air Heater Tests, First Parametric Series, SRI report to SCS, September 1, 1995.

**TESTING AND ANALYTICAL SERVICES FOR THE INNOVATIVE CLEAN COAL
TECHNOLOGY DEMONSTRATION OF SELECTIVE CATALYTIC REDUCTION
(SCR) TECHNOLOGY FOR THE CONTROL OF NITROGEN OXIDE (NO_x)
EMISSIONS FROM HIGH SULFUR COAL**

Final Report for Task 4

**Air Heater Tests
Reactors A, B, C
First Parametric Series**

Prepared by

**Kenneth M. Cushing
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**DOE Prime Contract Number DE-FC22-89-PC89652
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SRI Report Number SRI-ENV-94-744-7613**

September 1, 1995

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Section 1

INTRODUCTION

This is the fifth in a series of reports describing the results of testing and analytical services for the Innovative Clean Coal Technology Demonstration of Selective Catalytic Reduction (SCR) Technology for the control of nitrogen oxide (NO_x) emissions from high-sulfur, coal-fired boilers. The test results for the initial set of air heater tests are summarized in this report. The air heater tests were conducted during the last week of April and the first two weeks of May, 1994, following the completion of the first sequence of parametric testing under Task 4: Long-Term Parametric Tests.

The SCR test facility is located at Gulf Power Company's Plant Crist Unit 5 in Pensacola, Florida. The test facility includes three large SCR reactors, each designed to treat 5000 wscfm of flue gas, and six small reactors, each processing 400 wscfm of flue gas. An air heater capable of removing sufficient heat to reduce the flue gas temperature from a range of 600 to 750 °F down to 300 °F was included in the design of each of the three large reactors. The three large reactors are designated as reactors A, B, and C. Reactors A and B incorporate Ljungstrom-type air heaters and Reactor C's air heater incorporates a heat pipe design.

The original design of the SCR test facility also included bypass heat exchangers on each of the three large reactors. These units were intended for use during parametric testing of the reactors (Task 4: Long Term Parametric Tests) so that flue gas containing higher concentrations of slip ammonia could be diverted around the air heaters. However, the bypass heat exchangers did not function as planned and the practice of bypassing the air heaters on Reactors A, B and C during parametric testing was abandoned. Thus, the air heater test data presented in this report include the effects of day-long periods of exposure to ammonia concentrations normally ranging from a few parts per million by volume ppm(v) to 20 ppm(v), although brief excursions approached 100 ppm(v).

The air heater testing included the determinations of particulate mass concentration and measurements of the concentrations of sulfur dioxide, sulfur trioxide, hydrogen chloride and ammonia. All tests were conducted with manual sampling methods. Ammonia samples were segregated into solid and gas-phase fractions to characterize the gas/solid phase partitioning of ammonia across the air heaters. In general, simultaneous tests were performed at the inlet and outlet of each of the three air heaters (except for mass concentration measurements at the outlet of the Reactor C air heater). The fly ash (particulate) catches from the mass concentration measurements were further analyzed in the laboratory to determine particle size distributions, ash mineralogy, and ash resistivity.

This report is divided into several sections. Section 2 describes the test methods used for the air heater testing. Section 3 reviews all of the test results. The test data are briefly summarized in Section 4.

Section 2

TEST METHODS

A variety of test methods were used to characterize air heater performance. The following subsections briefly describe the test methods for ammonia (NH_3), sulfur dioxide (SO_2), sulfur trioxide (SO_3), hydrogen chloride (HCl), and particulate mass concentration, particle size, ash mineralogy, and ash resistivity. Figure 2-1 shows a sketch of a large reactor and its air heater (not to scale). The test port locations for air heater inlet and outlet measurements are indicated on the diagram. Test ports (three horizontal ports) upstream of catalyst layer 4 (the normal reactor outlet test location) were used to measure concentrations of air heater inlet particulate and gas constituents. Three test ports (horizontal) were installed in a transition piece in the outlet ducting of the Reactor A and B air heaters that were suitable for particulate and gas phase flue gas constituent testing. Test ports at this same location in the Reactor C air heater outlet were not installed. However, a single test port (horizontal) in a section of ducting at the inlet transition to the cyclone on Reactor C (downstream of the reactor bypass duct) was suitable for testing gas phase flue gas constituents. Turbulent flow at this location precluded mass concentration tests.

Air heater tests were performed with the reactors operating at Test Condition 22, the normal baseline or long-term operating condition. These operating conditions included a flue gas temperature of 700 °F, an NH_3/NO_x ratio of 0.8, and a flue gas flow rate of 5000 wscfm. During measurements of air heater inlet and outlet ammonia concentrations only, additional tests were performed at Test Condition 24 (except Reactor A). The operating parameters for these tests were a flue gas temperature of 700 °F, an NH_3/NO_x ratio of 1.0, and a flue gas flow rate of 5000 wscfm.

AMMONIA

Ammonia concentrations were measured simultaneously at the air heater inlet (upstream of catalyst layer 4) and at the air heater outlet on each of the large reactors. At each site and at each test condition, three independent tests were performed to determine an average ammonia concentration. Sampling ports located upstream of catalyst layer 4 were used to measure air heater inlet ammonia concentration. Flue gas was sampled at three equidistant points in each of the three test ports (a total of nine traverse points) to determine an average inlet ammonia concentration for each test. At the air heater outlet sampling location on Reactors A and B the probe also sampled at three equidistant points in each of the three test ports (a total of nine traverse points). At the single Reactor C outlet test port, the ammonia probe traversed three, equally-spaced, points (front to rear) during each test.

Figure 2-1. Schematic drawing of SCR reactor and air heater cross section (side view).

Figure 2-2. Schematic diagram of the ammonia sampling train.

A schematic drawing of the ammonia sampling train is shown in Figure 2-2. The flue gas sample is drawn through a glass-lined probe and a heated filter to remove particulate. The filter is maintained at the nominal flue gas temperature. The gas sample passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain a 0.1 normal solution of sulfuric acid. The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. A second glass wool plug is placed in the line between the last two impingers to collect any ammonia aerosols that may escape the second impinger. The remainder of the train consists of a silica gel column to remove the last traces of water from the flue gas sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The samples were segregated during sample collection into a solid-phase sample and a gas-phase sample. The solid-phase sample consisted of the heated filter, all of the collected particulate, and the probe-wash liquid. The gas-phase sample consisted of the impinger liquids and the wash and rinse liquids. The two samples were analyzed separately to characterize the ammonia partitioning between the gas and solid phases.

At the outset of the Task 4 parametric testing, modified Greenberg-Smith type impingers were used for ammonia sampling. However, 100 mL gas sampling impingers, each containing about 50 mL of solution, were used for the air heater ammonia testing to reduce the detection limit to less than 1 ppm(v) for a reasonable sample volume, about 3 ft³ of gas. The lower detection limit was needed for much of the air heater ammonia testing because ammonia concentrations are very low at the reactor exit when operating at the reactor design condition with an NH₃/NO_x ratio of 0.8.

The impinger solutions are made alkaline in the laboratory (converting the NH₄⁺ ion to free NH₃ in solution). The concentration of ammonia is then determined with an ammonia ion specific electrode, Orion Model 920A. As mentioned above, three individual measurements of ammonia concentration are made for each test condition. Two independent determinations of the ammonia concentration are then conducted on each sample.

SULFUR DIOXIDE AND SULFUR TRIOXIDE

During the air heater evaluation, SO₂ and SO₃ concentrations were measured simultaneously at the reactor outlet upstream of catalyst layer 4 (air heater inlet) and the air heater outlet. Sulfur dioxide (SO₂) and sulfur trioxide (SO₃) were collected in a controlled condensation sampling train. All

tests were conducted using single-point sampling in either the right-hand port (air heater inlet test location) or middle port (air heater outlet test location).

Figure 2-3. Schematic diagram of the controlled condensation sampling train for SO_2 and SO_3 .

A schematic drawing of the SO_2/SO_3 sampling train is shown in Figure 2-3. The flue gas sample is drawn through a heated, quartz-lined probe maintained above 550 °F. The sample then passes through a quartz filter housed in a heated quartz filter holder, also maintained at 550 °F. The next element in the train is the SO_3 condenser. The condenser is a length of quartz tubing packed with quartz wool and maintained between 120 °F and 130 °F in a heated water bath. The sample next passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain a 3% solution of hydrogen peroxide to oxidize SO_2 . The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. The remainder of the train consists of a silica gel column to remove the last traces of water from the sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The SO_3 is collected in the condensing element and the SO_2 is collected in the bubblers by oxidation with the hydrogen peroxide, converting it to H_2SO_4 in solution. In the condenser the SO_3 present begins a hydration reaction with the water vapor present making H_2SO_4 . The excess water vapor also condenses to produce a condensate of concentrated aqueous H_2SO_4 . Thus, two solutions of H_2SO_4 are collected; one a very concentrated solution of limited amount containing the original SO_3 and the other a relatively weak solution in far greater amount containing the original SO_2 . The concentrations of the sulfate ion are determined by ion chromatography using a DIONEX Model DX-100 Ion Chromatograph.

HYDROGEN CHLORIDE

Figure 2-4. Schematic diagram of the chloride sampling train.

A schematic sketch of the chloride sampling train is shown in Figure 2-4. The flue gas sample is drawn through a heated, quartz-lined probe maintained above 550 °F. The sample then passes through a quartz filter housed in a heated quartz filter holder, also maintained at 550 °F. The sample next passes through a length of PTFE tubing to a set of three impingers in an ice bath. The first two impingers contain type 1 de-ionized water for removal of HCl vapor. The third impinger serves as a trap to prevent the solutions from accidentally being drawn into the pump. The remainder of the train consists of a silica gel column to remove the last traces of water from the sample, a leak-free pump, and a dry gas meter to measure the volume of the sample.

The chloride ion concentration is determined by ion chromatography using a DIONEX Model DX-100 Ion Chromatograph. Hydrogen chloride concentrations were determined by single-point sampling (at the mid-point of the duct) at the air heater inlet (upstream of the 4th catalyst layer, right-hand port) and at the air heater outlet (middle port). Inlet and outlet tests were conducted simultaneously.

MASS CONCENTRATION

Mass concentration was measured using a sampling train similar to that defined in the EPA Reference Test Method 17. A schematic drawing of the train used for mass concentration sampling is shown below in Figure 2-5.

Figure 2-5. Schematic drawing of the mass concentration sampling train.

This train consists of a filter holder and nozzle that is designed to operate in the flue gas stream, followed by a heated sampling probe, a condenser, a drying column, a gas meter, a pump, and finally a flow control mechanism, usually a calibrated orifice. An S-type pitot and thermocouple located near the nozzle provide a means for sampling isokinetically during each test. The glass fiber thimbles are desiccated before and after sampling and then weighed on an electronic microbalance.

Particulate mass concentrations were measured simultaneously upstream of catalyst layer 4 (air heater inlet) and at each air heater outlet (except Reactor C). At the air heater inlet, the three sample ports immediately above the access door at the fourth catalyst layer were used to obtain the sample. During each test particulate was sampled at three, equally-spaced, positions within the three test ports (a total of nine traverse points). At the air heater outlets on reactors A and B, three ports are available for mass sampling. During each test particulate was sampled at

three, equally-spaced, positions within the three test ports (a total of nine traverse points). Three runs were made at each test site to give a meaningful statistical average.

ASH MINERALOGY, PARTICLE SIZE, ASH RESISTIVITY

After the particulate catches were weighed, they were transported to SRI's Birmingham laboratories where particle size distributions, ash mineralogy and ash resistivity were determined. The instrument used to determine particle size distribution of the fly ash was a Shimadzu Model SA-CP4 Centrifugal Particle Size Analyzer. It is able to size the particles into approximately 25 size intervals between 0.056 and 56.2 micrometers physical diameter (or Stokes diameter, based on assumed spherical shape and true, or actual, particle density). Ash resistivity was determined using the IEEE 548 (1984) ash resistivity test method, commonly referred to as a descending temperature method. The humidity of the atmosphere surrounding the resistivity test cell was controlled to a moisture content of 8.7%, approximately the same as that measured in the flue gases at the Plant Crist SCR Test Facility.

SAMPLING QUALITY ASSURANCE

The measures adopted to ensure that meaningful results were obtained during the various testing procedures can be divided into three categories; equipment maintenance and calibrations, operating techniques, and analytical techniques. New equipment was obtained for use in the SCR testing program and preventive maintenance and calibrations are performed at regular intervals. Due to the dynamic nature of the process, each measurement requires three replicate samples. The replicates are averaged to yield a representative value. Variability gives a means to discern any anomalies not revealed by other quality control checks.

The sample trains are leak-checked by drawing a vacuum of 15 in. Hg before and after each sampling run. No in-leakage is tolerated. If any loss of pressure is observed within one minute, the source of the leakage is found and eliminated prior to testing. During each run the oxygen level of the gas exiting the dry gas meter is measured to detect air in-leakage that could dilute the samples.

Ammonia samples are analyzed with an ion-specific electrode. The instrument is calibrated using 0.1 µg/mL, 0.5 µg/mL, 1.0 µg/mL, 5.0 µg/mL and 10 µg/mL standards. At the conclusion of each analytical session, a sample spiked with a known amount of ammonia is analyzed and the result compared with the predicted concentration. A blank sample is analyzed to detect zero drift and a mid-range standard is also analyzed to detect calibration drift.

Both the sulfate (SO₂ & SO₃) samples and the chloride (HCl) samples are analyzed by means of ion chromatography. This instrument is calibrated using a minimum of three points on the calibration curve. As with the ammonia analysis, spiked samples, blanks, and standards are analyzed to ensure that calibration drift has not occurred.

While gas-phase flue gas constituents are being sampled, both field blanks and sample blanks are run as quality control checks. The field blank consists of a container of type 1 de-ionized water that is exposed to ambient air at the sampling site. Sample blanks are obtained by passing ambient

air through the probe and through the impingers filled with the appropriate solution. These blanks would detect sample contamination should it occur.

Section 3

AIR HEATER TEST RESULTS

AMMONIA CONCENTRATIONS

Concentrations of ammonia were measured simultaneously at each of the three air heater inlet and outlet test locations. The tests were conducted so as to differentiate between gas-phase and solid-phase ammonia. All three air heaters were tested with the reactors operating at Test Condition 22, the baseline operating condition. The Reactor B and C air heaters were also tested at Test Condition 24 ($\text{NH}_3/\text{NO}_x = 1.0$), however, the Reactor A air heater was not tested at this condition because of concern that the air heater might become plugged during operation with the higher inlet ammonia concentration.

The test data are summarized in Table 3-1. Table 3-1 also gives estimated solid-phase ammonia concentrations on a mass of ammonia per mass of ash basis. The values shown in the table were derived both from solid-phase ammonia measurements made at the same time the gas-phase measurements were made and mass concentration measurements made earlier at the same location. The ammonia partitioning between the gas and solid phases shown in the table are similar for the air heaters on reactors B and C. These data show that ammonia partitioning at the inlet to the air heater is roughly equal (on a flue gas volumetric basis or ammonia mass basis) between the gas and solid phases on reactors B and C. On both reactors the partitioning shifts toward the solid phase at the air heater outlet. As expected, a significant increase in total ammonia concentration was measured during operation at the higher NH_3/NO_x ratio at Test Condition 24 for both the air heater inlet and air heater outlet on Reactors B and C. The test run on the Reactor A air heater at Test Condition 22 indicated a higher total ammonia concentration at the air heater inlet than at the air heater outlet with the ammonia strongly partitioned to the solid phase. The total ammonia measured at the Reactor A air heater outlet was less than one-half of the total ammonia measured at the air heater inlet with the ammonia more evenly partitioned between the gas and solid phases. The lack of ammonia mass balance closure suggests an error in the measurement so that this result should be disregarded until further testing is done.

SO_2 and SO_3 CONCENTRATIONS

Concentrations of sulfur dioxide and sulfur trioxide were measured simultaneously at the air heater inlet and outlet test locations of Reactors A, B, and C during operation at Test Condition 22. The test data are summarized in Table 3-2.

SO_3 inlet concentrations ranged from 8 to 19 ppm(v) dry @ 3% O_2 , while SO_3 concentrations at the reactor outlets ranged from 5 to 11 ppm(v) dry @ 3% O_2 . Reductions in SO_3 concentration across the air heaters ranged from 37.5% on Reactor B to 42.1% on Reactor A. Within the tolerance of the standard deviations given for average SO_2 concentrations in Table 3-2, there was no measurable change in SO_2 concentration across any of the three air heaters. Air heater inlet SO_2 concentrations averaged 2044 ± 41 ppm(v) dry @ 3% O_2 , while the air heater outlet SO_2 concentrations averaged 2095 ± 34 ppm(v) dry @ 3% O_2 . Since the standard deviations of the mean concentrations at the inlet and outlet overlap, it can be concluded that no measurable

Table 3-1. SCR Reactor Air Heater Tests: Ammonia Concentrations

Reactor	Date	$\frac{\text{NH}_3}{\text{NO}_x}$	Air Preheater Location	Gas-Phase NH_3 , ppm(v), @ 3% O_2 , dry	Solid-Phase NH_3 , Equivalent ppm(v), @ 3% O_2 , dry	Estimated Solid-Phase NH_3 , $\mu\text{g/g}$	Estimated Upper Limit Solid-Phase NH_3 , $\mu\text{g/g}$	Estimated Lower Limit Solid-Phase NH_3 , $\mu\text{g/g}$
A	11-May-94	0.8	Inlet	0.6 ± 0.1	3.3 ± 0.2	307	372	256
			Outlet	0.6 ± 0.1	1.0 ± 0.2	108	139	84
A	Not Tested	1.0	Inlet					
			Outlet					
B	10-May-94	0.8	Inlet	< 0.4	0.2 ± 0.01	22	24	20
			Outlet	< 0.3	0.5 ± 0.01	65	70	61
B	10-May-94	1.0	Inlet	1.8 ± 0.4	1.9 ± 0.5	186	239	137
			Outlet	< 0.4	1.3 ± 0.2	162	189	137
C	6-May-94	0.8	Inlet	0.7 ± 0.1	0.6 ± 0.1	56	68	45
			Outlet	< 0.3	1.0 ± 0.1	116	137	99
C	9-May-94	1.0	Inlet	3.3 ± 0.3	3.8 ± 0.5	332	390	280
			Outlet	< 0.3	6.7 ± 1.6	746	1003	525

Table 3-2. SCR Reactor Air Heater Tests: SO₂ and SO₃ Concentrations

Reactor	SO ₃		SO ₂	
	Inlet	Outlet	Inlet	Outlet
	ppm(v), dry @ 3% O ₂	ppm(v), dry @ 3% O ₂	ppm(v), dry @ 3% O ₂	ppm(v), dry @ 3% O ₂
A	19 ± 1.4	11 ± 1.2	2040 ± 23	2073 ± 19
B	8.0 ± 1.5	5.0 ± 0.4	2087 ± 17	2078 ± 15
C	17 ± 0.5	10 ± 3.2	2005 ± 50	2134 ± 144

difference in SO₂ concentration was observed.

HCl CONCENTRATION

Measurements of the concentration of hydrogen chloride were conducted simultaneously at the inlet and outlet of each of the three large reactor air heaters. Three independent tests were performed at each test location. Single point sampling was used. The test results are presented in Table 3-3 as the average concentration and the standard deviation. All HCl concentrations fell within the range of 221 to 233 ppm(v) dry @ 3% O₂. There was no measurable change in HCl concentration across any of the three air heaters.

PARTICULATE MASS CONCENTRATION

Particulate mass concentrations were measured simultaneously at the inlet and outlet of the air heaters on Reactors A and B. Particulate mass concentration was measured only at the inlet to the heat pipe air heater on Reactor C. No outlet measurement was possible on the Reactor C air heater because sampling ports in a location suitable for mass concentration measurements could not be installed on that reactor.

Table 3-4 presents the mass concentration data for the air heater tests. The average air heater inlet mass concentration ranged from 3.42 ± 0.12 (Reactor B) to 3.84 ± 0.17 gr/dscf (Reactor C). The Reactor A inlet and outlet average mass concentration values were within one standard deviation of each other, while the Reactor B average outlet mass concentration was 19% lower than the Reactor B average inlet mass concentration.

PARTICLE SIZE DISTRIBUTIONS

The particulate collected during the air heater inlet and outlet particulate mass concentration tests was analyzed using a Shimadzu Model SA-CP4 Centrifugal Particle Size Analyzer to determine particle size distribution. Table 3-5 presents the mass median diameter (MMD) as the Stokes' diameter (micrometers) for each of the ash samples. The Stokes' or physical diameter is based on the assumption of spherical particles and the true, or actual, particle density. The mass median diameters ranged from 9.9 to 10.4 micrometers at the air heater inlets, and were 10.5 and 11.2 micrometers at the outlet of the Reactor A and B air heaters, respectively. No significant change in MMD across the air heaters was evident from these data. Figures A-1 through A-5 in Appendix show the particle size distributions graphically on both a cumulative per cent mass basis and a differential mass basis. The figures visually demonstrate the similarity between the particle size distributions among the various ash samples.

ASH MINERALOGY

Fly ash samples collected during mass concentration tests at the air heater inlet and outlet test locations were submitted for ash mineralogy tests. Fly ash samples were available for air heater

Table 3-3. SCR Reactor Air Heater Tests: HCl Concentration

Reactor	HCl	
	Inlet	Outlet
	ppm(v), dry @ 3% O ₂	ppm(v), dry @ 3% O ₂
A	233 ± 9.8	227 ± 11
B	222 ± 5.7	221 ± 7.2
C	233 ± 2.8	221 ± 8.5

Table 3-4. SCR Reactor Air Heater Tests: Inlet and Outlet Mass Concentrations

Reactor	Air Heater Inlet	Air Heater Outlet
	gr/dscf	gr/dscf
A	3.52 ± 0.44	3.21 ± 0.32
B	3.42 ± 0.12	2.76 ± 0.14
C	3.84 ± 0.17	Not Tested

Table 3-5. SCR Reactor Air Heater Tests: Fly Ash Particle Size

Reactor	Air Heater Inlet	Air Heater Outlet
	Stokes' MMD, micrometers	Stokes' MMD, micrometers
A	9.9	11.2
B	10.4	10.5
C	10.0	Not Tested

inlets on Reactors A, B and C and for the air heater outlets on Reactors A and B. No fly ash samples were collected at the outlet of the Reactor C air heater. The test results are presented in Table 3-6. There were no significant differences in the chemical constituents of the fly ash between the air heater inlets and outlets or among the three air heaters.

ASH RESISTIVITY

Laboratory measurements of ash resistivity were conducted on fly ash samples collected at the inlet and outlet of the three large reactor air heaters, except the outlet of the Reactor C air heater. The test data are summarized in Figure 3-1. The resistivity/temperature relationships for the five ash samples are very similar. For all of the ash samples the peak in the resistivity ranging from 4.5×10^{11} to 1×10^{12} ohm-cm, occurs at about 300 °F. The ash resistivity measured during the Task 1 baseline testing at the location of the SCR test facility inlet scoop was 3.9×10^{11} ohm-cm @ 293 °F. Based on these two measurements, the SCR process appears to have little or no effect on fly ash resistivity.

Table 3-6. SCR Reactor Air Heater Tests: Ash Mineralogy

Chemical Constituent	Reactor A Inlet % weight	Reactor A Outlet % weight	Reactor B Inlet % weight	Reactor B Outlet % weight	Reactor C Inlet % weight
Li ₂ O	0.03	0.03	0.03	0.03	0.04
Na ₂ O	0.97	0.91	1.2	1.1	0.91
K ₂ O	2.4	2.2	2.4	2.5	2.5
MgO	0.98	0.97	1.0	0.97	1.0
CaO	3.5	3.7	2.9	2.9	3.3
Fe ₂ O ₃	18.7	20.3	18.5	18.2	18.5
Al ₂ O ₃	22.2	22.0	22.3	21.7	21.9
SiO ₂	48.0	47.4	49.3	49.3	49.0
TiO ₂	1.1	1.1	1.1	0.91	0.91
P ₂ O ₅	0.14	0.14	0.14	0.15	0.14
SO ₃	1.1	1.0	0.78	0.27	0.78
Loss on Ignition	14.7	16.5	16.3	16.9	16.3

LABORATORY DUST RESISTIVITY
PLANT CRIST SCR TEST FACILITY
LARGE REACTOR AIR HEATERS

Figure 3-1. Resistivity/temperature relationship for air heater inlet and outlet fly ashes.

Section 4

SUMMARY

This report describes the initial air heater performance characterization tests for the Innovative Clean Coal Technology Demonstration of Selective Catalytic Reduction (SCR) Technology for the control of nitrogen oxide (NO_x) emissions from high-sulfur, coal-fired boilers. The SCR test facility is located at Gulf Power Company's Plant Crist Unit 5 in Pensacola, Florida. The air heater evaluations were conducted during the last week of April and the first two weeks of May, 1994, following the completion of the first sequence of parametric testing under Task 4: Long-Term Parametric Tests.

The SCR test facility at the Plant Crist test site includes three large SCR reactors, each designed to treat 5000 wscfm of flue gas, and six small reactors, each processing 400 wscfm of flue gas. An air heater capable of removing sufficient heat to reduce the flue gas temperature from a range of 600 to 750 °F down to 300 °F was included in the design of each of the three large reactors. The three large reactors are designated as reactors A, B, and C. Reactors A and B incorporate Ljungstrom-type air heaters and Reactor C incorporates a heat pipe design.

The air heater testing included the determinations of particulate mass concentration and measurements of the concentrations of sulfur dioxide, sulfur trioxide, hydrogen chloride and ammonia with manual sampling methods. Ammonia samples were segregated into solid and gas-phase fractions to characterize the gas/solid phase partitioning of ammonia across the air heaters. Tests were performed at the inlet and outlet of each of the three air heaters (except for mass concentration measurements at the outlet of the Reactor C air heater). The fly ash (particulate) catches from the mass concentration measurements were further analyzed in the laboratory to determine particle size distributions, ash mineralogy, and ash resistivity.

AMMONIA CONCENTRATIONS

Concentrations of ammonia were measured simultaneously at each of the three air heater inlet and outlet test locations. The tests were conducted so as to differentiate between gas-phase and solid-phase ammonia. The majority of the ammonia was found in the solid phase. Most of the gas phase ammonia concentrations were below the detection limit, especially at the air heater outlets. A measurable increase in ammonia concentration in the solid phase was detected during operation at the higher NH_3/NO_x ratio during operation at Test Condition 24 (1.8 to 6.7 ppm(v) dry @ 3% O_2 at Test Condition 24 versus 0.2 to 1.0 ppm(v) dry @ 3% O_2 at Test Condition 22 on Reactors B and C).

SO_2 and SO_3 CONCENTRATIONS

Concentrations of sulfur dioxide and sulfur trioxide were measured simultaneously at the air heater inlet and outlet test locations of Reactors A, B, and C during operation at Test Condition 22.

SO_3 inlet concentrations ranged from 8 to 19 ppm(v) dry @ 3% O_2 , while SO_3 concentrations at the reactor outlets ranged from 5 to 11 ppm(v) dry @ 3% O_2 . Reductions in SO_3 concentration across the air heaters ranged from 37.5% on Reactor B to 42.1% on Reactor A. Within the

tolerance of the standard deviations for the average SO₂ concentrations, there was no measurable change in SO₂ concentration across any of the three air heaters. Air heater inlet SO₂ concentrations averaged 2044 ± 41 ppm(v) dry @ 3% O₂, while the air heater outlet SO₂ concentrations averaged 2095 ± 34 ppm(v) dry @ 3% O₂.

HCl CONCENTRATION

Measurements of the concentration of hydrogen chloride were conducted simultaneously at the inlet and outlet of each of the three large reactor air heaters. All HCl concentrations fell within the range of 221 to 233 ppm(v) dry @ 3% O₂. There was no measurable change in HCl concentration across any of the three air heaters.

MASS CONCENTRATION

Mass concentration was measured simultaneously at the inlet and outlet of the air heaters on Reactors A and B. Mass concentration was measured only at the inlet to the heat pipe air heater on Reactor C.

The average air heater inlet mass concentrations ranged from 3.42 ± 0.12 (Reactor B) to 3.84 ± 0.17 gr/dscf (Reactor C). The Reactor A inlet and outlet average mass concentration values were within one standard deviation of each other, while the Reactor B average outlet mass concentration was 19% lower than the Reactor B average inlet mass concentration.

PARTICLE SIZE DISTRIBUTIONS

The particulate collected in the Method 17 thimbles during the reactor inlet and outlet mass concentration tests was analyzed using a Shimadzu Model SA-CP4 Centrifugal Particle Size Analyzer to determine particle size distribution. The mass median diameters (Stokes' diameters) of the fly ashes ranged from 9.9 to 11.2 micrometers. The Stokes' or physical diameter is based on the assumption of spherical particles and the true, or actual, particle density. No significant change in MMD across the air heaters was evident in these data.

ASH MINERALOGY

Fly ash samples collected during mass concentration tests at the air heater inlet and outlet test locations were submitted for ash mineralogy tests. Fly ash samples were available for air heater inlets on Reactors A, B and C and for the air heater outlets on Reactors A and B. No fly ash samples were collected at the outlet of the Reactor C air heater. There were no significant differences in the chemical constituents of the fly ash between the air heater inlets and outlets or among the three air heaters.

ASH RESISTIVITY

Laboratory measurements of ash resistivity were conducted on fly ash samples collected at the inlet and outlet of the three large reactor air heaters, except the outlet of the Reactor C air heater. The

resistivity/temperature relationships for the five ash samples are very similar. For all of the ash samples the peak in the resistivity, ranging from 4.5×10^{11} to 1×10^{12} ohm-cm, occurs at about 300 °F. These ash resistivities are similar to those measured during Task 1 on ashes collected from the Unit 5 ESP inlet duct.

APPENDIX
ORIGINAL DATA SUMMARIES

Figure A-1. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor A air heater inlet.

Figure A-2. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor A air heater outlet.

Figure A-3. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor B air heater inlet.

Figure A-4. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor B air heater outlet.

Figure A-5. Particle size distributions (cumulative % mass, top; differential % mass, bottom) of fly ash collected at the Reactor C air heater inlet.